

# **Treatment of wastewater using an innovative biofilm reactor – the air suction flow-biofilm reactor**

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*by*

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*“Live as if you were to die tomorrow. Learn as if you were to live forever”*

- Mahatma Gandhi (1869-1948)

*“Be the change you want to see in the world”*

- Mahatma Gandhi (1869-1948)

*“Opportunity is missed by most people because it is dressed in overalls and looks like work”*

- Thomas Edison (1847-1931)

*“To improve is to change; to be perfect is to change often”*

- Winston Churchill (1874-1965)

*“The greatest danger for most of us is not that our aim is too high and we miss it, but that it is too low and we reach it”*

- Michelangelo (1475-1564)

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## ABSTRACT

The role of wastewater treatment infrastructure in protecting water resources and thus as a vital part of a country's infrastructure cannot be overemphasised. Failure to sufficiently treat wastewater threatens human health, ecosystems, biodiversity, food security and the sustainability of water resources. Population growth, increased economic activity, increasingly stringent discharge regulations and a need to reduce costs and emissions will increase pressure on the existing wastewater treatment infrastructure. Due to their size, large scale facilities may have the capacity to adapt to such challenges. However, small scale and decentralised wastewater treatment plants can pose particular problems including (i) lack of on-site maintenance staff (ii) inability to meet discharge limits and (iii) inefficient and costly operation regimes. Thus the development of innovative technologies that can meet the challenges of this sector is timely.

In this research study, an innovative wastewater treatment technology, the Air Suction Flow-Biofilm Reactor (ASF-BR), developed by a research team in Civil Engineering at the National University of Ireland, Galway was investigated as an alternative to existing technologies; in particular for decentralised applications. The ASF-BR is a batch biofilm process that comprises two sealed reactors (R1 and R2), solenoid valves, a vacuum pump, and plastic media. Biofilm, which has accumulated on the media in each reactor, is passively aerated as wastewater is moved between the two reactors using a partial vacuum created by the vacuum pump. This novel technology potentially offers a number of advantages over conventional treatment process; (i) the passive aeration strategy can eliminate the need for mechanical blowers or stirrers, (ii) the sealed nature of the reactors allows control of gaseous emissions traditionally associated with wastewater treatment (iii) as a batch reactor the ASF-BR can accommodate aerobic, anoxic and anaerobic phases during a treatment cycle and (iv) flexible operational regimes can accommodate periods of low flow which can be problematic in tourist areas.

The research investigated the efficacy of the ASF-BR in the treatment of a high strength synthetic wastewater (Study 1), municipal wastewater (Study 2) and landfill leachate (Study 3). Furthermore the emission of nitrous oxide ( $N_2O$ ) from the ASF-BR during the nitrification and denitrification of municipal wastewater and landfill leachate was analysed (Study 4).

Study 1 comprised two phases (Phase 1 - 166 days and Phase 2 - 264 days) with both configured to achieve the removal of organic matter, nitrification and denitrification from synthetic high strength wastewater. In Phase 2 the cycle configuration was adjusted resulting in a 66% reduction in the energy requirement of the treatment process while maintaining excellent performance. Maximum average removal efficiencies were 97% filtered chemical oxygen demand ( $COD_f$ ), 88% filtered total nitrogen ( $TN_f$ ) and 99% ammonium-nitrogen ( $NH_4-N$ )

Two phases (Phases 1 and 2) were conducted during Study 2, lasting 212 days and 117 days respectively. The objective of Phase 1 was to investigate the performance of the ASF-BR for the removal of organic carbon and ammonium-nitrogen from municipal wastewater. During the 8.5 h treatment cycle, removals averaged 68%  $COD_f$ , 35%  $TN_f$  and 89%  $NH_4-N$ . During Phase 2 an anoxic period was added to facilitate denitrification of the wastewater. Removals averaged 89%  $COD_f$ , 93%  $NH_4-N$  and 75%  $TN_f$  (augmenting the municipal wastewater with an external carbon source).

Study 3 investigated the performance of the ASF-BR in treating landfill leachate over two phases (Phase 1 and Phase 2). Phase 1 sought to evaluate the performance of the ASF-BR in treating landfill leachate through nitrification and the removal of organic carbon. Removal efficiencies achieved averaged 63% COD<sub>f</sub>, 38% TN<sub>f</sub> and 86% NH<sub>4</sub>-N. During Phase 2, a 90 minute anoxic period was added to the beginning of the treatment cycle to facilitate the denitrification of the leachate and the cycle duration was also shortened for optimisation. Early results found that a lack of biodegradable carbon was likely limiting denitrification. Following the addition of an external biodegradable carbon source, removal efficiencies averaged 33% COD<sub>f</sub>, 73% NH<sub>4</sub>-N and 7% TN<sub>f</sub>.

In Study 4, N<sub>2</sub>O emissions from the ASF-BR when treating municipal wastewater and landfill leachate were monitored. Maximum nitrogen removal was observed in Study 2 which resulted in 0.18 mg N<sub>2</sub>O produced/mg NO<sub>3</sub>-N removed. Nitrogen removal through denitrification was low in Study 3 and limited N<sub>2</sub>O emissions were observed.

The results obtained from this study show the ASF-BR could offer a promising alternative to conventional wastewater treatment systems. The treatment performance observed was comparable to, or in excess of, many current technologies. The ASF-BR can also potentially offer benefits including (i) low energy requirements, (ii) operational flexibility, and (iii) an ability to capture the gases produced during the wastewater treatment process. While the ASF-BR could offer a reliable and robust alternative to existing processes, further studies, at a field scale, would be required to evaluate in detail the operation and maintenance of the unit in larger scale applications.

# **CHAPTER 1**

## **Introduction**



## 1.1 Introduction

Water is essential for life and is required for virtually everything we produce and consume. The availability of unpolluted water resources is crucial to the environment, the global economy and the livelihood of the world's population. As such, legislation has been introduced around the world which aims to mitigate the impact of wastewater discharges on receiving water bodies. In Europe, the Water Framework Directive (WFD) (2000/60/EC) and the Urban Wastewater Treatment Directive (UWTD) (91/271/EEC) regulate discharges from wastewater treatment plants and protect water bodies which receive these wastewater discharges. In the United States the Clean Water Act 1977 regulates the discharges of wastewater in a similar manner. The discharge of wastewaters containing excessive levels of carbon, nitrogen and phosphorus results in the pollution of receiving water bodies.

Water pollution can also have significant economic and social impacts. In an EU context, failure to comply with the requirements set out in the legislation can result in substantial penalties being imposed on offending states; such as those recently imposed on Greece, Belgium, Poland, Ireland, Hungary, Austria and Italy. The pollution of water bodies also has a negative impact on the tourism industry and furthermore many industries require an abundant supply of clean water.

In Europe, the WFD stipulates that EU member states must strive to achieve an improvement in the quality of all water bodies by 2015. In particular, requirements for nitrogen removal will necessitate an increase in the use of combined nitrification and denitrification systems. This will inevitably result in an increase in the production of greenhouse gases associated with wastewater treatment – namely carbon dioxide (CO<sub>2</sub>), nitrous oxide (N<sub>2</sub>O) and methane (CH<sub>4</sub>) which will have implications for compliance with the agreements set out under the Kyoto Protocol. Of particular concern is the potential increase in the production of N<sub>2</sub>O which has a global warming potential 300 times that of CO<sub>2</sub> (Kampschreur *et al.*, 2009).

Many problematic systems are small scale facilities that do not have permanent operational staff. Such facilities can also be subject to stringent discharge licenses but are often inefficient and inflexible. Point sources of pollution are not confined to municipal and domestic sources; landfill operations and industrial, commercial and agricultural facilities may require on-site treatment plants.

Thus future wastewater treatment systems will be required to meet discharge requirements while simultaneously being sustainable, cost effective and flexible. The potential impact of the treatment process on the atmospheric environment will also be a key consideration.

The Air Suction Flow Biofilm reactor (ASF-BR) is an innovative wastewater treatment technology developed by a research team in Civil Engineering at the National University of Ireland, Galway (NUI Galway) as a potential solution to these challenges. In this study the ASF-BR was investigated for its efficiency in removing organic carbon, nitrogen and phosphorus from various types of wastewaters. The study also investigated N<sub>2</sub>O emissions associated with nitrogen removal in the ASF-BR. The potential of the ASF-BR to offer a low energy, low maintenance, flexible treatment system that can control all gases produced during wastewater treatment was also assessed.

## **1.2 Objectives**

The overall objectives of this study were:

1. To commission and operate two pilot-scale ASF-BR units.
2. To evaluate the efficacy of the ASF-BR, at a pilot scale in treating a range of wastewaters.
3. To assess the operational characteristics of the ASF-BR technology and study the passive aeration process employed.
4. To measure the production of nitrous oxide during biological treatment of municipal wastewater and landfill leachate in the ASF-BR.

## **1.3 Experimental procedures**

Two identical pilot scale ASF-BR treatment units were designed and commissioned in this study. The two ASF-BR units were located within a temperature controlled room (11°C) in the Environmental Engineering Laboratory (Civil Engineering, NUI Galway). Such temperatures are typical of ambient air and water temperatures in Ireland. The units each comprised two sealed reactors filled with plastic media, a number of valves and a vacuum pump. The ASF-BR was operated as a sequencing batch biofilm reactor.

The ASF-BRs were loaded (over a number of experimental phases) with synthetic and municipal wastewaters and landfill leachate. Influent and effluent samples, along with

samples taken during individual treatment cycles, were analysed to determine carbon, nitrogen and phosphorous removal efficiencies. Operational parameters such as dissolved oxygen and sludge production were also analysed.

#### **1.4 Structure of dissertation**

Chapter 2 presents a review of wastewater treatment, the regulations governing wastewater treatment, wastewater treatment kinetics and existing technologies. A detailed description of the pilot scale ASF-BR is presented in Chapter 3 along with the methodology used in carrying out the research. Chapter 4 details the performance of the ASF-BR when treating high strength synthetic wastewater. Results of the performance of the ASF-BR when treating municipal wastewater are presented in Chapter 5 and Chapter 6 presents the results of studies carried out while treating landfill leachate. Observations of N<sub>2</sub>O emissions during the treatment of the municipal wastewater and landfill leachate are presented in Chapter 7. Chapter 8 presents the conclusions of the research and recommendations for future research.

# **CHAPTER 2**

## **Literature Review**

## 2.1 Introduction

The need for improved management of our water resources has led to the introduction of legislation governing the quality of our water bodies. A key part in protecting water bodies is managing wastewater discharges which these water bodies receive. Most regulatory bodies around the world have introduced legislation and guidelines to classify permissible quality standards in water bodies and effluent exiting wastewater treatment plants. Such regulation includes the Water Framework Directive (WFD; 2000/60/EC) in the EU and the Clean Water Act 1977 in the US, which directly influence the permissible discharges from wastewater treatment facilities.

Organic contaminants, solids, nutrients, pathogens and other harmful compounds found in wastewater result in water pollution which can lead to: (i) human health issues, (ii) a deterioration of the aquatic environment, and (iii) a deterioration in the socio-economic status of the affected region.

A sustainable approach to wastewater management requires consideration of all aspects of the environment. While stringent legislation is used to control the permissible discharges from wastewater treatment plants, little consideration is given to the atmospheric pollution arising from the treatment of the wastewater. The emission of odorous gases from the wastewater treatment process is a common source of complaint to environmental protection agencies and local authorities internationally, and many of these odorous gases, for example, hydrogen sulphide ( $\text{H}_2\text{S}$ ), are also toxic (Philips, 2008). Greenhouse gases such as nitrous oxide ( $\text{N}_2\text{O}$ ), carbon dioxide ( $\text{CO}_2$ ), and methane ( $\text{CH}_4$ ) are also produced during wastewater treatment. It has further been identified that there is an increasing need to develop wastewater treatment systems capable of complying with legislative requirements while minimising operation and management costs and encompassing design flexibility (Kampschreur *et al.*, 2009; Bhunia *et al.*, 2010).

This chapter outlines legislation related to wastewater treatment, describes the removal of organic carbon, nitrogen and phosphorus from wastewaters, outlines the characteristics of wastewater produced from both domestic and industrial sources and summarises current technologies employed in the treatment of wastewater.

## 2.2 The history of wastewater treatment

One of the earliest wastewater management systems was discovered in Mohenjo-Daro on the banks of the River Indus in Pakistan and dates back to 1500 BC. Private and public houses were found to be equipped with toilets; a separate water system was used for washing and bathing while rain water was carried away from areas through sloped canals toward a local river. Such installations demonstrated a high level of awareness regarding hygiene and sanitation in early advanced civilisations (Wiesmann *et al.*, 2006).

Other civilisations succeeded in developing primary treatment processes for wastewater treatment; however, it was not until the development of microscopes, which enabled the detailed examination of microorganisms that secondary treatment processes began to develop (Wiesmann *et al.*, 2006). Initial treatment processes focused on the reduction in biochemical oxygen demand (BOD) through constant mixing, with the subsequent removal of solids during a settlement period; this was followed by the development of the trickling filter (Cooper, 2007). Many variations of treatment technologies were developed during the 20<sup>th</sup> century as a result of advances in environmental analytical chemistry, enabling an understanding of wastewater treatment processes and facilitating the development of more efficient and effective treatment systems. However many challenges still persist, e.g. effective operation, cost and maintenance, efficiency, etc.

## 2.3 Regulations relating to wastewater treatment

The development of methodologies to test the characteristics of wastewater, such as BOD, and a growing awareness of the health and environmental implications of inadequate wastewater treatment led to the implementation of the first standards for effluent wastewaters, like those contained in the Eighth Report (1912) of the Royal Commission on Sewage Disposal in the UK (Lofrano and Brown, 2010). Over time governments began to regulate permissible standards required for wastewater discharges from treatment plants. Today the majority of the developed world has stringent legislation governing wastewater treatment and the permissible levels of contaminants in effluent wastewater discharged from treatment plants.

### **2.3.1 Regulation in the European Union**

The member states of the European Union have adopted a common legislative approach to the regulation of acceptable standards in drinking water since the 1970s through directives such as:

- The Surface Waters Directive (75/440/EEC)<sup>a</sup>;
- The Freshwater Fish Directive (78/659/EEC)<sup>b</sup>;
- The Shellfish Waters Directive (79/923/EEC)<sup>b</sup>;
- The Drinking Water Directive (80/778/EEC, repealed by 98/83/EC).

Note: <sup>a</sup> directive repealed by the Urban Water Framework Directive (2000/60/EC, see Section 2.3.1.1); <sup>b</sup> directive due to be repealed by 2000/60/EC in 2013.

Between 1990 and 1996, a second round of European directives was introduced relating to the management of water:

- The Urban Wastewater Treatment Directive (91/271/EEC, amended as 98/15/EEC);
- Nitrates Directive (91/676/EEC);
- The Integrated Pollution and Prevention Control Directive (96/61/EC, amended as 2008/1/EC).

As part of the terms of various European Treaties, all countries in the European Economic Community (EEC) and the subsequent European Union (EU) adopted these directives. During the 1990s, it was decided to introduce one directive that incorporated all water-related directives – The Water Framework Directive (2000/60/EC), which was adopted in 2000. Currently there are two primary directives relating to wastewater, the Water Framework Directive (2000/60/EC) and the Urban Wastewater Treatment Directive (91/271/EEC).

#### *2.3.1.1 The Urban Wastewater Treatment Directive (UWWTD)*

The UWWTD has been one of the main standards for effluent wastewater discharges since it was adopted in May 1991. The objective of this directive is to protect the environment from the adverse effects of urban wastewater discharges and discharges from certain industrial sectors (Outlined in Annex III of the UWWTD).

The UWWTD set out discharge limits for agglomerations with a population equivalent (PE) over 2,000 (Tables 2.1 and 2.2), which applied to all member states. The 12 accession states (those who joined the EU in 2004 and 2007) were given an extended timeframe to meet these regulations, having only joined the EU between 2004 and 2007, and in many cases, having underdeveloped infrastructure (European Commission, 2011). Until recently, at their discretion,

local authorities have applied discharge limits for population centres under 2,000 PE using the UWWTD as a minimum standard.

**Table 2.1 Regulations concerning discharges from urban wastewater treatment plants <sup>a</sup> (91/271/EEC)**

<b>Parameter</b>	<b>Concentrations</b>	<b>Minimum % reduction <sup>b</sup></b>
5-day biochemical oxygen demand (BOD <sub>5</sub> at 20°C) without nitrification	25 mg O <sub>2</sub> /l	70-90%
Chemical oxygen demand (COD)	125 mg O <sub>2</sub> /l	75%
Total suspended solids (TSS)	35 mg/l <sup>c</sup>	90% <sup>c</sup>
	35 mg/l in high mountain regions for agglomerations with more than 10,000 PE	90% in high mountain regions for agglomerations with more than 10,000 PE
	60 mg/l in high mountain regions for agglomerations with a PE between 2,000 and 10,000	70% in high mountain regions for agglomerations with a PE between 2,000 and 10,000

Notes: <sup>a</sup> the values of concentration or percentage of the reduction can be chosen differently; <sup>b</sup> reduction in relation to the load of the influent; <sup>c</sup> this requirement is optional

**Table 2.2 Requirements for discharge from urban wastewater treatment plants to sensitive areas (91/271/EEC)**

<b>Parameter</b>	<b>Concentrations</b>	<b>Minimum % reduction <sup>a</sup></b>
Total phosphorus (TP)	2 mg/l (10,000-100,000 PE)	80%
	1 mg/l (>100,000 PE)	
Total nitrogen <sup>b</sup> (TN)	15 mg/l (10,000-100,000 PE)	70-80%
	10 mg/l (>100,000 PE)	

Notes: <sup>a</sup> reduction in relation to the load of the influent; <sup>b</sup> total nitrogen means the sum of the total kjeldahl nitrogen (organic nitrogen, ammonium-nitrogen (NH<sub>4</sub>) and ammonia-nitrogen (NO<sub>3</sub>))

Wastewater discharges from treatment plants continue to be regulated by the UWWTD; however, the Water Framework Directive (2000/60/EC) has an overriding stipulation that the ecological status of the receiving water body and associated river basin as a whole needs to be considered, resulting in the discharge limits applied often being more stringent than those set out in the UWWTD.

### 2.3.1.2 The Water Framework Directive (WFD)

The WFD divides each country in the EU into River Basin Districts (RBDs) which comprise one or more river basins together with associated ground, surface and coastal waters. The RBDs are defined based on the rivers' geographical, natural and hydrological formations, beginning at the



water body source and continuing through to the coast rather than on national or political boundaries. The RBD classifies the surface water ecological status as ‘high’, ‘good’, ‘moderate’, ‘poor’ or ‘bad’, as summarised in Table 2.3.

**Table 2.3 Summary of ecological quality standards as defined in the Directive 2000/60/EC (adapted from 2000/60/EC)**

<b>High status</b>	<b>Good status</b>	<b>Moderate status</b>
There are no, or only slight, alterations to the values of the quality for the surface water body type from those normally associated with that type under undisturbed conditions.	There are low levels of distortion of the biological qualities of water body resulting from human activity.	The values of the biological quality elements for the surface water body type deviate moderately from those normally associated with the surface water body type.

Waters achieving a status below moderate are classified as poor or bad.

The objectives of the WFD include: (i) to prevent further deterioration of water bodies and to protect water bodies with a ‘high status’, and (ii) to restore all water bodies to ‘good status’ by 2015. When setting the permissible discharge from a wastewater treatment plant, local authorities must consider the assimilation capacity of the entire RBD downstream of the point in question. The WFD does not prevent individual member states from implementing stricter national standards where they are deemed necessary (Moss, 2004).

### ***2.3.2 Regulation in Ireland***

In Ireland, the Water Services Act (2007) provides a comprehensive review of the water legislation through the consolidation and updating of the existing instruments. In 2007 discharge licences were introduced for wastewater treatment plants with a PE over 10,000; thereafter the initiative was expanded on a phased basis to include wastewater treatment plants with population equivalents as low as 500 PE under the Waste Water Discharge (Authorisation) Regulations, 2007 (ISB, 2012). Under this regulation, each local authority must apply to the EPA for a discharge licence or Certificate of Authorisation for each wastewater treatment plant in their jurisdiction. The EPA then determines the treatment plants requirements based on the assimilation capacity of the receiving water body, as described in the WFD (2000/60/EC).

### **2.3.3 Regulation in the United States, Canada and Australia**

In the US, the Clean Act 1977 outlines the regulations governing the discharge of wastewater into water bodies and regulates the quality standards required for surface water in a similar manner to the WFD in Europe (Davis, 2010). In Canada, water quality is regulated through the Fisheries Act 2012, which stipulates that all wastewater discharges exceeding 100 cubic meters per day must receive secondary treatment (screening, filtering and biological treatment) before entering a receiving water body (James and Leffler, 2013). Water quality in Australia is federally regulated under the Australian Drinking Water Guidelines (2004), with each state having agreed under the Intergovernmental Agreement on the National Water Initiative (NMI) to adequately manage the local water resources (Marques, 2010).

### **2.3.4 Regulation relating to industrial wastewater in Ireland**

Any industry seeking to discharge wastewater into a municipal sewerage collection system is required to characterise the wastes adequately, identifying the ranges of constituent contaminants and the resulting mass loading. This characterisation will determine if pre-treatment is required prior to discharge into the wastewater collection system.

Industries seeking to discharge wastewater to a sewage collection system must apply to the EPA for a discharge licence. The EPA then determines if pre-treatment is required and to what extent, based on the capacity of the collection system, the ability of the municipal wastewater treatment plant (MWTP) to sufficiently treat the increased load and the ability of the river basin to assimilate the increase in discharge from the wastewater treatment plant.

## **2.4 Wastewater characteristics**

Wastewater can be divided into a number of categories (Hall, 2002; Tchobanoglous *et al.*, 2004) comprising:

- **Municipal wastewater** – includes wastewater from household connections, small enterprises, schools and light industry.
- **Industrial wastewater** – refers to any wastewater which is discharged from premises used for carrying on any trade or industry other than domestic waste water and run-off rain water.
- **Storm water** – refers to water run-off resulting from rainfall or snowmelt.

### 2.4.1 Municipal wastewater

The primary sources of municipal wastewater in a community are the residential areas, schools, commercial premises and light industry. Municipal wastewater flow rates are determined by population equivalent (PE), based on the water usage within the area being served by the collection system (1 PE = 0.2m<sup>2</sup>/d or 60 g BOD<sub>5</sub>/d).

The average hydraulic load on a treatment plant treating municipal wastewater varies internationally, ranging from 150 l/PE/d in Ireland to 250 l/PE/d in the United States (Tchobanoglous *et al.*, 2004; EPA, 2009). Individual loading rates can have a broader range depending on local conditions.

Municipal wastewater typically consists of 99.9% water containing a mixture of organic and inorganic materials (Gray, 2004). The levels of contaminants present in the domestic/municipal wastewater can be evaluated using the following parameters:

- Solids – measured as suspended solids (SS) or TSS;
- Organic carbon – measured using BOD<sub>5</sub> and COD;
- Nitrogen (N) – measured as total nitrogen (TN), ammonium nitrogen (NH<sub>4</sub>), nitrate-nitrogen (NO<sub>3</sub>) and nitrite-nitrogen (NO<sub>2</sub>);
- Phosphorus – measured as TP and inorganic P in the form of orthophosphate-phosphorus (PO<sub>4</sub>-P).

Municipal wastewater can also contain substances such as fats, oils, greases, sulphates, chlorides and volatile organic compounds (Tchobanoglous *et al.*, 2004). Table 2.4 outlines a range of municipal wastewater constituent concentrations.

**Table 2.4 Characteristics of municipal wastewater**

Parameter	Concentration (mg/l unless otherwise stated)				
	Gray (2004)	Davis (2010) <sup>b</sup>	EPA (2009)	Tchobanoglous <i>et al.</i> , 2004 <sup>a</sup>	Henze <i>et al.</i> , (2000) <sup>a</sup>
BOD <sub>5</sub>	326	300	350-500	110-350 (190)	110-350 (190)
Total COD	650	1000	300-1000	250-800 (430)	250-800 (430)
Total N	66	80 <sup>c</sup>	-	20-70 (40)	20-70 (40)
Organic N	19	-	-	-	8-25 (15)
NH <sub>4</sub>	47	4-13	22-80	12-45 (25)	12-45 (25)
NO <sub>2</sub>	0	<1	-	0	0-0 (0)
NO <sub>3</sub>	-	<1	-	8-25 (15)	0-0 (0)
Total P	15	20	5-20	4-12 (7)	4-12 (7)
Organic P	3	-	-	-	1-4 (2)
Inorganic P	12	-	-	-	3-8 (5)
Total coliforms <sup>d</sup>	-	-	10 <sup>6</sup> -10 <sup>8</sup>	10 <sup>6</sup> -10 <sup>10</sup> (10 <sup>7</sup> -10 <sup>9</sup> )	10 <sup>6</sup> -10 <sup>10</sup> (10 <sup>7</sup> -10 <sup>9</sup> )
Faecal coliforms <sup>d</sup>	-	-	-	10 <sup>3</sup> -10 <sup>8</sup> (10 <sup>4</sup> -10 <sup>6</sup> )	10 <sup>3</sup> -10 <sup>8</sup> (10 <sup>4</sup> -10 <sup>6</sup> )
TSS	127	400	-	390-1230 (430)	120-450 (210)

Notes: <sup>a</sup> mean value in parenthesis; <sup>b</sup> Strong municipal wastewater; <sup>c</sup> Total kjeldahl nitrogen; <sup>d</sup> No./100ml

### 2.4.2 Industrial wastewater

There are many different kinds of human activities that generate industrial wastewater, including manufacturing, petrochemical, pharmaceutical, fertilizer and food industries, breweries, and leachates produced by solid waste disposal sites (Davis, 2010).

The composition and quantity of wastewater from industrial sources vary greatly depending on the operation and scale of a particular industry. Many industrial wastewaters will contain contaminants associated with a particular commercial process as well as elevated levels of organic matter, nitrogen and phosphorous (Galil and Levinsky, 2006). Part of this study focused on high-strength wastewaters typically arising from municipal landfills – i.e. landfill leachate.

#### 2.4.2.1 Landfill leachate

In most countries, sanitary landfilling is the most common method of disposing of municipal solid wastes (MSW) (Renou *et al.*, 2008). Landfill leachate is generated by excess rainwater percolating through the waste layers in a landfill, dissolving the by-products of the waste degradation (Kjeldsen *et al.*, 2002). Leachates are defined as the aqueous effluent generated as a consequence of rainwater percolation through wastes, biochemical processes in waste cells and the inherent water content of wastes. Landfill leachate is one of the most difficult industrial wastewaters to treat due to the complex chemical composition and seasonal variation in volumes (Bohdziewicz *et al.*, 2001). Leachates may contain large amounts of organic matter (biodegradable, but also refractory to biodegradation), where humic-type constituents are a predominant group, as well as ammonia-nitrogen, heavy metals, and chlorinated organic and inorganic salts (Kjeldsen *et al.*, 2002; Renou *et al.*, 2008).

There are many factors affecting the quality of landfill leachates, such as landfill age, precipitation, seasonal weather variation, and waste type and composition. In particular, the composition of landfill leachates varies greatly depending on the age of the landfill (Baig *et al.*, 1999).

In the EU, a common approach to landfill leachate treatment is to treat the leachate together with municipal wastewater in a MWTP (Kalka, 2012). However, this solution has been questioned due to concern over the effects of leachates on municipal wastewater treatment processes and the quality of the sludge generated (Welander *et al.*, 1998; Ceçen and Aktas, 2004). It has been found that inhibitory compounds in the leachates may disturb the biological sewage treatment process, and ultimately result in the accumulation of hazardous compounds, such as bioaccumulable organics and heavy metals. The sludge formed in the treatment process may be unsuitable for use as a fertilizer in agriculture (Lagerkvist, 1986; Linde, 1995; Ceçen and Aktas, 2004; Deng, 2007).

Numerous systems have been adopted and studied for use in leachate treatment, including suspended growth systems such as aerated lagoons, conventional activated sludge processes and sequencing batch reactors (SBRs) (Robinson and Grantham, 1988; Hosomi *et al.*, 1989).

These systems have been found to have significant limitations, such as sensitivity to low temperatures and loss of activity due to poor settleability of the biomass (Knox, 1985; US EPA, 2009). As a result, there has been a continuous search for other alternative viable treatment processes. Attached growth systems, such as trickling filters (Knox, 1985), submerged aerobic biological filters (Pedersen and Jansen, 1992) and rotating biological contactors (Spengel and Dzombak, 1991), have also been investigated.

Tables 2.5 give an overview of the ranges and composition of landfill leachate. These data show that the age of the landfill and consequentially the degree of solid waste stabilisation has a significant effect on leachate characteristics. Concentrations of COD vary from 70,000 mg/l in leachate samples obtained from the landfill serving the Thessaloniki Greater Area (Greece) to 500 mg/l in samples from a landfill operating for 10 years near Marseille (France). The ratio of BOD/COD generally decreases with the age of the landfill, from 0.70 to 0.04 (Chian and DeWalle, 1976; Pi *et al.*, 2009). The pH of leachate was generally found to be between 5.8 and 8.5, depending on the nature of the biological activity taking place inside the landfill. The majority of Total Kjeldahl Nitrogen (TKN; organic nitrogen + ammonia nitrogen (NH<sub>3</sub>) + ammonium nitrogen (NH<sub>4</sub>)) is in the form of ammonia, and can range from 0.2 to 13,000 mg N/l (Renou *et al.*, 2008). This is due to the release of the large recalcitrant organic molecules from the solid wastes. Consequently, old landfill leachate is characterized by its low ratio of BOD/COD and fairly high ammonium (NH<sub>4</sub>) (Kjeldsen *et al.*, 2002).

**Table 2.5 Reported leachate concentrations (Adapted from Renou *et al.*, 2008)**

Reference	Site location	COD mg/l	BOD mg/l	BOD/COD ratio	pH	SS mg/l	TKN mg/l	NH <sub>4</sub> mg/l	Age*
Henry <i>et al.</i> , 1987.	Canada	13,800	9,660	0.70	5.8	-	312	42	Y
Lau <i>et al.</i> , 2001	Hong Kong, China	15,700	4,200	0.27	7.7	-	-	2,260	Y
Tatsi <i>et al.</i> , 2003	Greece	70,000	26,800	0.38	6.2	950	3,400	3,100	Y
Di Palma <i>et al.</i> , 2002	Italy	19,900	4,000	0.20	8	-	-	3,917	Y
Im <i>et al.</i> , 2001	South Korea	24,400	10,800	0.44	7.3	2,400	1,766	1,682	Y
Timur and Ozturk, 2001	Turkey	16,200-20,000	10,800-11,000	0.55-0.67	7.3-7.8	-	-	1,120-2,500	Y
Baumgarten and Seyfried, 1996	Germany	3,180	1,050	0.2	7.9	480	1,100	940	M
Bohdziewicz <i>et al.</i> , 2001	Poland	1,180	331	0.28	8	-	-	743	M
Wu <i>et al.</i> , 2004	Taiwan	6,500	500	0.08	8.1	-	-	5,500	M
Tatsi <i>et al.</i> , 2003	Greece	5,350	1,050	0.20	7.9	480	1,100	940	M
Frascari <i>et al.</i> , 2004	Italy	5,050	1,270	0.25	8.38	-	1,670	1,330	M
Silva <i>et al.</i> , 2004	Brazil	3,460	150	0.04	8.2	-	-	800	O
Hoilijoki <i>et al.</i> , 2000	Finland	556	62	0.11	-	-	192	159	O
Aziz <i>et al.</i> , 2004	Malaysia	1,533-2,580	48-105	0.03-0.04	7.5-9.4	159-233	-	-	O
Trebouet <i>et al.</i> , 1999	France	500	7.1	0.01	7.5	130	540	430	O
Uygur and Kargi, 2004	Turkey	10,000	-	-	8.6	1,600	1,680	1,590	O
Kim <i>et al.</i> , 2007	Korea	-	2,000	-	7.5	-	1,000- 1,600	1,200	O

\* Age of landfill: Y: Young landfill, <5years; M: Mature landfill 5-10 years; O: Old landfill >10years

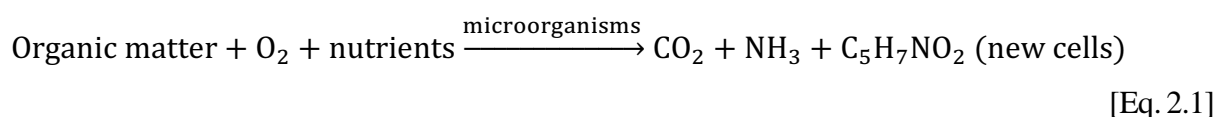
## 2.5 Biological wastewater treatment processes

### 2.5.1 Biodegradable organics

The most widely used parameters to quantify organic pollution are BOD<sub>5</sub>, COD and total organic carbon (TOC). Elevated levels of organic matter in waters can cause a high dissolved oxygen (DO) demand for oxidation processes, and can result in oxygen depletion in the receiving water body. The organic matter contained in wastewater is classified as either biodegradable or non-biodegradable organic matter. Only the biodegradable organic matter can be removed in biological treatment processes. This biodegradable organic matter is primarily composed of proteins, carbohydrates and fats (Tchobanoglous *et al.*, 2004). The removal of biodegradable organic matter in biological treatment systems is achieved through aerobic oxidation and heterotrophic denitrification (Tchobanoglous *et al.*, 2004; van Haandel and van der Lubbe, 2007; Henze *et al.*, 2008).

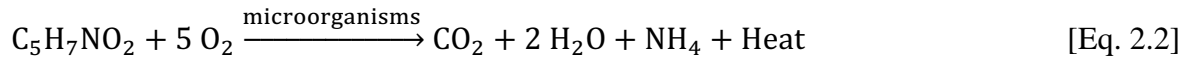
#### 2.5.1.1 Organic matter removal by aerobic oxidation biofilm systems

Microorganisms consume energy which can be obtained from the oxidation of organic matter by the microorganisms resulting in the production of CO<sub>2</sub>, ammonia, new cells and other minor end products. This process is known as assimilation. In addition to organic matter, the formation of new cells requires additional nutrients. The principal nutrients required are nitrogen and phosphorus. The ratio of organic matter to nitrogen and phosphorus during heterotrophic growth is generally accepted to be approximately 100 g BOD<sub>5</sub>: 5 g nitrogen: 1 g Phosphorus (Tchobanoglous *et al.*, 2004). During aerobic conditions, bacteria utilise free dissolved oxygen as the electron acceptor to oxidise organic matter. The first step in the aerobic oxidation process, called the “carbonaceous oxidation stage”, can be represented by the following equation:



However, the removal of organic matter by assimilation is not permanent as it can be released back into the wastewater when the organisms decay (Hoover and Porges, 1952). In the second step of the process, older cells are continually dying and the released substrate from these cells is used as an energy source in the production of new cells. This process is known as “endogenous respiration” and is denoted as follows:

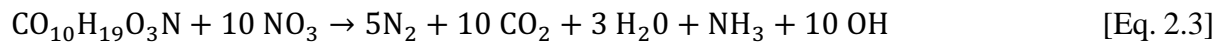




From these equations, we can see that: (i) energy is produced; (ii) the biomass is converted to simpler substances such as  $\text{CO}_2$ ,  $\text{H}_2\text{O}$ , and  $\text{NH}_3$ ; and (iii) a definite amount of oxygen is required (Tchobanoglous *et al.*, 2004).

### 2.5.1.2 Organic matter removal by heterotrophic denitrification

During the denitrification process, a source of organic matter is required to act as an electron donor to facilitate the reduction of nitrate nitrogen ( $\text{NO}_3$ ). Influent wastewater has conventionally been used as the source of this organic matter, resulting in the removal of organic matter, as per the following equation:



where:  $\text{CO}_{10}\text{H}_{19}\text{O}_3\text{N}$  is the term commonly used for organic matter in wastewater;  $\text{N}_2$  is nitrogen gas; and OH represents hydroxide.

It has been found that for each g  $\text{NO}_3$  reduced, approximately 4 g of BOD is required (Tchobanoglous *et al.*, 2004; van Haandel and van der Lubbe, 2007)

### 2.5.1.3 Factors affecting biological organic carbon removal

To achieve optimal organic matter removal, environmental factors such as the carbon source and concentration, temperature, DO concentration and pH must be monitored.

#### 2.5.1.3.1 Dissolved oxygen (DO), temperature and pH

During aerobic biological oxidation, DO is used as the electron acceptor. The oxygen dependency for aerobic processes can be described by the following Monod expression (Henze *et al.*, 2000):

$$\mu_{obs} = \mu_{max} \cdot \frac{S_{O_2,2}}{S_{O_2,2} + K_{S,O_2}} \quad [\text{Eq. 2.4}]$$

where:  $\mu_{obs}$  is the observed growth rate;  $\mu_{max}$  is the maximum growth rate constant ( $t^{-1}$ );  $S_{O_2,2}$  is the oxygen concentration in the reactor,  $K_{S,O_2}$  is the saturation constant for oxygen.

The growth rate of microorganisms is affected by temperature. The effect of temperature on the biological process can be described by the following equation (Tipadis, 1991):

$$\mu_{m,T} = \mu_{m,20} \theta^{(T-20)} \quad [\text{Eq. 2.5}]$$

where:  $\mu_{m,T}$  is the maximum growth rate of the heterotrophic bacteria at a temperature of  $T$  °C;  $\mu_{m,20}$  is the maximum growth rate at 20 °C; and  $\theta$  is a constant that has a value of between 1 and 1.14.

The pH of the biofilm system affects the enzymes that induce and control the rate of biochemical reactions. The aerobic conversions are pH dependent, and the kinetics of pH can be described as follows (Henze *et al.*, 2000):

$$\mu_{\max}(\text{pH}) = \mu_{\max}(\text{opt. pH}) \cdot \frac{K_{\text{pH}}}{K_{\text{pH}} + 1} \quad [\text{Eq. 2.6}]$$

where:  $\mu_{\max}(\text{pH})$  is  $\mu_{\max}$  for the pH condition; and  $\mu_{\max}(\text{opt. pH})$  is the  $\mu_{\max}$  for the optimum pH condition;  $K_{\text{pH}}$  is the pH constant;  $l = 10^{(\text{opt pH} - \text{pH})} - 1$

### 2.5.2 Biological removal of nitrogen

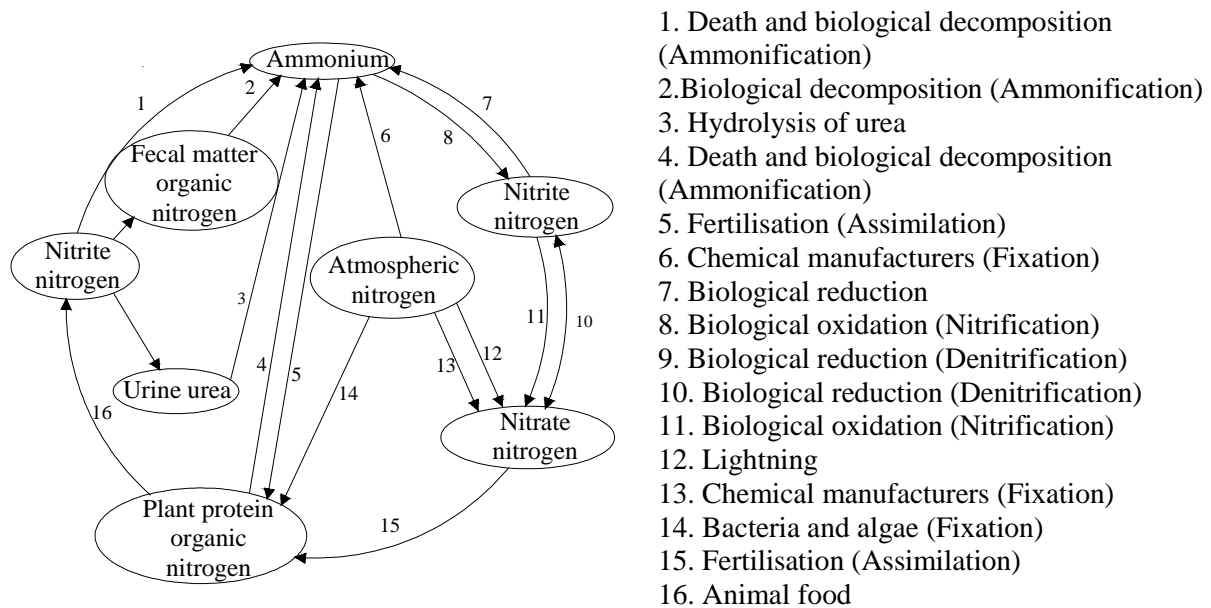
Nitrogen compounds discharged into the environment can cause serious problems, such as the eutrophication of rivers and deterioration of water sources, as well as being hazardous to human health.

Typically TKN, TN,  $\text{NH}_4$  and oxidised nitrogen ( $\text{NO}_3$  and nitrite nitrogen ( $\text{NO}_2$ )) are used to express the concentration of nitrogen in wastewater as follows (Tchobanoglous *et al.*, 2004):

$$\text{TKN} = N_{\text{org}} + \text{NH}_3 + \text{NH}_4 \quad [\text{Eq. 2.7}]$$

$$\text{TN} = N_{\text{org}} + \text{NH}_3 + \text{NH}_4 + \text{NO}_3 + \text{NO}_2 \quad [\text{Eq. 2.8}]$$

The relationship between nitrogen compounds and their transformations in nature are illustrated in Figure 2.1 (Tipadis, 1991). Recent research into the mechanisms of the nitrogen cycle has led to the development of alternative methods of nitrogen removal, such as the single reactor high activity ammonia removal over nitrite (SHARON) process, oxygen-limited autotrophic nitrification-denitrification (OLAND), anaerobic ammonium oxidation (ANAMMOX) and completely autotrophic nitrogen removal over nitrite (CANON) processes (Paredes *et al.*, 2007).

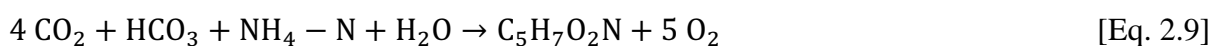


**Figure 2.1 Nitrogen cycle in nature (Tipadis, 1991)**

$\text{NH}_4$  is the main component of TN in influent wastewater, with the remainder mainly comprising organic N (Henze *et al.*, 2000). Oxidised nitrogen ( $\text{NO}_2$  and  $\text{NO}_3$ ) is rarely present in raw wastewater but is produced as  $\text{NH}_4$  is oxidised during the nitrification process. The transformation of  $\text{NH}_4$  to oxidised nitrogen is an oxidation process which can result in the depletion of oxygen in receiving waters and the subsequent death of marine life. The oxidation of  $\text{NO}_3$  has a lower oxygen demand than  $\text{NH}_4$  and  $\text{NO}_2$ ; however, it can still result in excess growth of algae and marine plants. The removal of nitrogen can be achieved by means of assimilation and conversion of nitrogen to nitrogen gas through the processes of nitrification and denitrification.

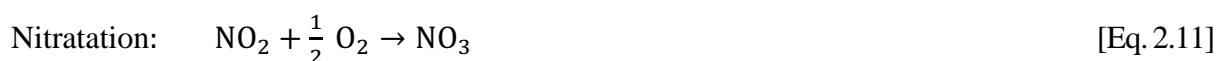
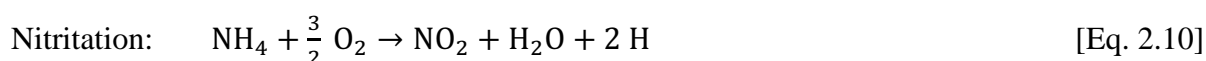
### 2.5.2.1 Nitrogen removal by assimilation

As microorganisms grow,  $\text{NH}_4$  in wastewater is used as a source of nitrogen for the new cells, thus contributing to overall nitrogen removal. Biomass can be represented by the formula  $\text{C}_5\text{H}_7\text{O}_2\text{N}$ , which shows the approximately 12.5% of the dry biomass is nitrogen. Nitrogen removed by assimilation is not a permanent removal mechanism as the nitrogen can be re-released into the wastewater as the organisms decay. The biomass synthesis reaction can be represented as follows (Tchobanoglous *et al.*, 2004):



### 2.5.2.2 Autotrophic nitrification

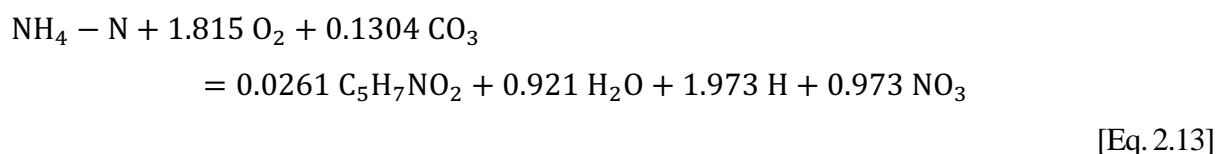
Nitrification is the term given to the process of oxidising  $\text{NH}_4$  to  $\text{NO}_2$  and its subsequent oxidation to  $\text{NO}_3$ . Nitrifying autotrophic bacteria carry out nitrification in a two-step process – nitritation followed by nitrataion. During the nitritation step,  $\text{NH}_4$  is oxidised to  $\text{NO}_2$  by ammonia oxidising bacteria (AOB), such as *Nitrosomanas* bacteria. Nitrite oxidising bacteria (NOB), such as *Nitrobacter* bacteria, then oxidise the  $\text{NO}_2$  to  $\text{NO}_3$  during the nitrataion step. The oxidisation of the  $\text{NH}_4$  to  $\text{NO}_3$  can be described by the following expressions (Tchobanoglous *et al.*, 2004):



The overall reaction for nitrification is:



From Equations 2.10 and 2.11, it can be seen that oxygen demands are 3.43 and 4.57 mg  $\text{O}_2/\text{mg}$   $\text{NH}_4$  for partial (nitritation) and complete (nitritation and nitrataion) nitrification respectively. If microbial assimilation is included, the nitrification equation becomes (Rittmann and McCarty, 2001):



In Equation 2.13, only 2.6% of the removal is used in biomass synthesis, which indicates the low net yield rate (0.21 g biomass/g N) of nitrifiers. Each gram of ammonia nitrogen converted during nitrification removes 7.14 g of alkalinity (Tchobanoglous *et al.*, 2004).

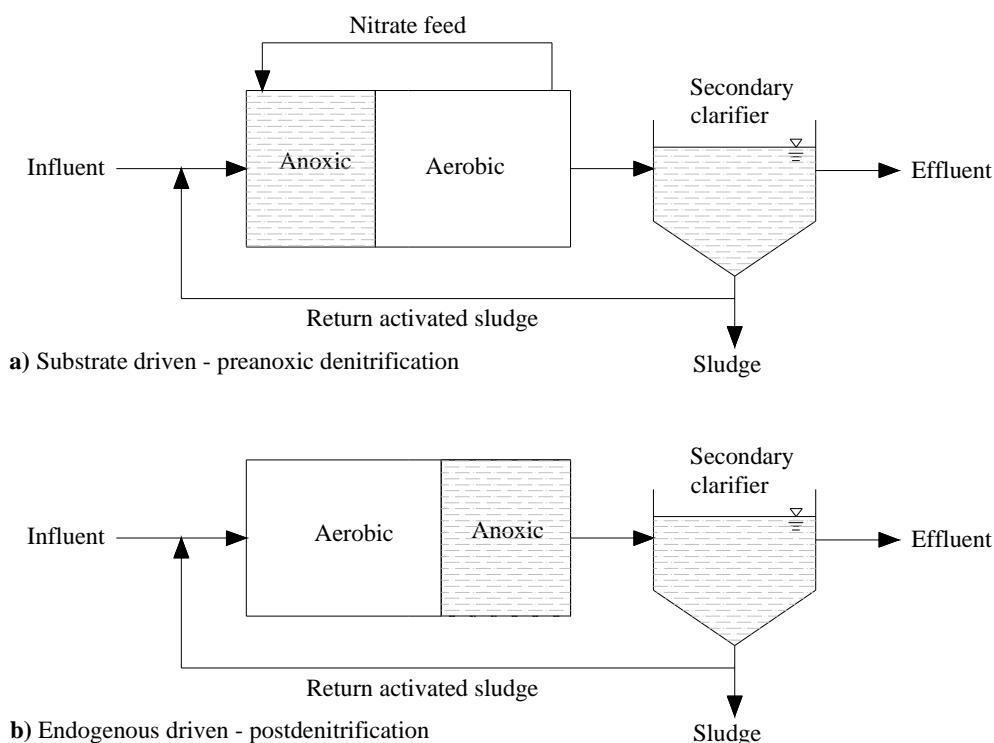
### 2.5.2.3 Heterotrophic denitrification

During denitrification, organic substrates are biologically oxidised using  $\text{NO}_2$  or  $\text{NO}_3$  as the electron acceptor rather than oxygen. The denitrification process uses a wide spectrum of heterotrophic bacteria to convert  $\text{NO}_2$  and  $\text{NO}_3$  into nitrogen gas. It is known that approximately 50% of all known bacteria can denitrify or at least carry out some partial denitrification reactions (Bhunja *et al.*, 2010). Low oxygen concentrations, a source of biodegradable organic matter and  $\text{NO}_2$  or  $\text{NO}_3$ , are prerequisites for the denitrification process (Baytshtok *et al.*, 2009). Following the formation of  $\text{NO}_3$  during the nitrification process,  $\text{NO}_3$  is removed from the wastewater through heterotrophic denitrification under anoxic conditions. Anoxic conditions occur when there is a low DO concentration combined with  $\text{NO}_3$  in an aqueous environment.  $\text{NO}_3$  acts as the final electron acceptor and can be reduced to nitrogen gas by facultative anoxic denitrifiers using organic carbon substrates as the electron donor for respiration. The reaction pathway can be expressed as follows (Tchobanoglous *et al.*, 2004):



where: NO is nitric oxide and  $\text{N}_2\text{O}$  is nitrous oxide

Denitrification can be classified as either pre-denitrification or post-denitrification. Pre-denitrification (substrate driven) occurs when organic carbon in the influent wastewater is used as the carbon source (Figure 2.2a), while post-denitrification (endogenous driven) occurs when externally added (influent wastewater or other carbon sources) or internally-stored organic substrates are used as the carbon source (Figure 2.2b) (Tchobanoglous *et al.*, 2004; Davis, 2010).



**Figure 2.2 Two configurations for biological nitrogen removal using activated sludge plants (Adapted from Tchobanoglous *et al.*, 2004)**

#### 2.5.2.4 Simultaneous nitrification and denitrification (SND)

The phenomenon of SND, in particular aerobic denitrification, has been observed in a number of studies (Hibiya *et al.*, 2003; Peng and Qi, 2007; Yang and Yang, 2011; Guo *et al.*, 2013). SND can occur where DO concentration gradients occur within microbial flocs or biofilms due to diffusional limitations creating anoxic zones within microbial floc. As a result, the nitrifiers exist in regions with high DO concentrations, whereas denitrifiers may be active in zones with very low DO concentrations (Munch *et al.*, 1996). Béline *et al.* (1999) and Peng and Qi (2007) observed the occurrence of SND within a biofilm with DO concentrations of 0.1 - 0.2 mg/l and 0.3 - 0.8 mg/l respectively. Munch *et al.* (1996) noted that SND can occur as a result of DO concentration gradients within microbial flocs and biofilms. In other cases, aerobic denitrifiers as well as heterotrophic nitrifiers have been shown to contribute to SND. Moreover, nitrification under fully anaerobic conditions has been shown to be possible (Robertson, 1994). SND offers a significant advantage over conventional separated nitrification and denitrification processes (Yang and Yang, 2011). The benefits of SND include: (i) SND can eliminate the need for two separate reactors, (ii) enhanced nitrogen removal efficiency can be achieved at lower C/N ratios as carbon substrates are available for nitrifiers (Hocaoglu *et al.*, 2011), (iii) it is estimated that the

SND process utilises 22-40% less carbon source and reduces sludge yield by 30% when compared with the conventional biological nitrogen removal (Seifi and Fazaelpoor, 2012).

#### 2.5.2.5 *Environmental factors affecting nitrification and denitrification*

Autotrophic nitrifying biomass is approximately 10 times more sensitive to variations in temperature and toxic compounds than their aerobic heterotrophic counterparts (Vanrolleghem *et al.*, 1996; Juliastuti *et al.*, 2003). Autotrophic nitrification is generally accepted to be the slowest step in the nitrogen removal process. It is more sensitive to un-ionized ammonia, pH, temperature variations, toxic compounds, oxygen and metals than denitrification processes (Hockenbury *et al.*, 1977; Vanrolleghem *et al.*, 1996). As a heterotrophic process, denitrification tends to be less sensitive to environmental parameters than nitrification; however, it is still affected by variations in dissolved oxygen (DO), the availability of organic carbon, and the presence or accumulation of NO<sub>2</sub> and NH<sub>4</sub> (Islam *et al.*, 2009).

##### 2.5.2.5.1 *DO*

The oxidation of 1 g of TAN (total ammonia nitrogen – NH<sub>4</sub> and NH<sub>3</sub>) requires approximately 4.57 g of oxygen (Tchobanoglous *et al.*, 2004). Low oxygenation rates result in reduced nitrification rates. At DO concentrations of less than 2 mg O<sub>2</sub>/l, the nitrification rate would be less than half the achievable rate if adequate DO was available (Henze, 2008).

The presence of dissolved oxygen in mixed liquor inhibits the development of denitrification by repressing the nitrate reduction enzyme (Tchobanoglous *et al.*, 2004; van Haandel and van der Lubbe, 2007). It is difficult to quantify this influence because concentration gradients of dissolved oxygen will develop in the flocs so the micro-environment in a floc may be different from the bulk of the liquid phase. However, it has been observed that a dissolved oxygen concentration of more than 0.2 to 0.5 mg O<sub>2</sub>/l reduces denitrification rates significantly (van Haandel and van der Lubbe, 2007).

##### 2.5.2.5.2 *Carbon*

The COD/N ratio is one of the main parameters in the design of nitrification and biological nitrogen removal processes in wastewater treatment (Carrera *et al.*, 2004). For optimisation of the nitrification process, the majority of the carbon must be removed before nitrifying organisms can become established (Tchobanoglous *et al.*, 2004).

In denitrification processes, an external biodegradable carbon source may need to be added to the wastewater to improve the nitrogen removal efficiency (Kargi and Uygyr, 2003; van Haandel and

van der Lubbe, 2007; Hayden and Gu, 2008). A readily biodegradable source of organic matter is preferred since this can give the highest denitrification rates (Æsøy *et al.*, 1998). Commonly used external carbon sources include methanol, ethanol, sugar and sludge digestion supernatant. Methanol is one of the most commonly used carbon sources as it is readily available, gives a high denitrification rate and excess amounts can easily be removed biologically (McCarty *et al.*, 1969; Christensen and Harremoës, 1977; Æsøy *et al.*, 1998). Other alternative carbon sources that have been proposed recently include MicroC<sup>TM</sup>, molasses, glycerol and corn starch (Hayden and Gu, 2008).

Determining the optimal ratio of carbon and nitrogenous substrates is critical due to the competition between autotrophic bacteria and heterotrophic bacteria, which are dominant members in the microbial ecology. The addition of excess carbon will manifest itself in the form of increased carbon content in the effluent wastewater known as “carbon breakthrough”, while if the addition carbon is insufficient, optimal denitrification will not be achieved (Xia *et al.*, 2008). In batch reactors, an anoxic period can be programmed at the beginning of the treatment cycle, thus utilising the influent wastewater as the source of carbon in many cases, eliminating the need for the addition of an external source of carbon. If additional carbon is required, the necessary carbon can be added to the influent wastewater. Table 2.6 presents the carbon:nitrogen (COD/N) ratios used by researchers when treating a variety of wastewaters.



Table 2.6 Carbon to nitrogen ratios reported in the literature and nitrogen removal performance

Study	Treatment process	Wastewater treated	Carbon: Nitrogen <sup>1</sup>	External carbon source <sup>2</sup>	Removal performance	Comments
Kumar <i>et al.</i> , 2012	Pilot scale MBR	Industrial wastewater	COD/N - 9.3	Sodium acetate	TN removal - 81%	-
Wang <i>et al.</i> , 2012	ASBR + SBR	Landfill leachate	COD/TN - 4	Raw leachate	TN removal - 50%	Temperate controlled at 25°C; 36 h treatment cycle
Wu <i>et al.</i> , 2009 b	UASB + A/O reactor + SBR	Landfill leachate	COD/TN - 4 BOD/TN - 2	-	TN removal - 80%	Study conducted at 32 - 35°C
Fu <i>et al.</i> , 2008	MMBR	SHSW	COD/N - 9.3	Sodium carbonate	TN removal - 83%	Study conducted at 25°C
Qingjuan <i>et al.</i> , 2008	AIC-MBR	Synthetic domestic WW	COD/TN - 10	-	TN removal - 73%	Study conducted at 25°C; COD/N ratio changed by altering the TN
Xia <i>et al.</i> , 2008	SCBR	Synthetic domestic WW	COD/N - 5	Glucose	TN removal - 78%	Study conducted at 15 - 22°C
Ryu <i>et al.</i> , 2007	BAF	Municipal WW	COD <sub>i</sub> /TKN - 4.3	-	TKN removal - 90%	Study conducted at 21°C
Fontenot <i>et al.</i> , 2007	SBR	Shrimp aquaculture wastewater	COD/TN - 10	Molasses	TN removal - 96%	Study conducted at >22°C
Petal <i>et al.</i> , 2006	CFBR	MWW	COD/TKN - 8	-	TKN removal - 95%	Study conducted at 25°C
Carrera <i>et al.</i> , 2004	Ludzack Ettinger pilot unit	HSA IWW	COD/N - 3.4	Ethanol	TN removal - 43%	Study conducted at 25°C
Gupta and Gupta, 2001	RBC	HS domestic WW	COD/N - 3.3	Sodium acetate	TN removal - 73%	Biofilm contained <i>Thiosphaera Pantotropha</i> bacterium; Study conducted at 27°C

<sup>1</sup> Optimal Carbon to nitrogen presented for studies in which the ratio was varied; <sup>2</sup> If used; AIC-MBR – Airlift internal circulation membrane bioreactor; WW – Wastewater; SCBR – Suspended carrier biofilm reactor; BAF – Biological aerated filter; SBR – Sequencing batch reactor; RBC – Rotating biological contactor; CFBR – Circulating fluidised bed bioreactor; HSA IWW - High strength ammonium industrial wastewater ; SHSW – Synthetic high strength wastewater; MMBR – Modified membrane bioreactor; ASF-BR – Air suction flow biofilm reactor; ASBR Anaerobic sequencing batch reactor; USAB – Up flow anaerobic sludge blanket; A/O reactor – Anoxic/aerobic reactor.

### 2.5.2.5.3 Alkalinity

The nitrification process consumes 7.14 g alkalinity (as CaCO<sub>3</sub>)/g NH<sub>4</sub> oxidised while the denitrification process produces 3.57 mg alkalinity (as CaCO<sub>3</sub>)/g nitrogen reduced (Tchobanoglous *et al.*, 2004; van Haandel and van der Lubbe, 2007). If a wastewater treatment plant is just operating a nitrification process, an external source of alkalinity may be required to ensure a sufficient supply of alkalinity to prevent the inhibition of the nitrification process. In a wastewater treatment plant aiming for complete nitrogen removal (nitrification and denitrification), in general the production of alkalinity (when combined with influent alkalinity) during the denitrification process should be sufficient for the nitrification process (van Haandel and van der Lubbe, 2007).

### 2.5.2.5.4 pH

The nitrification process is sensitive to the pH of wastewater. Generally, the optimum pH for nitrification ranges between 8 and 9 (US EPA, 1975). Research has shown that as pH decreases, the rate of nitrification also decreases (van Haandel and van der Lubbe, 2007). The pH of wastewater is related to the alkalinity of wastewater through a model developed by Loewenthal and Marais (1976) (van Haandel and van der Lubbe, 2007), which describes the interrelationship between alkalinity, acidity and pH in aqueous solutions as:

$$\text{Alk} = (\text{CO}_2)(10^{(\text{pH}-\text{pk}1)})(1 + 2 \times 10^{(\text{pH}-\text{pk}2)}) + 10^{(\text{pH}-\text{pk}3)} - 10^{-\text{pH}} \quad [\text{Eqn. 2.15}]$$

where: pk 1, pk2 and pk 3 are the negative logarithm of the dissociation constant of components of alkalinity (i.e. bicarbonate (HCO<sub>3</sub>), carbon dioxide (CO<sub>2</sub>) and hydroxide (OH)).

Using Equation 2.15, the alkalinity can be calculated for any pH value if the dissolved CO<sub>2</sub> is known.

### 2.5.2.5.5 Temperature

Like other biochemical reactions, nitrification activity is affected by temperature. Biological activity decreases with decreasing reaction temperature, and nitrification is limited when wastewater temperature is below 10°C (Water Environment Federation, 1998). On the other hand, high wastewater temperatures may inhibit nitrification by increasing the level of free ammonia. Temperature impacts can also be process dependant. Zhu and Chen (2002) found the impact of temperature on a fixed biofilm treatment process was less than that in suspended

growth processes, which was attributed to the domination of the mass diffusion transport process in a fixed film filter. Obaja *et al.* (2002) found that temperatures below 10°C have a negative impact on the rate of both nitrification and denitrification. Clifford *et al.* (2010) showed that the nitrification rate at 16 °C was three times the nitrification rate at 7°C.

The rate of denitrification also decreases with decreasing temperature (Henze, 2008). Denitrification process rates are generally less susceptible to temperature than nitrification processes; though a lack of oxidised nitrogen will obviously mean denitrification is limited. Illies and Mavinic (2001) found that nitrification and denitrification were seriously inhibited at a temperature of 10 °C when studying the performance of two 4-Stage Bardenpho process systems.

#### 2.5.2.5.6 *Nitrite*

NO<sub>2</sub> is an intermediary product in the nitrification and denitrification processes. If there is insufficient oxygen available for both nitrification and denitrification, complete nitrification will not be achieved, resulting in the accumulation of NO<sub>2</sub>. If NO<sub>2</sub> accumulation occurs during the nitrification process, NO<sub>3</sub> will not be produced. Thus denitrification will not occur. The accumulation of NO<sub>2</sub> has been observed in a number of studies (Hongwei *et al.*, 2009; Yang *et al.*, 2009; Ge *et al.*, 2012). Hongwei *et al.* (2009) attributed the accumulation NO<sub>2</sub> when treating a landfill leachate using a SBR to a difference in the rate of reduction of NO<sub>3</sub> and NO<sub>2</sub>.

#### 2.5.2.5.7 *Un-ionized compounds*

Free ammonia (FA) or un-ionized ammonia (NH<sub>3</sub>) and unionised nitrous acid or free nitrous oxide (FNA) can inhibit not only the activity of NOB but also the activity of AOB at high concentrations (Yoon and Kin, 2003). The degree of inhibition of AOB and NOB depends on the concentration of the nitrogen species present, the temperature and the pH (Tchobanoglous *et al.*, 2004). Studies have shown that FA can inhibit AOB at concentrations between 10 and 150 mg NH<sub>3</sub> /l and NOB at concentrations between 0.1 and 1.0 mg NH<sub>3</sub> /l (Anthonisen *et al.*, 1976); Turk & Mavinic, 1989). FNA inhibits both AOB and NOB at concentrations between 0.22 and 2.8 mg N<sub>2</sub>O/l (U.S. E.P.A., 1993).

#### 2.5.2.5.8 *Toxicity*

Nitrifying organisms are sensitive to a wide variety of organic and inorganic compounds at concentrations below those that would affect aerobic heterotrophic organisms (Tchobanoglous *et al.*, 2004). These toxic organic compounds include solvent organic chemicals, amines, proteins,

tannins, phenolic compounds, alcohols, cyanates, ethers, carbamates and benzene (Hockenbry and Grady, 1977; Vismara, 1982).

Organic compounds such as aniline, ortho-cresol and phenol have stronger inhibition influences on NOB than on AOB, so biological wastewater treatment of wastewater containing these organics might cause nitrite accumulation (Neufeld *et al.*, 1986). The extent of inhibition in the wastewater treatment process can depend on (i) the design and operation of the plant and (ii) the volume of wastewater containing the inorganic compounds relative to the volume of municipal wastewater being treated (Jonsson *et al.*, 2000).

#### 2.5.2.5.9 *Metals*

Metals have a dual effect on microbial growth; they can be either useful trace elements or process inhibitors (Juliastuti *et al.*, 2003). Heavy metals, such as chromium (Cr), nickel (Ni), copper (Cu), zinc (Zn), lead (Pb) and cadmium (Cd), have been shown to inhibit both AOB and NOB during nitrification (Vismara 1982; Grunditz *et al.*, 1998; Hu *et al.*, 2002; Kelly *et al.*, 2004). Metals such as Ni and Cr have been found to inhibit heterotrophic bacteria during denitrification processes (Dilek *et al.*, 1998; Stasinakis *et al.*, 2002).

## 2.6 Phosphorus removal

Phosphorus is a vital constituent of wastewater for cell synthesis in biological wastewater treatment systems, with a C:P ratio of 90-150:1 required for optimum microorganism growth (Gray, 2004). In domestic wastewaters, this ratio is often exceeded, with the remaining phosphorus discharged in the effluent unless the treatment process is configured to achieve phosphorus removal. The removal of phosphorus from wastewater involves the incorporation of phosphorus into total suspended solids (TSS) and the subsequent removal of these solids (Tchobanoglous *et al.*, 2004). Excess phosphorus in the aquatic environment can accelerate the growth of aquatic plants and can thus result in eutrophication occurring in the receiving water body (Tchobanoglous *et al.*, 2004). These nuisance growths of algae adversely affect water quality, impair downstream water treatment processes and restrict recreational activities in the vicinity (Greene and Hayes, 1981). The common forms of phosphorus found in aqueous solutions include orthophosphate, polyphosphate and organic phosphate. Phosphorus found in wastewater can be divided into the following components (Henze *et al.*, 2000):

$$C_{TP} = S_{PO_4-P} + S_{p-P} + S_{org,P} + X_{org,P} \quad [\text{Eq. 2.16}]$$

where:  $C_{TP}$  is the total phosphorus;  $S_{PO_4}$  is the dissolved inorganic orthophosphate phosphorus;  $S_{p-P}$  is the dissolved inorganic polyphosphate;  $S_{org,P}$  is the dissolved organic phosphate; and  $X_{org,P}$  is the suspended organic phosphorus.

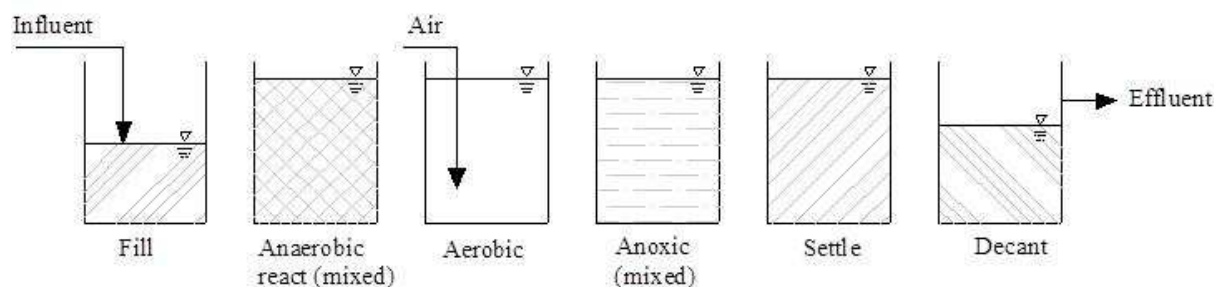
The concentrations of organic phosphorus in municipal wastewater tend to be small (though volumes are large), but it can be an important constituent of industrial wastewaters and waste sludge (Tchobanoglous *et al.*, 2004). Phosphorus removal techniques fall into three main categories: physical, chemical and biological. Biological phosphorous removal was the only phosphorus removal mechanism which could have been a factor in this study and thus is outlined briefly below.

### **2.6.1 Biological phosphorus removal**

Biological phosphorus removal systems are increasingly being incorporated in sewage treatment works (Stratful *et al.*, 1999). Biological phosphorus removal is a complex wastewater process, whereby net phosphorus removal is achieved when the aerobic phosphorus uptake used for growth and cell maintenance is greater than the sum of the phosphorus present in the influent and phosphorus released anaerobically (Brdjanovic *et al.*, 1998). Biological phosphorus removal (BPR) from wastewaters is based on the enrichment of activated sludge with phosphate accumulating organisms (PAOs) (Brdjanovic *et al.*, 1998; Wagner and Loy, 2002a). The BPR process is primarily characterised by circulation of activated sludge through anaerobic and aerobic phases, coupled with the introduction of influent wastewater into the anaerobic phase (Wagner and Loy, 2002b).

Such processes to achieve phosphorus removal are also known as enhanced biological phosphorus removal processes (EBPR). The process of EBPR is illustrated in Figure 2.4. Studies have reported phosphorus removals of up to 90% using EBPR (Water Environment Federation, 1998).

There are several plants configurations capable of achieving EBPR. One of the options for biological phosphorus removal using a SBR is illustrated in Figure 2.3.



**Figure 2.3 Biological phosphorus removal using a SBR (Adopted from Davis, 2010)**

### 2.6.2 Factors affecting biological phosphorus removal

The stability and efficiency of biological phosphorus removal (BPR) processes can be affected by several factors, such as the type of carbon substrate, pH, excessive aeration and the phosphorus load (Brdjanovic *et al.*, 1998; Mulkerrins *et al.*, 2004).

#### 2.6.2.1 DO and $NO_3$

DO and  $NO_3$  have been found to have both a positive and negative effect on EBPR (Janssen *et al.*, 2002). During the aerobic/anoxic period, DO and  $NO_3$  are needed to store the phosphorus biologically within the activated sludge organisms. In the anaerobic period, oxygen and nitrate input via the influent and/or the return streams will disturb EBPR. The magnitude of the impact of DO and  $NO_3$  is determined by the concentrations of these components in the influent and the return streams and the flows of these streams. The introduction of  $NO_3$  into the anaerobic tank can result in the growth of denitrifying bacteria which utilise the available carbon more efficiently, which will affect the amount of carbon available for EBPR (Janssen *et al.*, 2002). The DO present in the influent wastewater or the return stream will result in the oxidation of the available carbon, reducing the carbon available for EBPR (Janssen *et al.*, 2002; Tchobanoglous *et al.*, 2004).

#### 2.6.2.2 Carbon

EBPR relies on the enrichment of the activated sludge system with phosphorus accumulating organisms (PAO) (Lopez-Vázquez *et al.*, 2008). Under anaerobic conditions, PAO utilise carbon as a source of energy for growth during the subsequent aerobic/anoxic periods (Tchobanoglous *et al.*, 2004). Carbon is the primary source of volatile fatty acids (VFAs), which constitute the substrate for PAOs (Janssen *et al.*, 2002).

### 2.6.2.3 pH

Enriched PAO sludge's with a higher ambient pH can result in a higher anaerobic P release (Smolders *et al.*, 1994b; Liu *et al.*, 1996; Bond *et al.*, 1999; Filipe *et al.*, 2001a). Smolders *et al.* (1994a) found that the ratio of anaerobic P release to acetate uptake varied linearly from 0.25 to 0.75 (P:C) when pH rose from 5.5 to 8.5. A low pH results in a slow P release/acetate uptake rate, implying that at a low pH, more acetate is required to remove the released phosphorus (Janssen *et al.*, 2002). At higher pH values (>7.5), the solubility product of one or more of the phosphorus compounds will be exceeded, resulting in the precipitation of the phosphorus (Janssen *et al.*, 2002).

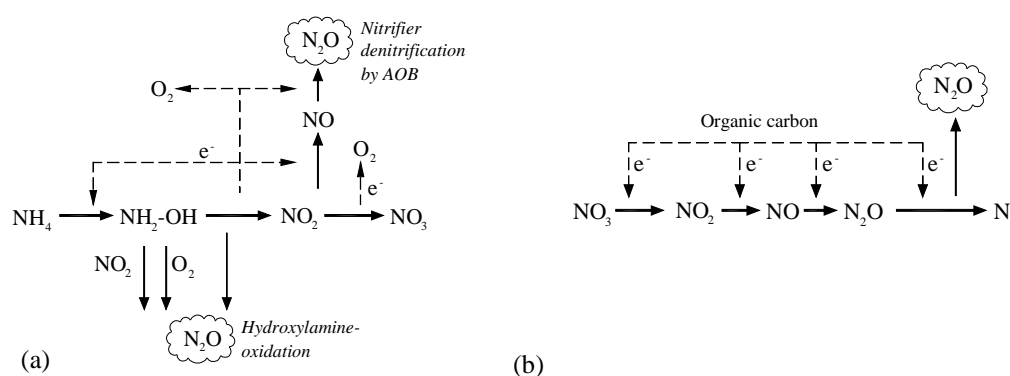
## 2.7 Production of gases during the biological treatment of wastewater

Biological wastewater treatment is responsible for the production of a variety of greenhouse gases in particular carbon dioxide (CO<sub>2</sub>), methane (CH<sub>4</sub>) and nitrous oxide (N<sub>2</sub>O), which are the principal greenhouse gases (IPCC, 2006). The global warming potential (GWP) is a relative measure of the amount of heat trapped by a certain mass of the gas in question to the amount of heat trapped by a similar mass of carbon dioxide. The GWP of CO<sub>2</sub>, CH<sub>4</sub> and N<sub>2</sub>O gases are 1, 23 and 296, respectively (European Commission, 2001, Houghton *et al.*, 2001). The significantly higher GWP of N<sub>2</sub>O means that even small emissions of N<sub>2</sub>O can have an adverse impact on the carbon footprint of a wastewater treatment plant. N<sub>2</sub>O can be produced by both heterotrophic denitrifying organisms (Kampschreur *et al.*, 2009; Lu and Chandran, 2010) and autotrophic AOB (Tallec *et al.*, 2006; Kampschreur *et al.*, 2009). According to Kampschreur *et al.* (2009), there are three main mechanisms which result in the production of N<sub>2</sub>O:

- *Hydroxylamine oxidation*: During nitrification, the first step in the nitrification process, hydroxylamine is produced as an intermediary product (Equation 2.17) (Bhunja *et al.*, 2010). N<sub>2</sub>O can be produced from: (i) the intermediates of biological hydroxylamine oxidation, which can be attributed to the imbalanced metabolic activity of AOB (Yu *et al.*, 2010), (ii) through the chemical decomposition of hydroxylamine, and (iii) chemical oxidation with NO<sub>2</sub> as an electron acceptor (Stüven *et al.*, 1992).



- Nitrifier denitrification:** During autotrophic nitrification,  $N_2O$  can be produced as a side-product in the catabolic pathway (oxidising ammonia to nitrite), or, alternatively, by denitrification of nitrite with ammonia, hydrogen or pyruvate as the electron donor (Colliver and Stephenson 2000; Wrage *et al.*, 2001). This reaction is carried out by only one group of microorganisms, namely autotrophic ammonia oxidizers (Wrage *et al.*, 2001). Könneke *et al.* (2005) found that ammonium oxidation can also be performed by ammonium oxidising archaea. Such ammonia oxidising archaea (AOA) were found to occur in WWTPs that were operated with low DO conditions and long solid retention times (Park *et al.*, 2006);
- Heterotrophic denitrification:**  $N_2O$  gas is produced as an intermediate product of the denitrification process. See Equation 2.14. The first reaction in the denitrification process converts  $NO_3$  to  $NO_2$  by nitrate reductase, the  $NO_2$  is subsequently catalysed by nitrite reductase. The conversion of  $NO$  to  $N_2O$  is catalysed by bacterial  $NO$  reductase (Yi *et al.*, 2008). In the final stage of the process,  $NO$  and  $N_2O$  are reduced to dinitrogen by  $NO$  reductase (Bhunia *et al.*, 2010). While some of the reactions which occur during the denitrification process are not fully understood, the emission of  $N_2O$  during denitrification can occur when the availability of oxidant exceeds the availability of reductant (most commonly organic carbon), resulting in the emission of  $N_2O$  (Firestone and Davidson, 1989; Itokawa *et al.*, 2001). The production of  $N_2O$  by heterotrophic denitrifiers can also occur due to oxygen inhibition (Lu and Chandran, 2010) or nitrite accumulation (von Schulthess *et al.*, 1994). It is generally believed that nitrifier denitrification by AOB is the main  $N_2O$  production pathway in biological wastewater treatment under aerobic conditions (Colliver and Stephenson, 2000).



**Figure 2.4  $N_2O$  production pathways in biological wastewater treatment. (a) – autotrophic nitrification with  $N_2O$  production via nitrifier denitrification or hydroxylamine oxidation. (b) – Heterotrophic denitrification,  $N_2O$  is an obligate intermediate in the nitrogen reduction process (Adapted from Wunderlin *et al.*, 2010)**



The IPCC (2006) guidelines state that 0.035% of the total nitrogen load of a wastewater treatment plant (WWTP) is emitted as N<sub>2</sub>O. Recent evidence indicates that emissions of N<sub>2</sub>O from some wastewater treatment plants may be significantly higher than previous estimates (Casey, 2010).

Kampschreur *et al.* (2008) found significant variations in the fraction of nitrogen that is emitted as N<sub>2</sub>O, both in laboratory scale (0 – 95% of the nitrogen load) and full-scale (0 – 14.6% of the nitrogen load). In a review of studies on N<sub>2</sub>O emissions from full-scale and lab-scale wastewater systems with BNR, Foley and Lant (2008) found that for low to medium strength municipal systems, the median emission rate was 0.01 kg N<sub>2</sub>O /kg N<sub>influent</sub> (range: 0.0003–0.03 kg N<sub>2</sub>O /kg N<sub>influent</sub>). Quan *et al.* (2012) found that an increase in the C/N ratio resulted in a decrease in the N<sub>2</sub>O emissions.

The emission of N<sub>2</sub>O from WWTPs is relatively small, accounting for an estimated 3% of anthropogenic N<sub>2</sub>O emissions, but accounting for a significant portion 26% of the greenhouse gas footprint, of the total water chain (Kampschreur *et al.*, 2009). Wunderlin *et al.* (2010) noted that N<sub>2</sub>O emissions dominate total greenhouse gas emissions from biological wastewater treatment.

The introduction of the WFD has resulted in a mandating for a reduction in the concentration of nitrogen emitted from wastewater treatment plants. To achieve the specified targets in an efficient manner will require the increased use of combined nitrification and denitrification, which will result in the inevitable and unavoidable increase in the amount of N<sub>2</sub>O arising from wastewater treatment (Kampschreur *et al.*, 2008; Bhunia *et al.*, 2010). Table 2.7 presents a sample of the N<sub>2</sub>O emissions reported in the literature.

**Table 2.7 N<sub>2</sub>O emission factors reported in literature**

Reference	Treatment process	Wastewater	C/N ratio	Scale	N <sub>2</sub> O emissions	
					Factor	Analysis method
Aboobakar <i>et al.</i> , 2013	Nitrifying plug flow ASP	Municipal wastewater	10	Operational treatment plant	0.036% influent TN	Continuous online monitoring of dissolved N <sub>2</sub> O
Daelman <i>et al.</i> , 2013	Nitrifying/denitrifying ASP	Municipal wastewater	-	Full scale	2.8% of influent nitrogen	Gaseous and aqueous N <sub>2</sub> O measured
Lotito <i>et al.</i> , 2012	Nitrifying/denitrifying ASP	Municipal wastewater	-	Pilot scale ASP	Daily averages 0.12-0.36% depending on operational configuration	N <sub>2</sub> O emission measured continuously online
Wunderlin <i>et al.</i> , 2012	Nitrifying/denitrifying ASP	Municipal wastewater	-	Pilot scale	Between 0.2% and 19% of nitrogen conversion rate, depending on operational conditions	N <sub>2</sub> O emission measured continuously online
Rassamee <i>et al.</i> , 2011	Anoxic/aerobic BNR	Municipal wastewater	1.2	Laboratory scale	Between 0 and 0.4% of TN, depending on DO	Online monitoring of liquid N <sub>2</sub> O and estimation of gaseous emissions using an equation derived from aqueous N <sub>2</sub> O
Quan <i>et al.</i> , 2012	Granular sludge SBR	Synthetic piggery wastewater	4.5 - 9	Laboratory scale	3.2 - 6.0% of influent nitrogen, depending in C/N ratio	Online measurement of dissolved N <sub>2</sub> O
Tallec <i>et al.</i> , 2006	Nitrifying ASP	Municipal wastewater	3.2*	Laboratory scale	0.4% of the reduced NO <sub>3</sub>	N <sub>2</sub> O emission measured continuously online

ASP – Activated sludge process; BNR – Biological nitrogen removal; \* C/TKN

## 2.8 Biofilm wastewater treatment technologies

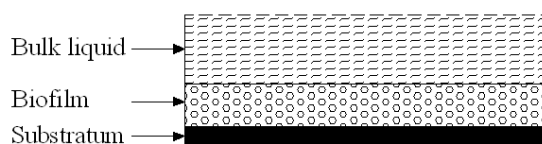
### 2.8.1 Biofilms in wastewater treatment

A biofilm refers to the colony of microorganisms attached to the surface of inert packaging material, such as rock, gravel or a variety of plastic or other synthetic materials (O'Toole *et al.*, 2000). Biofilms are complex coherent structures of cells and cellular products, such as extracellular polymers, which can grow on a static solid surface (static biofilms) or on suspended carriers (particle supported biofilms) (Heijnen, 1989; Characklis, 1982; Nicoletta, 2000). The organic contaminants and nutrients in the wastewater are removed by these microorganisms through the process of cell synthesis and biological conversion as the wastewater flows over and through the biofilm. Biofilms can be operated under aerobic, anoxic and anaerobic conditions (Henze, 2008).

Biofilm wastewater treatment systems have been used successfully used for the treatment of wastewater since the late 19<sup>th</sup> century (Lofrano & Brown, 2010).

### 2.8.2 Biofilm growth and structure

Biofilm growth occurs when microbial cells attach themselves to a surface of a solid support substratum in an aquatic environment (Figure 2.5).



**Figure 2.5 Biofilm system constituent compartments**

When sufficient quantities of the nutrients are available, they multiply in clusters and produce extracellular polymeric substances (EPS). For each 100 g of carbon utilised in the biofilm synthesis, approximately 7 g of nitrogen and 1 g of phosphorus are required (Tchobanoglous *et al.*, 2004). Biofilm development has a number of stages, including the transport of cells and nutrients to the substratum, attachment of cells to the substratum, growth and other metabolic processes, and detachment of portions of the biofilm. The substrates in the wastewater flowing through the biofilm are consumed by the microorganisms (Characklis, 1982; O'Toole *et al.*, 2000).

A biofilm will develop on most surfaces subject to a liquid flow and is a result of several chemical and biological factors, such as: (i) the transportation and absorption of the organic molecules to the surface, (ii) transport of the microbial cells to the surface, (iii) microorganism

attachment to the surface, (iv) microbial transformation (growth and exopolymer production) at the surface, resulting in the production of biofilm, and (v) the partial detachment of the biofilm caused by fluid shear stress. The rate of biofilm growth depends on the diffusion of nutrients into the biofilm followed by their synthesis into the attached biomass (Characklis, 1982). Hydrodynamic conditions can cause the biofilm to be sheared away from the supporting media and become entrained in the fluid. In thicker, less dense biofilms, sloughing can occur, where massive biofilm removal occurs and is attributed to a depletion of nutrients or oxygen in the inner layers of the biofilm (Characklis, 1982).

The overall biofilm substrate removal depends on the reaction rate associated with the microbiological metabolisms in the biofilm and the transfer of substrates and oxygen into, out of and within the biofilm (Rodgers, 1995). The substrate transfer into the biofilm can occur via advection or diffusion. Advection is defined as the horizontal or vertical transfer of a substrate due to the fluids bulk motion. Diffusion is defined as the non-advective migration of a substance in solution or in suspension in response to the concentration gradient of the substance. It happens at the molecular level and is described using Fick's first law of molecular diffusion (Tchobanoglous *et al.*, 2004). The substrate concentration inside a biofilm is determined by the amount of substrate transferred into the biofilm and the amount of the substrate consumed by the microbiological metabolism.

$$\text{Fick's first law of diffusion: } r = -D_m \frac{\partial C}{\partial x} \quad [\text{Eq. 2.18}]$$

where:  $r$  is the rate of mass transfer per unit area per unit time;  $D_m$  is the coefficient of molecular diffusion in the  $x$  direction;  $C$  is the concentration of constituent being transferred; and  $x$  is the distance.

The internal features of biofilms differ greatly with age, thickness, density, porosity and tortuosity (Hibiya *et al.*, 2004). In particular, the substrate and microbial distributions in biofilms change with biofilm thickness. These distributions are closely related to mass transfer phenomena (Nicolella *et al.*, 1999). A wastewater biofilm is highly stratified. As depth increases so too does the biofilm density, while porosity, metabolically active biomass and oxygen diffusivity all decrease (Bishop *et al.*, 1995). The diffusion coefficient in biofilms, which is lower than that in water, is influenced by their density, porosity, pore size, convection, type of EPS and minerals (Wimpenny *et al.*, 2000). The composition of the microorganisms in the biofilm is

determined by the type of available substrate while the structure of the biofilm is dictated by the constituent microorganisms.

### **2.8.3 Biofilm based technologies**

Biofilm processes can be divided into two categories: (i) fixed media systems, such as the trickling filter (TF), and submerged fixed bed biofilm reactors (SFBBR), where the media remain static within the reactor while the water is moved over the biofilm and (ii) moving media systems, such as rotating biological contactors (RBCs), moving bed biofilm reactors (MBBRs), fluidised bed reactors (FBRs), and moving bed biofilm reactors (MBBRs), where the media is kept in constant motion by aeration, mechanical stirrers and hydraulic forces. Sequencing batch biofilm reactor systems can be either fixed or moving bed reactors, depending on the configuration and media used in the unit (Tchobanoglous *et al.*, 2004). Because of their simple, reliable and stable characteristics, biofilm processes have been used in the removal of nutrients and other pollutants from both domestic and industry wastewater (Xia *et al.*, 2008).

Biofilm based technologies can offer several advantages over conventional treatment processes, including: (i) increased energy efficiency, (ii) better recovery from shock loads, and (iii) reduced sludge production when compared to activated sludge systems (O' Reilly *et al.*, 2011).

### **2.8.4 Development of Sequencing Batch Biofilm Reactors (SBBRs)**

The SBBR can be seen as a combination of a biofilm-based technology with an activated sludge process – the sequencing batch reactor (SBR). Aeration and sedimentation processes are carried out sequentially in the same reactor rather than simultaneously in separate reactors as occurs in the conventional activated sludge system. SBRs have been widely advocated as one of the most promising options for the biological treatment of wastewaters (Kurniawan *et al.*, 2010).

The SBBR combines the advantages of the SBR process with the benefits of biofilm systems. The result is a treatment process which: (i) has a higher biomass concentration in the reactor, (ii) is capable of treating a higher volumetric load, (iii) offers improved performance stability, and (iv) results in higher specific removal rates (Di Iaconi *et al.*, 2006; Xiao *et al.*, 2008).

The accumulation of biomass on the carrier media in SBBRs provides an additional biomass inventory as opposed to a conventional activated sludge where the only biomass consists of flocculent biomass. The high biomass concentrations that can be achieved by a SBBR system result in shorter hydraulic retention times (HRT) in comparison to ordinary SBR systems with

equivalent solid retention time (SRT) (Lessel, 1994; Goh *et al.*, 2009). The shorter HRT can result in smaller reactor size and greater treatment capacity at the same reactor size.

### 2.8.5 Main biofilm technologies used in secondary treatment of wastewater

Extensive research into wastewater treatment processes has led to the development of a number of different treatment processes, each with a number of variants, with the objective of increasing the efficacy of the treatment process. Table 2.6 outlines the advantages and disadvantages of some of the most commonly used secondary treatment processes.

**Table 2.8 The main secondary wastewater treatment processes (O'Reilly, 2011; Clifford, 2009; Cortez *et al.*, 2008; Rodgers *et al.*, 2006b; Tchobanoglous *et al.*, 2004; Di Iaconi *et al.*, 2003; Lekang and Kleppe, 1999; Ødegaard, 1999)**

	Advantages	Disadvantages
Trickling filters	<ul style="list-style-type: none"> <li>○ Low operating costs.</li> <li>○ Excellent SS, BOD removal and nitrification.</li> <li>○ Does not require skilled operators.</li> <li>○ Capable of handling shock and toxic loads.</li> <li>○ Moderate odour problems.</li> <li>○ Robust.</li> <li>○ Low sludge yields.</li> </ul>	<ul style="list-style-type: none"> <li>○ Prone to clogging if overloaded.</li> <li>○ High capital costs.</li> <li>○ Low flexibility of operation.</li> <li>○ Addition processes required to achieve denitrification and phosphorus removal.</li> <li>○ Can suffer from fly nuisance.</li> </ul>
Rotating biological contactors (RBCs)	<ul style="list-style-type: none"> <li>○ Low sludge production.</li> <li>○ Low operating costs.</li> <li>○ Process flexibility.</li> <li>○ Good sludge settleability.</li> <li>○ Good BOD removal and nitrification.</li> <li>○ Low plant footprint.</li> </ul>	<ul style="list-style-type: none"> <li>○ High capital costs.</li> <li>○ High maintenance requirement.</li> <li>○ Requires skilled operators.</li> <li>○ Addition processes required to achieve denitrification and phosphorus removal.</li> </ul>
Sequencing batch biofilm reactors (SBBRs)	<ul style="list-style-type: none"> <li>○ Can be configured to achieve nitrification, denitrification and phosphorus removal,</li> <li>○ Large operational flexibility.</li> <li>○ All stages of the treatment process occur in a single reactor saving space.</li> <li>○ Low maintenance requirement.</li> <li>○ Control over mixed liquor suspended solids and solids retention time.</li> </ul>	<ul style="list-style-type: none"> <li>○ Can be difficult to adjust cycle times for small communities.</li> <li>○ Potential for high energy requirement where forced aeration is used.</li> <li>○ Emission of noxious odours and greenhouse gases,</li> <li>○ Potentially high energy requirement for aeration and mixing.</li> <li>○ Balance tank required</li> </ul>

**Table 2.8 The main secondary wastewater treatment processes (cont'd)**

Moving bed biofilm reactors (MBBRs)	<ul style="list-style-type: none"> <li>○ Reduced area requirement.</li> <li>○ Stability under varying load.</li> <li>○ Single pass treatment.</li> <li>○ Flexible design that allows for increased capacity.</li> </ul>	<ul style="list-style-type: none"> <li>○ Clogging of the media.</li> <li>○ Sloughing of the biofilm from the media.</li> <li>○ Emission of noxious odours and greenhouse gases.</li> <li>○ Potentially high energy requirement for aeration and mixing.</li> </ul>
Horizontal flow biofilm reactors (HFBRs)*	<ul style="list-style-type: none"> <li>○ Low energy requirement</li> <li>○ Higher sludge age which benefits slow growing bacteria</li> <li>○ Reduced sludge production</li> </ul>	<ul style="list-style-type: none"> <li>○ Suitable for small scale wastewater treatment only</li> <li>○ Additional design/processes required for phosphorous removal</li> <li>○ Emission of noxious odours and greenhouse gases.</li> </ul>
Pumped flow biofilm reactors (PFBRs)*	<ul style="list-style-type: none"> <li>○ Low energy requirement</li> <li>○ Low maintenance requirement</li> <li>○ Low sludge yield</li> <li>○ Ease of operation</li> <li>○ Stable process</li> </ul>	<ul style="list-style-type: none"> <li>○ Emission of noxious odours and greenhouse gases.</li> <li>○ Suitable only for communities above 50 PE</li> <li>○ As with all batch systems a balance volume is required</li> </ul>

\* Treatment technology developed by the Environmental Engineering Department of NUI Galway

## 2.9 Challenges for the wastewater treatment sector

The European Commissioner for the Environment has stated the need for “*innovative solutions that contribute significantly to tackling water challenges at the European and global level, while stimulating sustainable economic growth*” (Potočnik, 2012). Such technologies will have to address the issues facing wastewater management in the future. Some of the main challenges that will be faced by researchers in the wastewater sector and by wastewater facility designers and operators are outlined below:

### 2.9.1 Increase in the wastewater treatment performance

The increasingly stringent legislation, such as the WFD, being enacted throughout many jurisdictions, has increased the quantity of wastewater requiring treatment while also increasing the standards of effluent wastewater required from wastewater treatment plants. Achieving these standards will require the increased usage of combined nitrification and denitrification treatment systems. As of 2012, almost half of Europe’s freshwaters have failed to meet the objectives set out in the WFD of good ecological status by 2015 (Potočnik, 2012). Failure to meet the targets

set out in the WTD results in multimillion euro financial penalties, such as those imposed in the past on Belgium, Italy, Sweden Poland, Portugal and Ireland (EU, 2013).

### **2.9.2 New regulation**

To date, wastewater treatment systems have been designed to reduce harmful emissions to receiving water bodies. Reductions in the consumption of energy originating from fossil fuels and greenhouse gas emissions to the atmosphere has attracted limited attention from facility designers and operators. Increasing the level of performance required from WWTPs will result in an unavoidable increase in the amount of N<sub>2</sub>O produced (Kampschreur *et al.*, 2009; Bhunia *et al.*, 2010). Under the Kyoto protocol, increased emissions of methane and N<sub>2</sub>O are regulated and any increase can result in financial penalties (Greenfield and Batstone, 2005). Thus new and innovative methods of achieving wastewater treatment are required which comply with legislative requirements while minimising the emissions of all gases associated with the treatment of wastewater.

### **2.9.3 Energy efficiency**

In 2006 the EU published a Green Paper “A European Strategy for Sustainable, Competitive and Secure Energy” (EU, 2006), which advocates measures to reduce energy demand through efficiency measures. It advocates energy efficiency as the most promising option for the current energy challenge.

Energy accounts for approximately 28% of the operating costs of a wastewater treatment plant in the United Kingdom, with aeration accounting for 55% of the energy used (KTN Environmental, 2008). As the energy bill is a substantial cost in the operation of a wastewater treatment plant, energy efficiency offers the opportunities for cost savings. Increasing energy prices and the continual need for local authorities to reduce their cost bases has increased interest in the possibility of saving energy in WWTP without compromising the level of treatment achieved. Efficient energy usage and procurement are crucial to improving the energy efficiency of wastewater treatment plants. A reduction in the energy used in a WWTP will also assist in the reduction of the carbon footprint of the wastewater treatment plant, which would assist in fulfilling the requirements of the Kyoto Protocol.

### **2.9.4 Technology design**

A key challenge given the need for increased plant performance and decreased operational costs, is the development of technologies that are robust and simple and facilitate flexibility in



operation and upgrade (O'Reilly *et al.*, 2008a; Geoff, 2009; Clifford *et al.*, 2009). The development of technologies that combine a modular, easily expandable design with flexibility of operation (e.g. systems that can switch from carbon removal and nitrification only to include denitrification) is a particular challenge.

### **2.9.5 The ASF-BR treatment unit**

The air suction flow biofilm reactor (ASF-BR) which was evaluated in this study, is a technology designed by a research team within the Civil Engineering Department of NUI Galway (Rodgers *et al.*, 2008a; Clifford *et al.*, 2012). The concept for the ASF-BR was derived from previous studies conducted in the Environment Engineering Laboratory of NUI Galway. The predecessors to the ASF-BR were the vertically moving biofilm reactor (Rodgers *et al.*, 2003; Rodgers and Zhan, 2004) and the pumped flow biofilm reactor (PFBR) (O'Reilly, 2005; Rodgers *et al.*, 2006a; O'Reilly *et al.*, 2008a; O'Reilly *et al.*, 2008b; Rodgers *et al.*, 2008b; Rodgers *et al.*, 2009; Wu *et al.*, 2009; Zhan *et al.*, 2006). The ASF-BR was the next step in the development of a sustainable versatile wastewater treatment unit suitable for the 21<sup>st</sup> century and aims to improve on the success of previously developed technologies.

The ASF-BR operates on similar principles as the PFBR technology, the difference being the mechanism used to move wastewater. With the PFBR, a hydraulic pump moves wastewater between two reactors, whereas the ASF-BR uses a partial vacuum to move the wastewater between two reactors. It is envisaged that the ASF-BR could offer a low energy requirement treatment process capable of offering a high level of treatment performance and an ability to control gaseous emissions. As a batch reactor, operational flexibility is a key advantage, and the system can be designed to achieve carbon removal, nitrification, denitrification and biological phosphorous removal (or required combinations thereof). The design of the ASF-BR as a modular batch biofilm reactor allows units to be prefabricated off site and added or removed to the wastewater treatment system to cater for a given PE demand. This study examines the performance of this new technology in treating a variety of wastewaters. The study also examines issues such as sludge production, operational optimisation and nitrous oxide production.

## **2.10 Summary**

This chapter presents the background to the research project. The legislation governing the treatment of wastewater is discussed. The characteristics of municipal wastewater and industrial wastewater are described, along with the processes used in the removal of contaminants from

wastewaters and the factors that affect technology performance. The chapter also summarises current biofilm based wastewater technologies. The challenges facing the wastewater sector are outlined and the ASF-BR is presented as a potentially innovative technology that can address some of these challenges.

## **CHAPTER 3**

# **Experimental apparatus and system operation**

### 3.1 Introduction

In this study, two pilot scale air suction flow biofilm reactors (ASF-BRs) were built to investigate the removal of organic carbon and nitrogen from synthetic and real wastewaters. Emissions of nitrous oxide from the ASF-BR were also monitored as was the overall potential of this new technology as an alternative to existing wastewater treatment technologies.

The ASF-BR is a dual-reactor biofilm based technology that uses a partial vacuum to circulate wastewater between the two reactors and thus achieving aeration of the biofilm. Each reactor contains plastic media to support the growth of biofilm. The sequential movement of wastewater between the two reactors allows biofilm to be alternately, and passively, exposed to air and wastewater; thus eliminating the need for mechanical aeration. The design of the unit as a batch biofilm reactor combines the advantages of a biofilm technology, with the potential of a batch treatment technology to remove: organic carbon, nitrogen and phosphorus.

This chapter details the design and construction of the experimental apparatus and the operational process of the ASF-BR employed in the pilot scale studies

### 3.2 Pilot-scale ASF-BRs

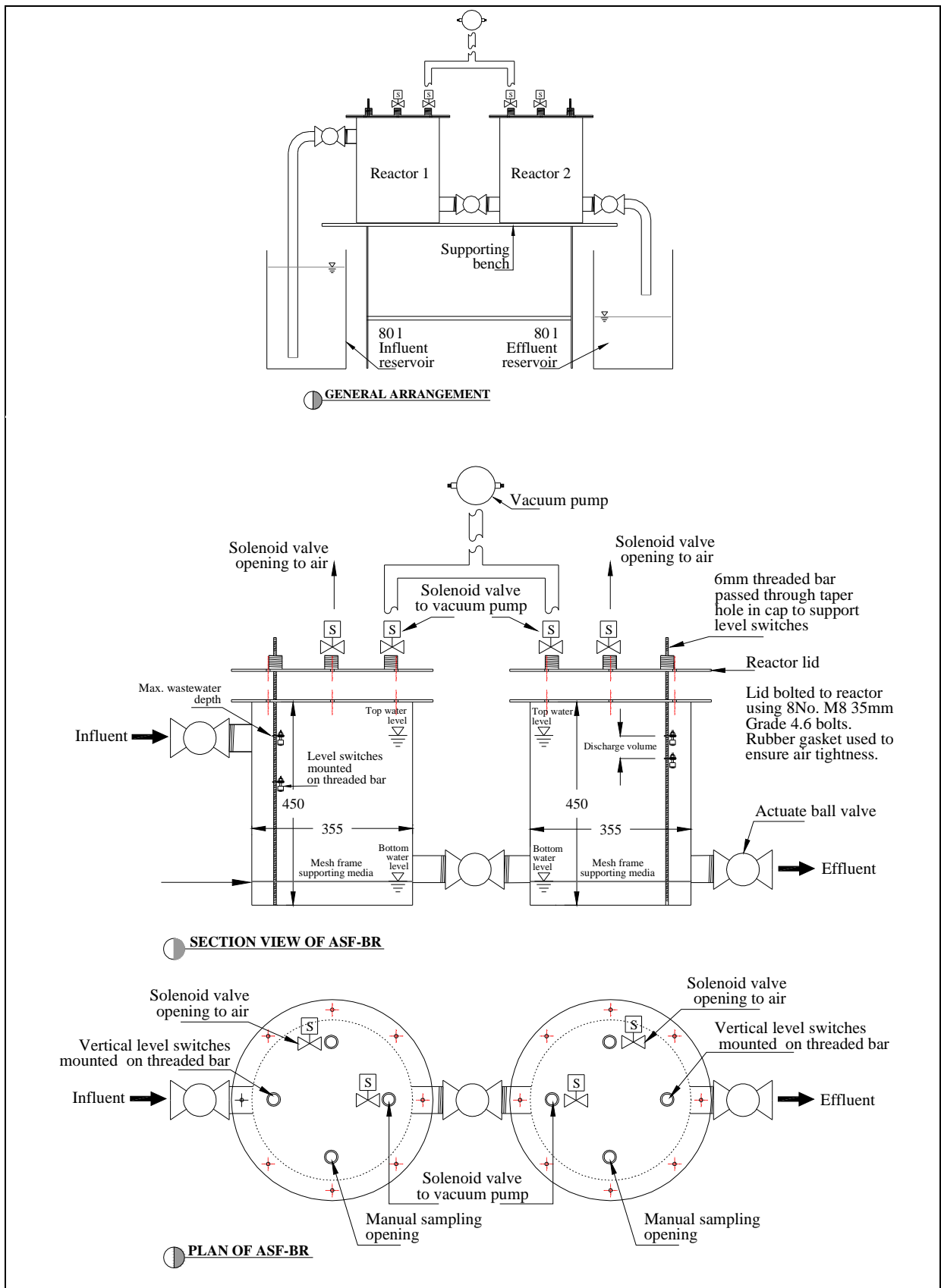
#### 3.2.1 Tank design

Each pilot scale ASF-BR comprised two cylindrical steel reactors (Alpha Tanks, Clones, Ireland) connected in series; Reactors 1 (R1) and 2 (R2) (the general arrangement of the unit is shown in Figures 3.1 & 3.2). Each reactor was manufactured from mild steel with 5 mm thick walls and measured 350 mm in diameter and 450 mm high. 5 mm thick mild steel lids, connected to the flanged rim of each reactor using 8 no. M8 bolts, were used to seal the reactors. A rubber gasket, attached to the flanged rim of each reactor, was used to ensure an air tight seal. The maximum working volume of each reactor was approximately 35 l but this was subject to change depending on the nature of the study at any given time. Two such ASF-BRs were commissioned for this study (ASF-BR 1 and ASF-BR 2)



**Figure 3.1 - Pilot scale ASF-BR commissioned within a temperature controlled room of the Environmental Engineering laboratory, NUI Galway.**

**Note:** Temperature was controlled at 11 °C which is typical of the ambient water temperature in Ireland



**Figure 3.2 Schematic of the ASF-BR**

**Note:** Drawings for illustrative purposes only, not to scale; media and level switch enclosure have been omitted for clarity.

### **3.2.2 Mechanical equipment**

#### *3.2.2.1 Vacuum pump*

In each ASF-BR wastewater was moved between R1 and R2 using a Bibus SV-50 vacuum pump (Bibus UK, Bucks, UK). For example, a partial vacuum created in R1 using the vacuum pump, caused wastewater to be drawn into R1 from R2 (and vice versa).

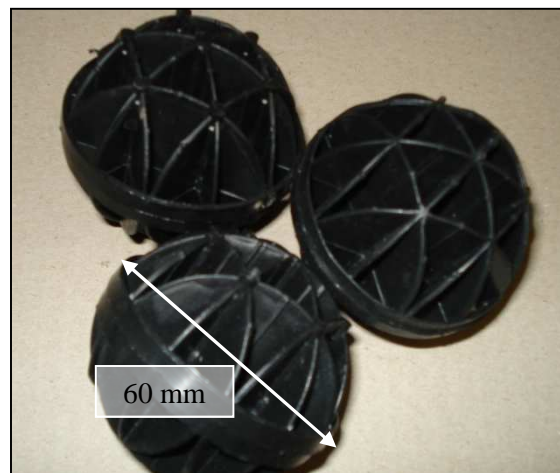
#### *3.2.2.2 Valves*

Each ASF-BR unit used seven automated valves to control the movement of wastewater and air within the unit. A 50 mm actuated ball valve (J & J Automation UK Ltd., Worcestershire, UK) was located on each of the influent and effluent lines and on the connection line between the two reactors. Two 25 mm solenoid valves (Irish Pumps and Valves Ltd., Carlow, Ireland), located on the lid of each of the reactors, were used to control air movement in and out of each reactor. A manual ball valve was located on the lid of each reactor to enable manual sampling during treatment cycles.

### **3.2.3 Plastic media**

Plastic media balls (Eloy-Fil, Belgium) were used as the biofilm support structure. The balls were chosen to allow a high surface area while simultaneously decreasing the likelihood of clogging. The spherical perforated balls each measured 60 mm in diameter (Figure 3.3). Each ball had a surface area of 0.041 m<sup>2</sup> and a specific surface area (SSA) of 310 m<sup>2</sup>/m<sup>3</sup>. A mesh frame, 50 mm in height was placed on the base of each reactor to prevent interaction between the plastic media and sludge settled at the base of each reactor.

The specific gravity of the clean plastic media was 0.9 and thus the media floated when bulk fluid was present in the reactor. It was expected this would allow the media to move and potentially decrease the risk of media clogging. The number of balls in each tank varied between experiments.



**Figure 3.3 Photograph of the spherical media balls**

### **3.2.4 System control**

The ASF-BRs were controlled using a custom program logic controller (PLC) and supervisory control and data acquisition (SCADA) system (© NUI Galway & Controlstar Systems Ltd.<sup>TM</sup>, Lincolnshire, UK). The system allowed full control over every aspect of ASF-BR operation.

#### **3.2.4.1 Program logic controller (PLC)**

A Controlstar System Ltd., CS3000 PLC was used to control the ASF-BRs. The operational program for the ASF-BR was developed by the NUI Galway research team in conjunction with Controlstar Systems Ltd. The SCADA system (including mechanical and electrical equipment control and event logging) could be accessed remotely in real-time. A graphical user interface facilitated ease of use when interacting with, and monitoring the system (Figure 3.4).

A control panel was built that allowed communication between the PLC system and the mechanical and electrical equipment controlling the ASF-BRs (e.g. opening and closing valves). The PLC was programmed to control ASF-BR units which were operated independently throughout the study.



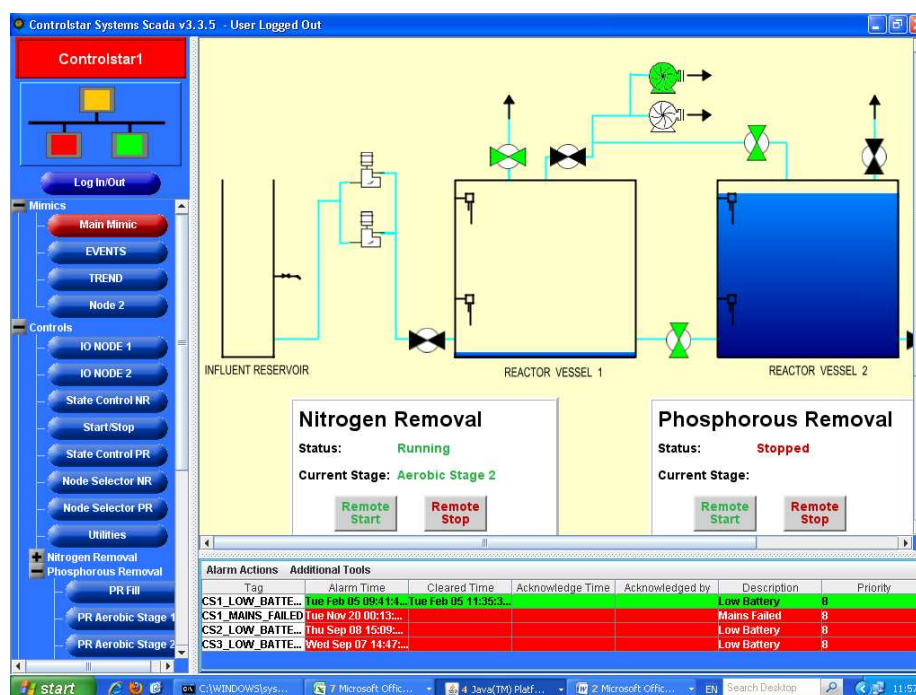


Figure 3.4 Controlstar Systems Ltd. SCADA user interface

### 3.2.4.2 SCADA programming

The PLC programme was designed to allow digital outputs (e.g. controlling valves, pumps and blowers) to be set in fixed states for a period of time, or until a certain condition was reached (such as a level switch being activated). Each state equated to a step in the treatment process.

Each of the input output (IO) function blocks consisted of a Node which in turn comprised two 4-channel analogue input modules, two 8-channel digital output modules, and one 16-channel digital input module. The initial IO configuration was as follows:

1. Digital outputs controlled the valves.
2. Digital outputs controlled the air blower.
3. Digital outputs controlled the inlet /outlet pump if required.
4. Digital inputs from the level switches.
5. Analogue inputs, which could be used in future work (for example if dissolved oxygen sensors are used to control aeration steps).

Full details on the programme developed for this study can be found in Appendix C.

### 3.2.4.3 Level switches

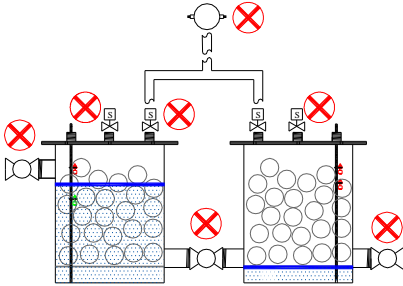
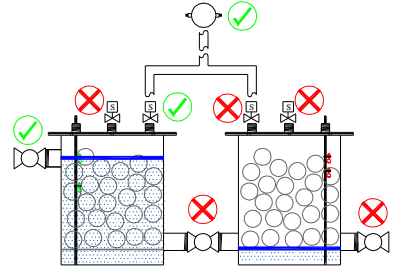
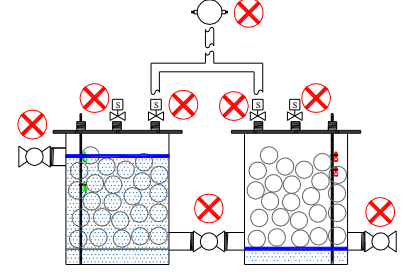
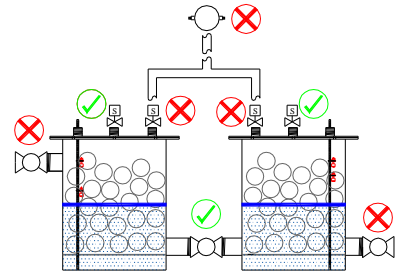
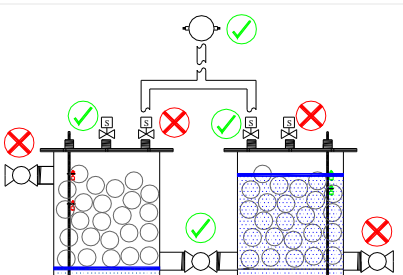

Two vertical level switches were placed at a desired depth within a perforated tube (to avoid potential damage from the moving plastic media) in each tank. In R1 the upper level switch controlled the influent volume drawn into the unit during the fill period. A “failsafe time” was used to terminate the fill period in the event of this level switch failing. The lower level switch in R1 was installed to allow for a lower water level to be set but was not required in the current study. The upper level switch in R1 and R2 was used to control maximum water levels in each tank during the aerobic period. When the upper level switches were activated, a signal was sent to the PLC to terminate the event (i.e. the vacuum pump drawing water into a given tank was switched off). This was also controlled by inputting a “failsafe time” which terminated the event if the level switch did not switch off the pump within a defined duration. The lower level switch in R2 was used to control the volume of wastewater discharged at the end of a treatment cycle.

### 3.2.5 Typical operational cycle

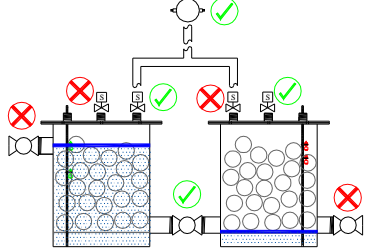
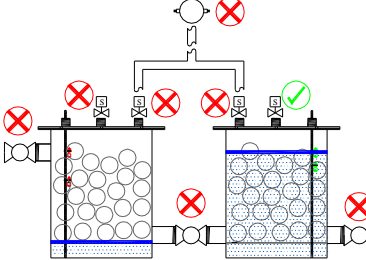
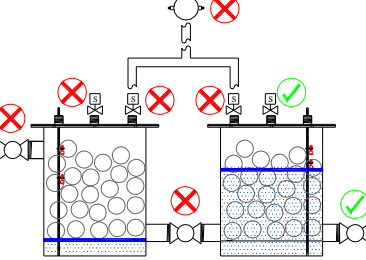

A typical ASF-BR wastewater treatment cycle is outlined in Table 3.1.

During the aerobic period, wastewater in one reactor was moved to the alternate reactor, thus exposing to the air, the biofilm accumulated on the plastic media in the empty reactor. This process was repeated as detailed above for the duration of the aerobic period. The duration for which the wastewater was held in a given reactor could be varied to effect the following for example: (i) when the wastewater was drawn and held in R1, the bulk fluid was in contact with the biofilm growing on the plastic media; (ii) the biofilm in R2 was exposed to atmospheric air and, (iii) the energy consumption of the system as energy was used in transferring the wastewater between R1 and R2. A minimum of about 5 l of wastewater was held in each reactor at all times. The overall duration of the aerobic period was defined using the aforementioned PLC and SCADA system.

**Table 3.1 Overview of a typical ASF-BR treatment cycle.**

Initial set-up	Following the discharge period, the ASF-BR equalised (Step 4). The remaining wastewater was then drawn to R1 (Step 6). Wastewater remaining in the ASF-BR following the completion of the previous treatment cycle was now in R1. The unit was then ready to begin a new cycle.	
Step 1 - Fill	Influent wastewater was drawn into the reactor under a partial vacuum until the upper level switch was tripped or a default time had elapsed.	
Step 2 – Anoxic (or anaerobic)	Wastewater was held quiescently in R1 for a pre-set time ( $t_1$ ), to allow the development of anoxic conditions suitable for denitrification to develop.	
Step 3 - Aerobic - equalisation	The valve connecting both reactors was opened along with the valve to air on the lid of each tank. Wastewater could then equalise between the two reactors by gravity.	
Step 4 - Aerobic – draw to R2	Wastewater was drawn into R2 under a partial vacuum until the top level switch was tripped. It was then held for a period of time ( $t_2$ ), in R2. This period could be set by time or controlled by a DO setpoint (or any other sensor input).	
Step 5 - Aerobic – equalisation	Repeat Step 3	

**Table 3.1 Overview of a typical treatment cycle in ASF-BR (cont'd).**

Step 6 - Aerobic – draw to R1	Wastewater was drawn into R1 until the top level switch was tripped. It was then held for a given period ( $t_3$ ) in R1. This was controlled as per Step 4. Steps 4 – 7 are then repeated a number of times to complete an aeration period. Steps 4 – 7 are considered one “pumping cycle”.	
Step 7 Settlement	Following the aeration period, the wastewater was held for a time ( $t_4$ ) under quiescent conditions in R2 to allow the settlement of biomass.	
Step 8 Decant	On completion of the settlement period the effluent valve was opened and discharge occurs until the lower levels switch was released.	
Symbol definition:		

### 3.3 Wastewaters tested

#### 3.3.1 Feed and effluent tanks

Plastic 80 l bins were used as influent and effluent wastewater tanks. The influent holding tank and the tubing which delivered the wastewater to the reactor were cleaned each time a new batch of wastewater was added, to prevent any accumulation of biofilm within the bin or the delivery tubing. The effluent holding tank and discharge tubing were cleaned each time there was a discharge of wastewater.

#### 3.3.2 High strength synthetic wastewater (HSWW)

In Study 1, HSWW was used to (i) ensure a reasonable consistency in the influent wastewater and (ii) ensure no toxins or chemicals were present which could inhibit the treatment process. The HSWW was prepared every second day in the laboratory in 40 l or 60 l batches using similar constituents to those used by Ødegaard and Rusten (1980) but, with higher

concentrations to replicate a high strength wastewater (Rodgers *et al.*, 2008c). The constituents of the HSWW are shown in Table 3.2.

HSWW was used as it: (i) is representative of raw sewage, (ii) offers consistent substrate loading and is not subject to dilution, (iii) avoids toxic pulses, and (iv) allows consistent comparison of various technologies. Using HSWW which can be produced within the laboratory removes the need to transport and store raw wastewater and avoids the issues relating to health and safety (Gray, 2004; Boeiji *et al.*, 1999). Studies conducted using the HSWW are detailed in Chapter 4.

**Table 3.2 HSWW constituents and concentrations**

Constituent	Conc.	Constituent	Conc.	Parameter	Average	Influent
	mg/l		mg/l		influent <sup>1</sup>	st. dev.
					mg/l	mg/l
Glucose	1600	KHCO <sub>3</sub>	400	COD**	2539	236
Yeast extract	240	NaHCO <sub>3</sub>	1040	COD <sub>f</sub>	2064	244
Dried milk	960	MgSO <sub>4</sub> .7H <sub>2</sub> O	400	TN**	217	32
Urea	240	FeSO <sub>4</sub> .7H <sub>2</sub> O	16	TN <sub>f</sub>	197	47
NH <sub>4</sub> Cl	480	MnSO <sub>4</sub> .H <sub>2</sub> O	16	NH <sub>4</sub> -N	167	30
Na <sub>2</sub> HPO <sub>4</sub> .12H <sub>2</sub> O	800	CaCl <sub>2</sub> .6H <sub>2</sub> O	24	PO <sub>4</sub> -P	50.3	37

<sup>1</sup>Average influent concentration over Phase 1 and Phase 2; All parameters marked \*\* were measured unfiltered, all other samples were filtered

### 3.3.3 Municipal wastewater

The municipal wastewater used in Study 2 was sourced from two municipal wastewater treatment plants. Initially the wastewater was sourced from Tuam Wastewater Treatment Plant (TWTP). This treatment plant accepts landfill leachate from a local municipal landfill site and to avoid adverse effects of landfill leachate on the treatment process, wastewater was subsequently sourced from Mutton Island Wastewater Treatment Plant, which serves Galway city and its environs. The municipal wastewater was collected on a weekly basis and stored within a temperature controlled room (11 °C) until required, as low temperatures inhibit the occurrence of biological reactions which result in the degradation of the wastewater (Tchobanoglous *et al.*, 2004). The characteristics of the municipal wastewaters used are presented in Table 3.3.

**Table 3.3 Concentrations of influent municipal wastewater**

Parameter	Phase 1a		Phase 1b		Phase 2	
	Average influent	Influent st. dev.	Average influent	Influent st. dev.	Average influent	Influent st. dev.
	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
COD**	223	114	366	103	359	82
COD <sub>f</sub>	97	41	182	86	79	29
TN**	35.6	10.0	54.3	12.1	23.2	5.1
TN <sub>f</sub>	26.8	8.1	50.8	13.0	21.5	4.5
NH <sub>4</sub> -N	20.8	5.8	49.2	15.1	21.9	5.2
PO <sub>4</sub> -P	1.6	0.7	3.4	1.0	1.1	0.5
Alk	-	-	346	39.6	219	71

All parameters marked \*\* were measured unfiltered, all other samples were filtered

### 3.3.4 Landfill leachate

The landfill leachate used in the Study 3 was sourced from a municipal landfill site located in east County Galway. The landfill has been operating since 2006 accepting 100,000 tonnes of municipal industrial and commercial waste. Samples of the landfill leachate were collected on a weekly basis. The leachate used in Study 3 is further detailed in Table 3.5.

**Table 3.4 Concentrations of influent landfill leachate**

Parameter	Phase 1		Phase 2	
	Average Influent	Influent st. dev.	Average Influent	Average Influent
	mg/l	mg/l	mg/l	mg/l
COD**	1756	443	1824	38
COD <sub>f</sub>	1512	362	1705	6
TN**	1288	157	1204	69.6
TN <sub>f</sub>	1223	153	1130	62
NH <sub>4</sub> -N	1156	180	807	83
PO <sub>4</sub> -P	5.8	2.8	7.5	0.43
Alk	5180	854	4837	478

All parameters marked \*\* were measured unfiltered, all other samples were filtered

## 3.4 Sampling and analysis

### 3.4.1 Wastewater analysis

#### 3.4.1.1 Wastewater sampling regime

Influent wastewater samples were taken each time a new batch of influent wastewater was added to the influent holding tank. Effluent samples were taken following the discharge of wastewater from the ASF-BR. All samples were taken directly from the influent or effluent

holding tank using 50 ml plastic containers. Samples were immediately filtered using Whatman 1.2  $\mu\text{m}$  micro-filters before being stored at a temperature of 4 °C until the samples were tested (within 1 to 2 days). Experimental procedures, outlined below, were carried out in accordance with standard procedures (APHA, 2005) and are detailed in Appendix B.

Regular Intensive Studies (IS) were carried out during each phase of the three studies. During an IS, wastewater samples were taken at regular intervals throughout a treatment cycle, via the manually operated valve on the lid of each reactor, to determine the changes in the concentrations of contaminants during a treatment cycle.

#### *3.4.1.2 Wastewater testing procedures*

Filtered wastewater samples were tested for ammonium nitrogen ( $\text{NH}_4\text{-N}$ ), nitrate nitrogen ( $\text{NO}_3\text{-N}$ ), nitrite nitrogen ( $\text{NO}_2\text{-N}$ ), total oxidised nitrogen (TON), ortho-phosphorus ( $\text{PO}_4\text{-P}$ ) and alkalinity using a Thermo Clinical Labsystems, Konelab 20 Nutrient Analyser. Total nitrogen (both filtered and unfiltered) was measured using Lovibond Total Nitrogen test kits in the initial stages of the study and using a BioTector analyser (BioTector, Cork) in the latter stages of the study.

Filtered and unfiltered COD was tested using Lovibond chemical oxygen demand (COD) test kits.

Dissolved oxygen and pH measurements were taken using a Wissenschaftlich Technische Werstätten (WTW) 350i multimeter with a Cellox 325 electrode probe used to measure dissolved oxygen and a WTW pH Electrode Sen Tix 41 probe to measure pH. Measurements were taken by inserting the probes into each reactor via a manually operated valve located on the lid of each of the reactors which was then sealed using a rubber bung. The probes used were robust enough to withstand the collision with the moving media within the reactor. All analysers and probes were regularly calibrated and maintained in accordance with the manufacturer's instructions.

### **3.4.2 Gas analysis**

#### *3.4.2.1 Gas sampling regime*

Monitoring of  $\text{N}_2\text{O}$  emissions was carried during the ISs as follows: In the anoxic period a tube connected to the  $\text{N}_2\text{O}$  meter (Section 3.4.2.2) was inserted into the head space in R1, where the wastewater was being held, via one of the manually operated valves and readings were taken every 5 minutes. In the aerobic period, gas samples were collected from the

effluent gas line of the vacuum pump using a gas sampling bag. This sampling bag was then connected to the nitrous oxide meter to measure N<sub>2</sub>O concentrations. During the settlement period, the tube connected to the N<sub>2</sub>O meter was inserted to the head space above the wastewater being held in R2 with readings taken every 5 minutes.

#### 3.4.2.2 *Testing procedure*

A ViaSensor G200 meter (Q Instrument Services Limited, Cork, Ireland) was used for the measurement of N<sub>2</sub>O. The ViaSensor G200 meter measures N<sub>2</sub>O in gaseous form. The readings on the meter were given in parts per million (ppm) of N<sub>2</sub>O present in the gas samples. The readings were subsequently converted to mg/l using the Ideal Gas Law derived from Boyles Law and Charles Law (Tchobanoglous *et al.*, 2004):

$$PV = nRT \quad [Eqn. 3.1]$$

where: P is the absolute pressure, atm; V is the volume occupied by the gas; n is the number of moles of the gas; R is the universal gas law constant (0.000082057 atm·m<sup>3</sup>/mole·K (where K is the gas temperature in degrees kelvin) and T is the temperature.

All N<sub>2</sub>O samples were tested at atmospheric pressure with an average temperature of 11 °C (284 K). The meter was calibrated regularly and maintained in accordance with the manufacturer's instructions.

#### 3.4.3 *Dissolved oxygen and pH profiling*

DO and pH profiling was conducted by inserting the probes into the bulk fluid in the reactor via the manually operated valve before sealing it with a rubber bung. Consequentially DO and pH profiling could not be conducted when an intensive study was being carried out. During the aeration process, as water was cycled between reactors DO concentrations rose and fell. Minimum DO conditions were observed in any given reactor just before wastewater was moved into the alternate reactor. Unless otherwise stated, DO readings presented refer to the minimum DO values when the probe was immersed in the bulk fluid. DO readings when the bulk fluid was held in the alternate reactor to where the probe was located (e.g. during parts of the aeration period and settlement) have been omitted as the probe was exposed to the air. The pH readings presented correspond to the pH value at the minimum DO readings.



### 3.4.4 Calculations

#### 3.4.4.1 Loading and removal rates

The biofilm media surface area loading rates presented were calculated as follows:

$$B_a = \frac{C_i \times Q_d}{A} \quad [Eqn.3.2]$$

where:  $B_a$  is the biofilm media surface area loading rate ( $\text{g}/\text{m}^2/\text{d}$ );  $C_i$  is the influent parameter concentration ( $\text{mg}/\text{l}$  or  $\text{g}/\text{m}^3$ );  $Q_d$  is the average daily volume treated ( $\text{m}^3/\text{d}$ ); and,  $A$  is the biofilm surface area ( $\text{m}^2$ ).

Removal rates were calculated using Eqn. 3.3

$$B_a = \frac{(C_i - C_e) \times Q_d}{A} \quad [Eqn. 3.3]$$

where:  $C_e$  is the effluent parameter concentration ( $\text{mg}/\text{l}$  or  $\text{g}/\text{m}^3$ ); all other terms are as outlined in equation 3.2.

#### 3.4.4.2 Excess sludge production

The production of excess sludge production is an important consideration in the design of any wastewater treatment system due to the economic, environmental and regulatory factors relating to the disposal of the sludge produced (Wei *et al.*, 2009).

The sludge yield was estimated using the Eqn. 3.4 (Wei *et al.*, 2009).

$$Y = \frac{TSS}{COD_{removed}} \quad [Eqn. 3.4]$$

where: TSS is the total suspended solids produced ( $\text{mg}/\text{l}$ );  $COD_{removed}$  refers to the COD removed ( $\text{mg}/\text{l}$ ).

#### 3.4.4.3 Cell synthesis

A portion of the substrate was utilised for cell synthesis, as outlined in Section 2.8.2. In this study, COD and  $COD_f$  were used to measure the carbon content of the wastewater but in estimating the fraction of substrate utilised for cell synthesis filtered samples were used (ie  $COD_f$ ,  $TN_f$ , and  $PO_4\text{-P}$ ) (Tchobanoglous *et al.*, 2004; Henze, 2008). Estimations of the substrate consumed in cell synthesis were calculated using the following ratio:

$$100 \text{ g } COD_f : 5 \text{ g } TN_f : 1 \text{ g } PO_4\text{-P} \quad [Eqn. 3.5]$$

#### 3.4.4.4 Hydraulic retention time (HRT)

The HRT was calculated as the average length of time for which the wastewater remained within the treatment unit. The HRT was calculated using Eqn. 3.5 (Tchobanoglous *et al.*, 2004):

$$HRT = \frac{V}{Q} \quad [Eqn. 3.6]$$

where: V is the volume of the reactor (l); Q is the volumetric flowrate (l/d).

#### 3.4.4.5 Dilution

The ASF-BR was operated as a batch biofilm treatment unit and as such only a portion of the wastewater treated in each cycle was discharged. The influent wastewater was therefore immediately diluted by wastewater remaining in the system after the discharge period. In the graphs detailing the concentration changes observed during the ISs an estimated post dilution concentration (or measured concentration if present) is depicted and calculations of removal rates are based on this post dilution concentration. The post dilution concentration ( $C_m$ ) within the system was calculated using Eqn. 3.7:

$$\frac{C_{mI}V_I + C_{mD}V_D}{V_I + V_D} = C_m \quad [Eqn. 3.7]$$

where:  $C_{mI}$  is the concentration of parameter m in the influent wastewater;  $V_I$  is the volume of influent wastewater at the start of the treatment cycle in question;  $C_{mD}$  is the concentration of parameter m remaining in the system at the start of the cycle in question (i.e. before the fill phase);  $V_D$  is the volume of wastewater remaining in the system at the start of the cycle in question.

#### 3.4.5 Biofilm analysis - DNA extraction and real-time PCR

Samples of the biofilm which had accumulation on the media were taken periodically during Study 1 from five specific balls, identified by tags. The biofilm specimens which were removed place in a 2 ml container and immediately frozen. On the conclusion of Study 1, the biofilm samples were sent to the Microbiology Department, NUI Galway. Following defrosting, DNA was extracted from biofilm (100 mg) samples using an automated Magstration GC Plus DNA/RNA Extraction and Purification System (Vh Bio, UK), and a Magstration Genomic DNA Extraction Kit (VhBio). DNA was visualised on 1% (w/v) agarose gels containing SYBR Gold (Molecular Probes), to determine purity and concentration. Real-

time, quantitative polymerase chain reactions (PCRs) were done in 20- $\mu$ l reactions using a LightCycler 480 (Roche) to quantify bacterial 16S rRNA, and the ammonia-oxidation (*amoA*), genes. Bacterial gene copy concentrations (GCCs; gene copies/g biofilm) were determined using primers 338f and 805r, and TaqMan probe 516f-ROX, as described by Yu *et al.*, (2005). Separately, GCCs of the functional nitrification gene in ammonia-oxidizing bacteria (AOB), *amoA*, were determined using primers *amoA*-1F and *amoA*-2R (Rotthauwe *et al.*, 1997), and Syber Green Mastermix (Roche), according to the protocol described by Geets *et al.* (2007). Absolute abundance of AOB in the biofilm samples (e.g.  $\text{AOB}_{\text{cells}} \text{g}[\text{biofilm}]^{-1}$ ) was determined assuming the frequency of two *amoA* gene copies per AOB cell (Dionisi *et al.*, 2002). Meanwhile, total bacterial cell densities are difficult to assess as it has been reported that different species may hold between one and fifteen 16S rRNA gene copies per cell (Kembel *et al.*, 2012); thus, the ratio of the *amoA*:16S rRNA GCC was determined and used.

### 3.5 Experimental studies

The overall working parameters of the pilot scale systems during each study are outlined below.

#### 3.5.1 Study 1 – Treatment of synthetic high strength wastewater (HSWW)

In Study 1, the influent wastewater to the ASF-BR comprised a synthetic high strength wastewater. The working volume of each reactor was 35 l with each reactor containing 100 balls to accommodate biofilm growth and the cycle length was approximately 12 h.

#### 3.5.2 Study 2 – Treatment of municipal wastewater (MWW)

In Study 2, the influent wastewater was sourced from two local municipal wastewater treatment plants. In order to achieve the required loading rates and operational configuration the working volume of the reactor was reduced to 25 l. This was achieved by lowering the level switches within the reactor and reducing the number of balls to 50 in each reactor. During the initial stage of the Study (Phase 1) the cycle duration was 8.5 h while the cycle duration during Phase 2 was 6 h.

#### 3.5.3 Study 3 – Treatment of landfill leachate (LL)

The landfill leachate used in Study 3 was sourced from a local municipal landfill site used for the disposal of domestic, commercial and industrial waste. The working volume of each reactor was 35 l with each reactor containing 100 balls to accommodate biofilm growth and the cycle length was approximately 8.5 h during Phase 1 and 6h during Phase 2. Due to the significantly lower hydraulic loading rates during treatment of the landfill leachate the level switches were not sensitive enough to control the fill and discharge of the leachate. Thus a Masterflex L/S Economy peristaltic pump (Cole-Parmer, Dublin), controlled with a timer (Sauter ZDR 102, Radionics, Dublin) that operated independently of the PLC, was employed for the addition of the influent leachate and the extraction of the effluent leachate. The timer controlling the independent peristaltic pump was set on a weekly basis to coincide with fill and empty events obtained from the events log on the SCADA system.

#### **3.5.4 Study 4 – Analysis of N<sub>2</sub>O emissions**

Study 4 was designed to quantify the amount of N<sub>2</sub>O produced during nitrogen removal from the treatment of both municipal wastewater and landfill leachate. N<sub>2</sub>O emitted during the wastewater treatment process was measured in gaseous form using a ViaSensor G200 meter during Phase 2 of Study 2 and Study 3.

### **3.6 Seeding**

#### **3.6.1 Study 1**

ASF-BR 1 was seeded using a mixture of 18 l of nitrifying activated sludge and 18 l HSWW. The activated sludge was taken from TWTP and had an average mixed liquor suspended solids (MLSS) of 5229 mg SS/l. This mixture was added to R1. The SCADA program was configured to alternate the seed mixture between R1 and R2 for a 48 hour period by which time a biofilm was visually observed to have begun to accumulate on the media. The seed mixture was discharged from the unit and the timing regime was revised to begin Study 1.

#### **3.6.2 Study 2**

ASF-BR 2 was seeded over a 48 hour period using 12 l of nitrifying activated sludge (MLSS = 4398 mg SS/l) sourced from TWTP mixed with 12 l of municipal wastewater. The SCADA program was configured to sequentially move the seed mixture between R1 and R2. After 48 hours the seed mixture was discharged and the SCADA program reconfigured to achieve the objectives of the study.

### 3.6.3 Study 3

Prior to commencing Study 3, ASF-BR 1 was seeded with a mixture of 25 l nitrifying activated sludge (average MLSS = 4306 mg SS/l) and 10 l landfill leachate. As the concentration of contaminants in landfill leachate was higher than municipal wastewater, less leachate was used in the seeding process. The SCADA program was configured to continuously move the leachate between R1 and R2. As a result of the significantly higher concentration in the landfill leachate, it took approximately 60 hours for biofilm to be visually observed on the plastic media. An extended acclimation period was expected as the return activated sludge was sourced from a municipal wastewater treatment plant and would therefore be expected to take longer to acclimatise to the landfill leachate.

After approximately 5 months of operation, following a period of reduced treatment performance, visual inspection of the media revealed that there was limited biofilm remaining and thus ASF-BR 1 was reseeded. On this occasion the seeding was carried out gradually over a 10 day period. Initially 20 l of nitrifying activated sludge (TWTP; average MLSS = 4522 mg SS/l) mixed with 10 l of diluted leachate (1 l leachate diluted with 9 l of tap water) was added to R1. The ASF-BR was programmed to constantly transfer the seed culture from R1 to R2 for 48 hours. After 48 hours 10 l of this seed culture was discharged and 10 l of influent (comprising of 4 l leachate diluted with 6 l of tap water) was added. The seed culture was then continuously transferred between R1 and R2 for 120 hours. Thereafter, 10 l of the seed culture was discharged and 10 l of leachate mix (comprising of 2.5 l leachate diluted with 7.5 l of tap water) was added to the ASF-BR and was transferred between R1 and R2 for a further 120 hours. After this time, a biofilm was observed on the media and approximately 60% of the seed culture was discharged and the ASF-BR was re-reprogrammed to continue normal operation.

## 3.7 Overview of ASF-BR studies

A summary of the studies carried out in evaluating the efficacy of the range of wastewaters is presented in Table 3.2.

**Table 3.5 Details of studies conducted on the ASF-BR**

Study	Waste-water	Objective	HLR	HLR	Cycle duration	Study duration	Chapter	
			l/cycle	l/d	min	d		
1	Phase 1	HSWW	D + N	6.4	11.8	726	166	4
	Phase 2		D + N	5.9	10.2	736	264	
2	Phase 1	MWW	N	12.1	36.1	510	212	5
	Phase 2		D + N	7.3	35.1	300	117	
3	Phase 1	LL	N	0.35	0.98	495	253	6
	Phase 2		D + N	0.34	1	480	74	
4	-	-	Analysis of N <sub>2</sub> O emission during Phase 2 of Study 2 and Study 3				7	

Nomenclature: HLR – Hydraulic loading rate; D – Denitrification; N – nitrification

### 3.8 Summary

This chapter describes the design, construction and operation of the pilot scale ASF-BR wastewater treatment unit. The chapter provides an overview of the various studies carried out during this study and summarises the wastewater types used in each study. The seeding regime is detailed as are daily system operation methods, sampling regime, and analyses methods.

## **CHAPTER 4**

### **Study 1 – Treatment of high strength synthetic wastewater**

## 4.1 Introduction

The performance of an experimental scale ASF-BR treating high strength synthetic wastewater (HSWW) was evaluated over a 430-day period. The study consisted of two phases: Phase 1 (166 days) and Phase 2 (264 days). During Phase 2, the frequency of the wastewater movements between the reactors was reduced. This reduced the operational energy requirement of 54%, when compared to Phase 1, due to a reduction in the number of pumping events. The objective of both phases was to achieve organic carbon removal, nitrification and denitrification while also investigating the performance of the unit with regard to energy consumption, cycle duration and loading rate.

## 4.2 Apparatus

The general set-up of the ASF-BR was described in Section 3.2.5. In this study, the unit was operated with a working volume of 35 l/reactor and a media surface area of 4.1 m<sup>2</sup>/reactor. Details of the cycle configuration used in each of the phases are outlined in the relevant subsections.

### 4.3 Phase 1: Application of the ASF-BR in the treatment of HSWW

#### 4.3.1 Cycle configuration

The treatment cycle in Phase 1 was configured to achieve nitrification and denitrification of HSWW, as outlined in Table 4.1.



**Table 4.1 Cycle configuration of the ASF-BR – Phase 1**

		Units	
Aerobic period	Study duration	d	166
	Step 1 – Fill*	s	45
	Step 2 – Anoxic period	min	90
	Step 3 – Equalisation	s	20
	Step 4 – Draw to R2*	s	20
	Step 5 – Hold in R2	s	30
	Step 6 – Equalisation	s	20
	Step 7 – Draw to R1*	s	20
	Step 8 – Hold in R1	s	30
	Repeat Steps 3 to 8		
	Total: Aerobic period	min	585
	Step 8 – Settlement period	min	45
	Step 9 – Discharge*	s	45
Total Cycle duration	min	725	
Average daily flow	l/d	11.6 (0.5)	
Average volume/cycle	l	6.5 (0.3)	
Reactor volume	l/reactor	35	
HRT	d	3	

\* Maximum duration, actual time may have been less if a signal was received from a level switch; Standard deviation shown in ( ).

### 4.3.2 Overall performance of the ASF-BR – Phase 1

Before the study commenced, the ASF-BR was seeded using return activated sludge (RAS) collected from a local municipal wastewater treatment plant, as described in Section 3.6.1. The unit reached steady state conditions after seven days. This steady state was taken from the time when relatively stable effluent concentrations were observed. All results presented refer to the ASF-BR operation during steady state conditions. The full set of results from Phase 1 are presented in Appendix D. The overall performance of the ASF-BR in Phase 1 is summarised in Table 4.2.

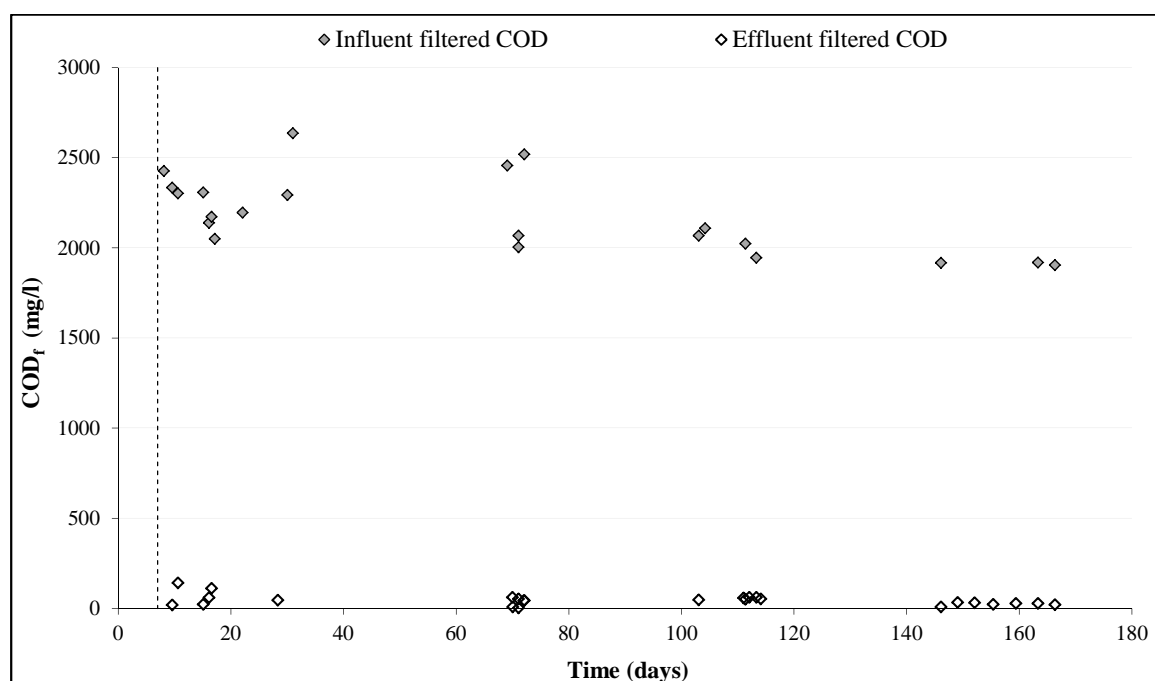
**Table 4.2 Overall performance of the ASF-BR – Phase 1**

Parameter	Average influent	Influent st. dev.	Loading rate*	Average effluent	Effluent st. dev.	Removal rate*	% Removal	n Inf/Eff
	mg/l	mg/l	g/m <sup>2</sup> media/d	mg/l	mg/l	g/m <sup>2</sup> media/d		
COD**	2506	267	3.6	198	68	3.3	92	18/42
COD <sub>f</sub>	2085	297	3.0	52	52	2.9	98	18/42
TN**	207	23.0	0.29	38.2	11.9	0.24	82	18/42
TN <sub>f</sub>	199	64.5	0.28	27.9	7.9	0.24	86	18/42
NH <sub>4</sub>	168	33.8	0.24	4.5	7.1	0.23	97	21/46
TON	0.2	1.8	0	11.3	8.1	-	-	21/46
NO <sub>2</sub>	0.2	0.33	0	2.6	3.9	-	-	21/46
NO <sub>3</sub>	0.1	0.70	0	8.7	6.0	-	-	21/46
PO <sub>4</sub> -P	48.9	16.8	0.07	32.0	13.0	0.02	35	21/46

\* Based on an average influent volume of 11.6 l/d and a surface area of 8.2m<sup>2</sup>; n is the number of samples; Inf – Influent; Eff – Effluent.

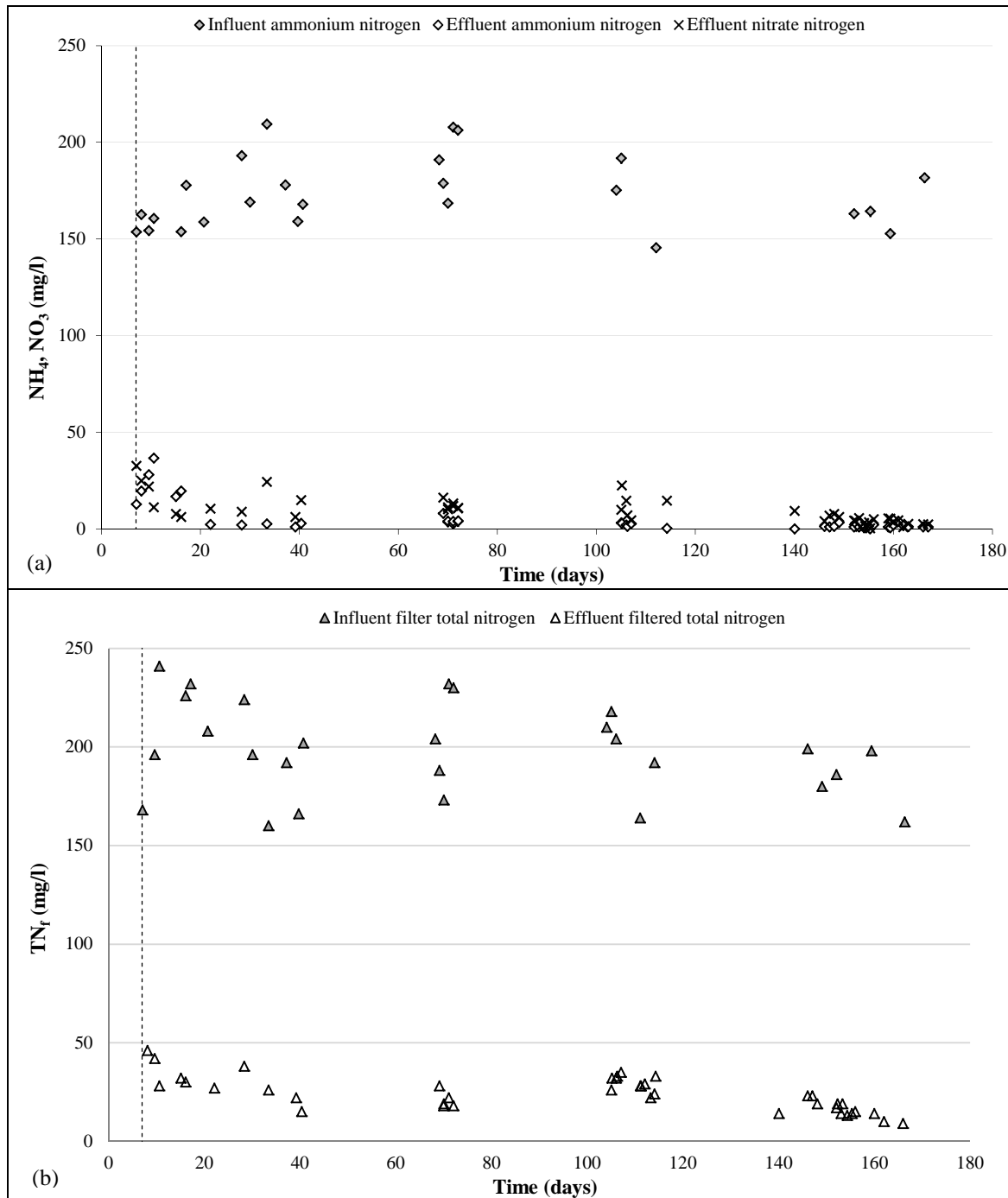
#### 4.3.2.1 Organic carbon

The use of synthetic wastewater resulted in a reasonably stable influent COD and COD<sub>f</sub> concentrations over the 166 days of Phase 1 (Figure 4.1). Average influent COD and COD<sub>f</sub> concentrations were 2506 mg/l and 2085mg/l respectively, resulting in an average surface area loading rate of 3.6 mg COD/m<sup>2</sup> media/d and 3.0 mg COD<sub>f</sub>/m<sup>2</sup> media/d. Average effluent COD and COD<sub>f</sub> concentrations were 198 mg/l and 52 mg/l respectively, equivalent to a removal efficiency of 92% and 98% respectively.

**Figure 4.1 Influent and effluent COD<sub>f</sub> – Phase 1**

### 4.3.2.2 Nitrogen

The nitrification performance of the ASF-BR remained reasonably consistent from Day 7 (Figure 4.2 (a)). The average  $\text{NH}_4$  removal during steady state conditions was 97%, equivalent to a removal rate of  $0.23 \text{ g NH}_4/\text{m}^2 \text{ media/d}$ . Effluent  $\text{TN}_f$  concentrations remained reasonably constant, with an average reduction of 86% being observed; a removal rate of  $0.24 \text{ g TN}_f/\text{m}^2/\text{d}$  was achieved.



**Figure 4.2 (a) Influent and effluent ammonium nitrogen and effluent nitrate nitrogen concentrations; (b) Influent and effluent filtered total nitrogen concentrations – Phase 1**

#### 4.3.2.3 Phosphorus

The removal of phosphorus was not a primary objective of this phase. However, phosphorus concentrations were monitored as phosphorus is utilised in the synthesis of new cells. The average influent  $\text{PO}_4\text{-P}$  concentration was 48.9 mg/l while the average effluent concentration was 32.0 mg/l. The average reduction in ortho-phosphorus of 17 mg/l is in line with the phosphorus requirements of cell synthesis (Section 3.4.4.3). Based on a  $\text{COD}_f$  reduction of 2032 mg/l (2084-52), an estimated 20.3 mg  $\text{PO}_4\text{-P/l}$  was used in the synthesis of new cells, which indicated that some  $\text{PO}_4\text{-P}$  may have been released from sludge accumulating at the bottom of the reactors.

#### 4.3.2.4 DO and pH

DO profiles were recorded periodically throughout Phase 1 (Figure 2a and Figure 2b), as described in Section 3.5.3. The DO profiles confirm DO conditions conducive to denitrification ( $< 0.2$  mg/l) were present during the anoxic period in R1 (Figure 4.3 (a)). During the aerobic period, DO concentrations were observed to increase in both R1 and R2, which is in line with the oxygen requirements of the nitrification process. A reduction in the oxygen requirement for carbonaceous oxidation was observed. As  $\text{NH}_4$  concentrations and the carbon content were reduced, oxygen demand decreased, thus resulting in an increase in the DO concentrations. A reduction in the DO concentration during the settlement period may have been the result of a release of carbon stored by glycogen resulting in carbonaceous oxidation (Berg *et al.*, 2002; Gregory *et al.*, 2002).

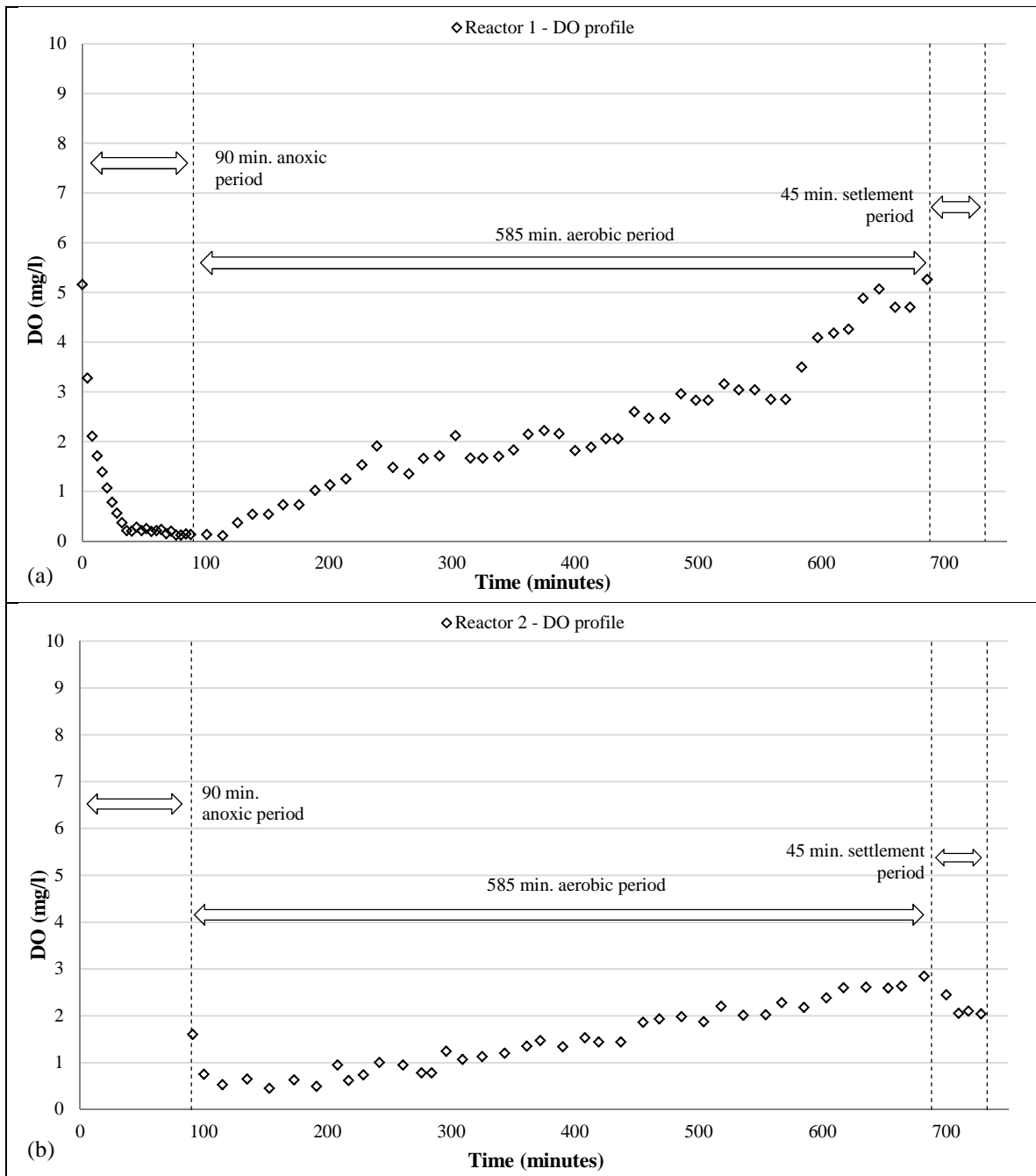


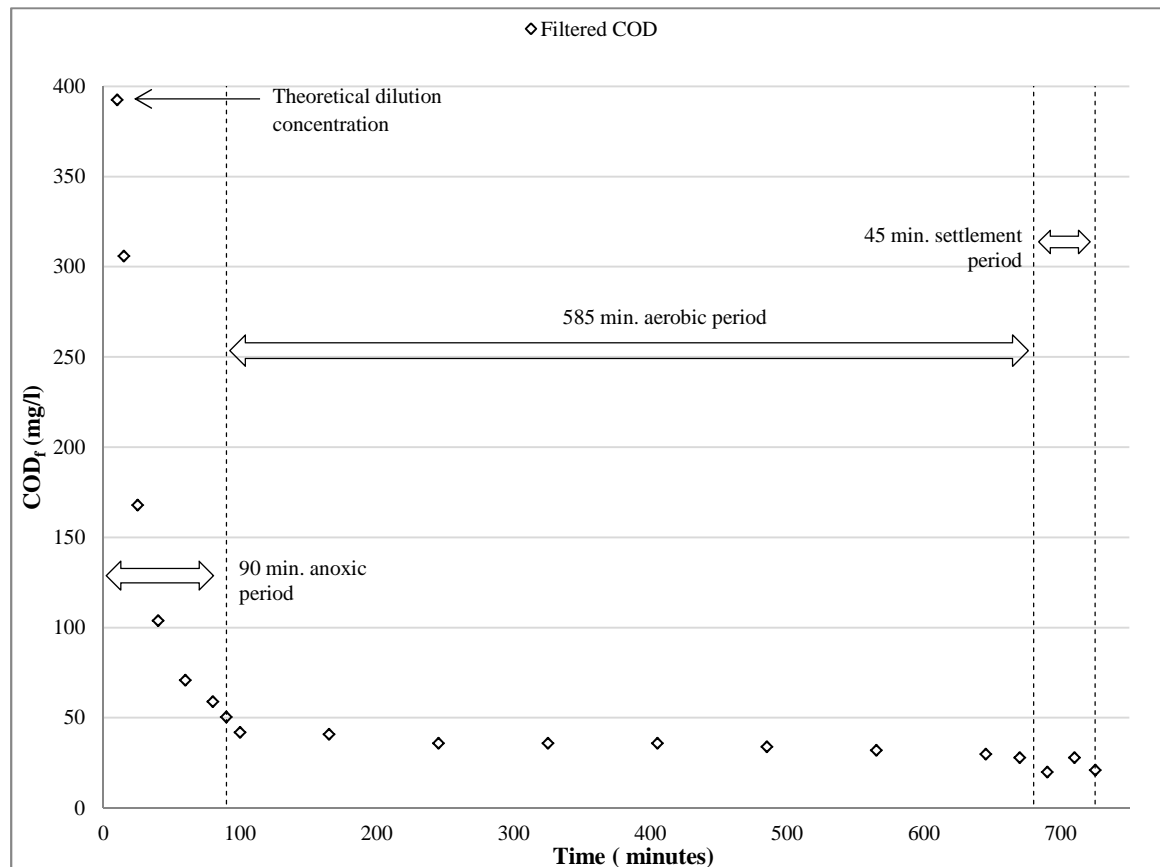
Figure 4.3 Typical DO – Phase 1: (a) - R1 (Day 98); (b) - R2 (Day 104)

#### 4.3.3 Intensive studies

During Phase 1, three intensive studies, HSWIS 1, HSWIS 2 and HSWIS 3, were conducted on Days 151, 158 and 164 respectively, as outlined in Section 3.5.1, during which carbon and nitrogen concentrations were monitored. As conditions compatible with phosphorus removal were not present,  $\text{PO}_4\text{-P}$  was not monitored as part of the ISs. The results of these ISs are summarised below, with full details of results available in Appendix D.

### 4.3.3.1 Organic carbon

The COD<sub>f</sub> concentrations observed in HSWIS 1 are presented in Figure 4.4. During HSWIS 1, the influent COD<sub>f</sub> concentration was reduced by 99%, with similar removal results and profiles observed during HSWIS 2 and HSWIS 3.



**Figure 4.4 COD<sub>f</sub> profile – HSWIS 1**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

Denitrification during the anoxic period was responsible for the highest COD<sub>f</sub> removal, with an average removal efficiency of 92% achieved. During the first 70 minutes of the aerobic period, COD<sub>f</sub> concentrations reduced further as carbon was utilised as a source of energy in the synthesis of biomass. Thereafter, the COD<sub>f</sub> concentration remained relatively stable at approximately 35 mg/l for the duration of the aerobic period. As there was no further reduction in of the remaining COD<sub>f</sub>, and considering the systems' operation, it is likely this portion was non-biodegradable

The observed rate of COD<sub>f</sub> removal was highest during HSWIS 2 at 95.8 mg COD<sub>f</sub>/m<sup>2</sup> media/min during the anoxic period, following dilution with a removal rate of 0.19 mg COD<sub>f</sub>/m<sup>2</sup> media/min in the aerobic period, resulting in a COD<sub>f</sub> reduction of 97% over the

treatment cycle. Table 4.3 presents the removal rates achieved in all three ISs conducted during Phase 1.

**Table 4.3 COD<sub>f</sub> removal rates – Phase 1**

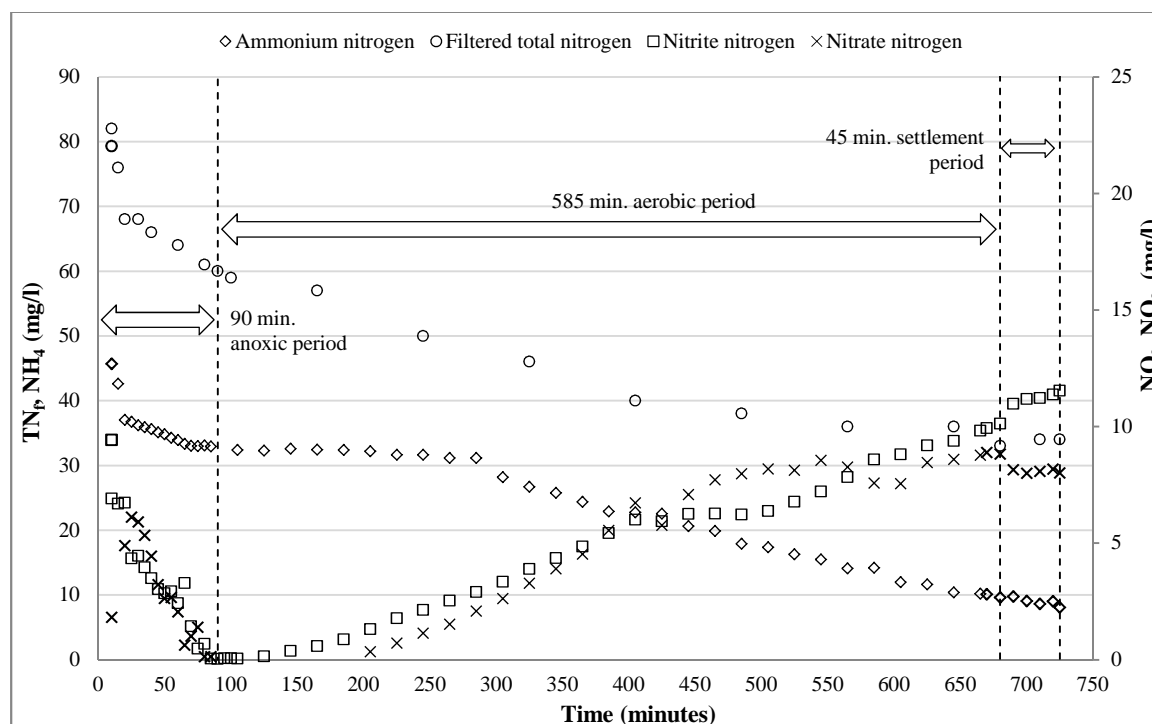
	Anoxic period <sup>1,2</sup>	Time	Aerobic period <sup>3</sup>	Time
	mg COD <sub>f</sub> /m <sup>2</sup> media/min	min	mg COD <sub>f</sub> /m <sup>2</sup> media/min	min
HSWIS 1	36.50	10-90	0.25	90-245
HSWIS 2	26.51	10-90	1.09	90-165
HSWIS 3	16.41	10-90	0.69	90-165

<sup>1</sup> Post dilution; <sup>2</sup> Surface area loading rate calculated based on surface area in R1 (4.1 m<sup>2</sup>);

<sup>3</sup> Surface area loading rate based on surface area of R1 and R2 (8.2 m<sup>2</sup>).

#### 4.3.3.2 Nitrogen

The nitrogen profile obtained during HSWIS 2 is presented in Figure 4.5. Nitrogen profiles observed during HSWIS 1 and HSWIS 3 were found to follow a similar trend. The average NH<sub>4</sub> reduction during the complete treatment cycle of the three ISs was found to be 96%.



**Figure 4.5 Nitrogen profiles – HSWIS 2**

Note: Initial concentrations are the theoretical post dilution concentrations; Influent concentrations omitted.

Nitrification during the aerobic period was found to be the principal NH<sub>4</sub> removal mechanism, with an average removal of 69% NH<sub>4</sub> removal through oxidation in the

aerobic period achieved over the three ISs, resulting in an average effluent  $\text{NH}_4$  concentration of 9.1 mg  $\text{NH}_4/\text{l}$ .

The rate of removal of  $\text{NH}_4$  as summarised in Table 4.4 was found to be similar in all three ISs. The highest rates were observed in HSWIS 2 when removal rates of 1.38 mg  $\text{NH}_4/\text{m}^2$  media/min and 0.14 mg  $\text{NH}_4/\text{m}^2$  media/min were achieved during the anoxic and aerobic periods respectively.

**Table 4.4  $\text{NH}_4$  removal rates – Phase 1**

	<b>Anoxic period</b> <sup>1, 2</sup>	<b>Time</b>	<b>Aerobic period</b> <sup>3</sup>	<b>Time</b>
	mg $\text{NH}_4/\text{m}^2$ media/min	min	mg $\text{NH}_4/\text{m}^2$ media/min	min
HSWIS 1	1.01	10-90	0.14	90-680
HSWIS 2	1.38	10-90	0.14	90-680
HSWIS 3	1.47	10-90	0.11	90-565

<sup>1</sup> Post dilution; <sup>2</sup> Surface area loading rate calculated based on surface area in R1 (4.1  $\text{m}^2$ ); <sup>3</sup> Surface area loading rate based on surface area of R1 and R2 (8.2  $\text{m}^2$ ).

Table 4.5 presents an average nitrogen mass balance based on the ISs conducted in Phase 1. The average mass of  $\text{NH}_4$  in the effluent wastewater implied that further nitrogen removal was possible using an optimised treatment cycle. Cell synthesis and the production of  $\text{N}_2\text{O}$  are likely to have resulted in the imbalance between the influent nitrogen and effluent mass.

**Table 4.5 Nitrogen mass balance – Phase 1 \***

		<b>TN<sub>f</sub></b>	<b>NH<sub>4</sub></b>	<b>NO<sub>2</sub></b>	<b>NO<sub>3</sub></b>
		g/d <sup>1</sup>	g/d <sup>1</sup>	g/d <sup>1</sup>	g/d <sup>1</sup>
<b>Post dilution mass</b> <sup>a</sup>	$M_{p, \text{dil}}$	4.86 (1.34)	2.88 (0.44)	0.67 (0.04)	0.65 (0.26)
<b>Mass change - Anoxic period</b>	$\Delta M_{\text{anx}}$	-1.29 (0)	-1.44 (0.96)	-0.66 (0.04)	-0.52 (0.08)
<b>Mass change - Aerobic period</b>	$\Delta M_{\text{aer}}$	-1.55 (0.56)	-1.43 (0.22)	0.72 (0.04)	0.69 (0.11)
<b>Effluent</b>	$M_{\text{eff}}$	2.12 (0.64)	0.53 (0.07)	0.82 (0.06)	0.79 (0.22)

\* Average results from ISs carried out in Phase 1; <sup>a</sup> – Theoretical average post dilution mass;  $M_{p, \text{dil}}$  – average post dilution mass;  $M_{\text{aer}}$  – average change in mass during aeration period;  $M_{\text{eff}}$  – average effluent mass; <sup>1</sup> 1.98 cycles/d; Standard deviation shown in ( ).

Linear trend between  $\text{NH}_4$  and  $\text{NO}_3$  concentrations and time in each of the ISs are presented in Table 4.6. The linear trend represents the relationship between the



independent variable, time and the dependent variable, the concentration of the contaminant. R-squared is a measure of the goodness of fit of the trendline to the data.

**Table 4.6 Summary of NH<sub>4</sub> and NO<sub>3</sub> linear trends during HSWIS 1, HSWIS 2 and HSWIS 3**

MWIS	N process	Period	Time	Linear trend	R <sup>2</sup>
			min	mg/l	
HSWIS 1	NH <sub>4</sub> decrease	Anoxic	10-90	NH <sub>4</sub> = - 0.105 t + 39.3	0.95
	NO <sub>3</sub> decrease	Anoxic	10-90	NO <sub>3</sub> = - 0.165 t +14.1	0.97
	NH <sub>4</sub> decrease	Aerobic	105-680	NH <sub>4</sub> = - 0.042 t + 40.1	0.94
	NO <sub>3</sub> increase	Aerobic	105-680	NO <sub>3</sub> = 0.021 t - 1.6	0.94
HSWIS 2	NH <sub>4</sub> decrease	Anoxic	10-90	NH <sub>4</sub> = - 0.070 t + 38.3	0.97
	NO <sub>3</sub> decrease	Anoxic	10-90	NO <sub>3</sub> = - 0.097 t + 8.0	0.92
	NH <sub>4</sub> decrease	Aerobic	105-680	NH <sub>4</sub> = - 0.046 t + 40.9	0.92
	NO <sub>3</sub> increase	Aerobic	165-680	NO <sub>3</sub> = 0.020 t + 2.8	0.97
HSWIS 3	NH <sub>4</sub> decrease	Anoxic	10-90	NH <sub>4</sub> = - 0.051 t + 27.5	0.98
	NO <sub>3</sub> decrease	Anoxic	10-90	NO <sub>3</sub> = - 0.060 t + 5.9	0.91
	NH <sub>4</sub> decrease	Aerobic	105-565	NH <sub>4</sub> = - 0.036 t + 26.5	0.96
	NO <sub>3</sub> increase	Aerobic	105-605	NO <sub>3</sub> = 0.014 t + 3.7	0.93

The average removal of the influent TN<sub>f</sub> during the complete treatment cycle over the three ISs was 88%. Denitrification occurring in the anoxic period was found to be responsible for the highest TN<sub>f</sub> removal as heterotrophic bacteria reduce NO<sub>2</sub> and NO<sub>3</sub> to nitrogen gases (Nitric oxide, NO; Nitrous oxide, N<sub>2</sub>O; Nitrogen gas, N<sub>2</sub>), resulting in an average TN<sub>f</sub> removal during the anoxic period of the three ISs of 57% of the TN<sub>f</sub> remaining following dilution.

During the aerobic period, the TN<sub>f</sub> concentration remaining after the anoxic period was reduced by an average of 41% over the three ISs, equivalent to a TN<sub>f</sub> reduction of 22.2 mg/l. Based on an average COD<sub>f</sub> reduction of 43.5 mg/l, cell synthesis accounted for a TN<sub>f</sub> reduction of approximately 2.18 mg/l (Section 3.4.4.3). The remainder of the TN<sub>f</sub> reduction is likely to have been a result of simultaneous nitrification and denitrification as a heavy, thick biofilm was observed throughout this study, as depicted in Figure 4.10. This dense biofilm is likely to have allowed anoxic conditions to develop within the inner layers of the biofilm, enabling denitrification to occur (Henze *et al.*, 2002; Biesterfeld *et al.*, 2003).



**Figure 4.6 Media with thick biofilm accumulation**

#### **4.4 Phase 2: Optimisation of performance of the ASF-BR**

Results from Phase 1 indicated that optimisation of the cycle configuration could significantly increase the operational efficiency of the treatment process. In Phase 2 (264 days), the operational efficiency of the treatment process was increased by reducing the frequency of wastewater movements between the reactors (Steps 4 and 6, Table 3.1); thus reducing the energy requirement. The results presented herein refer to the operation of the ASF-BR under steady state conditions, when effluent concentrations were deemed to be relatively constant. Steady state conditions were observed after Day 60. The delay in reaching steady state in this study was due to a faulty level switch that controlled the movement of wastewater within the reactors.

##### **4.4.1 Cycle configuration**

Table 4.7 presents the optimised cycle configuration employed during Phase 2.

**Table 4.7 Cycle configuration pilot scale ASF-BR – Phase 2**

		Units	
Aerobic period	Study duration	d	264
	Step 1 – Fill*	s	45
	Step 3–Anoxic period	min	90
	Step 2 - Equalisation	s	20
	Step 3 – Draw to R2*	s	20
	Step 4 – Hold in R2	s	90
	Step 5 - Equalisation	s	20
	Step 6 – Draw to R1*	s	20
	Step 7 – Hold in R1	s	90
	Repeat Steps 2 to 7		
	Total: Aerobic period	min	598
	Step 8 – Settlement period	min	45
	Step 9 – Discharge*	s	45
	Total Cycle duration	min	737
Average daily flow	l/d	10 (0.6)	
Average volume/cycle	l	5.2 (0.8)	
Reactor volume	l/reactor	35	
HRT	d	3.5	

\* Maximum duration, actual time may have been less if a signal was received from a level switch; Standard deviation shown in ( ).

#### 4.4.2 Overall performance of the ASF-BR – Phase 2

The average influent and effluent parameters from Phase 2 are summarised in Table 4.8. Full details of these results are available in Appendix D.

**Table 4.8 Overall performance of the ASF-BR – Phase 2**

Parameter	Average influent	Influent st. dev.	Loading rate*	Average effluent	Effluent st. dev.	Removal rate*	% Removal	n
	mg/l	mg/l	g/m <sup>2</sup> media/d	mg/l	mg/l	g/m <sup>2</sup> media/d		Inf/Eff
COD <sup>**</sup>	2571	205	3.1	399	178	2.6	84	32/45
COD <sub>f</sub>	2043	190	2.5	62	36	2.42	97	32/45
TN <sup>**</sup>	227	40.2	0.28	61.6	21.4	0.20	79	26/21
TN <sub>f</sub>	195	28.6	0.24	28.3	18.8	0.21	88	26/21
NH <sub>4</sub>	169 <sup>c</sup>	25.9	0.21	1.7 <sup>f</sup>	3.0	0.20	99	60/77
TON	0 <sup>c</sup>	0	0	11.0 <sup>f</sup>	7.8	-	-	60/77
NO <sub>2</sub>	0.1 <sup>c</sup>	0	0	1.3 <sup>f</sup>	1.9	-	-	60/77
NO <sub>3</sub>	0 <sup>c</sup>	0	0	9.8 <sup>f</sup>	7.3	-	-	60/77
PO <sub>4</sub> -P	50.6 <sup>c</sup>	24.8	0	12.2 <sup>f</sup>	10.2	0.05	76	60/77

\* based on an average influent volume of 10 l/d and a surface area of 8.2m<sup>2</sup>; n is the number of samples; Inf – Influent; Eff – Effluent.

#### 4.4.2.1 Organic carbon

Steady state influent and effluent COD and COD<sub>f</sub> sample results are presented in Figure 4.7. Influent COD and COD<sub>f</sub> concentrations were reduced at an average rate of 2.64 mg COD/m<sup>2</sup> media/d and 2.42 mg COD<sub>f</sub>/m<sup>2</sup> media/d, resulting in average effluent COD and COD<sub>f</sub> concentrations of 399 mg/l and 62 mg/l respectively.

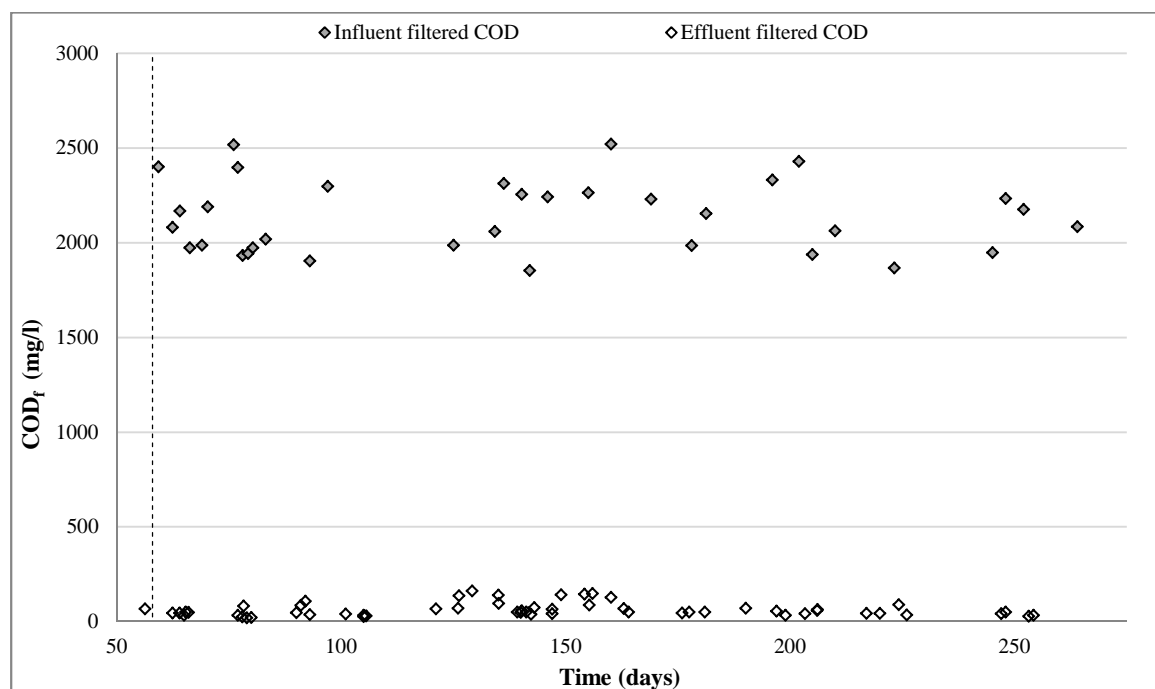
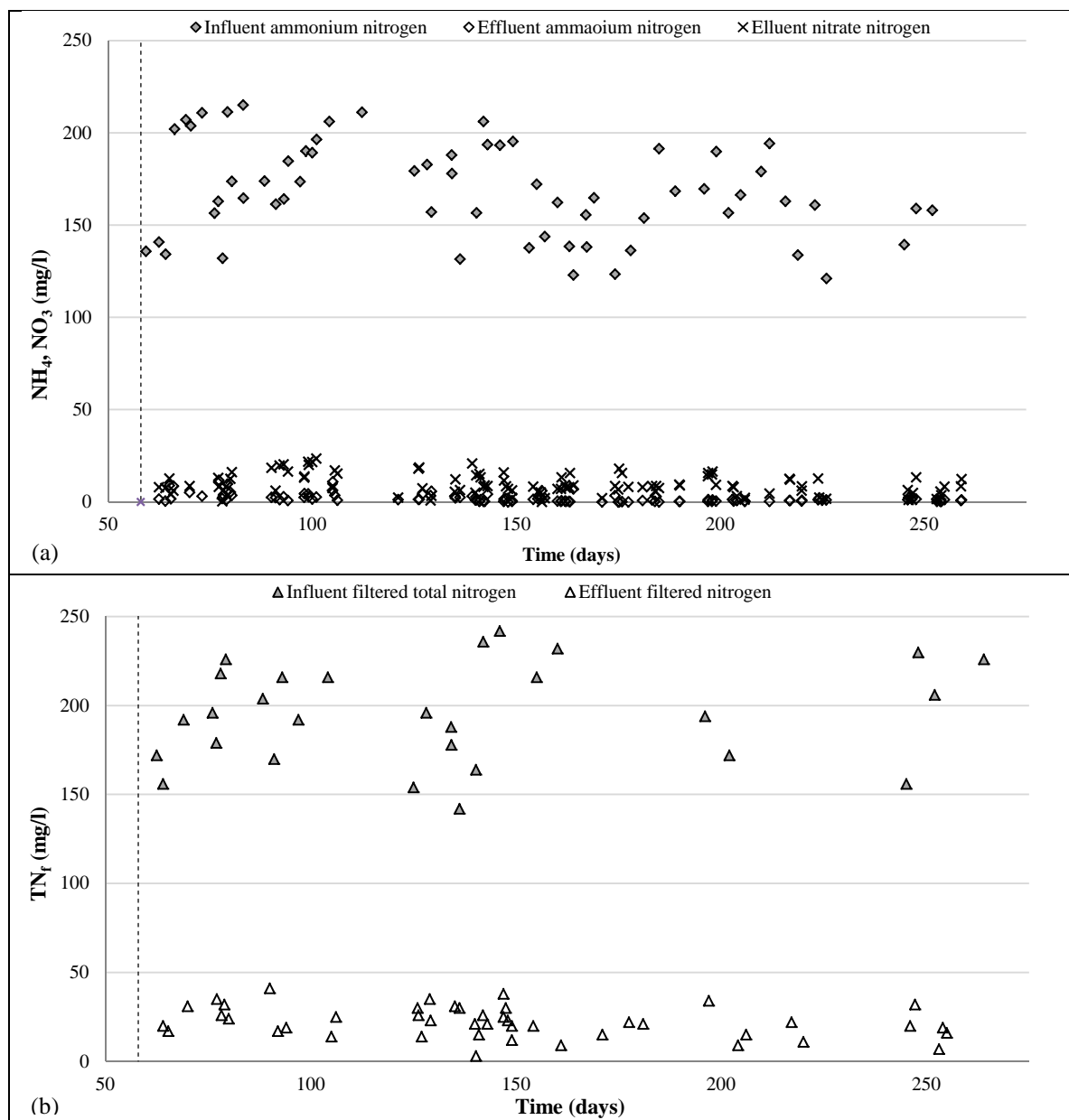


Figure 4.7 Influent and effluent COD<sub>f</sub> – Phase 2

#### 4.4.2.2 Nitrogen

Under steady state conditions, the average influent NH<sub>4</sub> concentration was 169 mg/l (Figure 4.8 (a)). The average NH<sub>4</sub> removal rate over the course of Phase 2 was 0.20 mg NH<sub>4</sub>/m<sup>2</sup> media/d, which resulted in an average effluent NH<sub>4</sub> concentration of 1.7 mg/l, equivalent to an average removal efficiency of 99%.

The average influent TN<sub>f</sub> in Phase 2 was 195 mg/l, with a corresponding effluent TN<sub>f</sub> of 28.3 mg/l, giving a rate of removal of 0.21 mg TN<sub>f</sub>/m<sup>2</sup> media/d and an average efficiency of 88% (Figure 4.8 (b)).



**Figure 4.8 (a) Influent and effluent ammonium nitrogen and effluent nitrate nitrogen; (b) Influent and effluent total nitrogen – Phase 2**

#### 4.4.2.3 Phosphorus

Phosphorus concentrations were monitored during Phase 2 as per Phase 1. The average influent  $\text{PO}_4\text{-P}$  in Phase 2 was 50.6 mg/l while the average effluent  $\text{PO}_4\text{-P}$  concentration of 12.2 mg/l.

#### 4.4.2.4 DO and pH

Figure 4.14 presents typical DO and pH profiles observed in Phase 2. As in Phase 1, DO profiles were taken periodically. The DO profiles (Figure 4.9 (a)) indicate that conditions conducive to denitrification (low DO;  $< 0.02$  mg  $\text{O}_2$ /l) were present during the anoxic

period. The low DO during the first 350 minutes of the aerobic period coincide with the nitrification process, the first stage of the nitrification, which has the highest oxygen demand, and the oxidation of the remaining biodegradable organic matter following the anoxic period. After this initial 350 minute period, the DO concentrations began to increase, which coincides with a reduction in the nitrification rate (Figure 4.11). The pH of the wastewater was also monitored periodically in Phase 2. During the anoxic period, denitrification resulted in the production of alkalinity, leading to an observed increase in the pH while, during the aerobic period, the nitrification process consumed alkalinity, resulting in the observed decrease in the pH.

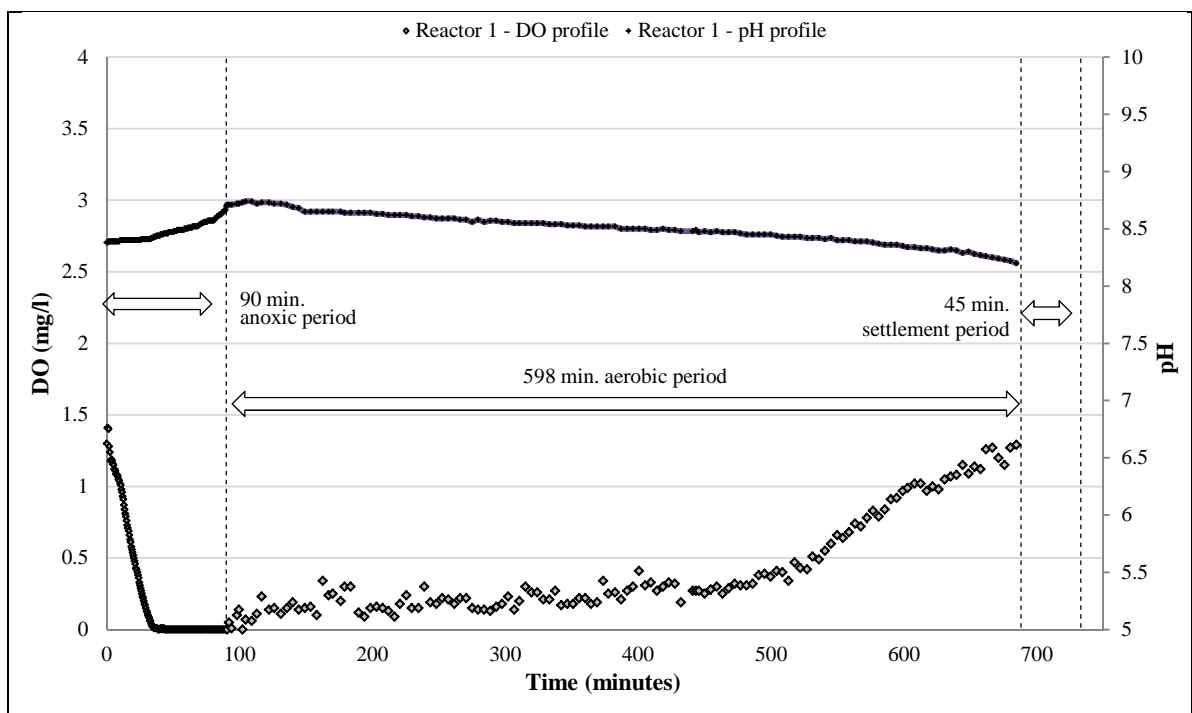


Figure 4.9 Typical DO and pH profiles – Phase 2: (a) – R1 (Day 132)

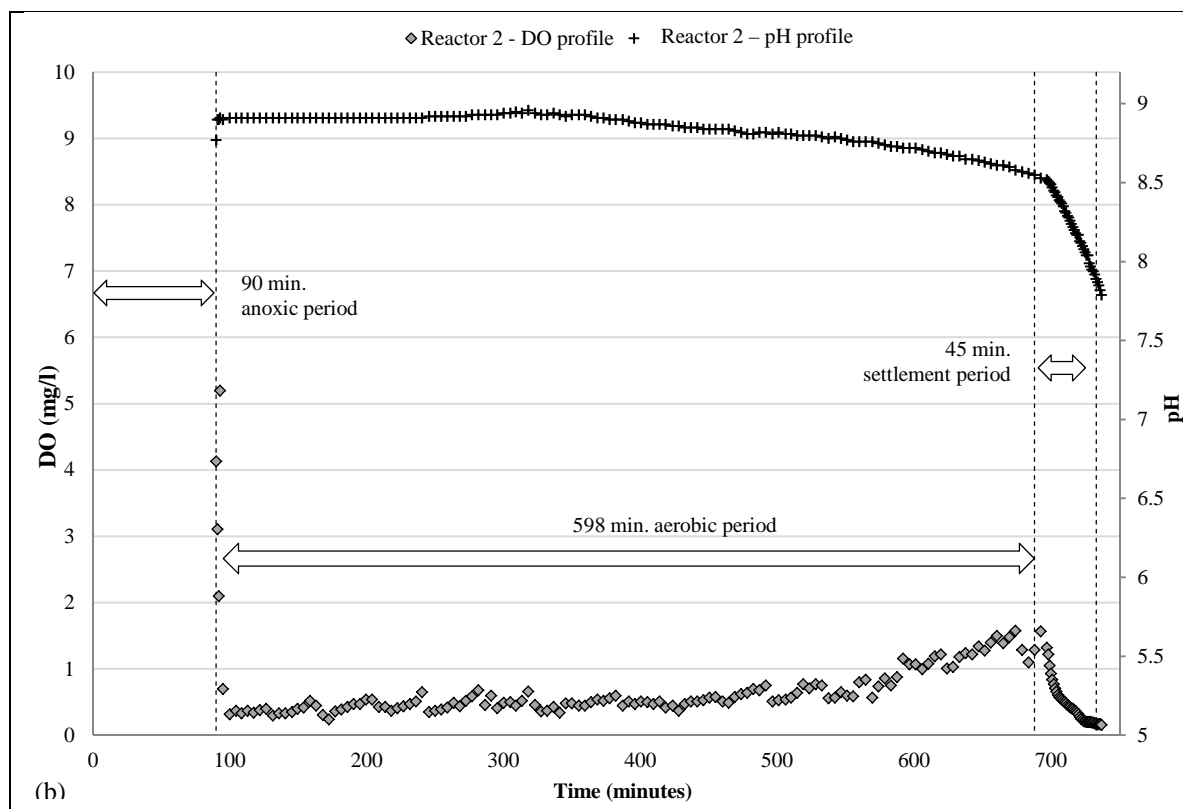


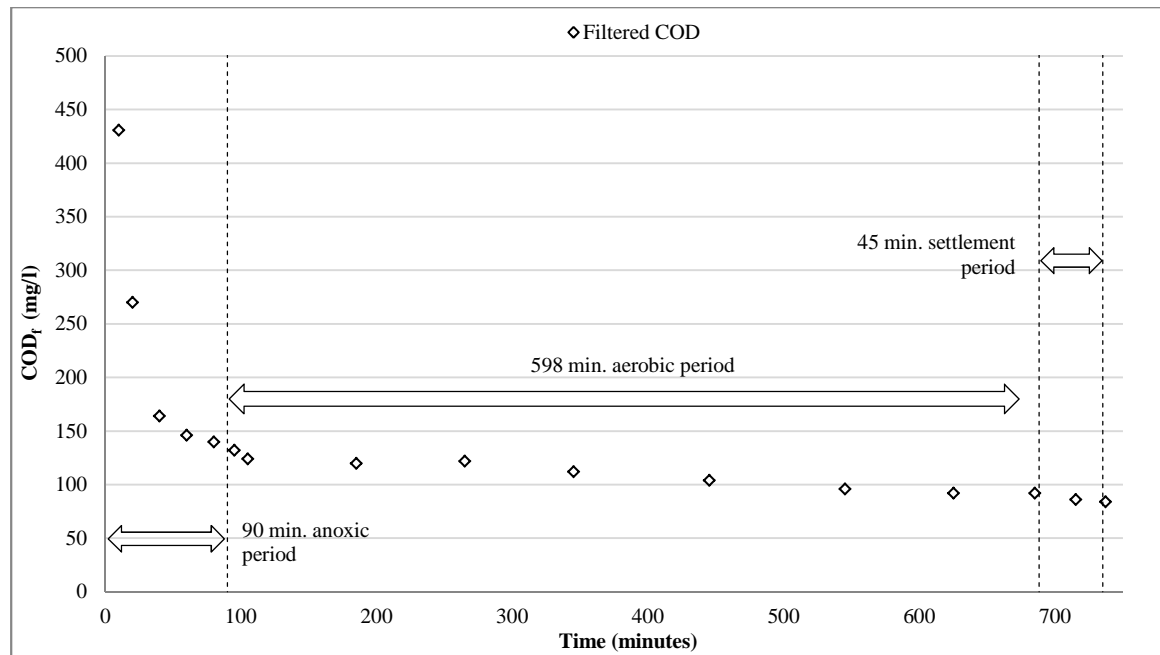
Figure 4.9 Typical DO and pH profiles – Phase 2: (b) – R2 (Day 154)

#### 4.4.3 Intensive studies

During Phase 2, three ISs, HSWIS 4, HSWIS 5 and HSWIS 6, were conducted on Days 105, 147 and 178 respectively. During these ISs, the concentrations of carbon and nitrogen in the wastewater were monitored. The results from these three ISs are summarised here, with full details contained in Appendix D.

##### 4.4.3.1 Organic carbon

The carbon profile, observed in HSWIS 4, is presented in Figure 4.10; a similar trend was observed in HSWIS 5 and HSWIS 6.



**Figure 4.10 COD<sub>f</sub> profile during – HSWIS 4**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

In all three of the ISs, the highest COD<sub>f</sub> reduction was found to occur during the anoxic period, when carbon is utilised as the electron donor in the denitrification process. The average COD<sub>f</sub> reduction, over the three ISs during the anoxic period, was 88% of the COD<sub>f</sub> following dilution. During the aerobic period, carbonaceous oxidation resulted in a further 23% reduction in the COD<sub>f</sub> remaining after the anoxic period. Over the complete treatment cycle, the COD<sub>f</sub> reduction was 95%.

The highest COD<sub>f</sub> removal rate observed during the ISs occurred during the anoxic period of HSWIS 5; a removal rate of 88.28 mg COD<sub>f</sub>/m<sup>2</sup> media/min was achieved. The COD<sub>f</sub> removal rates during the anoxic and aerobic periods of each of the ISs conducted in Phase 2 are outlined in Table 4.9.

**Table 4.9 COD<sub>f</sub> removal rates – Phase 2**

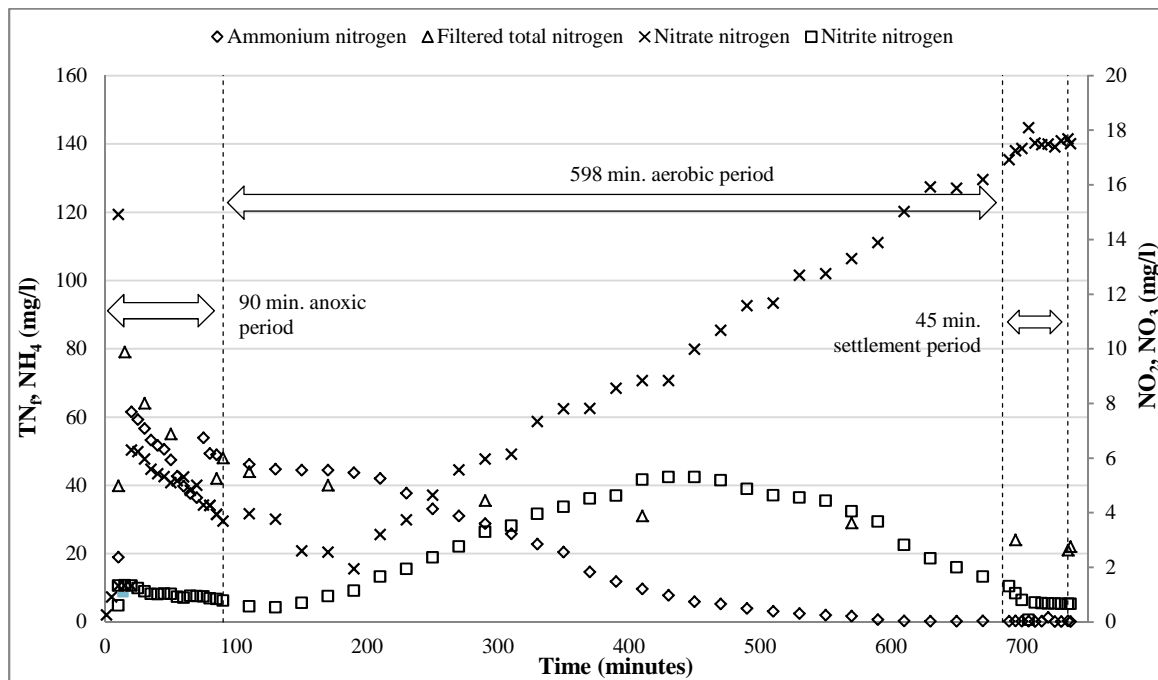
	Anoxic period <sup>1,2</sup>	Time	Aerobic period <sup>3</sup>	Time
	mg COD <sub>f</sub> /m <sup>2</sup> media/min	min	mg COD <sub>f</sub> /m <sup>2</sup> media/min	min
HSWIS 4	31.90	10-90	0.50	90-345
HSWIS 5	27.73	10-90	0.26	90-310
HSWIS 6	20.01	10-90	0.37	90-390

<sup>1</sup> Post dilution; <sup>2</sup> Surface area loading rate calculated based on surface area in R1 (4.1m<sup>2</sup>); <sup>3</sup> Surface area loading rate based on surface area of R1 and R2 (8.2 m<sup>2</sup>).



#### 4.4.3.2 Nitrogen

Nitrogen profiles in all three ISs were found to follow similar trends (Figure 4.11). Nitrification during the aerobic period was the primary  $\text{NH}_4$  removal mechanism. An average of 99% of the  $\text{NH}_4$  remaining after the anoxic period was removed through nitrification, with the highest level of nitrification occurring in HSWIS 6. After approximately 420 minutes of the aerobic period, the concentration was less than 2 mg  $\text{NH}_4/\text{l}$ .



**Figure 4.11 Nitrogen profiles – HSWIS 5**

Note: Initial concentrations is the theoretical post dilution concentrations; Influent concentrations omitted.

This compares very favourably with Phase 1 and would indicate that extending the period for which the wastewater was held in each reactor not only increased the efficiency of the treatment process but also increased treatment performance. Based on these results, further optimisation of the treatment process would be possible.

The  $\text{NH}_4$  removal rates observed during the anoxic and aerobic periods of the ISs conducted in Phase 2 are presented in Table 4.10.

**Table 4.10 NH<sub>4</sub> removal rates – Phase 2**

	<b>Anoxic period *,**</b>	<b>Time</b>	<b>Aerobic period ***</b>	<b>Time</b>
	mg NH <sub>4</sub> /m <sup>2</sup> media/min	min	mg NH <sub>4</sub> /m <sup>2</sup> media/min	min
HSWIS 4	1.92	10-90	0.42	90-445
HSWIS 5	2.70	10-90	0.26	90-530
HSWIS 6	2.41	10-90	0.43	90-530

\* Post dilution; \*\* Surface area loading rate calculated based on surface area in R1 (4.1m<sup>2</sup>); \*\*\*Surface area loading rate based on surface area of R1 and R2 (8.2 m<sup>2</sup>).

The average nitrogen mass balance observed in the ISs conducted during Phase 2 is presented in Table 4.11. The optimised treatment cycle resulted in an improvement in the nitrogen removal performance with the average NH<sub>4</sub> removal efficiency increasing significantly from 96% to 99% which was found to be highly statistically significant (P = 0.0003). The corresponding average effluent NH<sub>4</sub> mass of 0.3 mg/d in Phase 2 compared to an average of 5.3 mg/d in Phase 1 (Table 4.5). In the IS carried out in Phase 2 the percentage NH<sub>4</sub> removal was higher than those observed in Phase 1, however it should be noted that the influent daily mass loads also changed between the ISs carried out in Phase 1 and 2, a change which was statistically insignificant (P = 0.8). Thus daily mass removals of NH<sub>4</sub> were generally higher in Phase 1. In terms of meeting discharge limits Phase 2 had consistently lower effluent concentrations and would therefore meet more stringent regulations. Cell synthesis and the production of N<sub>2</sub>O are likely to have resulted in the imbalance between the influent nitrogen and effluent mass.

**Table 4.11 Nitrogen mass balance – Phase 2 \***

		<b>TN<sub>f</sub></b>	<b>NH<sub>4</sub></b>	<b>NO<sub>2</sub></b>	<b>NO<sub>3</sub></b>
		g/d <sup>1</sup>	g/d <sup>1</sup>	g/d <sup>1</sup>	g/d <sup>1</sup>
<b>Post dilution mass<sup>a</sup></b>	M <sub>p, dil</sub>	3.44 (2.17)	1.88 (2.10)	0.02 (0.01)	1.19 (0)
<b>Mass change - Anoxic period</b>	Δ M <sub>aer</sub>	-0.52 (0.64)	-1.44 (23.4)	- 0.01 (0.02)	-1.07 (0.25)
<b>Mass change - Aerobic period</b>	Δ M <sub>aer</sub>	-1.73 (0.21)	-3.29 (0.48)	0.02 (0.03)	1.26 (0.28)
<b>Effluent</b>	M <sub>eff</sub>	1.82 (0.28)	0.03 (0.03)	0.02 (0.02)	1.42 (0.23)

\* Average results from ISs carried out in Phase 1; <sup>a</sup> – Theoretical average post dilution mass; M<sub>p, dil</sub> – average post dilution mass; M<sub>aer</sub> – average change in mass during aeration period; M<sub>eff</sub> – average effluent mass; <sup>1</sup> 1.96 cycles/d; Standard deviation shown in ().

The linear trends observed between the changes in the concentrations of  $\text{NH}_4$  and  $\text{NO}_3$  and time in each of the ISs is presented in Table 4.12.

**Table 4.12 Summary of  $\text{NH}_4$  and  $\text{NO}_3$  linear trends during HSWIS 4, HSWIS 5 and HSWIS 6**

MWIS	N process	Period	Time	Linear relationship	$R^2$
			min	mg/l	
HSWIS 4	$\text{NH}_4$ decrease	Anoxic	10-90	$\text{NH}_4 = -0.174 t + 57.9$	0.91
	$\text{NO}_3$ decrease	Anoxic	10-90	$\text{NO}_3 = -0.065 t + 7.8$	0.94
	$\text{NH}_4$ decrease	Aerobic	90-445	$\text{NH}_4 = -0.099 t + 44.1$	0.93
	$\text{NO}_3$ increase	Aerobic	90-525	$\text{NO}_3 = 0.067 t - 11.3$	0.94
HSWIS 5	$\text{NH}_4$ decrease	Anoxic	15-90	$\text{NH}_4 = -0.443 t + 69.2$	0.96
	$\text{NO}_3$ decrease	Anoxic	10-90	$\text{NO}_3 = -0.033 t + 6.9$	0.93
	$\text{NH}_4$ decrease	Aerobic	90-530	$\text{NH}_4 = -0.092 t + 48.2$	0.97
	$\text{NO}_3$ increase	Aerobic	90-580	$\text{NO}_3 = 0.019 t - 1.3$	0.94
HSWIS 6	$\text{NH}_4$ decrease	Anoxic	10-90	$\text{NH}_4 = -0.331 t + 81.5$	0.96
	$\text{NO}_3$ decrease	Anoxic	10-90	$\text{NO}_3 = -0.093 t + 7.6$	0.92
	$\text{NH}_4$ decrease	Aerobic	90-530	$\text{NH}_4 = -0.108 t + 55.7$	0.97
	$\text{NO}_3$ increase	Aerobic	90-650	$\text{NO}_3 = 0.025 t - 4.4$	0.98

Heterotrophic denitrification during the anoxic period was found to be predominantly responsible for the reduction in the  $\text{TN}_f$  (Figure 4.17).

#### 4.5 Biofilm aeration

Aeration of the microorganisms which populate the biofilm was achieved in the ASF-BR each time the HSWW was moved to the alternate reactor. The volume of air to which the biofilm was exposed was equal to the volume of HSWW displaced, i.e. 35 l. Of the 35 l of air, approximately 7.35 l consists of oxygen (Hoeppe, 2010). During Phase 2, after approximately 420 minutes of the aerobic period, the concentrations of both COD and  $\text{NH}_4$  remained relatively stable for the duration of the aerobic period (Figure 4.15 and Figure 4.17). This coincides with an increase in the measured DO in the bulk fluid (Figure 4.14a and Figure 4.14b). The rate of DO increase after approximately 400 minutes of aerobic period was 0.005 mg DO/l/min in R1 and 0.007 mg DO/l/min in R2. Given the low levels of biological activity, the observed increase in the DO during this period was used to estimate the physical oxygen transfer capacity of the ASF-BR using the following equation (Tchobanoglous *et al.*, 2004):

$$K_L a = \frac{-\ln\left(\frac{C_S - C}{C_S - C_0}\right)}{t} \quad [Eqn. 4.1]$$

where:  $C_S$  is the saturation DO concentration;  $C_0$  is the initial DO concentration;  $C$  is the DO concentration at time  $t$ .

Thus the average  $K_L a$  of R1 and R2 in Phase 2 was approximately  $0.03 \text{ h}^{-1}$ . Jing *et al.* (2009) investigated the  $K_L a$  of a moving bed biofilm reactor (MBBR) with various fill ratios. The study found that the highest  $K_L a$  was  $0.3 \text{ h}^{-1}$  for optimal operating conditions using mechanical aeration. Jing *et al.* (2009) noted that the oxygen transfer rate can be increased by increasing the aeration intensity which, in the case of the ASF-BR, would require an increase in the frequency of the HSWW movements between the reactors, thus increasing the energy demand.

## 4.6 Suspended solids (SS) and desludging

### 4.6.1 Suspended solids

In Phases 1 and 2, influent SS averaged 203 and 273 mg SS/l while the corresponding average effluent SS were 341 and 721 mg/l respectively. The increase in the SS during the treatment cycle may have been caused by a number of factors: (i) the high strength of the synthetic wastewater, which would have promoted high levels of excess biofilm growth; (ii) collision of the media in suspension, which would have resulted in sloughing of the biofilm from the media, thus increasing the flocculent biomass and the sludge settled in the base of the reactor, and (iii) the discharge point was 50 mm from the base of the reactor to ensure maximum versatility in the volume discharged. Following the settlement period, the flocculent biomass could have settled to this region and may have been discharged with the effluent wastewater. The turbulence caused by the effluent discharge may also have agitated the settled sludge in the base, further increasing the suspended solids in the effluent wastewater.

The issue of high effluent SS can be resolved through a number of design alterations to future ASF-BRs: (i) the use of a static media would reduce the sloughing of the biofilm caused by collision; (ii) a vertical pipe could be installed on the discharge line. which would act as a weir when the wastewater was being discharged, thus eliminating the

discharge of flocculent biomass near the discharge point and preventing agitation of the settled sludge.

#### 4.6.2 Desludging

Desludging of the unit was carried out when a persistent increase in the average effluent SS concentrations was observed. During Phase 1, the unit was desludged on Day 126. All wastewater was drained from the unit and the sludge which had settled at the base of each reactor was removed. A volume of 4.5 l of settled sludge was removed from the base of R1. Settleability testing of the sludge yielded a sludge volume index (SVI) of 75 ml/g. A volume of 4.9 l of settled sludge, with a SVI of 64 ml/g, was removed from the base of R2. The mass of sludge removed from R1 was 39.2 g SS and from R2, 45.7 g SS. This equated to a total of 84.9 g SS sludge removed from the reactors.

On two occasions during Phase 2, the unit was desludged (Days 134 and 210). On Day 134, 3.5 l of settled sludge was removed from the base of R1 while 5.3 l of sludge was removed from the base of R2. The SVI of the sludge removed was 45 ml/g (R1) and 53 ml/g (R2). The mass of sludge removed was 41.7 g SS (R1) and 48.5 g SS (R2). Thus a total of 90.2 g SS was removed from the unit. On Day 210, 3 l of sludge was removed from the base of R1 and 2.8 l from the base of R2. Settleability testing of the sludge, from R1 and R2, gave SVIs of 38 ml/g and 42 ml/g respectively. The mass of sludge removed from R1 and R2 was 42.2 g SS and 43.9 g SS respectively, which equated to a total of 86.1 g SS removed from the reactors. Table 4.13 summarises the total excess sludge production in Study 1.

The SVI results indicate the sludge had excellent settling properties and would have been easily removed by clarification.

**Table 4.13 Estimation of sludge yield during Phase 1 and Phase 2**

	<b>Vol. treated</b>	<b>Desludging</b>	<b><math>\Delta</math>SS</b>	<b><math>\Delta</math>COD</b>	<b>Estimated sludge yield</b>
	l	g	g	g COD <sub>f</sub>	g SS/g COD <sub>f, removed</sub>
Phase 1	1926	84.9	271	3953	0.09
Phase 2	3062	176.3	1137	5507	0.24

#### 4.7 Biofilm analysis

Consistent long-term performance in bioreactors treating nitrogenous wastewater can only be ensured when the microbial community within the sludge functions optimally

(Geets *et al.*, 2007). In this study, real-time PCR was used to quantify the functional ammonia monooxygenase gene (*amoA*) in the reactor biofilm in order to investigate the stability of the ammonia oxidising bacteria (AOB) population over time and in response to a change in operational parameters. The DNA extraction and analysis were conducted as outlined in Section 3.5.5.

#### 4.7.1 Analysis of gene concentrations

The concentration (cells/g biofilm) of AOB on the plastic balls was found to be higher than in the sludge from the reactor tanks (Table 4.14). This was evident in R1 and R2 across all three sampling days (Table 4.14). Moreover, the relative abundance (ratio) of the *amoA*:16S rRNA GCC was, on average, 3.6 times higher in biofilm on the balls than in biomass from the tanks.

**Table 4.14 Biomass analysis results**

		Phase 1 – Day 167			
		R1		R2	
		Biofilm	Sludge	Biofilm	Sludge
GCC <sup>a</sup>	<i>amoA</i>	1.06x10 <sup>7</sup>	1.26 x10 <sup>6</sup>	1.70x10 <sup>7</sup>	9.05x10 <sup>6</sup>
	16S -rRNA	6.09x10 <sup>9</sup>	3.58 x10 <sup>9</sup>	5.55x10 <sup>9</sup>	1.03x10 <sup>10</sup>
	<i>amoA</i> :16S rRNA ratio	1.75x10 <sup>-3</sup>	3.52 x10 <sup>-4</sup>	3.06x10 <sup>-3</sup>	8.81x10 <sup>-4</sup>
	AOB cell numbers/g dry biofilm	5.31x10 <sup>6</sup>	6.31 x10 <sup>5</sup>	8.51x10 <sup>6</sup>	4.53x10 <sup>6</sup>
		Phase 2 – Day 132			
		R1		R2	
		Biofilm	Sludge	Biofilm	Sludge
GCC <sup>a</sup>	<i>amoA</i>	3.34x10 <sup>6</sup>	1.21x10 <sup>6</sup>	2.41x10 <sup>6</sup>	9.17x10 <sup>5</sup>
	16S -rRNA	5.65x10 <sup>9</sup>	3.69x10 <sup>9</sup>	3.46x10 <sup>9</sup>	4.43x10 <sup>9</sup>
	<i>amoA</i> :16S rRNA ratio	5.91x10 <sup>-4</sup>	3.28x10 <sup>-4</sup>	6.97x10 <sup>-4</sup>	2.07x10 <sup>-4</sup>
	AOB cell numbers/g dry biofilm	1.67x10 <sup>6</sup>	6.04x10 <sup>5</sup>	1.20x10 <sup>6</sup>	4.58x10 <sup>5</sup>
		Phase 2 – Day 209			
		R1		R2	
		Biofilm	Sludge	Biofilm	Sludge
GCC <sup>a</sup>	<i>amoA</i>	2.47x10 <sup>6</sup>	7.62x10 <sup>5</sup>	4.40x10 <sup>6</sup>	1.58x10 <sup>6</sup>
	16S -rRNA	3.82x10 <sup>9</sup>	1.53x10 <sup>9</sup>	3.06x10 <sup>9</sup>	3.62x10 <sup>9</sup>
	<i>amoA</i> :16S rRNA ratio	6.47x10 <sup>-4</sup>	4.98x10 <sup>-4</sup>	1.44x10 <sup>-3</sup>	4.38x10 <sup>-4</sup>
	AOB cell numbers/g dry biofilm	1.24x10 <sup>6</sup>	3.81x10 <sup>5</sup>	2.20x10 <sup>6</sup>	7.92x10 <sup>5</sup>

<sup>a</sup> Gene copy concentration (gene copies, g/dry biofilm)

The abundance of AOB in the system was relatively constant throughout the study, indicating that the passive aeration and partial vacuum conditions in the reactors provided a stable environment for the development of a nitrifying bacterial population. Furthermore, there were higher *amoA* gene concentrations in biofilm growing on the plastic media than in the sludge, which indicated that the support balls provided an advantageous surface for the growth of a nitrifying biomass. The amount of *amoA* present in the biofilm was in line with the findings of *Shore et al.* (2012) when treating a synthetic industrial wastewater. The inclusion of the plastic media in the reactor not only provides a large surface area for biofilm attachment but also increases the ammonia oxidizing potential of the system by increasing the concentration of AOB in the biomass.

#### **4.8 Maintenance**

Throughout Phase 1, there were no significant issues relating to the maintenance of the ASF-BR. Occasionally biofilm accumulation on the level switches affected the operational cycle. However, this was easily resolved and could be mitigated in full scale site units through the use of pressure transducers. In Phase 2, the switch used to control the discharge of the wastewater failed as a result of damage to the wire connecting the level switch to the control panel, which required the replacing of the level switch. Again this was an issue which can easily be mitigated through the use of pressure transducers in future units. All valves and the vacuum proved to be very robust and effective in carrying out their required functions.

#### **4.9 Comparison with alternative treatment systems**

During Studies 1 and 2, the ASB-FR unit performed excellently in terms of COD and nitrogen removal when compared with other technologies used in the treatment of high strength SWW, as summarised in Tables 4.15 and 4.16. These tables show that the ASF-BR offers comparable performance to alternative treatment units when operating under less amenable conditions for the biological treatment of wastewater.

**Table 4.15 Comparison of COD removal performance with alternative biofilm treatment systems**

Study	Treatment Process	Wastewater type	Organic loading rate	Removal rate	Comments
ASF-BR – Phase 1		HSWW	3.60 g COD/m <sup>2</sup> media/d 3.00 g COD <sub>T</sub> /m <sup>2</sup> media/d	3.31 g COD/m <sup>2</sup> media/d: 92% removal efficiency 2.93 g COD <sub>T</sub> /m <sup>2</sup> media/d; 98% removal efficiency	Unit operated at 11 °C; Excellent removal of both NH <sub>4</sub> and COD The ASF-BR is capable of capturing all gases produced during the treatment cycle
ASF-BR – Phase 2		HSWW	3.13 g COD/m <sup>2</sup> media/d 2.49 g COD <sub>T</sub> /m <sup>2</sup> media/d	2.64 g COD/m <sup>2</sup> media/d; 84% removal efficiency 2.42 g COD <sub>T</sub> /m <sup>2</sup> media/d; 97% removal efficiency	
Abdulgader <i>et al.</i> , 2009	SBBR	Synthetic dairy wastewater	0.17 g COD/m <sup>2</sup> media/d	97% COD removal efficiency	Study carried out at 22 °C No nitrogen removal presented. Media specific surface area of 2200m <sup>2</sup> /m <sup>3</sup>
Fu <i>et al.</i> , 2009	MBR	HSWW	-	96% COD removal efficiency	Operated at 25 °C
Rodgers <i>et al.</i> , 2008b	APSBBR	Dairy SWW	1.8 g COD/m <sup>2</sup> media/d	96% COD removal efficiency	Unable to capture gaseous emissions
Sirianuntapiboon <i>et al.</i> , 2005	SBBR	Synthetic whey processing wastewater	1.6 g COD <sub>T</sub> /m <sup>2</sup> media/d	97% COD removal efficiency	Hydraulic retention time of 3 days

Nomenclature: SBBR – Sequencing batch biofilm reactor; MBR – Membrane bioreactor; APSBBR – Alternating pumped sequencing batch biofilm reactor



**Table 4.16 Comparison of ammonium removal performance with alternative biofilm treatment systems**

Study	Treatment Process	Wastewater type	Organic loading rate	Removal rate	Comments
	ASF-BR – Phase 1	HSWW	0.24 g NH <sub>4</sub> /m <sup>2</sup> media/d	0.23 g NH <sub>4</sub> /m <sup>2</sup> media/d; 97% removal efficiency	Unit operated at 11°C; Excellent removal of both NH <sub>4</sub> -
	ASF-BR – Phase 2	HSWW	0.21 g NH <sub>4</sub> /m <sup>2</sup> media/d	0.20 g NH <sub>4</sub> /m <sup>2</sup> media/d; 99% removal efficiency	N and COD; Potential to capture of all gases produced
Fu <i>et al.</i> , 2009	MBR	HSWW	-	74% NH <sub>4</sub> removal efficiency	Unit operated at 25°C
Yu <i>et al.</i> , 2007	MBBR	HSWW	0.81 g NH <sub>4</sub> /m <sup>2</sup> media/d	95 % NH <sub>4</sub> removal	Study carried out at 23-29°C; No COD removal presented Forced aeration
Gupta and Gupta, 2001	RBC	HSWW	15 - 48 g NH <sub>4</sub> /m <sup>3</sup> /d	60 – 99% NH <sub>4</sub> removal	15 – 24 HRT; Unit operated at 23°C
Brar and Gupta, 2000	RBC	SWW +trichloro-ethylene	0.39 g NH <sub>4</sub> /m <sup>2</sup> media/d	99.8% NH <sub>4</sub> removal	Cultivated biofilm capable of simultaneous nitrification and denitrification

Nomenclature: MBR – Membrane bioreactor; MBBR – Moving bed biofilm reactor; RBC- Rotating biological contactor.

#### 4.10 Conclusions

This chapter presented the results from an investigation into the efficacy of the ASF-BR in the treatment of high strength synthetic wastewater. The study consisted of two phases: Phase 1 (166 days) and Phase 2 (264 days). In this second phase, the energy requirement was reduced by 66%. During Phase 1, the average COD<sub>f</sub>, TN<sub>f</sub> and NH<sub>4</sub> removals were 98%, 86% and 97% respectively. During Phase 2, the energy requirement of the unit was reduced by 66% while still achieving removal rates of 97% COD<sub>f</sub>, 88% TN<sub>f</sub> and 99% NH<sub>4</sub>. Results obtained from the intensive studies conducted during Study 1 enabled optimisation of the treatment process in Study 2. Meanwhile, ISs conducted in Study 2 indicated that further optimisation of the treatment cycle would be possible in future studies. This initial study on the performance of the ASF-BR has shown that it offers an efficient and effective alternative wastewater treatment process capable of treating HSWW.

# **CHAPTER 5**

## **Study 2 – Treatment of municipal wastewater**

## 5.1 Introduction

In this chapter, the performance of the ASF-BR in treating municipal wastewater is presented. The ASF-BR was operated for a 329-day period during which the carbon, nitrogen and phosphorus removal efficiencies were monitored. During this period, the unit was operated with three different cycle configurations, Phase 1 (212 days), Phase 2a (32 days) and Phase 2b (35 days), each of which are detailed here. The chapter presents overall performance results for each phase and associated results from intensive studies, dissolved oxygen and pH profiles, sludge production, energy requirements and maintenance.

## 5.2 Apparatus

The general set-up of the pilot-scale ASF-BR was described in Section 3.3.2. For logistical purposes, the working volume of each of the reactors was set at 25 l with a media surface area of 2.05 m<sup>2</sup>/reactor. Details of the cycle configuration used in each of the phases are outlined in the relevant subsections.

## 5.3 Phase 1: Nitrification of municipal wastewater

### 5.3.1 Cycle configuration

In Phase 1, the efficacy of the ASF-BR in the removal of organic carbon and ammonium nitrogen from municipal wastewaters through nitrification was analysed. The treatment cycle in Phase 1 was configured to achieve nitrification of the municipal wastewater as outlined in Table 5.1.

**Table 5.1 Cycle configuration pilot scale ASF-BR – Phase 1**

		Units	
Aerobic period	Study duration	d	212
	Step 1 - Fill*	s	30
	Step 2 - Equalisation	s	12
	Step 3 – Draw to R2*	s	20
	Step 4 – Hold in R2	s	90
	Step 5 - Equalisation	s	12
	Step 6 – Draw to R1*	s	20
	Step 7 – Hold in R1	s	90
	Repeat Steps 2 to 7		
	Total: Aerobic period	min	465
Step 8 – Settlement period	min	45	
Step 9 – Discharge*	s	30	
Total: Cycle duration	min	510	
Average daily flow	l/d	36.1 (4.4)	
Average volume/cycle	l	12.1 (1.5)	
Reactor volume	l/reactor	25	
HRT	d	2.1	

\* Maximum duration, actual time may have been less if a signal was received from a level switch; Standard deviation shown in ( ).

### 5.3.2 Influent characteristics

The municipal wastewater used in Phase 1 was sourced on a weekly basis from the influent wastewater channel, following screening and grit removal, before the wastewater entered the primary settlement tank at the Tuam town wastewater treatment plant (TWTP)

Tuam is a town located in north County Galway, which has a population of approximately 8,842 people (WDC, 2012), three schools, and a number of light industrial and commercial facilities, with a resulting population equivalent of 11,000. The TWTP, which serves Tuam and its environs, has a design capacity of 24,834 PE to cater for proposed expansion. As part of its operations, TWTP also accepts sludge from other regional wastewater treatment facilities (additional PE  $\approx$  3,940), along with leachate from a local municipal landfill site (additional PE  $\approx$  7,500) (Ryan Hanley, 2009). The characteristics of the wastewater are outlined in Table 5.2.

### 5.3.4 Overall performance of the ASF-BR – Phase 1

Prior to the commencement of this study, the unit was seeded, as outlined in Section 3.6.2. During the course of this 212-day study, the unit was operating under pseudo steady-state conditions for 175 days. Steady-state conditions (from Day 22) were taken from when the

unit was achieving reasonably consistent effluent results. On Day 111, a fault in the treatment process at TWTP resulted in agitation of the settled sludge in the primary settlement tank, which caused a significant increase in SS taken from site. Hence, for a two-week period, synthetic wastewater was used until this issue was resolved and the plant had regained steady state operating conditions. The results from this period have been omitted from all calculations.

Between Days 0 and 106 (Phase 1a), leachate was added to the TWTP as shock loads overnight when influent flows to the treatment plant were lowest. Thereafter the TWTP management began adding the leachate over a longer period at a gradual rate. This resulted in the presence of residual leachate (albeit diluted) in the wastewater used in this latter stage of the study (Phase 1b). This change had a significant effect on effluent results (in particular nitrogen) and thus results from Phase 1a and Phase 1b are presented.

Influent and effluent samples were taken on a daily basis and analysed for the parameters listed in Table 5.2. Results are presented for the steady state period (Day 22 to Day 212), with the exception of Days 111 to Day 127 due to the issues discussed above. Overall results are presented in Table 5.2. Full detailed results are presented in Appendix E.

Table 5.2 Overall performance of the ASF-BR – Phase 1

Parameter	Phase 1a (Day 21 - 106)						Phase 1b (Day 107 - 212)					
	Average influent	Influent st. dev.	Average Effluent	Effluent st. dev.	n	% Removal	Average influent	Influent st. dev.	Average effluent	Effluent st. dev.	n	% Removal
	mg/l	mg/l	mg/l	mg/l	Inf/Eff		mg/l	mg/l	mg/l	mg/l	Inf/Eff	
COD**	223	114	55	22.2	28/28	75	366	103	108	18	30/29	71
COD <sub>f</sub>	97	41	26	13.3	28/28	63	182	86	51	8	30/29	72
TN**	35.6	10.0	28.1	5.1	25/24	21	54.3	12.1	26.8	6.3	47/52	51
TN <sub>f</sub>	26.8	8.1	24.7	6.0	25/24	8	50.8	13.0	23.1	4.3	47/52	55
NH <sub>4</sub>	20.8	5.8	2.0	1.9	44/43	90	49.2	15.1	6.0	2.6	29/56	88
TON	1.1	1.3	22.9	7.9	44/43	-	0.4	0.8	19.8	6.0	29/56	-
NO <sub>2</sub>	0.2	0.6	0.7	0.8	44/43	-	0.3	0.6	0.9	0.9	29/56	-
NO <sub>3</sub>	0.9	1.4	22.2	7.8	44/43	-	0.1	0.5	18.9	6.2	29/56	-
PO <sub>4</sub> -P	1.6	0.7	0.7	0.5	44/43	57	3.4	1.0	2.2	0.5	29/56	36
Alk	-	-	-	-	44/43	-	346	39.6	149.3	32.2	29/56	57
SS	96	61	49	27	44/43	-	106	72	86	74	29/56	-

All samples marked \*\* were measured unfiltered, all other samples were filtered; n number of samples; n is the number of samples; Inf – Influent; Eff – Effluent.

### 5.3.4.1 Organic carbon

The change in the management of leachate (changed from batch-type addition to a more continual method of delivery) at the TWTP resulted in an increase in the average influent COD and COD<sub>f</sub> of 66% and 88% respectively to 366 mg COD/l and 182 mg COD<sub>f</sub>/l (Phase 1b) from 223 g COD/l and 97 g COD<sub>f</sub>/l (Phase 1a). This was a result of the addition of the landfill leachate to the municipal wastewater. The leachate had typical COD and COD<sub>f</sub> concentrations of 1756 mg/l and 1512 mg /l respectively. The observed increase in the COD and COD<sub>f</sub> loading, due to the revised operating procedures was found to a statistically significant ( $P < 0.05$ ).

Average effluent COD and COD<sub>f</sub> concentrations increased by 97% and 44% in Phase 1b when compared to Phase 1a. However COD and COD<sub>f</sub> removal efficiencies in Phase 1b remained relatively unchanged, with removals similar to those of observed in Phase 1a.

**Table 5.3 Organic carbon loading and removal rates – Phase 1**

	<b>COD loading</b> g COD/m <sup>2</sup> media/d	<b>COD removal</b> g COD/m <sup>2</sup> media/d	<b>% COD removal</b>	<b>COD<sub>f</sub> loading</b> g COD <sub>f</sub> /m <sup>2</sup> media/d	<b>COD<sub>f</sub> removal</b> g COD <sub>f</sub> /m <sup>2</sup> media/d	<b>% COD<sub>f</sub> removal</b>
Phase 1a <sup>a, c</sup>	1.87	1.41	75	1.19	0.52	63
Phase 1b <sup>b, c</sup>	3.35	2.38	71	1.67	1.20	72

<sup>a</sup> average influent volume of 34.6 l/d; <sup>b</sup> average influent volume of 37.6 l/d; <sup>c</sup> surface area of 4.1 m<sup>2</sup>.



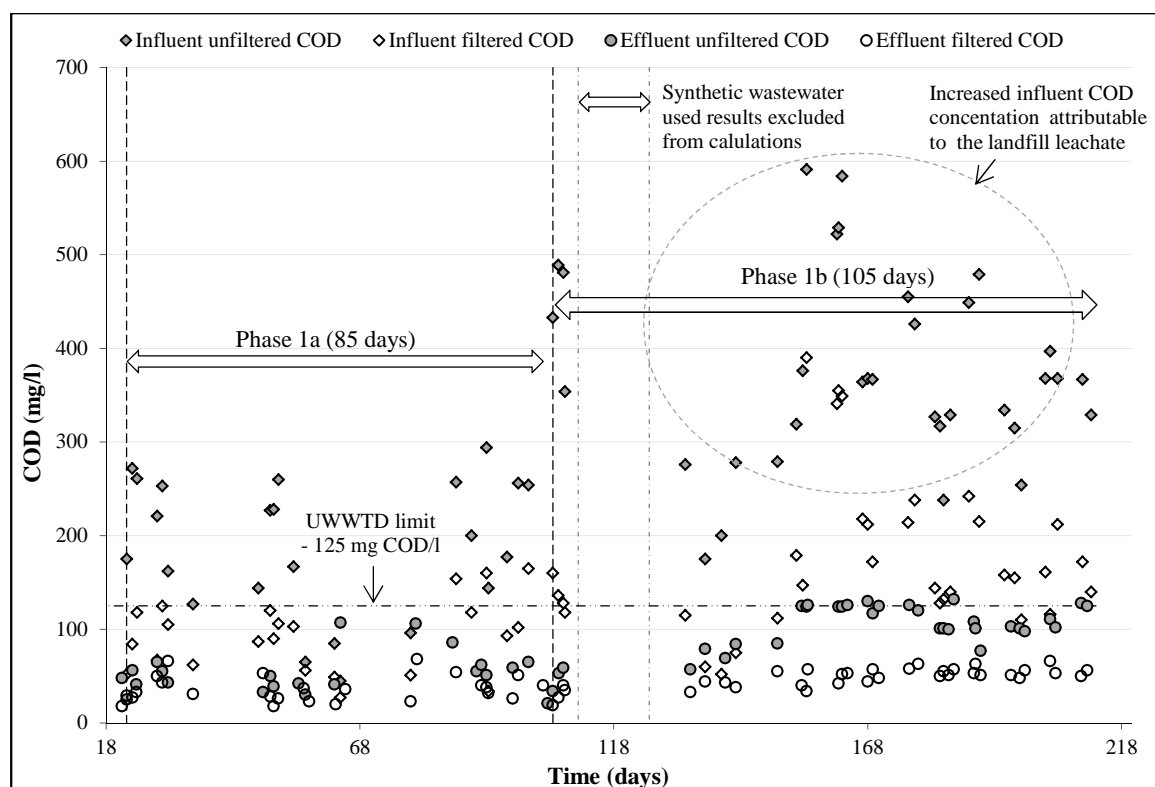


Figure 5.1 Influent and effluent COD and  $COD_f$  – Phase 1

#### 5.3.4.2 Nitrogen

In Phase 1a, influent concentrations averaging 20.8 mg  $NH_4$ /l were reduced by 90% to an average effluent concentration of 2.0 mg  $NH_4$ /l. In Phase 1b, the average influent  $NH_4$  concentrations increased by 136% to 49.2 mg  $NH_4$ /l when compared to Phase 1a. Effluent concentrations in Phase 1b averaged 6.0 mg  $NH_4$ /l, equating to an average reduction of 88%. Area specific removals increased significantly between Phases 1a and 1b, from 0.16 g/m<sup>2</sup>/day to 0.40 g/m<sup>2</sup>/day, indicating the ASF-BR adapted well to increased  $NH_4$  loads. The increase in the  $NH_4$  mass loading rate and the effluent  $NH_4$  mass, as a result of the revised leachate addition methodology were both found to be insignificant ( $P > 0.05$ ). The  $NH_4$  loading and removal rates during Phase 1 are summarised in Table 5.4 and Figure 5.2.

Table 5.4  $NH_4$  loading and removal rates – Phase 1

	NH <sub>4</sub> loading	NH <sub>4</sub> removal	% NH <sub>4</sub> removal
	rate	rate	
	g NH <sub>4</sub> /m <sup>2</sup>	g NH <sub>4</sub> /m <sup>2</sup>	
	media/d	media/d	
Phase 1a <sup>a, c</sup>	0.18	0.15	90
Phase 1b <sup>b, c</sup>	0.45	0.40	88

<sup>a</sup> average influent volume of 34.6 l/d; <sup>b</sup> average influent volume of 37.6 l/d; <sup>c</sup> surface area of 4.1m<sup>2</sup>

Influent  $\text{NO}_3$  concentrations averaged 0.23 mg  $\text{NO}_3$  /l in Phase 1a and 0.11 mg  $\text{NO}_3$ /l in Phase 1b. During the aerobic period, the oxidation of the  $\text{NH}_4$  resulted in an increase in the  $\text{NO}_3$  concentration, which resulted in an average effluent  $\text{NO}_3$  concentration of 22.2 mg  $\text{NO}_3$ /l in Phase 1a and 18.9 mg  $\text{NO}_3$ /l in Phase 1b.

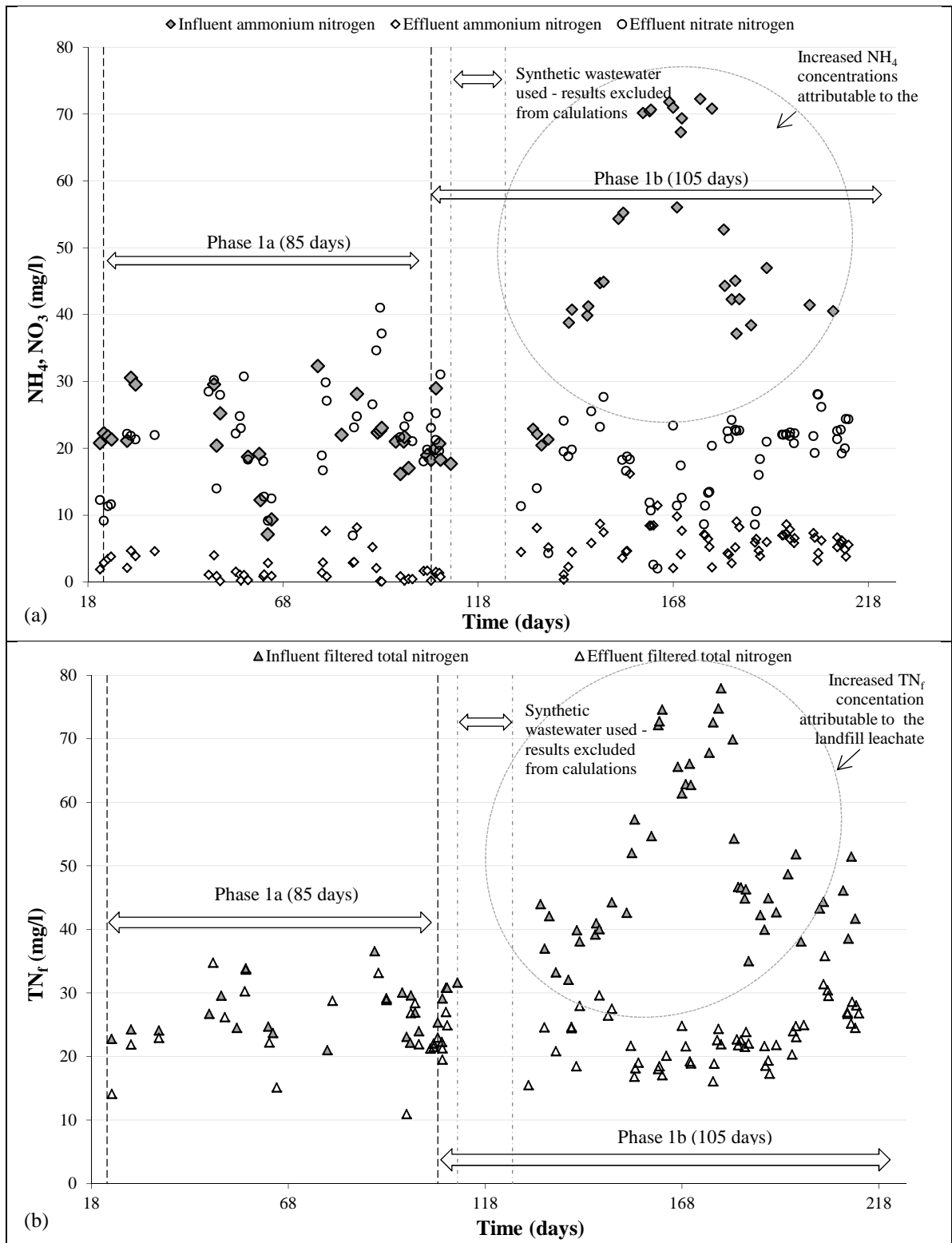
Prior to the implementation of a revised strategy for the addition of the leachate on Day 106 (Phase 1a), the average TN and  $\text{TN}_f$  concentrations were 35.6 mg TN/l and 26.8 mg  $\text{TN}_f$ /l respectively. The corresponding average effluent concentrations were 31.8 mg TN/l and 26.5 mg  $\text{TN}_f$ /l. In Phase 1b, the average influent TN and  $\text{TN}_f$  concentrations increased by 35% and 47% respectively to 54.3 mg TN/l and 50.8 mg  $\text{TN}_f$ /l, with corresponding effluent concentrations of 26.8 mg TN/l and 23.1 mg  $\text{TN}_f$ /l (Figure 5.2 (b)). It is clear from these results that the landfill leachate had a significant effect on influent loads to TWTP. Similar to the  $\text{NH}_4$  results, it was noted that removal rates increased significantly between Phases 1a and 1b. The loading and removal rates are summarised in Table 5.5.

**Table 5.5 TN loading and removal rates – Phase 1**

	<b>TN loading rate</b>	<b>TN removal rate</b>	<b>% TN removal</b>	<b><math>\text{TN}_f</math> loading rate</b>	<b><math>\text{TN}_f</math> removal rate</b>	<b>% <math>\text{TN}_f</math> removal</b>
	g $\text{TN}/\text{m}^2$ media/d	g $\text{TN}/\text{m}^2$ media/d		g $\text{TN}_f/\text{m}^2$ media/d	g $\text{TN}_f/\text{m}^2$ media/d	
Phase 1a <sup>a, c</sup>	0.30	0.06	20	0.22	0.02	8
Phase 1b <sup>b, c</sup>	0.50	0.25	50	0.47	0.25	55

<sup>a</sup> average influent volume of 34.6 l/d; <sup>b</sup> average influent volume of 37.6 l/d; <sup>c</sup> surface area of 4.1m<sup>2</sup>

In Phase 1a, nitrogen assimilation for cell synthesis (Section 3.4.4.3) accounted for an estimated  $\text{TN}_f$  reduction of 3.6 mg  $\text{TN}_f$ /l. The observed  $\text{TN}_f$  reduction was slightly lower at 2.1 mg  $\text{TN}_f$ /l. However, during Phase 1b, less than 25% or approximately 6.6 mg  $\text{TN}_f$ /l of the  $\text{TN}_f$  removed was due to cell synthesis. The balance of the 28 mg  $\text{TN}_f$ /l was possibly due to simultaneous nitrification and denitrification (SND) occurring within the biofilm. Henze *et al.* (2002) and Biesterfeld *et al.* (2003) found that within a biofilm, anoxic layers can develop even where the bulk fluid DO concentrations were as high as 5 mg  $\text{O}_2$ /l in the bulk fluid.



**Figure 5.2 (a) Influent and effluent ammonium nitrogen and effluent nitrate nitrogen – Phase 1 (b) Influent and effluent filtered total nitrogen – Phase 1**

#### 5.3.4.3 Alkalinity

The alkalinity of the influent wastewater was monitored from Day 181 to ensure the availability was not inhibiting the nitrification process. Over this period, the alkalinity

consumption rate was 4.6 g CaCO<sub>3</sub>/g NH<sub>4</sub>. The average effluent alkalinity was 149 g CaCO<sub>3</sub>/l, which indicated that alkalinity was not limiting the nitrification process.

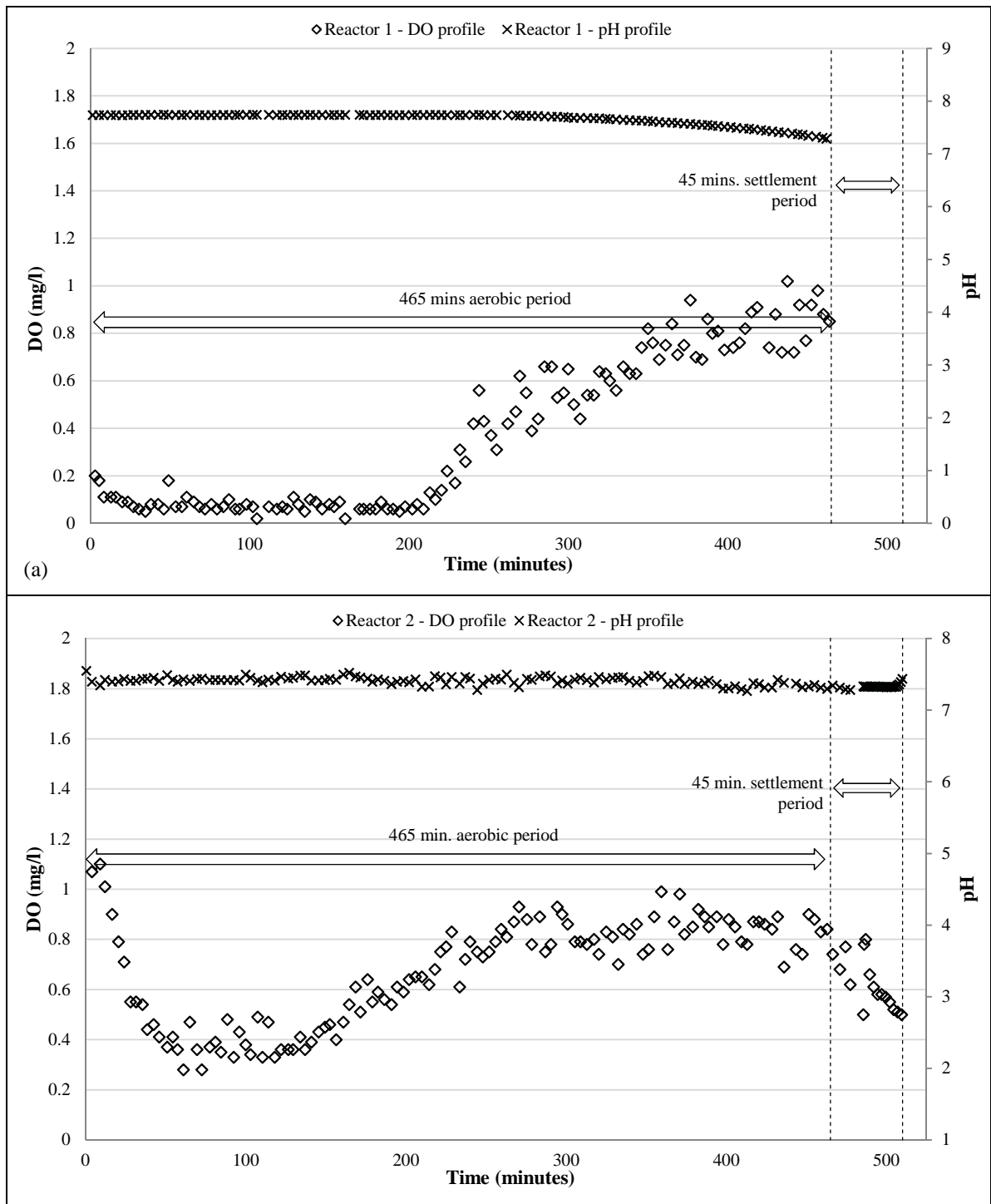
#### 5.3.4.4 Phosphorus

The removal of phosphorus was not one of the objectives of this study as conditions compatible with biological phosphorus removal were not present (Section 2.6). However, as part of the study, phosphorus concentrations were monitored (Figure 5.3). In Phase 1a, the average influent PO<sub>4</sub>-P concentration was 1.6 mg/l while the corresponding average effluent concentration was 0.7 mg/l. In Phase 1b, the average influent PO<sub>4</sub>-P increased to 3.4 mg/l with the average effluent concentration increasing to 2.2 mg/l. Influent phosphorous concentrations generally increased throughout Phase 1b with effluent concentrations increasing correspondingly. This may be attributable to landfill leachate although seasonal variations in the municipal wastewater may also have had an impact.

#### 5.3.4.5 DO and pH

Figure 5.3 depicts typical DO and pH profiles obtained from R1 (Figure 5.3a) and R2 (Figure 5.3b) (Phase 1b - Days 163 and 189 respectively). The DO profiles observed in from both reactors are consistent with the oxygen demands which occur during the nitrification process and carbon removal. Oxidation of NH<sub>4</sub> is a more oxygen demanding process than the oxidation of the NO<sub>2</sub>. Thus as nitrification proceeds, NH<sub>4</sub> is converted to NO<sub>2</sub> and the oxygen demand is reduced, and there was an observed increase in the DO in the bulk fluid. DO concentrations remained relatively low throughout the aerobic period in both reactors. This could be alleviated by increasing the number of pumping cycles in an aerobic period. This, in turn, increases energy costs; however, control of the pumping cycles using DO sensors could form part of future optimisation studies.

pH values decreased towards the end of the aerobic period as a result of the nitrification process (which results in CO<sub>2</sub> production and can lower pH).



**Figure 5.3 Typical DO and pH profile R1 – Phase 1: (a) – R1 (Day 163);  
Phase 1: (b) - R2 (Day 189)**

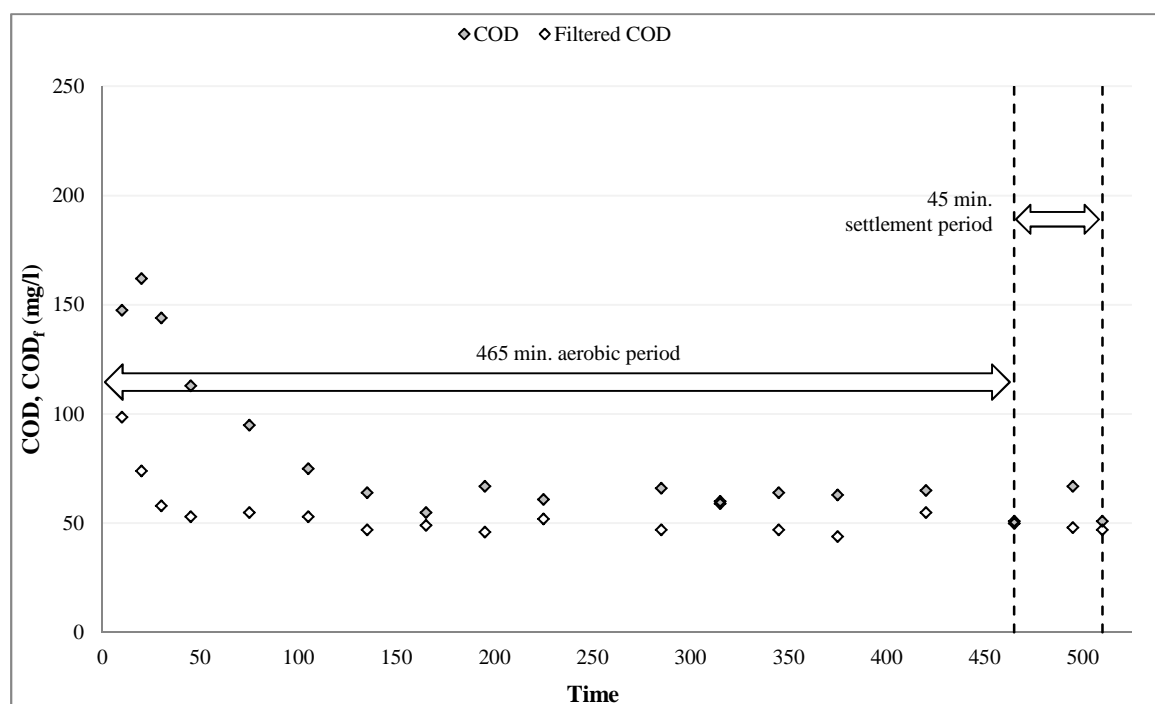
### 5.3.5 Phase 1 - Intensive studies

Three ISs, MWIS 1, MWIS 2 and MWIS 3, were carried out over the course of this phase on Days 55, 140 and 191 respectively. During each IS, samples were taken every five minutes for the first 45 minutes and every 15 minutes thereafter. Results of the organic

carbon and nitrogen removals are presented herein. Detailed results of all ISs are summarised below, with full details available in Appendix E.

### 5.3.5.1 Organic Carbon

COD and COD<sub>f</sub> removals were 68% COD and 79% COD<sub>f</sub>, 72% and 71% COD and COD<sub>f</sub> and 79% COD and 58% during MWIS 1, MWIS 2 and MWIS 3 respectively. The carbon profile observed during MWIS 1 is presented in Figure 5.4, with profiles of MWIS 2 and MWIS 3 contained in Appendix E.



**Figure 5.4 COD and COD<sub>f</sub> profiles – MWIS 1**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

Following the initial dilution of the influent wastewater with the wastewater remaining in the reactor following the previous treatment cycle, COD and COD<sub>f</sub> concentrations were reduced through aerobic oxidation over a period of approximately 100 to 150 minutes. After this time, the COD and COD<sub>f</sub> concentrations were found to remain relatively stable throughout the remainder of all ISs, with the remaining COD and COD<sub>f</sub> possibly consisting of the non-biodegradable COD and COD<sub>f</sub>.

Figure 5.6 presents the percentage COD<sub>f</sub> removal achieved during each stage of the treatment cycle during MWIS 1, MWIS 2 and MWIS 3.

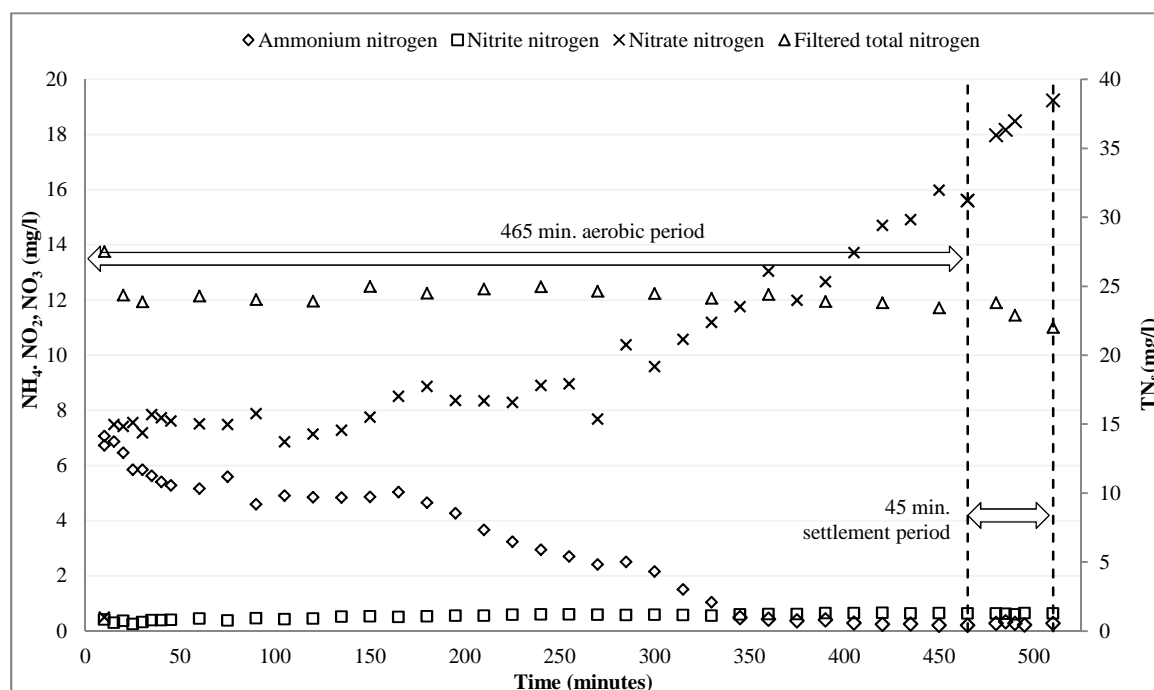
**Table 5.6 COD<sub>f</sub> oxidation rates – Phase 1**

	Aerobic period <sup>1,2</sup>		Time
	mg COD/m <sup>2</sup> media/min**	mg COD <sub>f</sub> /m <sup>2</sup> media/min**	min
MWIS 1	3.63	1.95	10 - 120
MWIS 2	5.38	2.35	10 - 165
MWIS 3	3.51	1.58	10 - 155

<sup>1</sup> Post dilution; <sup>2</sup>Surface area loading rate based on surface area of R1 and R2 (4.1 m<sup>2</sup>).

### 5.3.5.2 Nitrogen

Figure 5.9 presents the nitrogen profiles from MWIS 2, which is similar to those from MWIS 1 and MWIS 3 (Appendix E). The oxidation of NH<sub>4</sub> resulted in an average NH<sub>4</sub> removal rate of 91% over the three ISSs. During the first approximately 60 minutes of the aerobic period in MWIS 2 (Figure 5.5), the NH<sub>4</sub> concentrations were found to decrease without a corresponding increase in the NO<sub>2</sub> or NO<sub>3</sub> concentrations. During this time, the concentration of COD and COD<sub>f</sub> (Figure 5.4) were also found to decrease. These trends combined with the low DO in both R1 and R2 (Figure 5.3 (a) and Figure 5.3 (b)) tend to indicate that simultaneous nitrification and denitrification (SND) occurred at the beginning of the treatment cycles.

**Figure 5.5 Nitrogen profiles – MWIS 2**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

Evidence of nitrification in MWIS 2 was observed after 165 minutes into the treatment cycle, when NH<sub>4</sub> began to decrease and NO<sub>3</sub> began to increase. After approximately 180

minutes, the  $\text{NH}_4$  stabilised at  $> 1 \text{ mg NH}_4/\text{l}$ . A similar trend was observed in MWIS 1, while in MWIS 3, nitrification continued throughout the aerobic period. Nitrification rates from the three ISs are presented in Table 5.7.

**Table 5.7  $\text{NH}_4$  removal rates – Phase 1**

	Aerobic period <sup>1,2</sup>	Time
	$\text{mg NH}_4/\text{m}^2 \text{ media}/\text{min}$	min
HSWIS 1	0.23	10-330
HSWIS 2	0.21	10-330
HSWIS 3	0.15	10-465

<sup>1</sup> Post dilution; <sup>2</sup> Surface area loading rate based on surface area of R1 and R2 ( $4.1 \text{ m}^2$ ).

In all three ISs, the oxidation of the  $\text{NH}_4$  resulted in a linear increase in the  $\text{NO}_3$ , which commenced approximately 45 minutes after the  $\text{NH}_4$  oxidation began. The production rates of  $\text{NO}_3$  as a result of the nitrification process in MWIS 1, MWIS 2 and MWIS 3 were  $0.21 \text{ mg NO}_3/\text{m}^2 \text{ media}/\text{min}$  (MWIS 1),  $0.19 \text{ mg NO}_3/\text{m}^2 \text{ media}/\text{min}$  (MWIS 2) and  $0.15 \text{ mg NO}_3/\text{m}^2 \text{ media}/\text{min}$  (MWIS 3).  $\text{NO}_2$  concentrations during all three ISs were consistently low ( $< 0.8 \text{ mg NO}_2/\text{l}$ ), indicating that nitrification was not inhibited by a lack of oxygen or alkalinity (such inhibition would likely have resulted in partial nitrification and accumulation of nitrite). The  $\text{NH}_4$  and  $\text{NO}_3$  production and removal trends are summarised in Table 5.8.

**Table 5.8 Summary of the  $\text{NH}_4$  and  $\text{NO}_3$  linear trends during MWIS 1, MWIS 2 and MWIS 3**

MWIS	N process	Period	Time	Linear trend	$R^2$
			min	$\text{mg/l}$	
MWIS 1	$\text{NH}_4$ decrease	Aerobic	105-315	$\text{NH}_4 = -0.028 t + 8.9$	0.97
	$\text{NO}_3$ increase		155-465	$\text{NO}_3 = 0.030 t + 5.2$	0.93
MWIS 2	$\text{NH}_4$ decrease	Aerobic	165-330	$\text{NH}_4 = -0.022 t + 8.5$	0.97
	$\text{NO}_3$ increase		225-465	$\text{NO}_3 = 0.034 t + 0.1$	0.94
MWIS 3	$\text{NH}_4$ decrease	Aerobic	10-465	$\text{NH}_4 = -0.020 t + 11.3$	0.98
	$\text{NO}_3$ increase		20-465	$\text{NO}_3 = 0.025 t + 3.6$	0.96

Table 5.9 presents the average nitrogen mass balance observed in the ISs conducted in Phase 1. The imbalance between the observed decrease in the  $\text{NH}_4$  and the increase in the  $\text{NO}_3$  may have been caused by the utilisation of nitrogen in cell synthesis.



**Table 5.9 Nitrogen mass balance – Phase 1 \***

		TN <sub>f</sub>	NH <sub>4</sub>	NO <sub>2</sub>	NO <sub>3</sub>
		g/d <sup>1</sup>	g/d <sup>1</sup>	g/d <sup>1</sup>	g/d <sup>1</sup>
<b>Post dilution mass</b> <sup>a</sup>	M <sub>p, dil</sub>	2.01 (0.52)	0.92 (0.08)	0.03 (0.02)	0.70 (0.07)
<b>Mass change - Aerobic period</b>	Δ M <sub>aer</sub>	-0.30 (0.07)	-0.86(0.05)	0.01 (0)	0.51 (0.15)
<b>Effluent</b>	M <sub>eff</sub>	1.66 (6.5)	0.04 (0.03)	0.04 (0.01)	1.31 (0.6)

\* Average results from ISs carried out in Phase 1; <sup>a</sup> – Theoretical average post dilution mass; M<sub>p, dil</sub> – average post dilution mass; M<sub>aer</sub> – average change in mass during aeration period; M<sub>eff</sub> – average effluent mass; <sup>1</sup> 2.8 cycles/d; Standard deviation shown in ().

The TN and TN<sub>f</sub> removal efficiency was similar in all three ISs, with an average TN and TN<sub>f</sub> removal efficiencies of 32% and 31% respectively, with dilution accounting for a significant portion of the concentration change (Figure 5.9). Following dilution, the majority of the reduction in TN and TN<sub>f</sub> in all three ISs occurred within a 75-minute period, with concentrations remaining stable thereafter. In MWIS 1, cell synthesis accounted for a TN<sub>f</sub> reduction of approximately 1.3 mg TN<sub>f</sub>/l (Section 3.4.4.3) of the 4.4 mg TN<sub>f</sub>/l reduction observed in a 60-minute period following dilution, with SND likely to be responsible for the removal of the remainder. In MWIS 2 and MWIS 3, the estimated TN<sub>f</sub> utilised in cell synthesis accounted for the majority of the observed TN<sub>f</sub> reduction.

#### 5.4 Phase 2: Nitrification and denitrification of municipal wastewater

Phase 2 followed on directly from Phase 1; however, the source of the municipal wastewater was changed to Mutton Island Wastewater Treatment Plant (MIWTP), in Galway city. This change was as a result of the addition of landfill leachate to the influent stream in the TWTP, and the resulting potential for inhibition due to compounds within the leachate. While Phase 1 focused on organic carbon removal and nitrification, the objective of Phase 2 was to evaluate the efficacy of the ASF-BR in achieving organic carbon removal, nitrification and denitrification of municipal wastewater.

The results presented for Phase 2 have been divided into Phase 2a, when the municipal wastewater was treated without any amendments, and Phase 2b, during which glucose was added to the municipal wastewater. A pseudo steady state, where effluent contaminant concentrations remained relatively steady, was observed in Phase 2a after about 30 days. Between Days 30 and 62, during which time three ISs were conducted, it was found that expected levels of denitrification were not being achieved due to insufficient biodegradable carbon. As such, it was decided to add an external carbon

source to optimise the denitrification process during Phase 2b. Glucose was chosen as the source of external carbon. The quantity added was determined experimentally over a period of 20 days.

#### 5.4.1 Cycle configuration

The operational configuration of Phases 2a and 2b is outlined in Table 5.10. By reviewing the IS results from Phase 1, it was found that the treatment process could be optimised by shortening the duration of the aerobic period from 465 minutes to 185 minutes. An anoxic period of 90 minutes was scheduled at the beginning of the treatment cycle to facilitate denitrification. The configuration of the steps during the shortened aerobic period remained the same as those in Phase 1.

**Table 5.10 Cycle configuration pilot scale ASF-BR – Phase 2**

	Units	Time	
Study duration	d	117	
Step 1 – Fill*	s	30	
Step 2 – Anoxic period	min	90	
Aerobic period	Step 3 - Equalisation	s	12
	Step 4 – Draw to R2*	s	20
	Step 5 – Hold in R2	s	90
	Step 6 - Equalisation	s	12
	Step 7 – Draw to R1*	s	20
	Step 8 – Hold in R1	s	90
	Repeat Steps 3 to 7		
	Aerobic period	min	185
Step 9 – Settlement period	min	25	
Step 10 – Discharge*	s	30	
Total cycle duration	min	300	
Average daily flow	l/d	35 (0.82)**	
Average volume/cycle	l	7.3 (0.2)	
Reactor volume	l/reactor	25	
HRT	d	3.4	

\* Maximum duration, actual time may have been less if a signal was received from a level switch; Standard deviation shown in ( ).

#### 5.4.2 Influent characteristics

MIWTP serves the wastewater treatment needs of the Galway city agglomerate and, as such, the wastewater would originate primarily from municipal, residential, commercial and light industrial sources. The treatment plant was opened in 2004 and was designed to

cater for a PE of 91,600. As a result of the unprecedented population growth of Galway city and its environs, the plant is currently operating above this design capacity and tenders are currently being sought to upgrade the plant to cater for a peak PE of 170,000 (PC, 2012). The wastewater used in Phase 2 was collected before the wastewater entered the primary settlement tanks following grit removal and screening on a weekly basis. The characteristics of the wastewater used as the influent during Phase 2 are presented in Table 5.10.

#### ***5.4.3 Overall performance of the ASF-BR – Phase 2***

The performance of the ASF-BR during Phase 2 is summarised in Table 5.11. The performance has been divided into phases: Phase 2a refers to Days 0 to 62, when raw wastewater was used without any amendments, while during Phase 2b (Days 83-117), an external carbon source in the form of glucose was added at a rate of 0.8 g glucose ( $C_6H_{12}O_6$ )/l.

Table 5.11 Overall performance of the ASF-BR – Phase 2

Parameter	Phase 2a <sup>1</sup> (Day 32-68)						Phase 2b <sup>2</sup> (Day 88-118)					
	Average Influent	Influent st. dev.	Average effluent	Effluent st. dev.	n	% Removal	Average Influent	Influent st. dev.	Average effluent	Effluent st. dev.	n	% Removal
	mg/l	mg/l	mg/l	mg/l	Inf/Eff		mg/l	mg/l	mg/l	mg/l	Inf/Eff	
COD**	359	82	90	36	12/9	75	870	106	223	46	9/8	75
COD <sub>f</sub>	79	29	39	9	12/9	50	554	121	62	14	9/8	89
TN**	23.2	5.1	10.9	3.0	39/30	53	16.5	3.6	4.5	0.7	12/19	73
TN <sub>f</sub>	21.5	4.5	9.2	2.4	39/30	57	15.9	3.6	3.9	0.6	12/19	75
NH <sub>4</sub>	21.9	5.2	1.4	1.1	40/30	95	16.0	3.7	0.6	0.6	12/19	93
TON	0.9	1.3	5.8	2.6	40/30	-	0.3	0.2	0.3	0.2	12/19	-
NO <sub>2</sub>	0.1	0.2	0.5	0.6	40/30	-	0.1	0.2	0	0.1	12/19	-
NO <sub>3</sub>	0.8	1.3	5.3	2.3	40/30	-	0.1	0.1	0.2	0.1	12/19	-
PO <sub>4</sub> -P	1.1	0.5	0.5	0.2	40/30	55	0.8	1.1	0.2	0.2	12/19	70
Alk	219	71	117	27.7	40/30	-	269	25	166	13.7	12/19	-
SS	380	147	61.0	24.1	40/30	84	210	70	123	28.9	12/19	41

All parameters marked \*\* were measured unfiltered, all other samples were filtered; n is the number of samples; Inf – Influent; Eff – Effluent.

### 5.4.3.1 Organic carbon

Influent and effluent COD and COD<sub>f</sub> recorded during Phase 2a and Phase 2b are presented in Figure 5.6. During Phase 2a, the average influent COD and COD<sub>f</sub> concentrations were 359 mg COD/l and 79 mg COD<sub>f</sub>/l respectively. The removal rates were relatively stable: 75% of the influent COD and 50% of the influent COD<sub>f</sub> were removed resulting in effluent concentrations of 90 mg COD/l and 39 mg COD<sub>f</sub>/l.

In Phase 2b, the addition of glucose at a rate of 0.8 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l resulted in an increase in average influent concentrations to 870 mg COD/l and 554 mg COD<sub>f</sub>/l respectively which was found to be significant ( $P > 0.05$ ), which would be expected given the addition of an external source of carbon. The removal efficiencies again remained reasonably consistent, at 74% COD and 89% COD<sub>f</sub>, resulting in average effluent concentrations of 223 mg COD/l and 62 mg COD<sub>f</sub>/l. This increase in effluent COD was likely due to the addition of glucose to the influent, resulting in a carbon breakthrough, as noted by Chui and Chung (2003). The loading rates and removal rates achieved during Phase 2a and Phase 2b are summarised in Table 5.12.

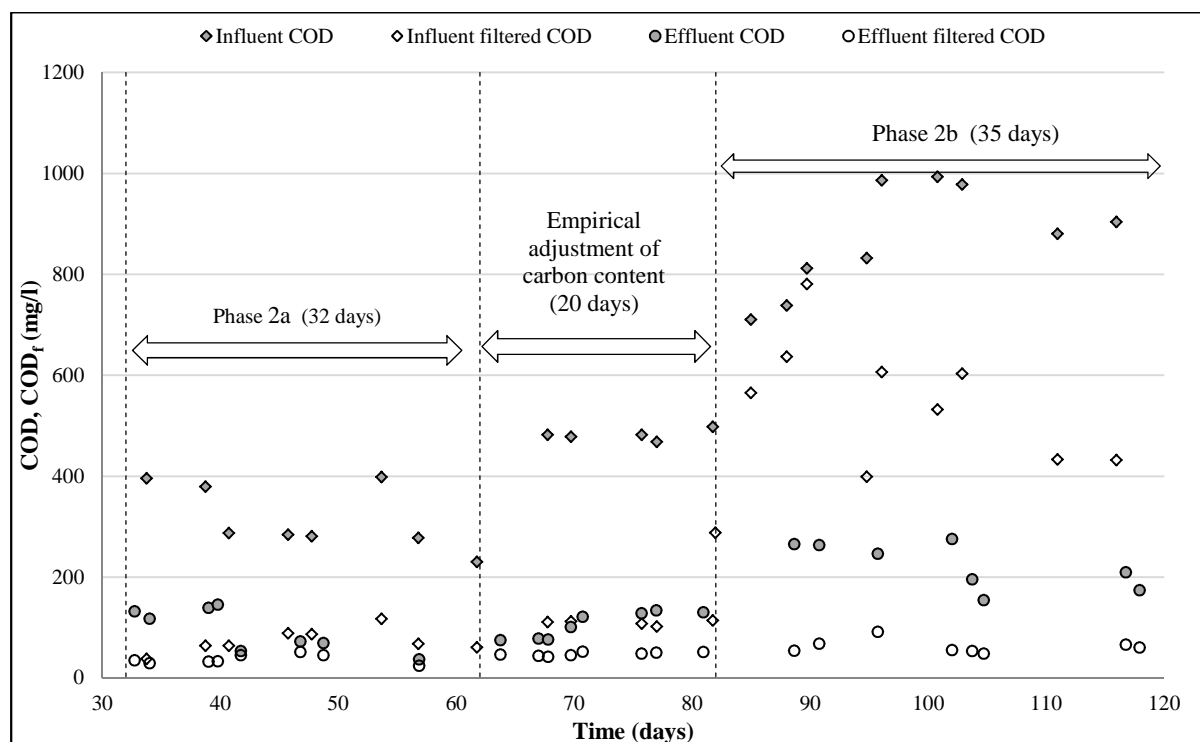


Figure 5.6 Influent and effluent COD and COD<sub>f</sub> – Phase 2

**Table 5.12 Organic carbon loading and removal rates – Phase 2**

	<b>COD loading</b>	<b>COD removal</b>	<b>% COD removal</b>	<b>COD<sub>f</sub> loading</b>	<b>COD<sub>f</sub> removal</b>	<b>% COD<sub>f</sub> removal</b>
	$\frac{\text{g COD/m}^2}{\text{media/d}}$	$\frac{\text{g COD/m}^2}{\text{media/d}}$		$\frac{\text{g COD}_f/\text{m}^2}{\text{media/d}}$	$\frac{\text{g COD}_f/\text{m}^2}{\text{media/d}}$	
Phase 2a <sup>a, c</sup>	3.07	2.30	75	0.68	0.34	50
Phase 2b <sup>b, c</sup>	7.41	5.51	74	4.72	4.19	89

<sup>a</sup> average influent volume of 35.1 l/d; <sup>b</sup> average influent volume of 34.9 l/d; <sup>c</sup> surface area of 4.1m<sup>2</sup>.

#### 5.4.3.2 Nitrogen

Figure 5.7 presents the nitrogen data from Phase 2a and 2b. During Phase 2a, the average influent NH<sub>4</sub> concentration was 21.9 mg NH<sub>4</sub>/l, resulting in a surface area loading rate of 0.19 g NH<sub>4</sub>/m<sup>2</sup> media/d. The influent NH<sub>4</sub> was reduced by an average of 95%, resulting in an average effluent concentration of 1.4 mg NH<sub>4</sub>/l. The average influent NH<sub>4</sub> concentration during Phase 2b was 16.0 mg NH<sub>4</sub>/l, which was reduced by an average of 93%, resulting in an effluent concentration 0.6 mg NH<sub>4</sub>/l. The lower concentrations of influent NH<sub>4</sub> observed during Phase 2b, which were found to be significant ( $P < 0.05$ ), were possibly a result of high levels of rainfall during the winter months and reduced tourism in Galway city. The NH<sub>4</sub> loading and removal rates achieved during Phase 2 are presented in Table 5.13. There was a slight decrease in removal efficiency between Phases 2a and 2b, possibly due to a combination of lower NH<sub>4</sub> loading rates and the impact of additional carbon on nitrification activity in the system.

**Table 5.13 NH<sub>4</sub> loading and removal rates – Phase 2**

	<b>NH<sub>4</sub> loading rate</b>	<b>NH<sub>4</sub> removal rate</b>	<b>% NH<sub>4</sub> removal</b>
	$\frac{\text{g NH}_4/\text{m}^2}{\text{media/d}}$	$\frac{\text{g NH}_4/\text{m}^2}{\text{media/d}}$	
Phase 2a <sup>a, c</sup>	0.19	0.18	95
Phase 2b <sup>b, c</sup>	0.14	0.13	93

<sup>a</sup> average influent volume of 35.1 l/d; <sup>b</sup> average influent volume of 34.9 l/d; <sup>c</sup> surface area of 4.1m<sup>2</sup>.

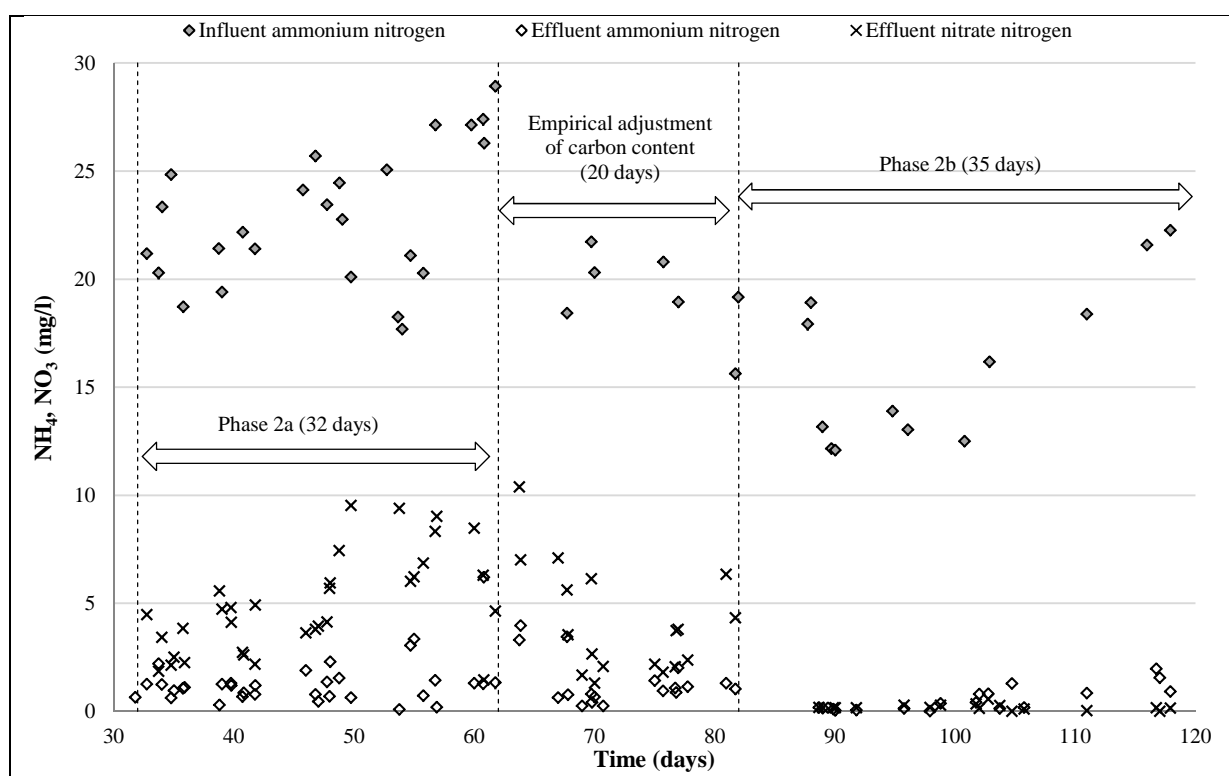
The average influent TN and TN<sub>f</sub> concentrations in Phase 2a were 23.2 mg TN/l and 21.5 mg TN<sub>f</sub>/l, which were reduced by 53% and 57% respectively. It was felt that the relatively low nitrogen removal rates may have been due to low C:N ratio in the influent wastewater. As is detailed further below, dissolved oxygen concentrations were near 0 mg/l during the anoxic stage and thus it is unlikely that oxygen inhibited the denitrification process.

As a result, it was decided to amend the influent wastewater with glucose, which increased the biodegradable organic carbon content in the wastewater for Phase 2b. During Phase 2b, the average effluent TN and  $TN_f$  concentrations were 4.5 mg TN/l and 3.9 mg  $TN_f$ /l, an average reduction of 73% and 75% respectively. TN and  $TN_f$  loading and removal rates are summarised in Table 5.14. The increase in the percentage reduction of the TN and  $TN_f$  indicates that the addition of the external carbon source facilitated an improvement in the denitrification process.

**Table 5.14 TN loading and removal rates – Phase 2**

	TN loading	TN loading	% $TN_f$	TN <sub>f</sub> loading	TN <sub>f</sub> loading	% $TN_f$
	rate	rate		rate	rate	
	g TN/m <sup>2</sup>	g TN/m <sup>2</sup>	removal	g TN <sub>f</sub> /m <sup>2</sup>	g TN <sub>f</sub> /m <sup>2</sup>	removal
	media/d	media/d		media/d	media/d	
Phase 2a <sup>a, c</sup>	0.20	0.11	53	0.18	0.11	57
Phase 2b <sup>b, c</sup>	0.14	0.10	73	0.14	0.10	75

<sup>a</sup> average influent volume of 35.1 l/d; <sup>b</sup> average influent volume of 34.9l/d; <sup>c</sup> surface area of 4.1m<sup>2</sup>.



**Figure 5.7 (a) Influent and effluent ammonium nitrogen and effluent nitrate nitrogen – Phase 2**

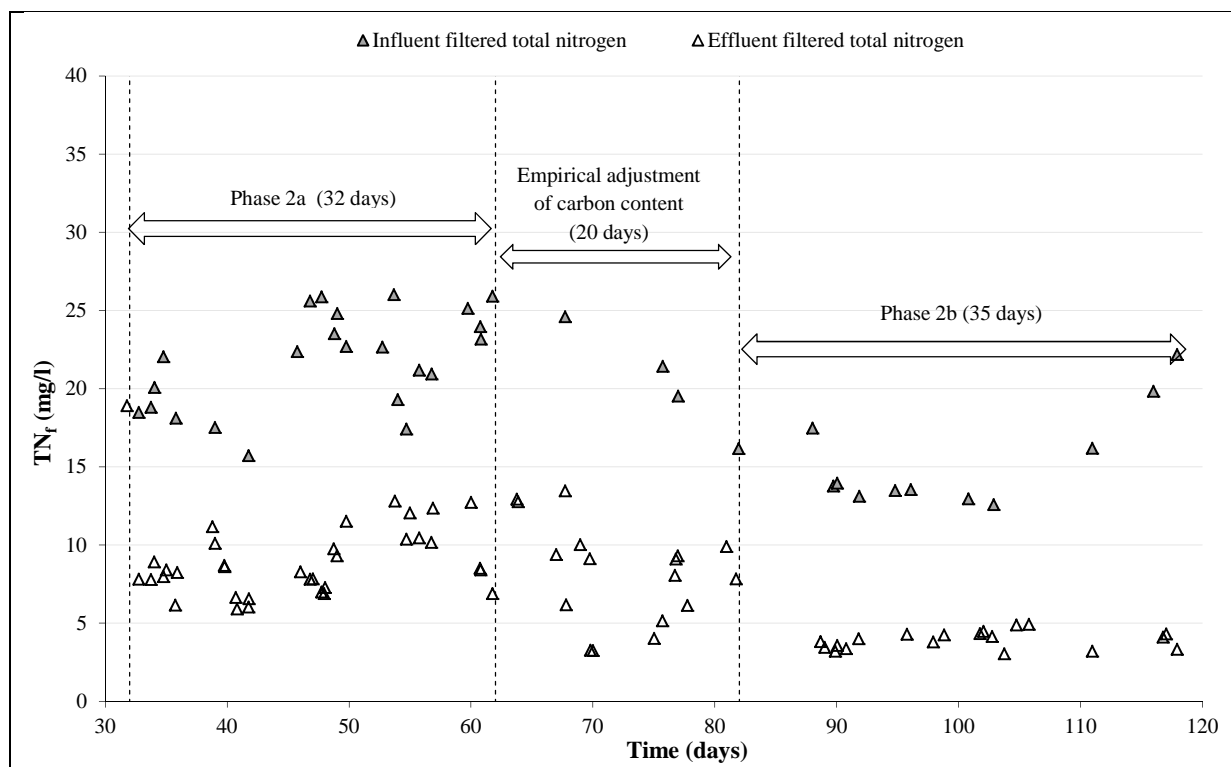


Figure 5.7 (b) Influent and effluent filtered total nitrogen – Phase 2

#### 5.4.3.3 Phosphorus

In Phase 2a, the influent  $\text{PO}_4\text{-P}$  was 1.1 mg  $\text{PO}_4\text{-P/l}$ , which is already below the limit required by UWWTD (91/271/EEC) of 2 mg  $\text{PO}_4\text{-P/l}$  (Table 2.2). The  $\text{PO}_4\text{-P}$  concentration was reduced by an average of 52% at a rate of 0.005 g  $\text{PO}_4\text{-P/m}^2$  media/d.

During Phase 2b, the average influent  $\text{PO}_4\text{-P}$  was 0.8 mg  $\text{PO}_4\text{-P/l}$ , resulting in a loading rate of 0.0076 g  $\text{PO}_4\text{-P/m}^2$  media/d with an average effluent concentration of 0.2 mg  $\text{PO}_4\text{-P/l}$ .

#### 5.4.3.4 Alkalinity

During Phase 2a, the average influent alkalinity was 219 mg  $\text{CaCO}_3\text{/l}$ , with an average effluent concentration of 117 mg  $\text{CaCO}_3\text{/l}$ . During Phase 2b, the average influent alkalinity was 269 mg  $\text{CaCO}_3\text{/l}$ , with a corresponding effluent concentration of 166 mg  $\text{CaCO}_3\text{/l}$ . The initial denitrification in Phase 2 would have resulted in the production of alkalinity, which was subsequently consumed during nitrification. As such meaningful removal rates cannot be determined, however, the presence of alkalinity in the effluent indicates that there was sufficient alkalinity for the nitrification process.

#### 5.4.3.5 DO and pH profiles

Figure 5.8 presents typical DO and pH profiles measured during Phase 2b. At the beginning of the treatment cycle, the addition of the influent wastewater provided the source of carbon



necessary for denitrification. The wastewater was then held quiescently in R1, which allowed the development of anoxic conditions (i.e. low DO in the presence of  $\text{NO}_3$ ), which is a prerequisite for denitrification. After 22 minutes, the DO was reduced to 0 mg DO/l, providing the conditions conducive to the denitrification process. Observations during the anoxic period in R2 have been omitted as the wastewater was being held in R1.

As was observed in Phase 1, on commencement of the aerobic period, DO concentrations began to increase and gradually escalated as biological reactors involving organic carbon and ammonium nitrogen decreased.

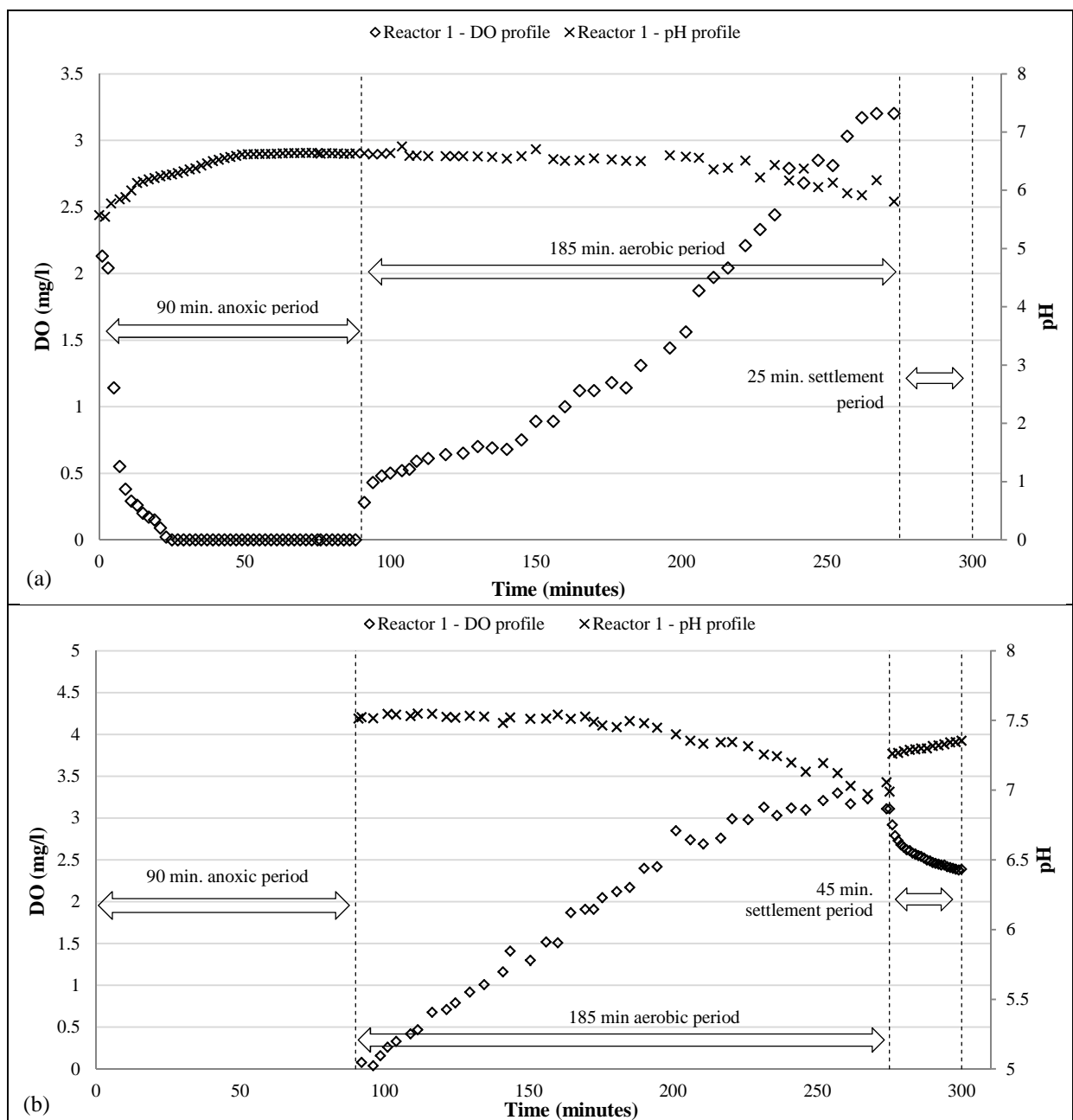


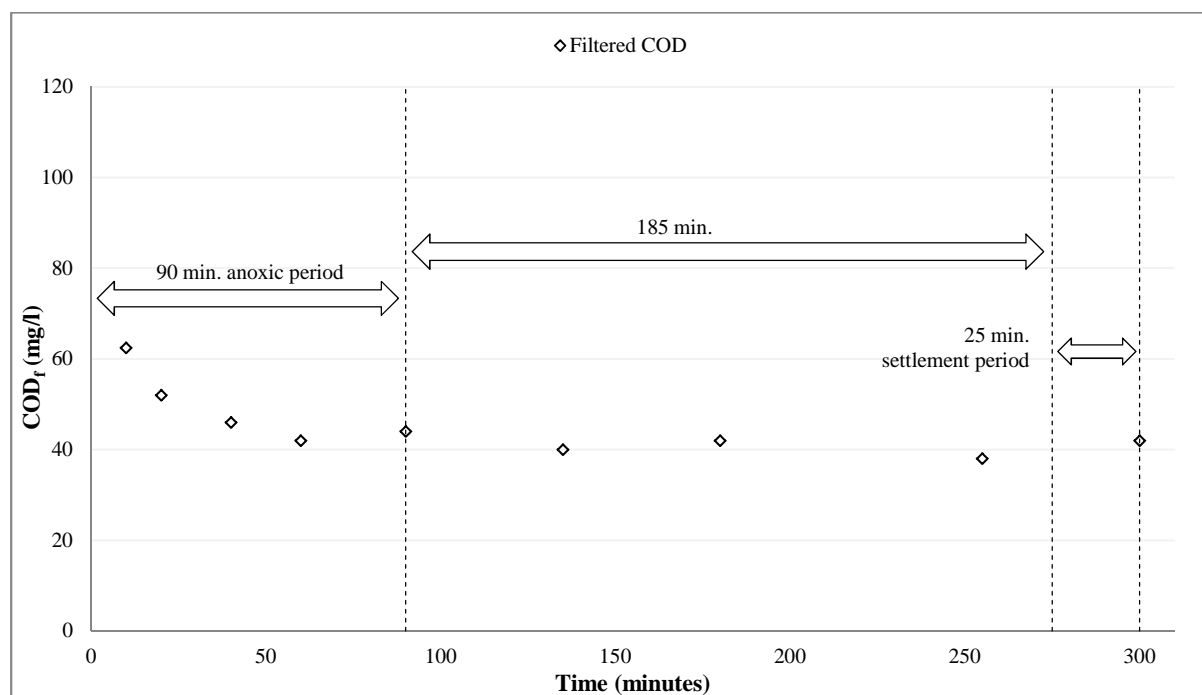
Figure 5.8 Typical DO and pH profile –Phase 2: (a) – R1 (Day 54); (b) – R2 (Day 53)

#### 5.4.4 Intensive studies – Phase 2a

Three ISs were conducted during Phase 2a on Day 21 (MWIS 4), Day 40 (MWIS 5) and Day 50 (MWIS 6). COD<sub>f</sub> analysis was only conducted during MWIS 4 and MWIS 6. As the phosphorus levels in the wastewater were so low, phosphorus was not monitored during these ISs. Detailed results of the ISs summarised below are presented in Appendix E.

##### 5.4.4.1 Organic carbon

The influent COD<sub>f</sub> removal efficiency in MWIS 4 was 63% (Figure 5.9) and 69% in MWIS 6. COD was not measured during MWIS 5 but can be assumed to be similar to the performance achieved during MWIS 4 and MWIS 6 given the consistency of the effluent results.



**Figure 5.9 COD<sub>f</sub> profiles – MWIS 4**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

Heterotrophic denitrification during the anoxic period was responsible for the 35% and 23% COD<sub>f</sub> removal during MWIS 4 and MWIS 6 respectively. A further 14% and 11% of the remaining COD<sub>f</sub> was removed during the aerobic period of MWIS 4 and MWIS 6 respectively through aerobic oxidation of the organic matter. The COD<sub>f</sub> removal rates achieved during the anoxic period and the aerobic periods of MWIS 4 and MWIS 6 are summarised in Table 5.15.

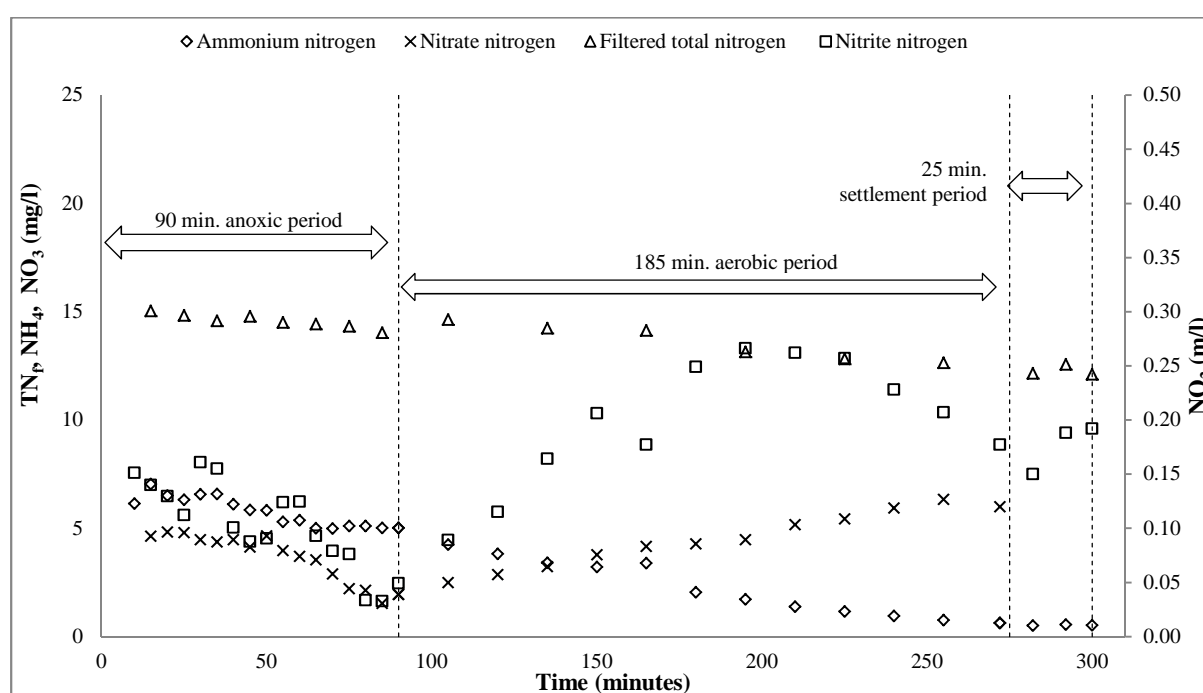
**Table 5.15 COD<sub>f</sub> removal rates – Phase 2a**

	Anoxic period <sup>1,2</sup>	Time	Aerobic period <sup>3</sup>	Time
	mg COD <sub>f</sub> /m <sup>2</sup> media/min	min	mg COD <sub>f</sub> /m <sup>2</sup> media/min	min
MWIS 4	2.81	10-90	0.81	90-135
MWIS 6	2.44	15-90	0.45	90-195

<sup>1</sup> Post dilution; <sup>2</sup> Surface area loading rate calculated based on surface area in R1 (2.05m<sup>2</sup>); <sup>3</sup> Surface area loading rate based on surface area of R1 and R2 (4.1 m<sup>2</sup>).

#### 5.4.4.2 Nitrogen

The nitrogen profile obtain in MWIS 5 is presented in Figure 5.10 (Similar results were obtained in MWIS 4 and MWIS 6 and are available in Appendix E).

**Figure 5.10 Nitrogen profiles – MWIS 5**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

NH<sub>4</sub> removal rates of 91%, 97% and 97% respectively were achieved during the treatment cycle in MWIS 4, MWIS 5 and MWIS 6.

The NH<sub>4</sub> removal rates during the anoxic period and the aerobic periods of MWIS 4, MWIS 5 and MWIS 6 are presented in Table 5.16. The removal rate during the anoxic period of MWIS 6 was higher at 0.56 mg NH<sub>4</sub> /m<sup>2</sup> media/min possibly as a result of this IS having the highest influent concentration. The rate of NH<sub>4</sub> during the aerobic period varied between 0.15 mg NH<sub>4</sub>/m<sup>2</sup> media/min (MWIS 5) and 0.20 mg NH<sub>4</sub>/m<sup>2</sup> media/min (MWIS 6). Table 5.16 summarises the removal rates for each period of MWIS 4, MWIS 5 and MWIS 6.

**Table 5.16 NH<sub>4</sub> removal rates – Phase 2a**

	<b>Anoxic period</b> <sup>1,2</sup>	<b>Time</b>	<b>Aerobic period</b> <sup>3</sup>	<b>Time</b>
	mg NH <sub>4</sub> /m <sup>2</sup> media/min	min	mg NH <sub>4</sub> /m <sup>2</sup> media/min	min
MWIS 4	0.13	10-90	0.19	90-275
MWIS 5	0.44	5-90	0.15	90-275
MWIS 6	0.56	5-90	0.20	90-255

<sup>1</sup> Post dilution; <sup>2</sup> Surface area loading rate calculated based on surface area in R1 (2.05m<sup>2</sup>); <sup>3</sup> Surface area loading rate based on surface area of R1 and R2 (4.1 m<sup>2</sup>).

The average nitrogen mass balance over the ISs conducted in Phase 2a is presented in Table 5.17.

**Table 5.17 Nitrogen mass balance – Phase 2a \***

		<b>TN<sub>f</sub></b>	<b>NH<sub>4</sub></b>	<b>NO<sub>2</sub></b>	<b>NO<sub>3</sub></b>
		g/d <sup>1</sup>	g/d <sup>1</sup>	g/d <sup>1</sup>	g/d <sup>1</sup>
<b>Post dilution mass</b>	M <sub>p, dil</sub>	2.29 (0.42)	1.36 (0.16)	0.01 (0.02)	0.73 (0.02)
<b>Mass change - Anoxic period</b>	Δ M <sub>anx</sub>	-0.26 (0.14)	-0.49 (0.18)	-0.14 (0.25)	-0.23 (0.18)
<b>Mass change - Aerobic period</b>	Δ M <sub>aer</sub>	-0.15 (0.11)	-0.68 (0.10)	0.14 (0.26)	0.32 (0.30)
<b>Effluent</b>	M <sub>eff</sub>	1.84 (0.37)	0.13 (0.08)	0.015 (0.01)	0.84(0.13)

\* Average results from ISs carried out in Phase 1; M<sub>p, dil</sub> – average post dilution mass; M<sub>aer</sub> – average change in mass during aeration period; M<sub>eff</sub> – average effluent mass; <sup>1</sup> 1.96 cycles/d; Standard deviation shown in ( ).

The influent TN and TN<sub>f</sub> removal efficiencies in MWIS 4, MWIS 5 and MWIS 6 were 42% and 42%, 35% and 33%, and 32% and 26% respectively. Denitrification during the anoxic period was predominantly responsible for the reduction in the TN and TN<sub>f</sub>.

During the aerobic period, TN<sub>f</sub> was utilised in the synthesis of new biomass. Based on COD<sub>f</sub> reductions of approximately 6 mg COD<sub>f</sub>/l and 4 mg COD<sub>f</sub>/l during the aerobic periods of MWIS 4 and MWIS 6 respectively, cell synthesis accounted for approximately 60% of the TN<sub>f</sub> reduction. The balance of the observed TN<sub>f</sub> reduction may be attributable to the occurrence of SND within the biofilm.

The objective of denitrification is to reduce NO<sub>3</sub> to N<sub>2</sub> gas (See 2.5.1.3). The maximum denitrification rate was observed during MWIS 4, at 0.48 mg NO<sub>3</sub>/m<sup>2</sup> media/min, corresponding to a 52% reduction. The denitrification rates during MWIS 5 and MWIS 6 were lower, at 0.43 mg NO<sub>3</sub>/m<sup>2</sup> media/min and 0.17 mg NO<sub>3</sub>/m<sup>2</sup> media/min respectively, corresponding to reductions of 59% and 24%. However, after the anoxic period, an average

of 2.8 mg NO<sub>3</sub>/l was found to be remaining in the wastewater. NO<sub>3</sub> increased throughout the aerobic period at a steady rate in line with the reduction in the NH<sub>4</sub>.

**Table 5.18 Summary of NH<sub>4</sub> and NO<sub>3</sub> linear trends during MWIS 4, 5 and 6**

MWIS	N process	Period	Time	Linear trend	R <sup>2</sup>
			min	mg/l	
MWIS 4	NH <sub>4</sub> decrease	Anoxic	10-90	NH <sub>4</sub> = - 0.001 t + 7.9	N.T
	NH <sub>4</sub> decrease	Aerobic	90-275	NH <sub>4</sub> = - 0.026 t + 10.5	0.98
	NO <sub>3</sub> decrease	Anoxic	15-90	NO <sub>3</sub> = - 0.042 t + 6.5	0.81
	NO <sub>3</sub> increase	Aerobic	90-275	NO <sub>3</sub> = 0.018 t + 0.61	0.91
MWIS 5	NH <sub>4</sub> decrease	Anoxic	10-90	NH <sub>4</sub> = - 0.080 t + 8.7	0.90
	NH <sub>4</sub> decrease	Aerobic	90-275	NH <sub>4</sub> = - 0.026 t + 6.8	0.96
	NO <sub>3</sub> decrease	Anoxic	15-90	NO <sub>3</sub> = - 0.044 t + 6.0	0.86
	NO <sub>3</sub> increase	Aerobic	90-275	NO <sub>3</sub> = 0.024 t + 0.01	0.98
MWIS 6	NH <sub>4</sub> decrease	Anoxic	10-90	NH <sub>4</sub> = - 0.016 t + 7.6	NT
	NH <sub>4</sub> decrease	Aerobic	90-275	NH <sub>4</sub> = - 0.038 t + 10.4	0.94
	NO <sub>3</sub> decrease	Anoxic	10-90	NO <sub>3</sub> = - 0.016 t + 5.3	0.88
	NO <sub>3</sub> increase	Aerobic	90-275	NO <sub>3</sub> = 0.022 t + 3.2	0.83

Nomenclature: N.T – No trend

The NO<sub>3</sub> remaining after the anoxic period was an indicator that the ASF-BR was failing to adequately denitrify the wastewater. DO and pH profiling conducted periodically throughout the course of the study showed that DO and pH were not inhibiting the denitrification process. As such, it was decided to investigate if carbon was the limiting factor in the denitrification process; this was the reasoning behind Phase 2b, during which an external source of carbon was added to the wastewater.

#### 5.4.5 Intensive studies – Phase 2b

Two ISs were conducted during Phase 2b, as per Phase 2a on Day 104 (MWIS 7) and Day 117 (MWIS 8). During Phase 2b, glucose was added at a rate of 0.8 g/l influent wastewater.

##### 5.4.5.1 Organic carbon

Following initial dilution (by the water remaining in R1 after discharge), the COD<sub>f</sub> was reduced by 56% and 42% during the anoxic period of MWIS 7 and MWIS 8 respectively. During the aerobic period, aerobic oxidation resulted in the removal of approximately 22% of the remaining COD<sub>f</sub> in MWIS 7 and MWIS 8.

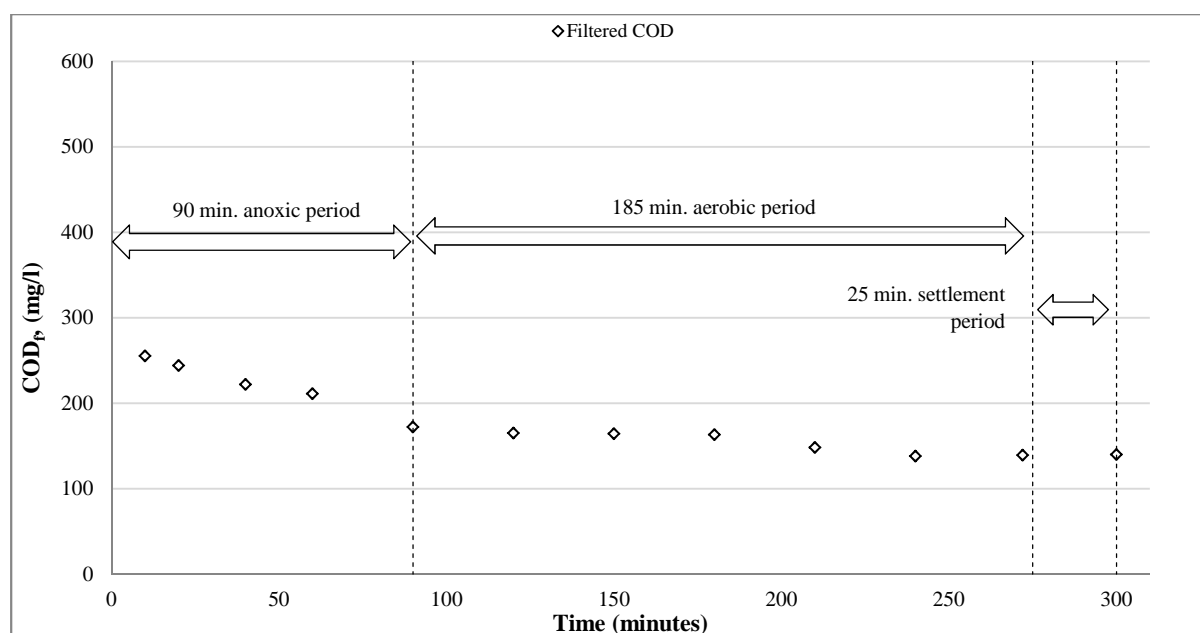
The COD<sub>f</sub> removal rates during the anoxic period and aerobic periods of MWIS 7 and MWIS 8 are presented in Table 5.19. The higher removal rate during the anoxic period of MWIS 8 may have resulted from a higher influent COD<sub>f</sub> concentration. During the aerobic period, the removal rates were 1.08 mg COD<sub>f</sub>/m<sup>2</sup> media/min and 2.24 mg COD<sub>f</sub>/m<sup>2</sup> media/min MWIS 7 and MWIS 8 respectively.

**Table 5.19 COD<sub>f</sub> removal rates – Phase 2b**

	Anoxic period <sup>1,2</sup>	Time	Aerobic period <sup>3</sup>	Time
	mg COD <sub>f</sub> /m <sup>2</sup> media/min	min	mg COD <sub>f</sub> /m <sup>2</sup> media/min	min
MWIS 7	12.66	10-90	1.08	90-275
MWIS 8	12.93	10-90	2.24	90-225

<sup>1</sup> Post dilution; <sup>2</sup> Surface area loading rate calculated based on surface area in R1 (2.05m<sup>2</sup>); <sup>3</sup> Surface area loading rate based on surface area of R1 and R2 (4.1 m<sup>2</sup>).

Figure 5.11 presents a profile of the COD<sub>f</sub> measured during MWIS 7. Full details of the results from MWIS 7 and MWIS 8 are presented in Appendix E.

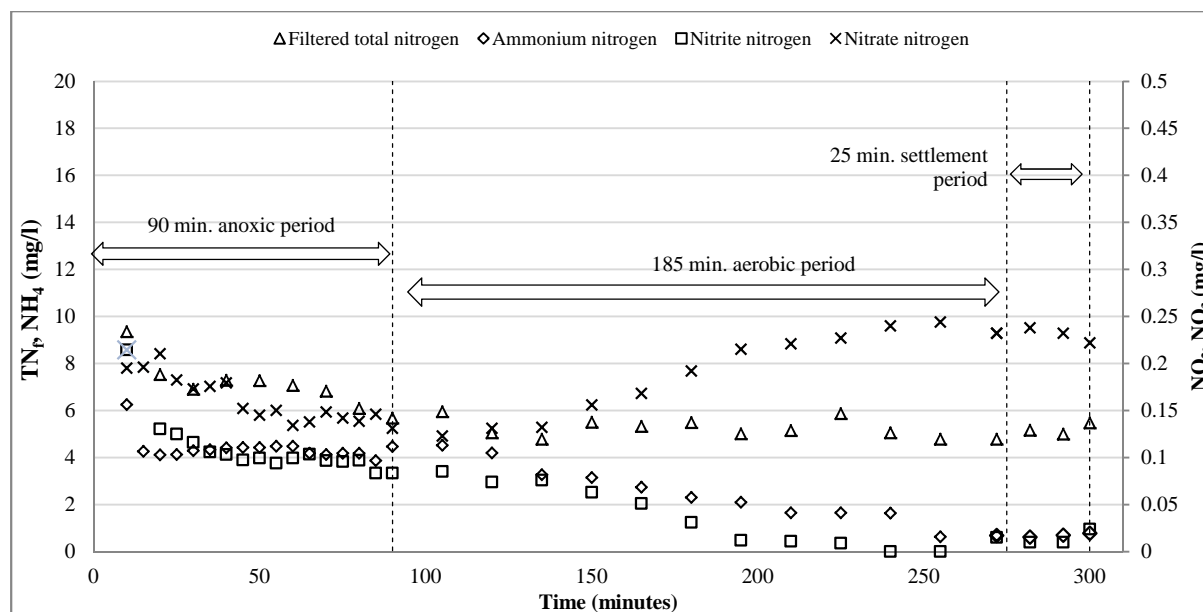


**Figure 5.11 COD<sub>f</sub> profiles – MWIS 7**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

#### 5.4.5.2 Nitrogen

Influent NH<sub>4</sub> concentrations were reduced by 99% and 91% during the course of MWIS 7 and MWIS 8 (Figure 5.12) respectively.



**Figure 5.12 Nitrogen profiles – MWIS 7**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

Nitrification of the wastewater during the aerobic period was responsible for the majority of the  $\text{NH}_4$  removal.

During MWIS 7 (Figure 5.12), nitrification occurred at a steady rate for approximately 150 minutes of the aerobic period, remaining relatively constant thereafter; while in MWIS 8, nitrification continued throughout the aerobic period. The nitrification rates in MWIS 7 and MWIS 8 were  $0.11 \text{ mg NH}_4/\text{m}^2 \text{ media}/\text{min}$  and  $0.13 \text{ mg NH}_4/\text{m}^2 \text{ media}/\text{min}$  respectively. Table 5.20 presents a summary of the removal during both the anoxic period and the aerobic period of MWIS 7 and MWIS 8.

**Table 5.20  $\text{NH}_4$  removal rates – Phase 2b**

	Anoxic period <sup>1,2</sup>	Time	Aerobic period <sup>3</sup>	Time
	$\text{mg NH}_4/\text{m}^2 \text{ media}/\text{min}$	min	$\text{mg NH}_4/\text{m}^2 \text{ media}/\text{min}$	min
MWIS 7	0.07	5-90	0.10	90-240
MWIS 8	0.03	5-90	0.13	90-275

<sup>1</sup> Post dilution; <sup>2</sup> Surface area loading rate calculated based on surface area in R1 ( $2.05 \text{ m}^2$ ); <sup>3</sup> Surface area loading rate based on surface area of R1 and R2 ( $4.1 \text{ m}^2$ ).

The average mass balance from the ISs conducted in Phase 2 b is presented in Table 5.21. The reduction in the mass of  $\text{NH}_4$  and  $\text{NO}_3$  in the effluent wastewater tends to indicate that the addition of an external source of carbon resulted in an improvement in the nitrogen removal performance.

**Table 5.21 Nitrogen mass balance – Phase 2b \***

		TN <sub>f</sub>	NH <sub>4</sub>	NO <sub>2</sub>	NO <sub>3</sub>
		g/d <sup>1</sup>	g/d <sup>1</sup>	g/d <sup>1</sup>	g/d <sup>1</sup>
<b>Post dilution mass</b> <sup>a</sup>	M <sub>p, dil</sub>	0.79 (0.26)	0.59 (0.23)	0.01 (0.01)	0.01 (0.1)
<b>Mass change - Anoxic period</b>	Δ M <sub>aer</sub>	-0.37 (0.04)	-0.36 (0.02)	0 (0.02)	0 (0.01)
<b>Mass change - Aerobic period</b>	Δ M <sub>aer</sub>	-0.05 (0.02)	-0.21 (0.08)	0.01 (0)	0 (0)
<b>Effluent</b>	M <sub>eff</sub>	0.33 (0.07)	0.03 (0.03)	0 (0)	0.2 (0)

\* Average results from ISs carried out in Phase 1; <sup>a</sup> – Theoretical average post dilution mass; M<sub>p, dil</sub> – average post dilution mass; M<sub>aer</sub> – average change in mass during aeration period; M<sub>eff</sub> – average effluent mass; <sup>1</sup> 1.96 cycles/d; Standard deviation shown in ().

The addition of the glucose improved the denitrification performance, as evident in the improved TN and TN<sub>f</sub> removal. During the anoxic period of MWIS 7 and MWIS 8, TN was reduced by an average of 34% while the average TN<sub>f</sub> reduction was 33%.

Denitrification rates during the anoxic period of MWIS 7 and MWIS 8 were similar at 0.01 mg NO<sub>3</sub>/m<sup>2</sup> media/min, corresponding to an average NO<sub>3</sub> reduction of 29% during the anoxic period of MWIS 7 and MWIS 8. The NO<sub>3</sub> concentration following the anoxic period was 0.10 mg NO<sub>3</sub>/l and 0.13 mg NO<sub>3</sub>/l for MWIS 7 and MWIS 8 respectively, which supports the assumption that an inadequate supply of readily available carbon was inhibiting the denitrification process during Phase 2a. Full details of the nitrogen results from MWIS 7 and MWIS 8 are available in Appendix E.

Table 5.22 presents the linear trends which represent the changes in NH<sub>4</sub> and NO<sub>3</sub> concentrations during the anoxic and aerobic periods of MWIS 7 and MWIS 8.

**Table 5.22 Summary of NH<sub>4</sub> and NO<sub>3</sub> linear trends during MWIS 7 and MWIS 8**

MWIS	N process	Period	Time min	Linear relationship mg/l	R <sup>2</sup>
MWIS 7	NH <sub>4</sub> decrease	Anoxic	10-90	NH <sub>4</sub> = - 0.007 t + 2.2	NR
	NH <sub>4</sub> decrease	Aerobic	105-240	NH <sub>4</sub> = - 0.016 t + 4.3	0.96
	NO <sub>3</sub> decrease	Anoxic	10-90	NO <sub>3</sub> = - 0.001 t + 0.2	0.88
	NO <sub>3</sub> increase	Aerobic	105-270	NO <sub>3</sub> = 0.001 t + 0.1	0.89
MWIS 8	NH <sub>4</sub> decrease	Anoxic	10-90	NH <sub>4</sub> = - 0.002 t + 4.4	NR
	NH <sub>4</sub> decrease	Aerobic	105-270	NH <sub>4</sub> = - 0.022 t + 6.6	0.96
	NO <sub>3</sub> decrease	Anoxic	10-90	NO <sub>3</sub> = - 0.001 t + 0.2	0.82
	NO <sub>3</sub> increase	Aerobic	105-270	NO <sub>3</sub> = 0.001 t + 0.1	0.93

Nomenclature: NR – No relationship



## 5.5 Desludging

The suspended solids (SS) measured in the effluent wastewater during Phase 1 and Phase 2a and Phase 2b were relatively constant throughout the duration of the study period. Visual observations of the interior of the tanks found a limited sludge build-up at the base of each reactor. Thus it was not deemed necessary to desludge the reactors. Estimates were made of excess sludge production (Y) based on the observed effluent suspended solids, effluent volumes of treated wastewater and the COD removal achieved during this study (Section 3.5.1.2). The sludge yields during Phase 1, Phase 2a and Phase 2b are summarised in Table 5.23.

**Table 5.23 Estimation of sludge yield during Phase 1, Phase 2a and Phase 2b**

	<b>Vol. treated</b> (l)	<b><math>\Delta</math>SS</b> (g)	<b><math>\Delta</math>COD</b> (g COD)	<b>Estimated sludge yield</b> (g SS/g COD <sub>f, removed</sub> )
Phase 1	7653	56.4	275.2	0.20
Phase 2a	2380	49.0	70.7	0.69
Phase 2b	998	97.2	78.2	0.13

Phase 1 and Phase 2b compare favourably with the finding of O'Reilly (2011) and Ginestet and Camacho (2007). O'Reilly (2011) found a sludge yield of 0.13 g SS/g COD, when treating municipal wastewater using a Pumped Flow Biofilm Reactor. In a study of 30 activated sludge plants, Ginestet and Camacho (2007) found that the average sludge yield was 0.315 g SS/g COD. The higher sludge yield in Phase 2a resulted from a higher effluent SS concentration, possibly caused by an increase in biological activity occurring following the addition of supplementary biodegradable carbon. A low sludge yield is a recognised advantage of attached growth biofilm reactors.

## 5.6 Energy requirement

These studies were conducted on a pilot scale unit and, as such, calculations of energy requirements based on this pilot scale unit could not be used to determine the energy requirements of a full scale site unit. A recent field study on the ASF-BR found the energy requirements of a larger pilot scale ASF-BR (~35 PE) ASF-BR were 0.5 kWh/m<sup>3</sup> wastewater treated, which is equivalent to an energy demand of less than 35 kWh/PE/year (McNamara, 2012). This energy demand was achieved without specifically aiming to achieve energy efficiency in the operation of the field scale ASF-BR and compares favourably with the

energy requirements of oxidation ditches, which have been reported to have an energy requirement of 0.44 – 2.07 kWh/m<sup>3</sup> wastewater treated, and activated sludge plants, which have a reported energy requirement of 0.30 – 1.89 kWh/m<sup>3</sup> wastewater treated (Mizuta and Shimada, 2010).

A similar treatment unit which operates on the same principles as the ASF-BR, the Pumped Flow Biofilm Reactor (PFBR), developed by members of project team in which the wastewater is pumped between the two reactors, proved to be highly efficient when it was scaled up for commercial applications, with operating costs of between 0.46 – 0.63 kWh/m<sup>3</sup> wastewater treated (O'Reilly, 2011). Thus as the ASF-BR achieves treatment of the wastewater through the movement of air as opposed to water, it can potentially offer higher efficiencies than the PFBR treatment system.

## **5.7 Maintenance**

The ASF-BR is a relatively simple treatment technology, consisting of valves and a single vacuum pump. There were no issues with any of these mechanical components of the unit over the course of the study. The only maintenance issue related to the level switches. The levels switches used were float switches, which were located inside a protective casing within the reactor to prevent contact with the moving media. The build-up of biofilm on the level switches resulted in their failure to respond to the movements of the wastewater, which disrupted the sequence of the treatment programme. This issue could be resolved through the use of pressure transducers.

## **5.8 Performance comparison to alternative treatment processes**

During Phases 1, 2a and 2b, the ASF-BR unit performed excellently in terms of COD and ammonium nitrogen removal when compared with alternative technologies used in the treatment of municipal strength SWW, as summarised in tables 5.24 and 5.25.

**Table 5.24 Comparison of COD removal performance with alternative biofilm treatment systems**

Study	Treatment Process	Wastewater	Organic loading rate	Removal rate	Comments
	ASF-BR – Phase 1a	MWW	1.87 g COD/m <sup>2</sup> media/d 1.19 g COD <sub>p</sub> /m <sup>2</sup> media/d	1.41 g COD/m <sup>2</sup> media/d; 75% removal efficiency 0.52 g COD <sub>p</sub> /m <sup>2</sup> media/d; 63% removal efficiency	
	ASF-BR – Phase 1b	MWW	3.35 g COD/m <sup>2</sup> media/d 1.19 g COD <sub>p</sub> /m <sup>2</sup> media/d	1.41 g COD/m <sup>2</sup> media/d; 71% removal efficiency 0.52 g COD <sub>p</sub> /m <sup>2</sup> media/d; 72% removal efficiency	Excellent removal of both NH <sub>4</sub> and COD Potential to capture of all gases produced; Unit operated at 11 °C
	ASF-BR – Phase 2a	MWW	3.07 g COD/m <sup>2</sup> media/d 0.68 g COD <sub>p</sub> /m <sup>2</sup> media/d	2.30 g COD/m <sup>2</sup> media/d; 75% removal efficiency 0.34 g COD <sub>p</sub> /m <sup>2</sup> media/d; 50% removal efficiency	
	ASF-BR – Phase 2b	MWW	7.41 g COD/m <sup>2</sup> media/d 4.72 g COD <sub>p</sub> /m <sup>2</sup> media/d	5.51 g COD/m <sup>2</sup> media/d; 74% removal efficiency 4.19 g COD <sub>p</sub> /m <sup>2</sup> media/d; 89% removal efficiency	
Ding <i>et al.</i> , 2011	SBBR	MWW	1.64 g COD/m <sup>2</sup> media/d	96% COD removal efficiency	SBBR operated at ~ 25 °C
Kermani <i>et al.</i> , 2008	MBBR	MWW	2.5 g COD <sub>p</sub> /m <sup>2</sup> media/d	98% COD removal efficiency	Four separate reactors used mechanical aeration
Zafarzadeh <i>et al.</i> , 2010	MBBR	SMWW	6.11 g COD/m <sup>2</sup> media/d	6.0 g COD/m <sup>2</sup> media/d; 99% COD removal efficiency	Unit operated at 25 °C; Used partial nitrification and denitrification
Patel <i>et al.</i> , 2006	CFBB	MWW	4.69 g COD/m <sup>2</sup> media/d	91% COD removal efficiency	Unit operated at 25 °C

Nomenclature: SBBR – Sequencing batch biofilm reactor; MBBR – Moving bed biofilm reactor; CFBB – Circulating fluidized bed bioreactor; MWW – Municipal wastewater; SMWW – Synthetic municipal wastewater;

**Table 5.25 Comparison of ammonium removal performance with alternative biofilm treatment systems**

Study	Treatment Process	Wastewater	Organic loading rate	Removal rate	Comments
	ASF-BR – Phase 1a	MWW	0.18 g NH <sub>4</sub> /m <sup>2</sup> media/d	0.16 g NH <sub>4</sub> /m <sup>2</sup> media/d; 89% removal efficiency	Excellent removal of both NH <sub>4</sub> and COD Potential to capture of all gases produced; Unit operated at 11 °C
	ASF-BR – Phase 1b	MWW	0.45 g NH <sub>4</sub> /m <sup>2</sup> media/d	0.40 g NH <sub>4</sub> /m <sup>2</sup> media/d; 89% removal efficiency	
	ASF-BR– Phase 2a	MWW	0.19 g NH <sub>4</sub> /m <sup>2</sup> media/d	0.19 g NH <sub>4</sub> /m <sup>2</sup> media/d; 95% removal efficiency	
	ASF-BR– Phase 2b	MWW	0.14 g NH <sub>4</sub> /m <sup>2</sup> media/d	0.13 g NH <sub>4</sub> /m <sup>2</sup> media/d; 93% removal efficiency	
Bassin <i>et al.</i> , 2012	MBBR	MWW	0.30 g NH <sub>4</sub> /m <sup>2</sup> media/d	85% NH <sub>4</sub> removal efficiency	Operated at 24°C; 24 h cycle
	SBBR		0.20 g NH <sub>4</sub> /m <sup>2</sup> media/d	95% NH <sub>4</sub> removal efficiency	
Ding <i>et al.</i> , 2011	SBBR	MWW	0.51 g NH <sub>4</sub> /m <sup>2</sup> media/d	99% NH <sub>4</sub> removal efficiency	SBBR operated at ~ 25 °C
Zaharzadeh <i>et al.</i> , 2010	MBBR	MWW	0.89 g NH <sub>4</sub> /m <sup>2</sup> media/d	0.66 g NH <sub>4</sub> /m <sup>2</sup> media/d; 74% NH <sub>4</sub> removal efficiency	MBBR operated at ~ 28.5 °C; Mechanical aeration
Kermani <i>et al.</i> , 2007	MBBR	MWW	0.31 g NH <sub>4</sub> /m <sup>2</sup> media/d	0.31 g NH <sub>4</sub> /m <sup>2</sup> media/d; 99% NH <sub>4</sub> removal efficiency	Four separate reactors used to create an anaerobic, anoxic and aerobic conditions with mechanical aeration
Patel <i>et al.</i> , 2006	CFBB	MWW	0.32 g NH <sub>4</sub> /m <sup>2</sup> media/d	96% NH <sub>4</sub> removal efficiency	Unit operated at 25 °C
Hiras <i>et al.</i> , 2004	RBC	HSWW	0.99 g NH <sub>4</sub> /m <sup>2</sup> media/d	93% NH <sub>4</sub> removal efficiency	Unit operated at > 15°C

Nomenclature: <sup>1</sup> Average values; MBBR – Moving bed biofilm reactor; SBBR – Sequencing batch biofilm reactor; CFBB - Circulating fluidized bed bioreactor; RBC- Rotating biological contactor; MWW – Municipal wastewater; SMWW – Synthetic municipal wastewater; HSMWW – High strength municipal wastewater.

## 5.9 Conclusions

A pilot scale ASF-BR was evaluated over a 329-day period while treating municipal wastewater. During the course of this evaluation, two studies, Phase 1 and Phase 2, were conducted. The objective of Phase 1 was to evaluate the performance of the ASF-BR in nitrifying municipal wastewater. Excellent nitrification rates were achieved, with an average 89% reduction in the  $\text{NH}_4$  concentration at an average loading rate of  $0.32 \text{ mg NH}_4/\text{m}^2 \text{ media/d}$ . These results were achieved despite the fact that the wastewater treatment plant from which the wastewater was sourced was co-treating landfill leachate from a local municipal landfill site.

The objective of Phase 2 was to achieve nitrification and denitrification of municipal wastewater. The wastewater was sourced from an alternate municipal wastewater treatment plant to mitigate the variability of the wastewater. Maximum denitrification rates were observed during Phase 2b and averaged  $0.01 \text{ g NO}_3 /\text{m}^2 \text{ media/min}$ , with a  $\text{NH}_4$  removal rate of 93%. The estimated sludge yields during Phase 1, Phase 2a and Phase 2b were 0.20, 0.69 and  $0.13 \text{ g SS/g COD}_f$  respectively. The only operational issues during this pilot scale study concerned biofilm accumulation on the level switches.

## **CHAPTER 6**

### **Study 3 – Treatment of landfill leachate**

## 6.1 Introduction

Landfill leachate (LL) is generally more problematic to treat in comparison to municipal sewage and other industrial wastewaters. It poses a particular problem for landfill operators and municipal authorities. In many cases, LL is disposed of at municipal wastewater treatment plants, where it is treated in combination with the municipal wastewater. This can have significant and adverse effects on the treatment efficiency of municipal wastewater treatment plants, effects which can be costly due to the large amounts of organic matter and ammonium nitrogen as well as heavy metals and chlorinated organic and inorganic salts present (Renou *et al.*, 2008). Thus in-situ treatment of LL is desirable. This chapter describes an investigation into the treatment of LL using the ASF-BR. It was decided to use LL as it is an especially challenging wastewater that is of particular concern for regulatory authorities (e.g. the Irish Environmental Protection Agency).

The study, conducted over a 329-day period, comprised three phases (Phase 1, Phase 2a and Phase 2b), during which the performance of the ASF-BR in removing carbon and nitrogen from LL was evaluated. This chapter describes the operational configuration during of each phase, results from ISs conducted during the study and details of the sludge production.

## 6.2 Apparatus

The apparatus used in both Phase 1 and Phase 2 was previously described in Section 3.3.3. The cycle configurations for Phase 1 and Phase 2 are presented in the relevant sections below.

## 6.3 Landfill leachate (LL)

The LL used in this study was sourced from a municipal landfill site located approximately three kilometres west of Kilconnel, in east County Galway. The landfill serves the Connacht region of Ireland, and has an annual licensed capacity of 100,000 tons per year. The site, operational since 2006, has an overall area of 60 hectares, 15 hectares of which are used for the disposal of municipal, industrial and commercially produced solid waste, with the remainder acting as a buffer between the landfill and surrounding properties (EPA, 2010). The LL is collected through a network of slotted

pipes located above a flexible impermeable membrane layer before being pumped from a series of centrally located sumps within each cell to a holding lagoon with a capacity of 90,000 m<sup>3</sup>. The LL is then transferred to a holding tank with a capacity of 5,000 m<sup>3</sup>, from which it is collected by tankers and transported to one of several municipal wastewater treatment plants, where it is treated in combination with the municipal wastewater. The LL used in this study was taken from this holding tank.

## 6.4 Phase 1: Nitrification of landfill leachate

### 6.4.1 Cycle configuration

The treatment cycle in Phase 1 was configured to achieve organic carbon removal and nitrification. The timing of each period during a treatment cycle is outlined in Table 6.1.

**Table 6.1 Cycle configuration pilot scale ASF-BR – Phase 1**

		Units	
Aerobic period	Phase duration	d	254
	Step 1 – Fill*	s	60
	Step 2 – Equalisation	s	15
	Step 3 – Draw to R2*	s	35
	Step 4 – Hold in R2	s	120
	Step 5 – Equalisation	s	15
	Step 6 – Draw to R1*	s	35
	Step 7 – Hold in R1	s	120
	Repeat Steps 2 to 7		
	Aerobic period duration	min	450
Step 9 – Settlement period	min	45	
Step 9 – Discharge*	s	30	
Total cycle duration	min	498	
Average daily flow	l/d	0.98 (0.06)	
Average volume/cycle	l	0.35 (0.02)	
Reactor volume	l/reactor	35	
HRT	d	36	

\* Maximum duration, actual time may have been less if a signal was received from a level switch; Standard deviation shown in ( ).

### 6.4.2 Overall performance of the ASF-BR – Phase 1

The ASF-BR was seeded, as described in Section 3.6.3, using return activated sludge (RAS) collected from a local municipal wastewater treatment plant. The unit reached steady state conditions (from when relatively stable effluent concentrations were



observed) after 14 days. All results presented refer to the ASF-BR operation during steady state conditions. On Day 36 and again on Day 86, the peristaltic pump used for the addition of the influent LL and the discharge effluent LL suffered a mechanical failure. During the intervening periods (Day 86 to Day 102 and Day 194 to Day 208), the addition of influent LL and extraction of effluent LL was carried out manually to maintain the biological activity within the unit. Results from both of these periods have been omitted from calculations.

At steady state operating conditions, the performance remained reasonably constant until Day 161. After Day 161, the treatment performance of the unit began to deteriorate. On Day 194, a visual observation of the media found that there was little biofilm remaining. Thus it was decided to reseed the unit with a view to improving the treatment performance. This seeding was carried out, as detailed in Section 3.6.3, over a 10-day period. The results for the remaining 51 days of the study remained reasonably constant. Table 6.2 presents a summary of the average performance of the ASF-BR during Phase 1.

**Table 6.2 Overall performance of the ASF-BR – Phase 1**

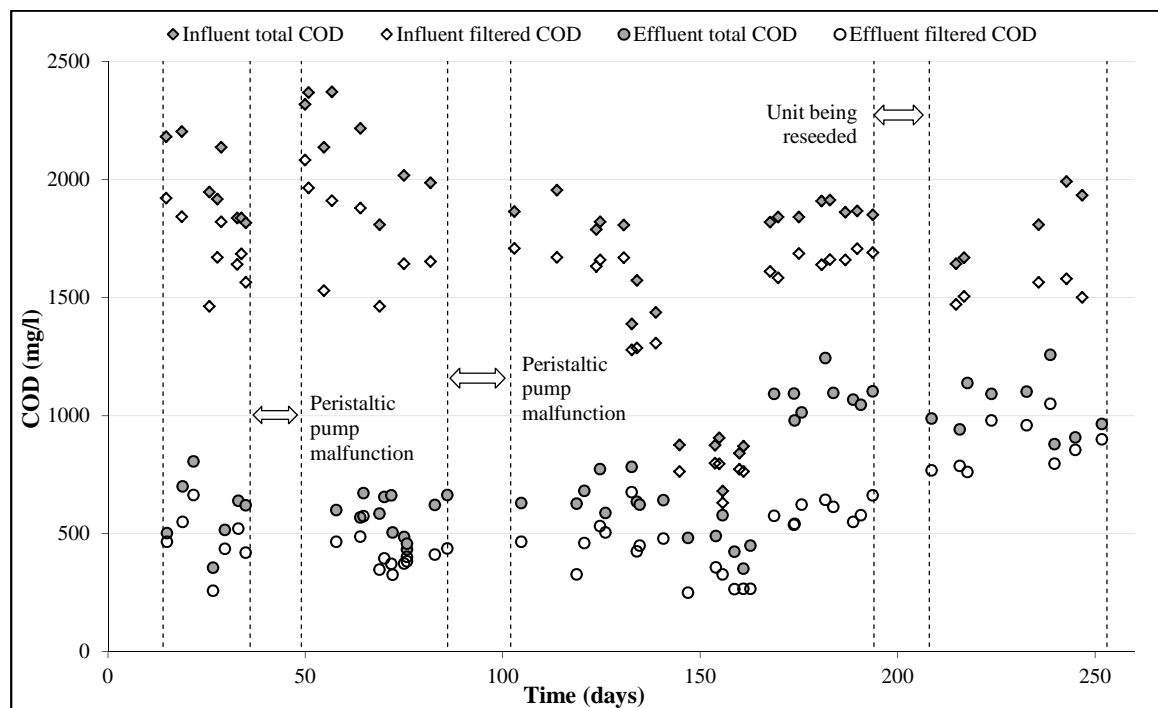
Phase 1 (253 Days)						
Parameter	Average Influent	Influent st. dev.	Average effluent	Effluent st. dev.	n	% Removal
	mg/l	mg/l	mg/l	mg/l	Inf/Eff	
COD <sup>**</sup>	1756	443	794	252	41/52	57
COD <sub>f</sub>	1512	362	559	200	41/52	63
TN <sup>**</sup>	1288	157	968	353	72/114	25
TN <sub>f</sub>	1223	153	898	345	72/114	27
NH <sub>4</sub>	1156	180	165	119	78/126	86
TON	2.1	5.7	706	313	78/126	-
NO <sub>2</sub>	1.0	3.7	49.0	2.1	78/126	-
NO <sub>3</sub>	1.2	2.9	657	290	78/126	-
PO <sub>4</sub> -P	5.8	2.8	1.7	2.9	78/126	71
Alk	5180	854	259	105	78/126	95
SS	161	105	155	63	78/126	-

All parameters marked \*\* were measured unfiltered, all other samples were filtered; n is the number of samples; Inf – Influent; Eff – Effluent.

#### 6.4.2.1 Organic carbon

Issues with the peristaltic pump and the need to reseed the unit resulted in the ASF-BR being off line on three occasions; however, the performance of the ASF-BR remained reasonably consistent during the first 134 days of steady state conditions, as shown in

Figure 6.1. In this period, the average influent COD and COD<sub>f</sub> were 2017 mg COD/l and 1718 mg COD<sub>f</sub>/l, with average removal efficiencies of 66% and 71% respectively.



**Figure 6.1 Influent and effluent COD and COD<sub>f</sub> – Phase 1**

Following Day 134, influent COD and COD<sub>f</sub> concentrations began to decline, ultimately dropping by 52% and 54% respectively, both of which were found to be significant ( $P < 0.05$ ). This reduction in the COD and COD<sub>f</sub> coincided with a period of heavy rainfall, which would have resulted in a dilution of the LL percolating through the landfill. The influent COD and COD<sub>f</sub> of the LL remained low for a period of 35 days, with average concentrations of 1050 mg/l and 934 mg/l respectively. After this period, influent COD and COD<sub>f</sub> concentrations increased by 57% and 58% respectively to an average of 1843 mg COD/l and of 1606 mg COD<sub>f</sub>/l, an increase which was found to significant ( $P < 0.05$ ). This may have been a result of the LL collection regime, in which LL was collected on a weekly basis. Thus the increase could have occurred over a number of days on site. These shock loads, which may have caused the degradation of the biofilm.

A visual inspection of the media on Day 193 confirmed that the biofilm on the media had deteriorated. This reduction in biomass coincided with a reduction in performance. As such, it was decided to reseed the unit. Following the reseeded, the ASF-BR failed to achieve the level of performance achieved in the earlier stages of Phase 1. During the final 43 days of the phase, the average COD and COD<sub>f</sub> removal efficiencies were 41%

and 43% respectively. Table 6.3 presents the COD and COD<sub>f</sub> loading and removal rates in Phase 1. Issues relating to the deterioration of the biofilm resulted in a reduction in treatment performance in the latter stages of Phase 1.

**Table 6.3 Organic carbon loading and removal rates – Phase 1**

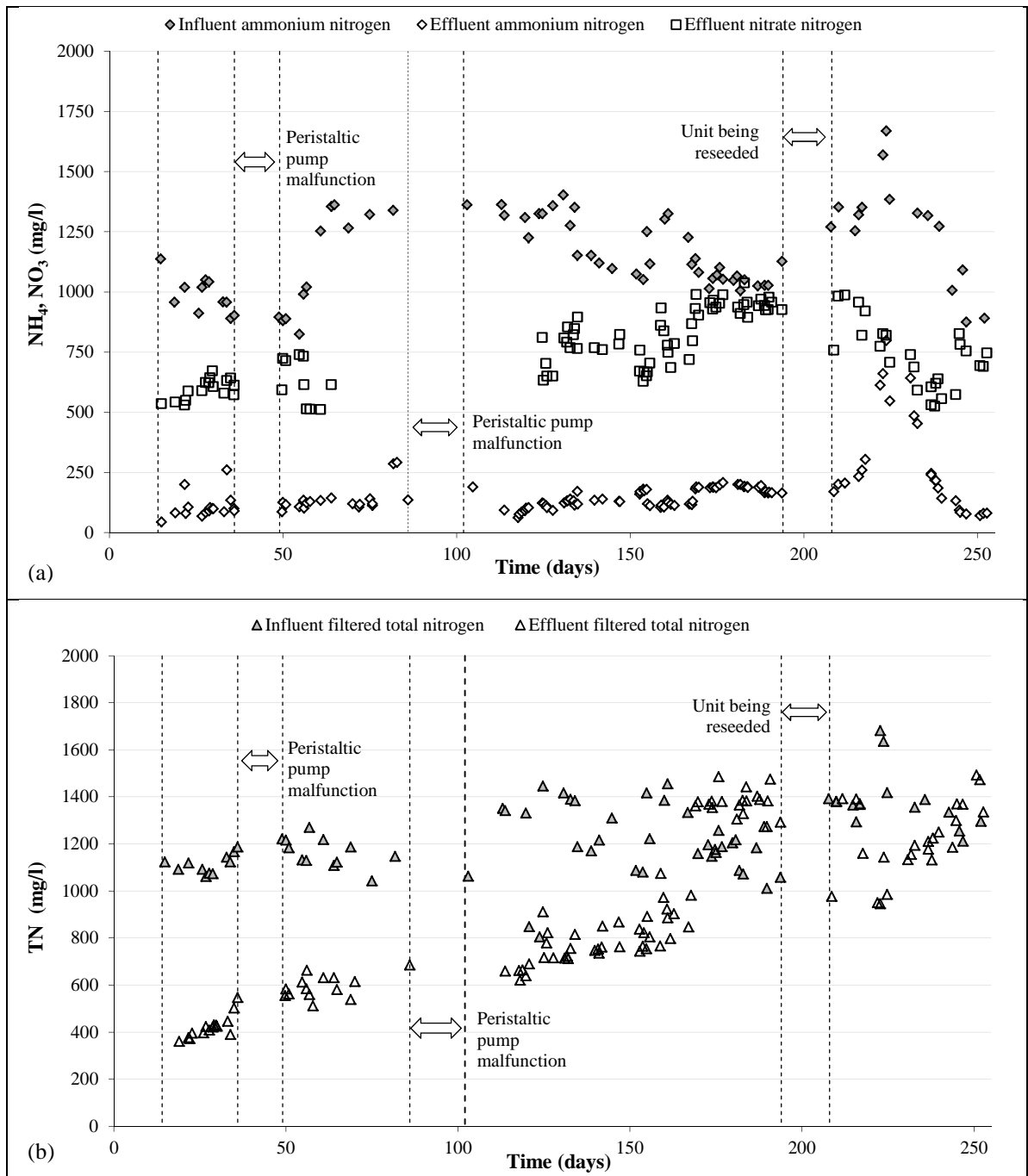
	<b>COD loading rate</b>	<b>COD removal rate</b>	<b>% COD removal</b>	<b>COD<sub>f</sub> loading rate</b>	<b>COD<sub>f</sub> removal rate</b>	<b>% COD<sub>f</sub> removal</b>
	<small>g COD/m<sup>2</sup> media/d</small>	<small>g COD/m<sup>2</sup> media/d</small>		<small>g COD<sub>f</sub>/m<sup>2</sup> media/d</small>	<small>g COD<sub>f</sub>/m<sup>2</sup> media/d</small>	
Day 14-36 <sup>a, e</sup>	0.24	0.17	71	0.21	0.15	71
Day 49-86 <sup>b, e</sup>	0.26	0.19	73	0.22	0.16	73
Day 102-194 <sup>c, e</sup>	0.18	0.08	44	0.16	0.10	63
Day 208-255 <sup>d, e</sup>	0.22	0.09	41	0.19	0.08	43

<sup>a</sup> average influent volume 0.91 l/d; <sup>b</sup> average influent volume 0.96 l/d; <sup>c</sup> average influent volume 0.99 l/d; <sup>d</sup> average influent volume 1.0 l/d; <sup>e</sup> surface area of 8.2m<sup>2</sup>.

#### 6.4.2.2 Nitrogen

Nitrification performance during the initial 134-day period of steady state conditions was relatively consistent, with an average 89% NH<sub>4</sub> removal efficiency and an average effluent NH<sub>4</sub> concentration of 120 mg/l. As with COD measurements, heavy rainfall resulted in dilution of the influent NH<sub>4</sub> concentrations from Day 134 (Figure 6.2). Influent concentrations remained relatively low for approximately 24 days before increasing by approximately 20%. The average effluent NH<sub>4</sub> concentrations then increased by 32% to an average of 185 mg/l. This increase coincides with the increase in the effluent COD and COD<sub>f</sub> and biofilm deterioration, and may be attributable to the shock load discussed previously. On Day 222, influent NH<sub>4</sub> concentrations increased by an average of 22%. This was accompanied by a 74% increase in effluent NH<sub>4</sub> concentrations. However, the unit recovered to steady state performance after approximately 20 days. The increased influent and effluent NH<sub>4</sub> concentrations observed after days 134 and 222 were significant on both occasions ( $P < 0.05$ ).

It would appear that the system was susceptible to significant variations in influent concentrations. Although no change in landfill operations was observed, it is possible that inhibitory compounds present in the influent affected the system's performance. Small changes in concentrations of metals, for example, could have been detrimental to performance.



**Figure 6.2 (a) Influent and effluent ammonium nitrogen and effluent nitrogen; (b) Influent and effluent filtered total nitrogen – Phase 1**

Table 6.4 presents the average loading and removal rates along with the percentage removal rates during Phase 1.

**Table 6.4 NH<sub>4</sub> loading and removal rates – Phase 1**

	<b>NH<sub>4</sub> loading rate</b>	<b>NH<sub>4</sub> removal rate</b>	<b>% NH<sub>4</sub> removal</b>
	g NH <sub>4</sub> /m <sup>2</sup> media/d	g NH <sub>4</sub> /m <sup>2</sup> media/d	
Day 14-36 <sup>a, e</sup>	0.12	0.11	92
Day 49-86 <sup>b, e</sup>	0.14	0.12	86
Day 102-194 <sup>c, e</sup>	0.14	0.12	86
Day 208-255 <sup>d, e</sup>	0.15	0.12	80

<sup>a</sup> average influent volume 0.91 l/d; <sup>b</sup> average influent volume 0.96 l/d; <sup>c</sup> average influent volume 0.99 l/d; <sup>d</sup> average influent volume 1.0 l/d; <sup>e</sup> surface area of 8.2m<sup>2</sup>.

It is notable that unlike carbon removal, NH<sub>4</sub> removal rates remained relatively stable. The results may also indicate that under these conditions, the maximum removal rate was approximately 0.12 g NH<sub>4</sub>/m<sup>2</sup> media/d.

The average influent TN and TN<sub>f</sub> concentrations in Phase 1 were 1290 mg/l and 1224 mg/l respectively. Effluent TN and TN<sub>f</sub> concentrations increased at a constant rate throughout the study period (Figure 6.2 (b)). The higher TN and TN<sub>f</sub> removal efficiencies of 46% and 45% respectively, in the early stage of Phase 1 may have been a result of high populations of heterotrophic bacteria in the biofilm following the seeding process, which may have enabled SND to occur.

TN and TN<sub>f</sub> loading and removal rates are summarised in Table 6.5 along with the percentage removal during each period.

**Table 6.5 Total nitrogen and filtered total nitrogen loading and removal rates – Phase 1**

	<b>TN loading rate</b>	<b>TN removal rate</b>	<b>% TN removal</b>	<b>TN<sub>f</sub> loading rate</b>	<b>TN<sub>f</sub> removal rate</b>	<b>% TN<sub>f</sub> removal</b>
	g TN/m <sup>2</sup> media/d	g TN/m <sup>2</sup> media/d		g TN/m <sup>2</sup> media/d	g TN <sub>f</sub> /m <sup>2</sup> media/d	
	Day 14-36 <sup>a, e</sup>	0.14		0.08	57	
Day 49-86 <sup>b, e</sup>	0.15	0.07	47	0.14	0.08	57
Day 102-194 <sup>c, e</sup>	0.16	0.03	19	0.15	0.03	20
Day 208-255 <sup>d, e</sup>	0.18	0.02	11	0.17	0.02	12

<sup>a</sup> average influent volume 0.91 l/d; <sup>b</sup> average influent volume 0.96 l/d; <sup>c</sup> average influent volume 0.99 l/d; <sup>d</sup> average influent volume 1.0 l/d; <sup>e</sup> surface area of 8.2m<sup>2</sup>.

Based on the average carbon removal, a mass balance for cell synthesis shows that approximately 10% of the TN and TN<sub>f</sub> removal was attributable to cell synthesis. The remainder of the removal observed may be attributable to the occurrence of SND.

On Day 167, there was a sudden increase in the effluent TN and  $TN_f$ , which coincided with the observed increase in the effluent COD,  $COD_f$  and  $NH_4$ .

Following the reseeded of the ASF-BR effluent, TN and  $TN_f$  were observed to remain reasonably constant for approximately 16 days, with cell synthesis accounting for the removal of approximately 41.2 mg  $TN_f/l$  of the average  $TN_f$  reduction of 286 mg  $TN_f/l$ , indicating that renewed biofilm may have facilitated SND. After this period, the TN returned to its increasing pattern, with negligible TN reduction in the final days of the study.

It was noted during the course of this study that there was a difference in the TN and the sum of the  $NH_4$  and TON. The variation can be attributed to the organic nitrogen in the LL, which was not measured in this study.

#### 6.4.2.3 Alkalinity

Alkalinity (measured as  $CaCO_3$ ) was monitored from Day 73 in Phase 1 (Figure 6.3). The average influent alkalinity concentration during the first 120 days of monitoring was 4998 mg  $CaCO_3/l$ , while the average effluent concentration was 70.7 mg  $CaCO_3/l$ , which may have resulted in the inhibition of nitrification as the nitrification process consumes alkalinity at a rate of 7.14 g  $CaCO_3/g$   $NH_4$  removed. Following the reseeded of the unit, additional alkalinity was added to the influent LL at a rate of 3 g  $CaCO_3/l$  to eliminate any possible inhibition of the nitrification process. As a result, the average influent concentrations increased to 5593 mg  $CaCO_3/l$  and the average effluent concentrations during the final 47 days increased to 258 mg  $CaCO_3/l$ . The increase in the average influent and effluent alkalinity were both found to be statistically insignificant ( $P > 0.05$ ).

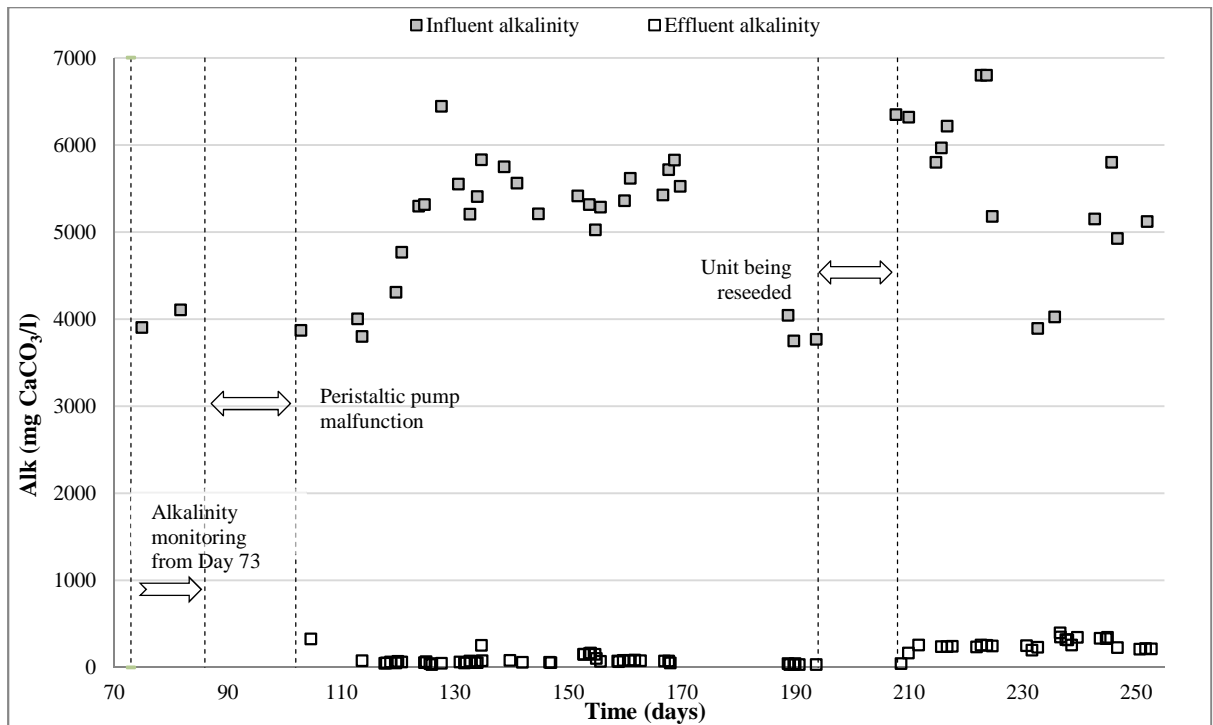


Figure 6.3 Influent and effluent alkalinity – Phase 1

#### 6.4.2.4 Phosphorus

Influent PO<sub>4</sub>-P averaged 5.83 mg PO<sub>4</sub>-P/l was reduced by an average of 82% resulting in average effluent concentrations of 1.67 mg PO<sub>4</sub>-P/l. The reduction in PO<sub>4</sub>-P can most likely be attributed to the nutrient requirement for cell synthesis.

#### 6.4.2.5 DO and pH profiles

DO and pH profiles were taken periodically throughout Phase 1. Figure 6.4 illustrates a typical DO and pH obtained from R1 (Figure 6.4(a)) and R2 (Figure 6.4(b)) (Days 135 and 156 respectively). DO concentrations remained relatively steady throughout treatment cycles in both reactors, varying between 1 mg DO/l and about 2.5 mg DO/l. It is likely DO concentrations in the biofilm were lower and thus oxygen may have inhibited the system's performance. The nitrification of the LL resulted in a steady reduction in the pH throughout the aerobic period.

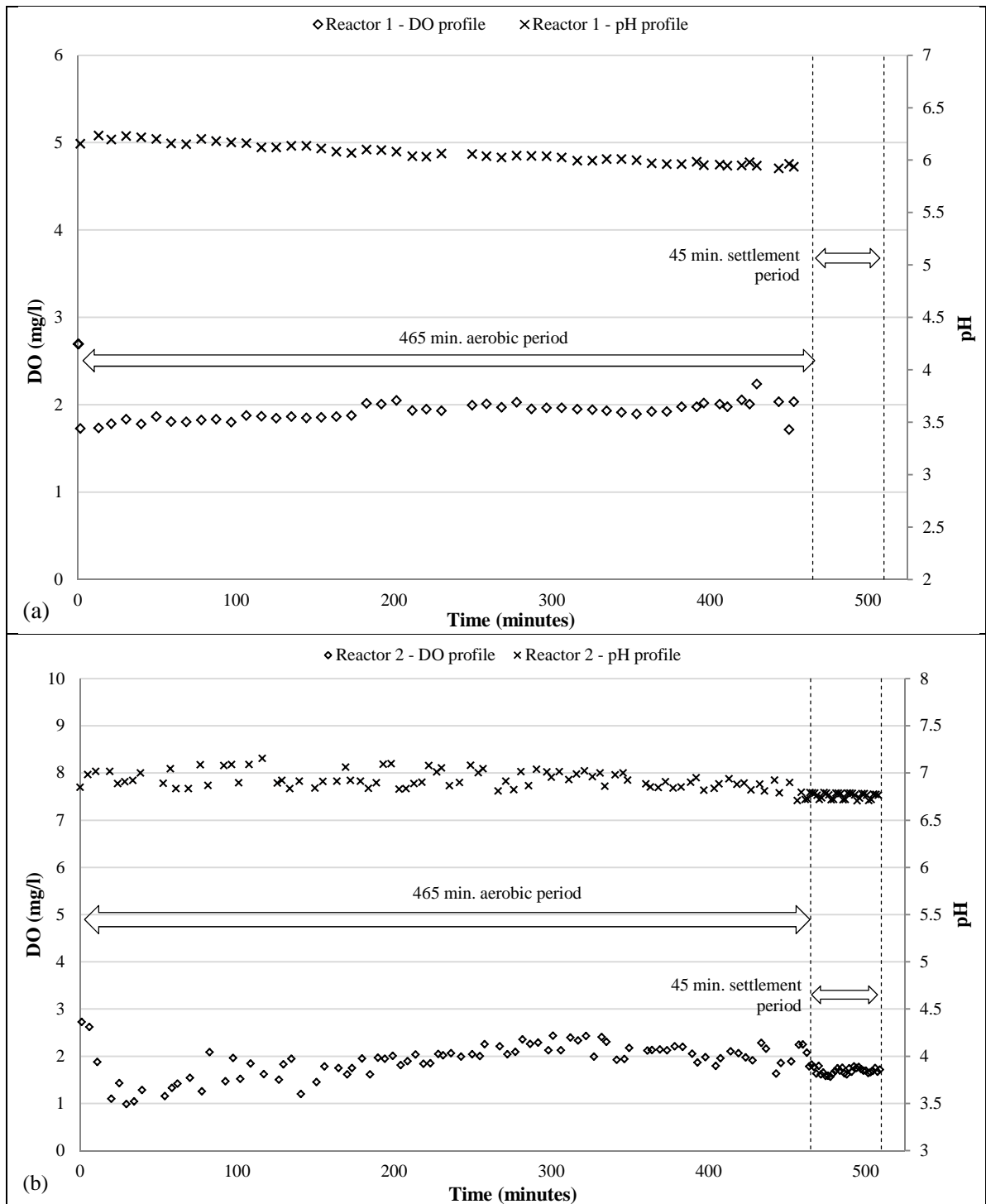


Figure 6.4 Typical DO and pH profiles – Phase 1: (a) – R1 (Day 135); (b) – R2 (Day 156)

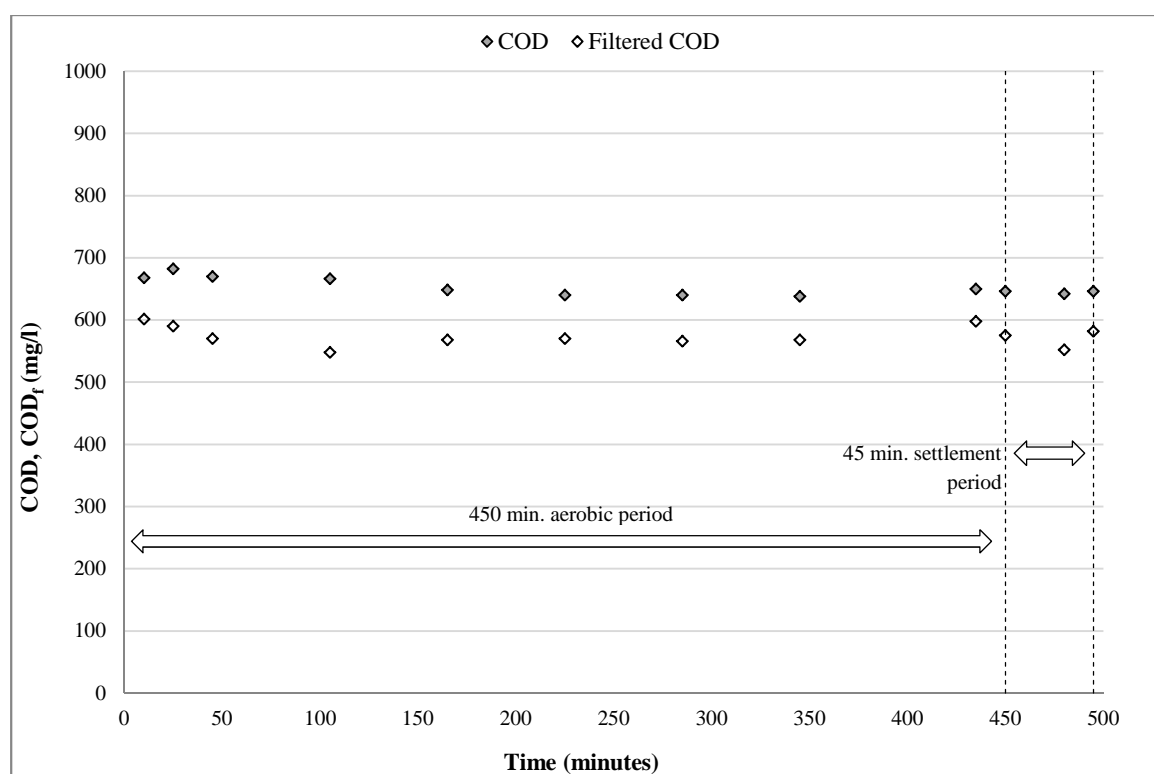
### 6.4.3 Intensive studies – Phase 1

During the course of Phase 1, five ISs – LLIS 1, LLIS 2, LLIS 3, LLIS 4 and LLIS 5 (Section 3.5.1) – were conducted on Days 139, 160, 174, 190 and 236 respectively. The results of the ISs conducted during Phase 1 are summarised below and detailed fully in Appendix F. During the ISs, carbon and nitrogen removal were evaluated along with alkalinity. Phosphorous was not monitored as part of the ISs conducted.



### 6.4.3.1 Organic carbon

COD and COD<sub>f</sub> profiles were measured in four of the five ISs (LLIS 1, LLIS 2, LLIS 4 and LLIS 5). These profiles were consistent and it would be expected that LLIS 3 (where COD was not measured) would have a similar profile. On entering R1 at the start of a cycle, influent wastewater was significantly diluted by wastewater remaining from the previous cycle. After this initial dilution, limited COD removal was observed during the aerobic period (Figure 6.5). This may be due to the presence of significant quantities of non-biodegradable carbon in the influent, which is a characteristic of LL originating from older landfill (operational for greater than five years).



**Figure 6.5 COD and COD<sub>f</sub> profiles – LLIS 1**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

While the percentage removal of COD and COD<sub>f</sub> was low in comparison to the reduction in concentrations due to dilution, COD and COD<sub>f</sub> removal was observed during the aerobic period of each of the ISs, with a maximum removal of 11% COD and 11% COD<sub>f</sub> recorded during the aerobic period of LLIS 2.

Table 6.6 outlines the rate of COD and COD<sub>f</sub> removal during each of the ISs.

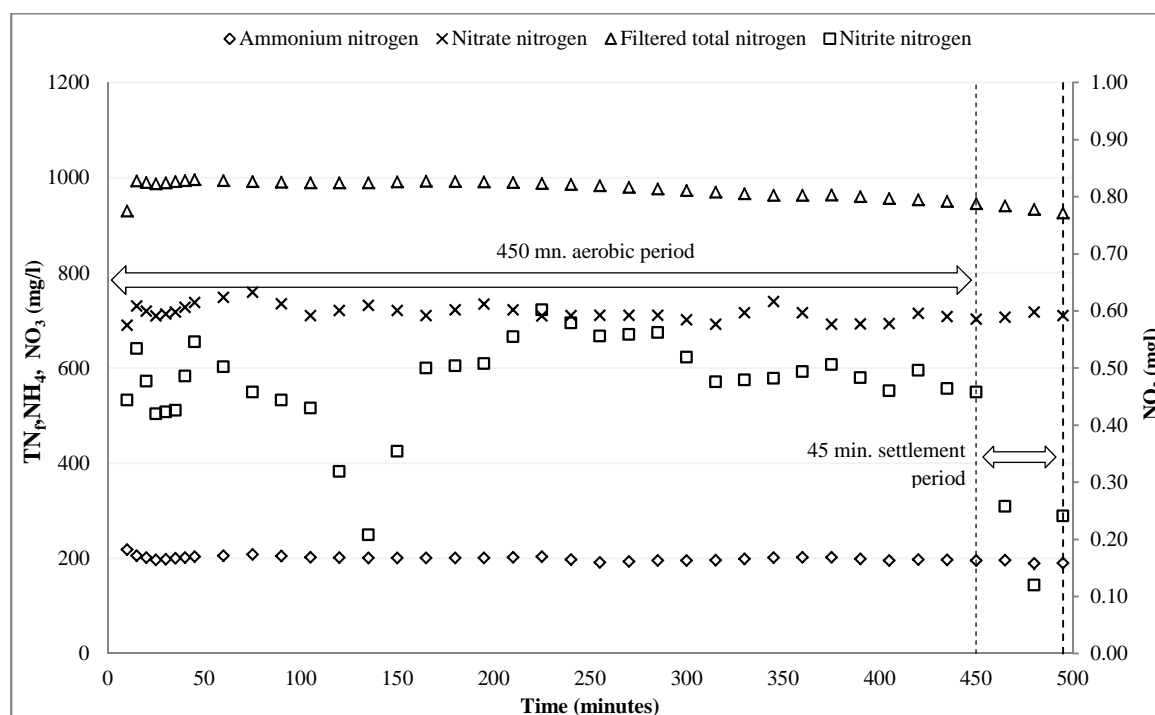
**Table 6.6 COD and COD<sub>f</sub> removal rates – Phase 1**

	Aerobic period <sup>1,2</sup>		Time
	mg COD/m <sup>2</sup> media/min	mg COD <sub>f</sub> /m <sup>2</sup> media/min	min
LLIS 1	0.12	0.10	10-450
LLIS 2	0.13	0.17	10-450
LLIS 4	0.15	0.14	10-450
LLIS 5	0.24	0.16	10-450

<sup>1</sup> Post dilution; <sup>2</sup> Surface area loading rate based on surface area of R1 and R2 (8.2 m<sup>2</sup>).

### 6.4.3.2 Nitrogen

An initial reduction in nitrogen concentrations was observed due to dilution with LL remaining following the previous treatment cycle. Figure 6.6 presents the nitrogen profile from LLIS 2. Full details of the ISs are available in Appendix F.

**Figure 6.6 Nitrogen profiles – LLIS 2**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

The maximum NH<sub>4</sub> removal was observed during LLIS 5, when 87% of the influent NH<sub>4</sub> was removed during the treatment cycle. The highest autotrophic nitrification rate was observed in LLIS 1, when the NH<sub>4</sub> concentration was reduced by 16% during the aeration period. The NH<sub>4</sub> removal efficiencies during each period of the ISs are presented in Figure 6.8.

Nitrification was found to continue throughout the aerobic period in four of the five ISs. In LLIS 4, the concentration of  $\text{NH}_4$  and  $\text{NO}_3$  remained relatively consistent at approximately 193 mg  $\text{NH}_4$ /l and 1281 mg  $\text{NO}_3$ /l. In all of the ISs, the  $\text{NH}_4$  at the end of the aerobic period was in the range of 120 – 190 mg  $\text{NH}_4$ . It is possible a significant portion of this  $\text{NH}_4$  may have been refractory in nature and further physical or chemical treatment may be necessary (Fiss and Stein, 2013; Cortez *et al.*, 2010; Bowers and Eckenfelder, 2003; Loukidou and Zouboulis, 2001). The highest autotrophic nitrification rate occurred in LLIS 5, when a nitrification rate of 0.27 mg  $\text{NH}_4$  /m<sup>2</sup> media/min was observed. The nitrification rates for each of the ISs are presented in Table 6.7.

**Table 6.7  $\text{NH}_4$  removal rates – Phase 1**

	<b>Aerobic period</b> <sup>1,2</sup>	<b>Time</b>
	mg $\text{NH}_4$ /m <sup>2</sup> media/min	min
LLIS 1	0.17	10-450
LLIS 2	0.21	10-450
LLIS 3	0.16	10-450
LLIS 4	0.18	10-360
LLIS 5	0.24	10-450

<sup>1</sup> Post dilution; <sup>2</sup> Surface area loading rate based on surface area of R1 and R2 (8.2 m<sup>2</sup>).

As can be seen from Figure 6.6, following the initial dilution, there was a slight yet constant reduction in the concentration of the  $\text{NO}_3$ , a trend observed in three of the five ISs.

Table 6.8 presents the average nitrogen mass balance from the ISs conducted in Phase 1.

**Table 6.8 Nitrogen mass balance – Phase 1 \***

		<b>TN<sub>f</sub></b>	<b>NH<sub>4</sub></b>	<b>NO<sub>2</sub></b>	<b>NO<sub>3</sub></b>
		g/d <sup>1</sup>	g/d <sup>1</sup>	g/d <sup>1</sup>	g/d <sup>1</sup>
<b>Post dilution mass</b> <sup>a</sup>	$M_{p, \text{dil}}$	122.8 (31.4)	21.2 (6.26)	0.21 (0.26)	95.9 (28.5)
<b>Mass change - Aerobic period</b>	$\Delta M_{\text{aer}}$	-0.89 (1.12)	-0.61 (0.89)	0.01 (0.01)	1.01 (1.01)
<b>Effluent</b>	$M_{\text{eff}}$	122.5 (31.1)	20.1 (6.0)	0.22 (0.27)	97.1 (28.8)

\* Average results from ISs carried out in Phase 1; <sup>a</sup> – Theoretical average post dilution mass;  $M_{p, \text{dil}}$  – average post dilution mass;  $M_{\text{aer}}$  – average change in mass during aeration period;  $M_{\text{eff}}$  – average effluent mass; <sup>1</sup> 2.9 cycles/d; Standard deviation shown in ( ).

The  $TN_f$  was also found to gradually decline in all five of the ISs. The steady reduction in both  $NO_3$  and  $TN_f$  combined with the low DO in the bulk fluid tends to indicate that SND was occurring during the treatment cycle.

TN and  $TN_f$  had similar profiles to that of  $NH_4$ , with dilution accounting for the majority of the observed removal (Figure 6.6).

Table 6.9 presents the observed linear trend of the  $NH_4$  and  $NO_3$  concentrations over time during each of the ISs.

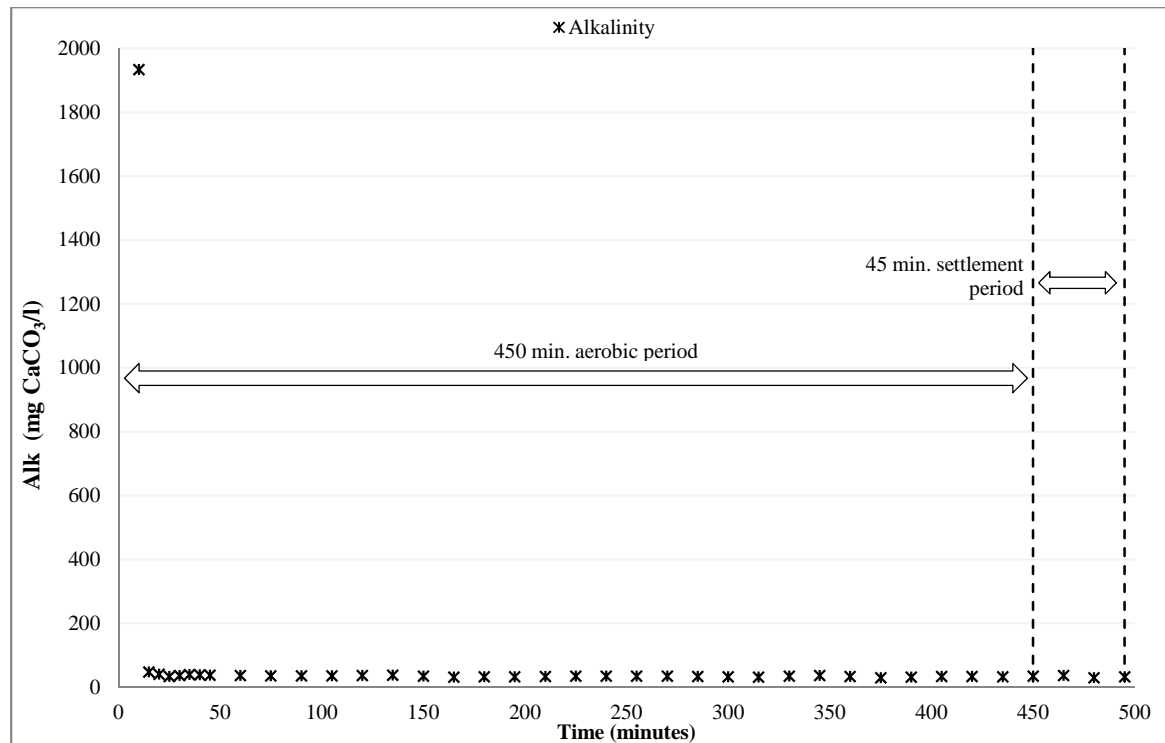
**Table 6.9 Summary of the  $NH_4$  and  $NO_3$  linear trends during LLIS 1, LLIS 2, LLIS 3, LLIS 4 and LLIS 5**

MWIS	N process	Period	Time min	Linear relationship mg/l	$R^2$	% $D_{avg}$
LLIS 1	$NH_4$ oxidation	Aerobic	10-450	$NH_4 = -0.010 t + 123.6$	0.04	2.7
	$NO_3$ increase		10-450	$NO_3 = 0.077 t + 608.0$	0.42	-0.1
LLIS 2	$NH_4$ oxidation	Aerobic	10-450	$NH_4 = -0.010 t + 201.7$	0.09	0
	$NO_3$ increase		10-450	$NO_3 = -0.090 t + 739.1$	0.46	-0.4
LLIS 3	$NH_4$ oxidation	Aerobic	10-450	$NH_4 = -0.028 t + 224.5$	0.89	-4.6
	$NO_3$ increase		10-450	$NO_3 = -0.061 t + 223.3$	0.25	-0.1
LLIS 4	$NH_4$ oxidation	Aerobic	10-360	$NH_4 = -0.033 t + 194.8$	0.04	2.5
	$NO_3$ increase		10-360	$NO_3 = 0.076 t + 1237.1$	0.23	0.1
LLIS 5	$NH_4$ oxidation	Aerobic	10-450	$NH_4 = -0.013 t + 290.8$	0.04	0
	$NO_3$ increase		10-450	$NO_3 = -0.105 t + 1015.9$	0.27	-0.2

**Note:** In general, the  $R^2$  generated for all of the linear relationships was low, implying that the linear relationship of the concentrates provided a poor representation of the actual concentrations. However, the differences between the linear relationship and the actual concentrations have been calculated ( $D_{avg}$ ) and presented in Table 6.8. The low  $R^2$  was caused by the high concentrations and the minimal changes observed in the concentrations.

#### 6.4.3.3 Alkalinity

Nitrification of the LL resulted in the consumption of alkalinity. Figure 6.7 presents the alkalinity profile from LLIS 3, with profiles from each of the ISs available in Appendix F. The ISs conducted over the first 194 days support the theory that alkalinity may have inhibited the nitrification process. The rate of alkalinity consumption during each of the ISs along with the alkalinity consumption as a function of the  $NH_4$  reduction is presented in Table 6.10. The reduction in the alkalinity as a function of the  $NH_4$  was found to be lower than the stoichiometric consumption in LLIS 1 of 7.14 g  $CaCO_3/g NH_4$  in all of the ISs.



**Figure 6.7 Alkalinity profile - LLIS 3**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

**Table 6.10 Alkalinity removal rates – Phase 1**

	Aerobic period * **		Time min
	mg CaCO <sub>3</sub> /m <sup>2</sup> media/min	mg CaCO <sub>3</sub> /mg NH <sub>4</sub>	
LLIS 1	1.57	11.34	10-450
LLIS 2	1.71	8.12	10-450
LLIS 3	1.76	8.44	10-450
LLIS 4	1.74	9.88	10-360
LLIS 5	1.33	8.35	10-450

\* Post dilution; \*\*Surface area loading rate based on surface area of R1 and R2 (8.2 m<sup>2</sup>).

## 6.5 Phase 2: Nitrification and denitrification of landfill leachate

On the completion of Phase 1, Phase 2 began. The objective of Phase 2 was to evaluate the efficacy of the ASF-BR in achieving biological nitrification, denitrification and organic carbon removal. The ASF-BR operated at steady state conditions for 74 days. Batch testing conducted during Phase 1 of Study 3 revealed that a high portion (> 95%) of the carbon contained in LL was non-biodegradable. This is in line with the finding of Bohdziewicz *et al.* (2001), Loukidou and Zouboulis (2001) and Ushikoshi *et al.* (2002), who noted that the majority of the carbon contained in LL from older landfills (> 5 years) is non-biodegradable.

To increase organic carbon/nitrogen ratios, it was decided to amend the LL with a source of external carbon in the form of glucose 10 days after starting Phase 2. The mass of glucose was increased incrementally over a 40-day period. Results are presented for the initial 18 days (Phase 2a), without any addition of an external carbon source, Phase 2b (32 days), during which glucose was added at a rate of 2.5 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l, and Phase 2c (18 days), during which glucose was added at rate of 20 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l, along with the results from ISs conducted during Phase 2b and Phase 2c (IS conducted in Phase 1a omitted due to negligible denitrification). The observed increase in the COD of the LL did not reflect the mass of glucose added. This may have been a result of a combination of the settlement of the glucose in the absence of mixing and solubility of glucose in the landfill leachate at 11 °C. Alves *et al.* (2007) found that the solubility of glucose decreases with temperate in water. The calculations based on the observed increase in the mass of glucose added and the observed increase in the COD showed that continuing to add glucose to optimise denitrification was not a viable or feasible option due to the quantity which would be required. The ratios of COD<sub>f</sub>/TN<sub>f</sub> presented were calculated based on the portion of the influent COD<sub>f</sub> and TN<sub>f</sub> removed, which was taken as the biodegradable portion of the COD<sub>f</sub> and TN<sub>f</sub>.

### **6.5.1 Cycle configuration**

The cycle configuration during Phase 2 is outlined in Table 6.11. A 90-minute anoxic period was added to the beginning of the cycle and the settlement period was shortened to 25 minutes. The timing of each step during the aerobic period remained unchanged.

**Table 6.11 Cycle configuration pilot scale ASF-BR – Phase 2**

		Units	
Aerobic period	Phase duration	d	75
	Step 1 – Fill*	s	60
	Step 2 Anoxic period	min	90
	Step 3 – Equalisation	s	15
	Step 4 – Draw to R2*	s	35
	Step 5 – Hold in R2	s	120
	Step 6 – Equalisation	s	15
	Step 7 – Draw to R1*	s	35
	Step 8 – Hold in R1	s	120
	Repeat Steps 2 to 7		
Aerobic period duration	min	365	
Step 9 – Settlement period	min	25	
Step 10 – Discharge*	s	30	
Total cycle duration	min	485	
Average daily flow	l/d	1.02 (0.05)	
Average volume/cycle	l	0.34 (0.02)	
Reactor volume	l/reactor	35	
HRT	d	35	

\* Maximum duration, actual time may have been less if a signal was received from a level switch; Standard deviation shown in ( ).

### 6.5.2 Overall performance of the ASF-BR – Phase 2

The performance of the ASF-BR during Phase 2 is summarised in Table 6.12. The performance has been divided into three phases, as outlined in Section 6.5.

**Table 6.12a Performance of the ASF-BR – Phase 2a (Day 0-10)**

Parameter	Average Influent	Influent st. dev.	Average effluent	Effluent st. dev.	n Inf/Eff	% Removal
	mg/l	mg/l	mg/l	mg/l		
COD <sup>**</sup>	1824	38	1778	26	5/5	3
COD <sub>f</sub>	1705	6	1664	4	5/5	2
TN <sup>**</sup>	1204	69.6	1190	67	9/9	1
TN <sub>f</sub>	1130	62	1118	55	9/9	1
NH <sub>4</sub>	807	83	161	23	7/10	80
TON	45.4	6.0	1146	40	7/10	-
NO <sub>2</sub>	12.3	13.5	647	76	7/10	-
NO <sub>3</sub>	33.1	12.0	399	120	7/10	-
PO <sub>4</sub> -P	7.5	0.43	0.4	0.34	7/10	94
Alk	4837	478	397	119	7/10	92
SS	383	144	170	104	7/10	56

COD<sub>f</sub>/TN<sub>f</sub> – 3.6; All parameters marked \*\* were measured unfiltered, all other samples were filtered; n is the number of samples; Inf – Influent; Eff – Effluent.

**Table 6.12b Performance of the ASF-BR – Phase 2b\* (Day 18-50)**

Parameter	Average Influent	Influent st. dev.	Average effluent	Effluent st. dev.	n Inf/Eff	% Removal
	mg/l	mg/l	mg/l	mg/l		
COD <sup>**</sup>	2107	59	2026	231	5/4	4
COD <sub>f</sub>	1938	42	1857	108	5/4	3
TN <sup>**</sup>	1175	21.5	1140	85.0	9/18	3
TN <sub>f</sub>	1085	26.9	1060	55.4	9/18	2
NH <sub>4</sub>	710	59.2	156	24.6	9/18	78
TON	49.7	16.5	1042	110	9/18	-
NO <sub>2</sub>	27.9	7.9	770	101	9/18	-
NO <sub>3</sub>	21.2	14.7	273	39.8	9/18	-
PO <sub>4</sub> -P	6.1	3.6	1.5	4.0	9/18	75
Alk	5197	1282	276	61.7	9/18	95
SS	260	70.2	174	51.2	9/18	34

\* 2.5 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l added; COD<sub>f</sub>/TN<sub>f</sub> – 2.4; Parameters marked \*\* were measured unfiltered all other samples were filtered; n is the number of samples; Inf – Influent; Eff – Effluent.



**Table 6.12c Performance of the ASF-BR – Phase 2c\* (Day 56-74)**

Parameter <sup>a</sup>	Average	Influent	Average	Effluent	n	% Removal
	Influent	st. dev.	effluent	st. dev.		
	mg/l	mg/l	mg/l	mg/l	Inf/Eff	
COD <sup>**</sup>	3440 <sup>a</sup>	169	2265 <sup>d</sup>	296	4/4	34
COD <sub>f</sub>	2973 <sup>a</sup>	229	1982 <sup>d</sup>	158	4/4	33
TN <sup>**</sup>	1229 <sup>b</sup>	12.3	1065 <sup>e</sup>	88.3	4/8	13
TN <sub>f</sub>	1080 <sup>b</sup>	28.8	1003 <sup>e</sup>	74.1	4/8	7
NH <sub>4</sub>	879 <sup>b</sup>	150.2	176 <sup>e</sup>	3.44	4/8	80
TON	51.1 <sup>c</sup>	24.2	840 <sup>f</sup>	40.2	4/8	-
NO <sub>2</sub>	40.7 <sup>c</sup>	19.4	494 <sup>f</sup>	34.8	4/8	-
NO <sub>3</sub>	10.3 <sup>c</sup>	5.71	345 <sup>f</sup>	9.1	4/8	-
PO <sub>4</sub> -P	1.0 <sup>c</sup>	0.2	0.5 <sup>f</sup>	0.2	4/8	45
Alk	4557 <sup>c</sup>	137.8	1302 <sup>f</sup>	155	4/8	71
SS	325 <sup>c</sup>	90.2	227 <sup>f</sup>	166.0	4/8	30

\* 20 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l added; COD<sub>f</sub>/TN<sub>f</sub> – 8.2; Parameters marked \*\* were measured unfiltered all other samples were filtered; n is the number of samples; Inf – Influent; Eff – Effluent.

### 6.5.2.1 Organic carbon

Figure 6.8 presents the influent and effluent carbon measured during Phase 2. A readily available source of organic carbon is one of the prerequisites for the denitrification process (Section 2.5.1.3). To achieve the objectives of this phase, it was decided to amend the LL with a source of external carbon in the form of glucose to compensate for the absence of a readily available source of biodegradable carbon. Under ideal conditions, a COD/N ratio of about 3.7 is required for denitrification for domestic wastewater (Chui and Chung, 2003). However, the optimal COD/N has been found to be dependent on the wastewater characteristics and the treatment process employed, as illustrated in Table 2.5.

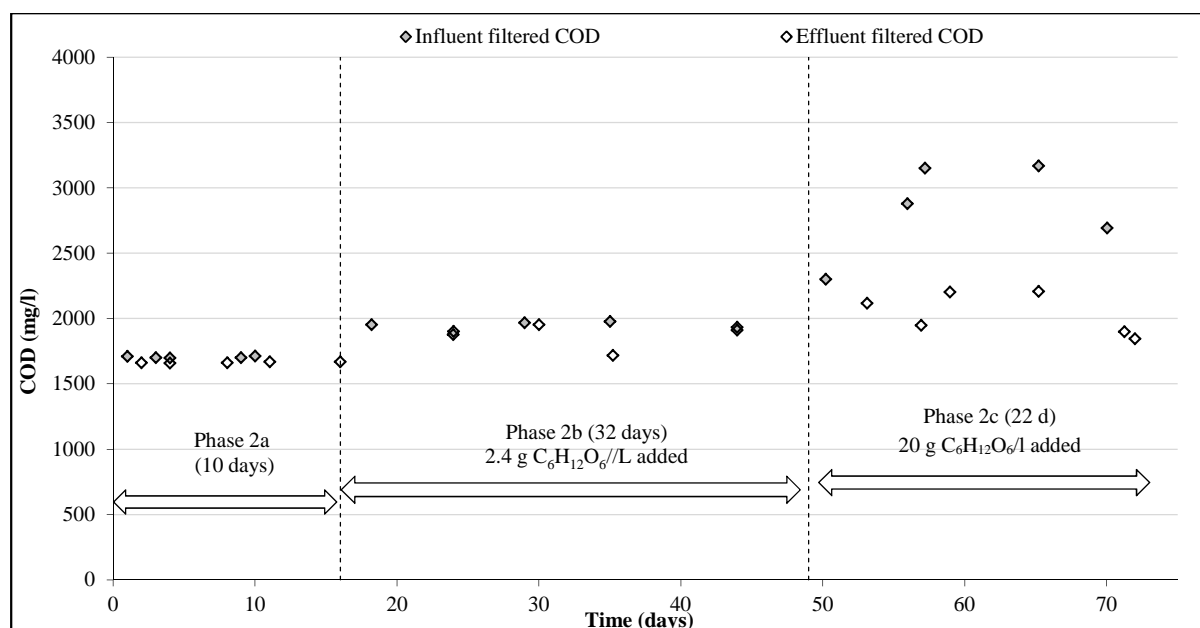


Figure 6.8 Influent and effluent COD and  $COD_f$  – Phase 2

During Phase 2a, the average influent COD concentration was 1824 mg COD/l and the average influent  $COD_f$  concentration was 1705 mg  $COD_f$ /l. Batch tests conducted on the LL found that over 95% of the carbon contained in the LL was non-biodegradable, resulting in a removal rate of just 4% COD and 5%  $COD_f$ . The addition of glucose during Phase 2b and 2c resulted in an increased removal of the COD and  $COD_f$ , as outlined in Table 6.13. The increase in the influent COD and  $COD_f$  were found to be significant ( $P < 0.05$ ) which was expected given the addition of the external source of carbon. However, much of this increased percentage removal resulted from the addition of the external carbon as effluent COD, and  $COD_f$  concentrations remained relatively constant.

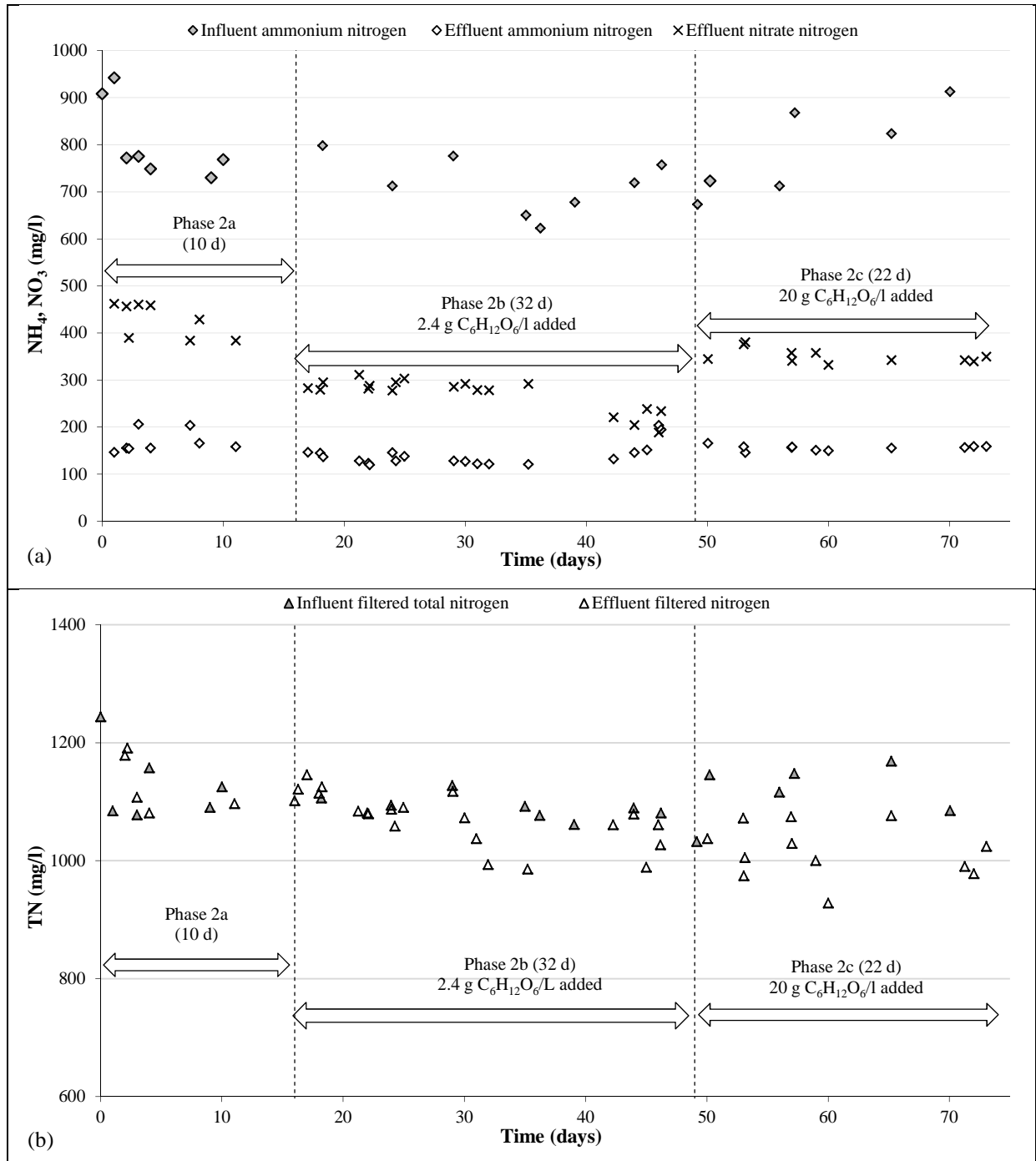
Table 6.13 Organic carbon loading and removal rates – Phase 2

	<b>COD loading</b> g COD/m <sup>2</sup> media/d	<b>COD removal</b> g COD/m <sup>2</sup> media/d	<b>% COD removal</b>	<b>COD<sub>f</sub> loading</b> g COD <sub>f</sub> /m <sup>2</sup> media/d	<b>COD<sub>f</sub> removal</b> g COD <sub>f</sub> /m <sup>2</sup> media/d	<b>% COD<sub>f</sub> removal</b>
Phase 2a <sup>a</sup>	0.23	0.01	3	0.21	0.01	2
Phase 2b <sup>b</sup>	0.26	0.01	4	0.24	0.01	4
Phase 2c <sup>c</sup>	0.42	0.14	33	0.36	0.12	33

<sup>a</sup> average influent volume 1.03 l/d, no glucose,  $COD_f/TN_f$  – 3.6; <sup>b</sup> average influent volume 1 l/d, 2.4g  $C_6H_{12}O_6$ /l added,  $COD_f/TN_f$  – 2.4; <sup>c</sup> average influent volume 1.05 l/d, 20 g  $C_6H_{12}O_6$ /l added,  $COD_f/TN_f$  – 8.2.

### 6.5.2.2 Nitrogen

Influent and effluent nitrogen parameters are presented in Figure 6.9. Overall influent concentrations averaged 765 mg NH<sub>4</sub>/l and remained relatively constant throughout Phase 2.



**Figure 6.9 (a) Influent and effluent ammonium nitrogen and effluent nitrate; (b) Influent and effluent filtered total nitrogen – Phase 2**

The average NH<sub>4</sub> loading and removal and removal rates are summarised in Table 6.14. The removal achieved during each of the phases was approximately 80%, resulting in a consistent

average effluent concentration of 150 mg NH<sub>4</sub> /l. The variations in the influent NH<sub>4</sub> between Phase 2a and 2b and between 2b and 2c were significant (P < 0.05 and P < 0.05 respectively).

**Table 6.14 NH<sub>4</sub> loading and removal rates – Phase 2**

	<b>NH<sub>4</sub> loading</b>	<b>NH<sub>4</sub> removal</b>	<b>% removal</b>
	g NH <sub>4</sub> /m <sup>2</sup> media/d	g NH <sub>4</sub> /m <sup>2</sup> media/d	
Phase 2a <sup>a</sup>	0.10	0.08	80
Phase 2b <sup>b</sup>	0.09	0.07	78
Phase 2c <sup>c</sup>	0.11	0.08	73

<sup>a</sup> average influent volume 1.03 l/d, no glucose, COD<sub>f</sub>/TN<sub>f</sub> – 3.6; <sup>b</sup> average influent volume 1 l/d, 2.4g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l added, COD<sub>f</sub>/TN<sub>f</sub> – 2.4; <sup>c</sup> average influent volume 1.05 l/d, 20 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l added, COD<sub>f</sub>/TN<sub>f</sub> – 8.2.

In Phase 2a, LL was treated without amendments. Thus the majority of the carbon present was non-biodegradable. As a result, there was little observed TN or TN<sub>f</sub> removal in Phase 2a. The addition of an external source improved the denitrification performance, particularly in Phase 2c. Glucose was added to the influent LL at a rate of 20 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l in Phase 2c, which provided improved conditions of denitrification, resulting in TN and TN<sub>f</sub> removals of 13% and 7% respectively. The TN and TN<sub>f</sub> loading and removal rates during Phase 2a, Phase 2b and Phase 2c are summarised in Table 6.15. The changes in the average influent TN and TN<sub>f</sub> were not significant (P > 0.05).

**Table 6.15 TN and TN<sub>f</sub> loading and removal rates – Phase 2**

	<b>TN loading</b>	<b>TN removal</b>	<b>% TN removal</b>	<b>TN<sub>f</sub> loading</b>	<b>TN<sub>f</sub> removal</b>	<b>% TN<sub>f</sub> removal</b>
	g TN/m <sup>2</sup> media/d	g TN/m <sup>2</sup> media/d		g TN <sub>f</sub> /m <sup>2</sup> media/d	g TN <sub>f</sub> /m <sup>2</sup> media/d	
Phase 2a <sup>a</sup>	0.15	0.002	1	0.14	0.001	1
Phase 2b <sup>b</sup>	0.14	0.004	3	0.13	0.003	2
Phase 2c <sup>c</sup>	0.16	0.02	13	0.14	0.01	7

<sup>a</sup> average influent volume 1.03 l/d, no glucose, COD<sub>f</sub>/TN<sub>f</sub> – 3.6; <sup>b</sup> average influent volume 1 l/d, 2.4g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l added, COD<sub>f</sub>/TN<sub>f</sub> – 2.4; <sup>c</sup> average influent volume 1.05 l/d, 20 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l added, COD<sub>f</sub>/TN<sub>f</sub> – 8.5.

A further increase in the mass of glucose added may have resulted in further improvements in performance; however, such an approach would not be viable from a financial or logistical perspective in practical applications of the ASF-BR. Calculations based on the results from Phase 2b and Phase 2c found that approximately 170 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l would be required to achieved complete denitrification.

### 6.5.2.3 Alkalinity

Figure 6.10 presents the influent and effluent alkalinity as measured throughout Phase 2. As in Phase 1, the LL was amended with 3 g  $\text{CaCO}_3/\text{l}$  to ensure sufficient alkalinity for the nitrification process. The influent amended LL had an average alkalinity of 5078 mg  $\text{CaCO}_3/\text{l}$  over the course of Phase 2. During Phase 2a and the first 24 days of Phase 2b, effluent alkalinity concentrations averaged 349.2 mg  $\text{CaCO}_3/\text{l}$  (94% reduction) and 275 mg  $\text{CaCO}_3/\text{l}$  (95% reduction) respectively. Following Day 42, effluent alkalinity concentrations began to increase, reaching 1563 mg  $\text{CaCO}_3/\text{l}$  on a Day 65. This increase is possibly attributable to the alkalinity production during the denitrification process.

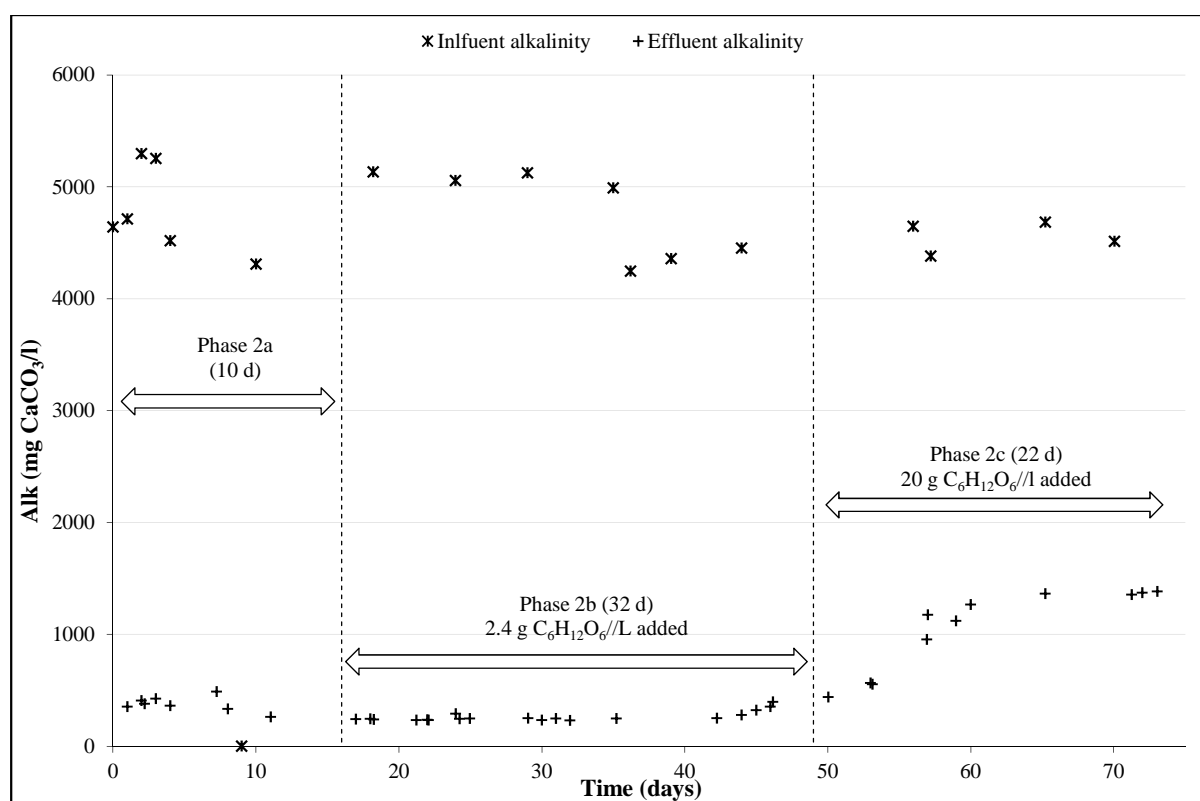


Figure 6.10 Influent and effluent alkalinity – Phase 2

### 6.5.2.4 Phosphorus

During Phase 2a, the average influent  $\text{PO}_4\text{-P}$  concentration was 7.49 mg  $\text{PO}_4\text{-P}/\text{l}$ , with an average effluent concentration of 0.44 mg  $\text{PO}_4\text{-P}/\text{l}$ , a 96% removal. For Phase 2b, influent concentrations averaged 6.08 mg  $\text{PO}_4\text{-P}/\text{l}$  76%, with effluent concentrations averaging 1.52 mg  $\text{PO}_4\text{-P}/\text{l}$ . During Phase 2c the influent concentrations averaging 0.95 mg  $\text{PO}_4\text{-P}/\text{l}$  were reduced by 45% resulting in average effluent concentrations of 0.5 mg  $\text{PO}_4\text{-P}/\text{l}$ . The treatment cycle of the ASF-BR was not conducive to phosphorus removal. However, the observed

reduction in phosphorus may be attributable to a combination of phosphorus uptake and cell synthesis.

#### 6.5.2.5 DO and pH profiles

Figure 6.11 (a) presents typical DO profiles (Section 3.5.3) from R1 (Day 41) while Figure 6.11 (b) presents a typical DO and pH profile from R2 (Day 35). The observed DO in the anoxic period was less than 0.2 mg DO/l, which indicates that excessive DO was not inhibiting the denitrification process. In the aerobic period, the minimum DO was observed to increase during the first 100 minutes and remained relatively constant thereafter – this may indicate a decrease in the oxygen demand due to organic carbon and ammonium oxidation. The reduction in the carbon via carbon oxidation was minimal and so this had a negligible impact on the DO. In the anoxic period, the pH increased, possibly due to the production of alkalinity during the denitrification process, while during the aerobic period, the oxidation of  $\text{NH}_4$  coincided with a reduction in pH.

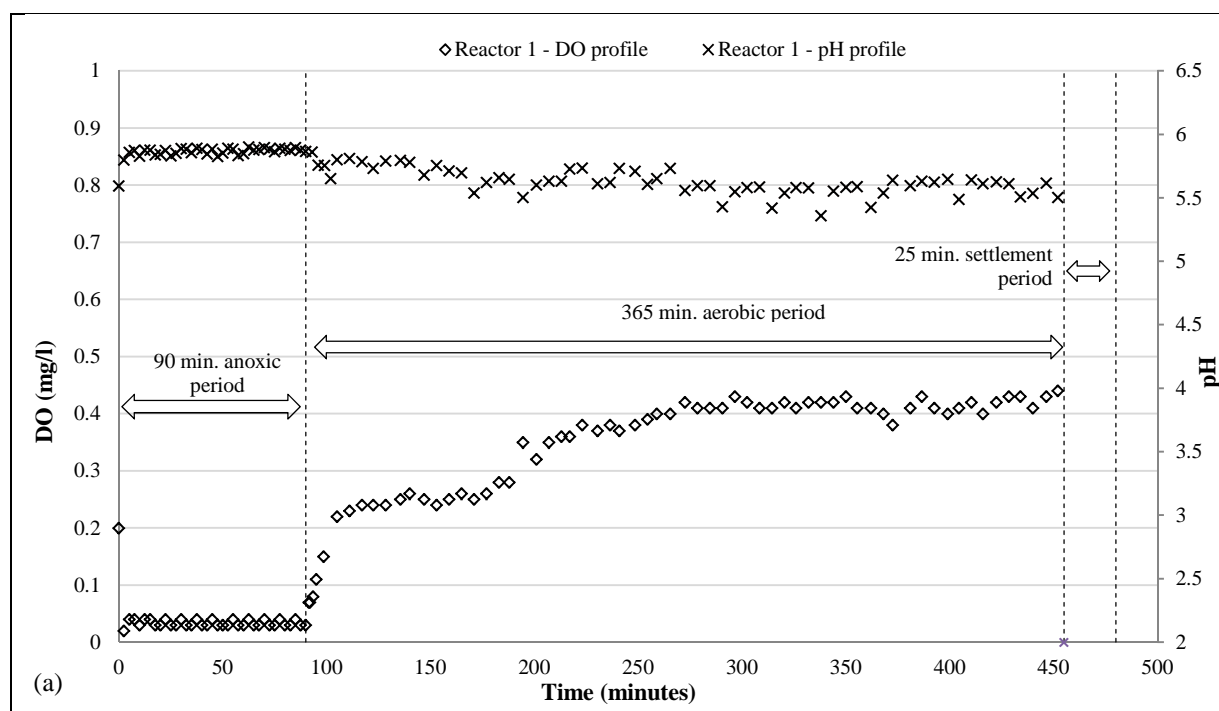


Figure 6.11 (a) Typical DO and pH profiles – Phase 2 – R1

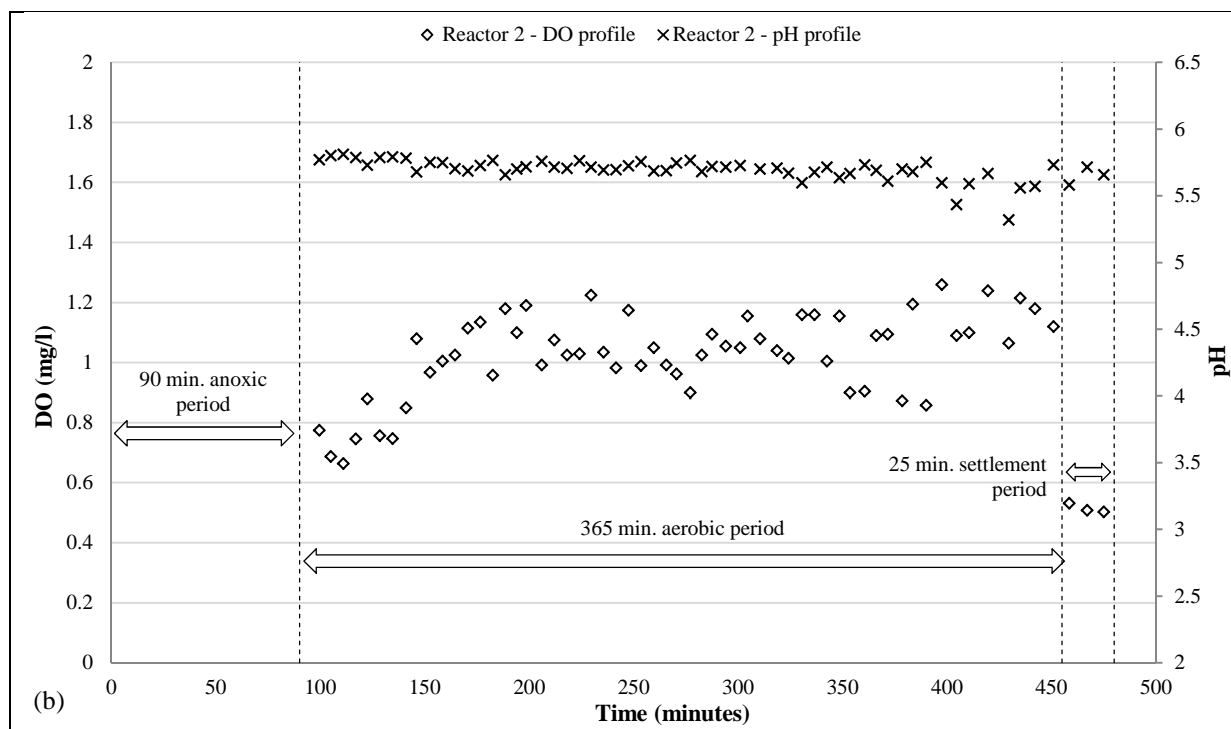


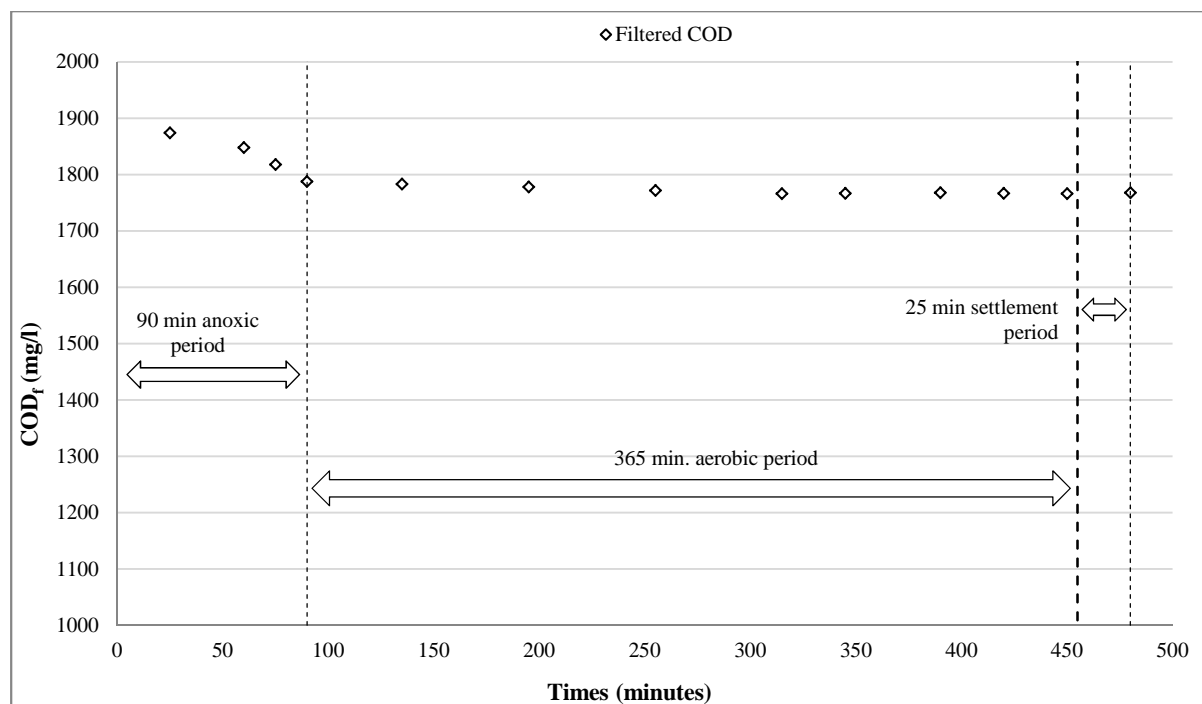
Figure 6.11 (b) Typical DO and pH profiles – Phase 2 – R2

### 6.5.3 Intensive studies – Phase 2b

During Phase 2b, two ISs were conducted (LLIS 7 and LLIS 8). Results of these intensive studies are summarised here, and the full results are available in Appendix F.

#### 6.5.3.1 Organic carbon

Figure 6.12 presents the  $COD_f$  profile obtained during LLIS 7 (a similar profile was obtained during LLIS 8). Average reductions of 6%, from influent concentrations of 1894 mg  $COD_f/l$  and 2116 mg  $COD_f/l$  during LLIS 7 and LLIS 8 respectively were observed.



**Figure 6.12** COD<sub>f</sub> profile - LLIS 7

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

Heterotrophic denitrification was responsible for an average COD<sub>f</sub> reduction of 5% during the anoxic period of LLIS 7 and LLIS 8. Carbonaceous oxidation during the aerobic period of LLIS 7 and LLIS 8 resulted in an average 1% reduction of the remaining COD<sub>f</sub>.

The rate of COD<sub>f</sub> removal during each period of the treatment cycle is summarised in Table 6.16. The low levels of carbon removal and low removal rates are a result of the low portion of biodegradable carbon contained in the LL.

**Table 6.16** COD<sub>f</sub> removal rates – Phase 2b <sup>a</sup>

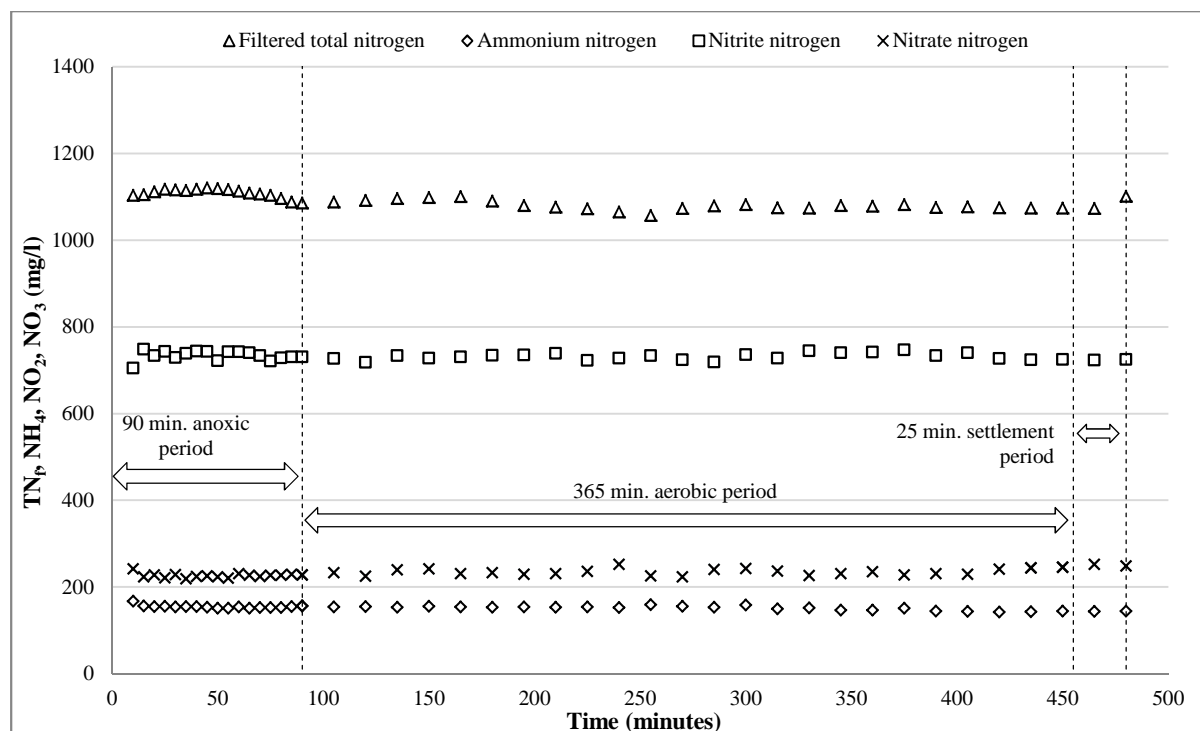
	Anoxic period <sup>1,2</sup>	Time	Aerobic period <sup>3</sup>	Time
	mg COD <sub>f</sub> /m <sup>2</sup>	min	mg COD <sub>f</sub> /m <sup>2</sup>	min
	media/min		media/min	
LLIS 7	0.88	10-90	0.26	90-450
LLIS 8	0.68	10-90	0.19	90-450

<sup>a</sup> Glucose added at a rate of 2.4 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l; <sup>1</sup> Post dilution; <sup>2</sup> Surface area loading rate calculated based on surface area in R1 (4.1m<sup>2</sup>); <sup>3</sup> Surface area loading rate based on surface area of R1 and R2 (8.2 m<sup>2</sup>).

### 6.5.3.2 Nitrogen

Figure 6.13 presents the nitrogen profile from LLIS 8, which is similar to that obtained during LLIS 7.





**Figure 6.13 Nitrogen profiles – LLIS 8**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

Nitrogen profiling showed that dilution accounted for a large proportion of the reduction in the concentration of the influent NH<sub>4</sub>, with biological nitrification and denitrification accounting for an average reduction 4% and 10% respectively, a relatively small portion of the overall observed nitrogen reduction.

Table 6.17 summarises the NH<sub>4</sub> removal rates during the anoxic and aerobic period of each of LLIS 7 and LLIS 8.

**Table 6.17 NH<sub>4</sub> removal rates – Phase 2<sup>a</sup>**

	Anoxic period <sup>1,2</sup>	Time	Aerobic period <sup>3</sup>	Time
	mg NH <sub>4</sub> /m <sup>2</sup> media/min	min	mg NH <sub>4</sub> /m <sup>2</sup> media/min	min
LLIS 7	0.52	5-90	0.12	90-450
LLIS 8	0.61	5-90	0.14	90-450

<sup>a</sup> Glucose added at a rate of 2.4 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l; <sup>1</sup> Post dilution; <sup>2</sup> Surface area loading rate calculated based on surface area in R1 (4.1m<sup>2</sup>); <sup>3</sup> Surface area loading rate based on surface area of R1 and R2 (8.2 m<sup>2</sup>).

Based on an average nitrification rate of 0.13 mg NH<sub>4</sub>/m<sup>2</sup> media/min, the ASF-BR removed approximately 389 mg NH<sub>4</sub> in a typical aerobic period. The average influent NH<sub>4</sub> concentration in Phase 2 was approximately 765 mg NH<sub>4</sub>/l, with an average HLR of 0.33 l, which resulted in the introduction of 252 mg NH<sub>4</sub> in each batch of influent wastewater. This indicates that the NH<sub>4</sub> removal was being achieved.

The nitrogen mass balance from the ISs conducted during Phase 2b is presented in Table 6.18.

**Table 6.18 Nitrogen mass balance – Phase 2b \***

		<b>TN<sub>f</sub></b>	<b>NH<sub>4</sub></b>	<b>NO<sub>2</sub></b>	<b>NO<sub>3</sub></b>
		<b>g/d<sup>1</sup></b>	<b>g/d<sup>1</sup></b>	<b>g/d<sup>1</sup></b>	<b>g/d<sup>1</sup></b>
<b>Post dilution mass<sup>a</sup></b>	$M_{p, dil}$	111.0 (1.25)	14.4 (1.42)	73.8 (1.34)	29.7 (6.61)
<b>Mass change - Anoxic period</b>	$\Delta M_{anx}$	-0.07(2.24)	-0.44 (0.02)	0.67 (0.88)	-2.41(0.71)
<b>Mass change - Aerobic period</b>	$\Delta M_{aer}$	-1.47 (0.39)	-1.11 (0.11)	0.13 (1.03)	2.47 (0.96)
<b>Effluent</b>	$M_{eff}$	110.9 (1.29)	13.6 (1.37)	74.5 (1.35)	29.9 (6.6.5)

\* Average results from ISs carried out in Phase 1; <sup>a</sup> – Theoretical average post dilution mass;  $M_{p, dil}$  – average post dilution mass;  $M_{aer}$  – average change in mass during aeration period;  $M_{eff}$  – average effluent mass; <sup>1</sup> 3 cycles/d; Standard deviation shown in ().

Similar results were observed for TN and TN<sub>f</sub> with initial dilution of the influent LL, with the treated LL remaining following the previous treatment cycle accounting for a significant proportion of the observed TN and TN<sub>f</sub> concentration reduction. The highest denitrification performance was achieved during the anoxic period in LLIS 7, with a TN removal of 5% and a TN<sub>f</sub> removal of 3%. Cell synthesis (Section 3.4.4.2) accounted for approximately 6% of the observed reduction in TN and TN<sub>f</sub> during the aerobic period of LLIS 7 and LLIS 8, with the remainder of the TN and TN<sub>f</sub> removal possibly due to SND.

The low TN and TN<sub>f</sub> removal was a result of the poor denitrification. During the anoxic period of LLIS 7 and LLIS 8, NO<sub>3</sub> removals averaged 3%. Following initial dilution, the average denitrification rate in the anoxic periods of LLIS 7 and LLIS 8 was 0.36 mg NO<sub>3</sub>/m<sup>2</sup> media/min. Table 6.19 presents the change in the nitrogen parameters observed during the anoxic and aerobic periods of LLIS 7 and LLIS 8.

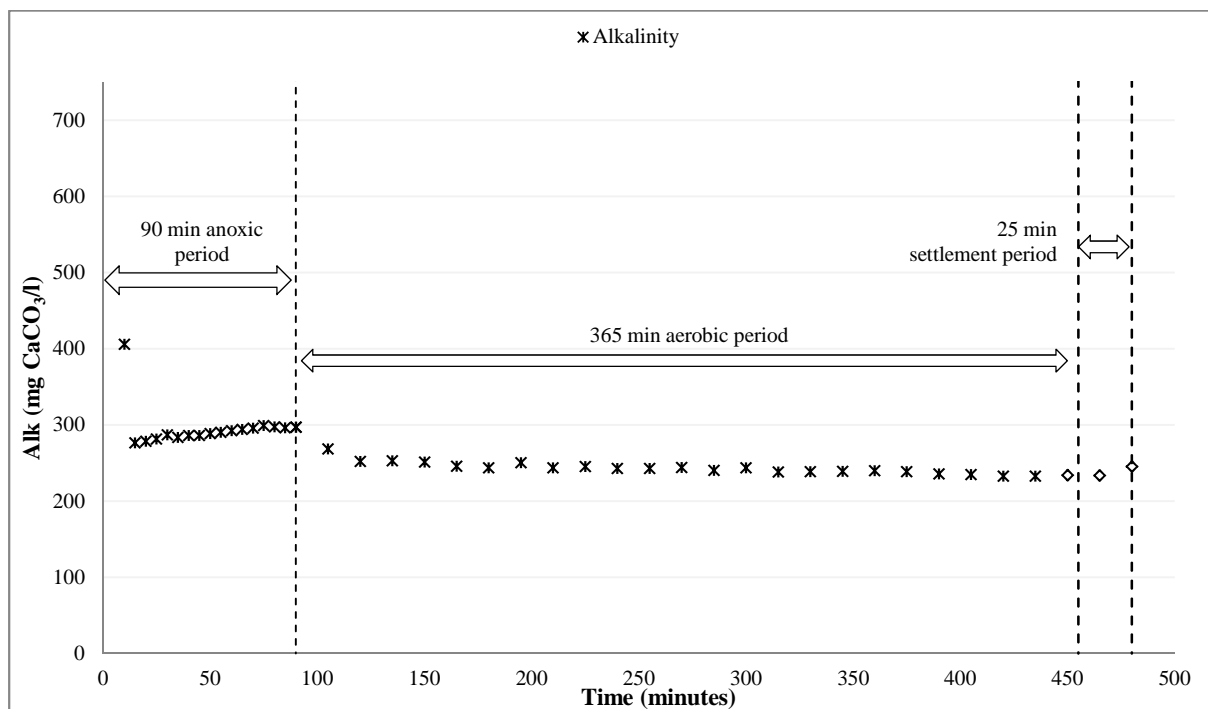
**Table 6.19 Summary of NH<sub>4</sub> and NO<sub>3</sub> linear trends during LLIS 7 and LLIS 8**

MWIS	N process	Period	Time	Linear trend	R <sup>2</sup>	% D <sub>avg</sub>
			min	mg/l		
MWIS 7	NH <sub>4</sub> reduction	Anoxic	5-90	NH <sub>4</sub> = - 0.0002 t + 137.2	3x10 <sup>-6</sup>	0
	NH <sub>4</sub> oxidation	Aerobic	90-450	NH <sub>4</sub> = - 0.024 t + 138.2	0.50	0.1
	NO <sub>3</sub> reduction	Anoxic	5-90	NO <sub>3</sub> = - 0.343 t + 314.6	0.27	-0.2
	NO <sub>3</sub> increase	Aerobic	90-450	NO <sub>3</sub> = 0.169 t + 259.6	0.58	-0.7
MWIS 8	NH <sub>4</sub> reduction	Anoxic	5-90	NH <sub>4</sub> = - 0.013 t + 152.2	0.03	0.6
	NH <sub>4</sub> oxidation	Aerobic	90-450	NH <sub>4</sub> = - 0.032 t + 160.2	0.55	0
	NO <sub>3</sub> reduction	Anoxic	5-90	NO <sub>3</sub> = - 0.063t + 222.3	0.16	0
	NO <sub>3</sub> increase	Aerobic	90-450	NO <sub>3</sub> = 0.015 t + 230.4	0.05	-0.2

**Note:** The low R<sup>2</sup> implies a poor relationship between the nitrogen parameters and time, a result of the high nitrogen concentrations and low nitrogen removal rates. However, the percentage difference between the linear relationship equation and the actual concentration has been determined, an average of which (D<sub>avg</sub>) is presented for each linear regression in Table 6.18, which shows that contrary to the inference of a poor relationship by the R<sup>2</sup>, the linear relationship can accurately present the nitrogen concentrations.

### 6.5.3.3 Alkalinity

The alkalinity profile observed during LLIS 7 is presented in Figure 6.14 (a similar profile was observed in LLIS 8). An increase in alkalinity concentrations occurred in the anoxic periods of both LLIS 7 and LLIS 8 indicating that denitrification was taking place. During the aerobic period nitrification of the LL resulted in an observed reduction of the alkalinity concentrations.

**Figure 6.14 Alkalinity profile – LLIS 7**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

Table 6.20 presents the rate of alkalinity production during the anoxic period and rate of alkalinity consumption during the aerobic period of LLIS 7 and LLIS 8.

**Table 6.20 Alkalinity production and removal rates – Phase 2<sup>a</sup>**

	Anoxic period ** **		Aerobic period ***	
	mg CaCO <sub>3</sub> produced /m <sup>2</sup> media/min <sup>b</sup>	mg CaCO <sub>3</sub> produced /mg NO <sub>3</sub> removed <sup>b</sup>	mg CaCO <sub>3</sub> consumed /m <sup>2</sup> media/min <sup>b</sup>	mg CaCO <sub>3</sub> consumed /mg NH <sub>4</sub> oxidised <sup>b</sup>
LLIS 7 <sup>a</sup>	5.88	4.94	0.75	6.21
LLIS 8 <sup>a</sup>	3.16	6.07	0.72	5.25

<sup>a</sup> Glucose added at a rate of 2.4 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l; <sup>b</sup> Timing as per Table 6.16; \* Post dilution; \*\* Surface area loading rate calculated based on surface area in R1 (4.1m<sup>2</sup>); \*\*\* Surface area loading rate based on surface area of R1 and R2 (8.2 m<sup>2</sup>).

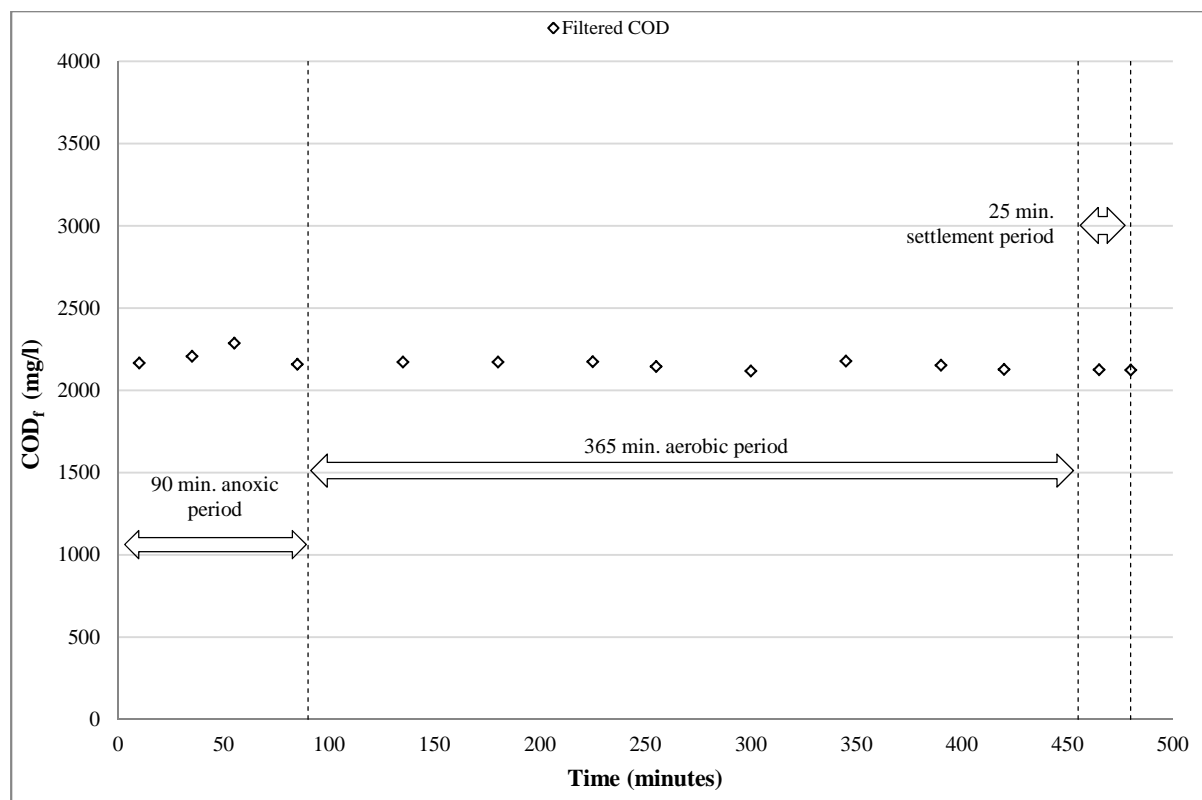
The production of alkalinity (measured as CaCO<sub>3</sub>) at an average rate of 3.38 mg CaCO<sub>3</sub>/mg NO<sub>3</sub> consumed in LLIS 7 and LLIS 8 is in line with the theoretical alkalinity production during the denitrification process. The reduction in the alkalinity in the aerobic period at an average rate of 5.73 mg CaCO<sub>3</sub>/mg NH<sub>4</sub> indicates that nitrification of the LL was occurring.

#### 6.5.4 Intensive Studies – Phase 2c

In Phase 2c, two ISs (LLIS 9 and LLIS 10) were conducted on Day 57 and Day 73 respectively, the results of which are summarised here. Full results are available in Appendix F.

##### 6.5.4.1 Organic carbon

Figure 6.15 presents the COD<sub>f</sub> profile measured during LLIS 9 (a similar profile was observed during LLIS 10). The addition of 20 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l resulted in an influent COD<sub>f</sub> concentration of 3588 mg COD<sub>f</sub>/l in LLIS 9 and 3256 mg COD<sub>f</sub>/l during LLIS 10. The COD<sub>f</sub> removal efficiencies in LLIS 9 and LLIS 10 averaged 39%, a significant improvement on an average of 6% in LLIS 7 and LLIS 8. However, the increase in the mass of glucose added to the influent LL is likely to have resulted in this improvement in the removal efficiency.



**Figure 6.15** COD<sub>f</sub> profile – LLIS 9

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

Following the initial dilution, heterotrophic denitrification during the anoxic period resulted in an average COD<sub>f</sub> removal of 8% during LLIS 9 and LLIS 10. During the aerobic periods of LLIS 9 and LLIS 10, an average of 2% of the COD<sub>f</sub> remaining after the anoxic period was removed through carbonaceous oxidation.

The improvement in the overall COD<sub>f</sub> removal in both LLIS 9 and LLIS 10 resulted in an increase in the removal rates in both the anoxic and aerobic periods. Table 6.21 presents the removal rate during the anoxic and aerobic periods of LLIS 9 and LLIS 10.

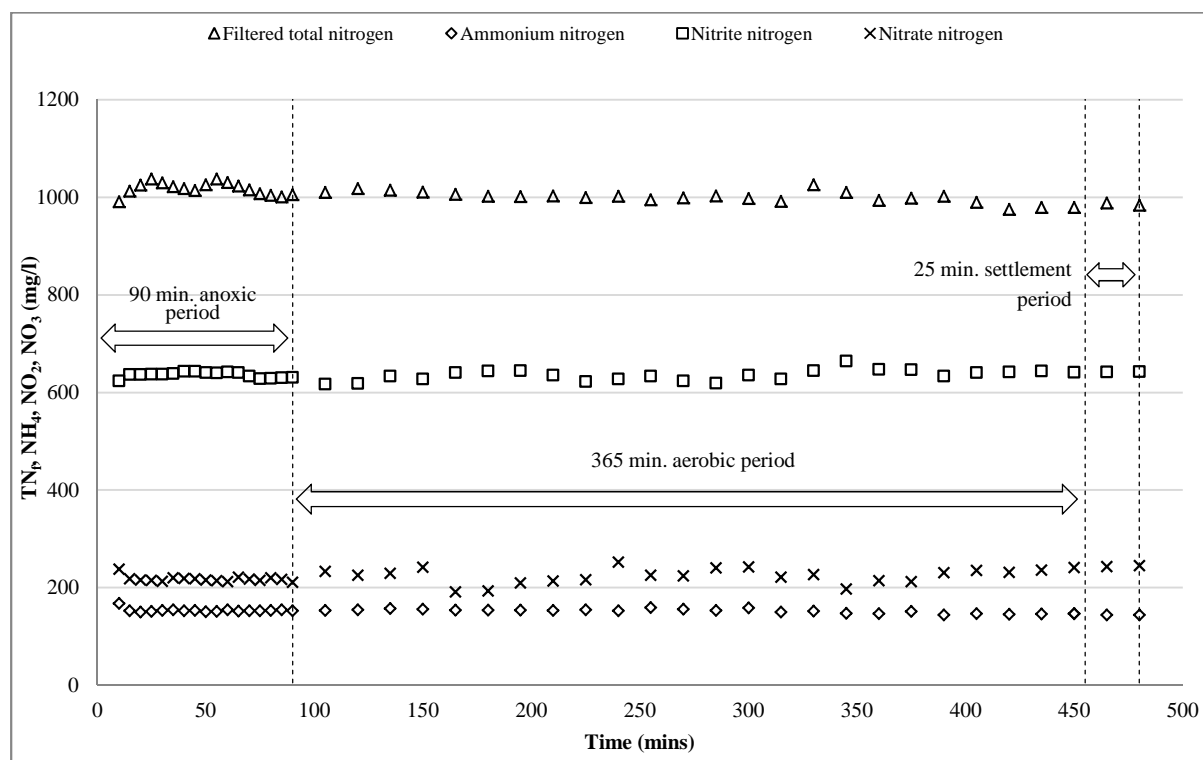
**Table 6.21** COD<sub>f</sub> removal rates – Phase 2c<sup>a</sup>

	Anoxic period <sup>1,2</sup>	Time	Aerobic period <sup>3</sup>	Time
	mg COD <sub>f</sub> /m <sup>2</sup> media/min	min	mg COD <sub>f</sub> /m <sup>2</sup> media/min	min
LLIS 9	0.92	10-90	0.37	90-450
LLIS 10	1.02	10-90	0.53	90-450

<sup>a</sup> Glucose added at a rate of 20 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l; <sup>1</sup> Post dilution; <sup>2</sup> Surface area loading rate calculated based on surface area in R1 (4.1 m<sup>2</sup>); <sup>3</sup> Surface area loading rate based on surface area of R1 and R2 (8.2 m<sup>2</sup>).

### 6.5.4.2 Nitrogen

The nitrogen removal profile from LLIS 10 is presented in Figure 6.16 (a similar profile was obtained in LLIS 9). Over the course of LLIS 9 and LLIS 10, 82% of the influent  $\text{NH}_4$  was removed, with dilution accounting for a significant portion of the observed reduction in the  $\text{NH}_4$  concentration.



**Figure 6.16 Nitrogen profiles – LLIS 10**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

$\text{NH}_4$  removal in each period of LLIS 9 and LLIS 10 is presented in Table 6.22. The highest nitrification rate ( $0.12 \text{ mg NH}_4/\text{m}^2 \text{ media}/\text{min}$ ) was observed during the aerobic period of LLIS 9.

**Table 6.22  $\text{NH}_4$  removal rates – Phase 2c <sup>a</sup>**

	Anoxic period <sup>1,2</sup>	Time	Aerobic period <sup>3</sup>	Time
	$\text{mg NH}_4/\text{m}^2 \text{ media}/\text{min}$	min	$\text{mg NH}_4/\text{m}^2 \text{ media}/\text{min}$	min
LLIS 9	0.82	10-90	0.12	90-450
LLIS 10	1.37	10-90	0.07	90-450

<sup>a</sup> Glucose added at a rate of  $20 \text{ g C}_6\text{H}_{12}\text{O}_6/\text{l}$ ; <sup>1</sup> Post dilution; <sup>2</sup> Surface area loading rate calculated based on surface area in R1 ( $4.1 \text{ m}^2$ ); <sup>3</sup> Surface area loading rate based on surface area of R1 and R2 ( $8.2 \text{ m}^2$ ).

Based on an average removal rate of  $0.10 \text{ mg NH}_4/\text{m}^2 \text{ media}/\text{min}$  during the aerobic period, approximately  $300 \text{ mg}$  of  $\text{NH}_4$  was removed in each treatment cycle. The average influent  $\text{NH}_4$  concentration in Phase 2c was  $830 \text{ mg NH}_4/\text{l}$  and the average HLR  $0.33 \text{ l}$  resulted in the addition of  $274 \text{ mg NH}_4$ , which would infer that the ASF-BR was capable of removing the influent  $\text{NH}_4$  in each treatment cycle. Table 6.23 presents the nitrogen mass balance from the ISs carried out in Phase 2c.

**Table 6.23 Nitrogen mass balance – Phase 2c \***

		$\text{TN}_f$	$\text{NH}_4$	$\text{NO}_2$	$\text{NO}_3$
		$\text{g/d}^1$	$\text{g/d}^1$	$\text{g/d}^1$	$\text{g/d}^1$
<b>Post dilution mass</b>	$M_{p, \text{dil}}$	94.1 (8.58)	15.9 (0.82)	58.3 (8.92)	24.3 (0.48)
<b>Mass change - Anoxic period</b>	$\Delta M_{\text{anx}}$	1.46 (0.74)	0.54 (0.61)	-1.04 (0.65)	-2.62 (0.86)
<b>Mass change - Aerobic period</b>	$\Delta M_{\text{aer}}$	-1.72 (1.34)	-0.89 (3.8)	1.10 (0.02)	2.88 (0.33)
<b>Effluent</b>	$M_{\text{eff}}$	93.8 (8.55)	15.3 (0.94)	58.8 (9.00)	24.5 (0.49)

\* Average results from ISs carried out in Phase 1;  $M_{p, \text{dil}}$  – average post dilution mass;  $M_{\text{aer}}$  – average change in mass during aeration period;  $M_{\text{eff}}$  – average effluent mass; <sup>1</sup> 3 cycles/d; Standard deviation shown in ().

The increase in the mass of glucose added resulted in an improvement in the denitrification observed. The overall TN and  $\text{TN}_f$  removal averaged 21% and 19% respectively over LLIS 9 and LLIS 10. Denitrification resulted in average TN and  $\text{TN}_f$  reductions of 3% and 5% respectively. During the aerobic period of LLIS 9 and LLIS 10,  $\text{TN}_f$  was reduced by  $7.6 \text{ mg/l}$  and  $26 \text{ mg/l}$  respectively. Cell synthesis accounted for approximately 20% of the  $\text{TN}_f$  reduction in LLIS 9 and 8% of the  $\text{TN}_f$  reduction in LLIS 10.

A reduction of 4% in the  $\text{NO}_3$  concentration during the anoxic period of LLIS 9 was observed. The average removal rate during this period was  $1.1 \text{ mg NO}_3/\text{m}^2 \text{ media}/\text{min}$ . During the anoxic period of LLIS 10, 4%  $\text{NO}_3$  removal was observed at a rate of  $1.23 \text{ mg NO}_3/\text{m}^2 \text{ media}/\text{min}$ . The nitrogen trends during the anoxic and aerobic periods of LLIS 9 and LLIS 10 are summarised in Table 6.24.

**Table 6.24 Summary of NH<sub>4</sub> and NO<sub>3</sub> linear trends during LLIS 9 and LLIS 10**

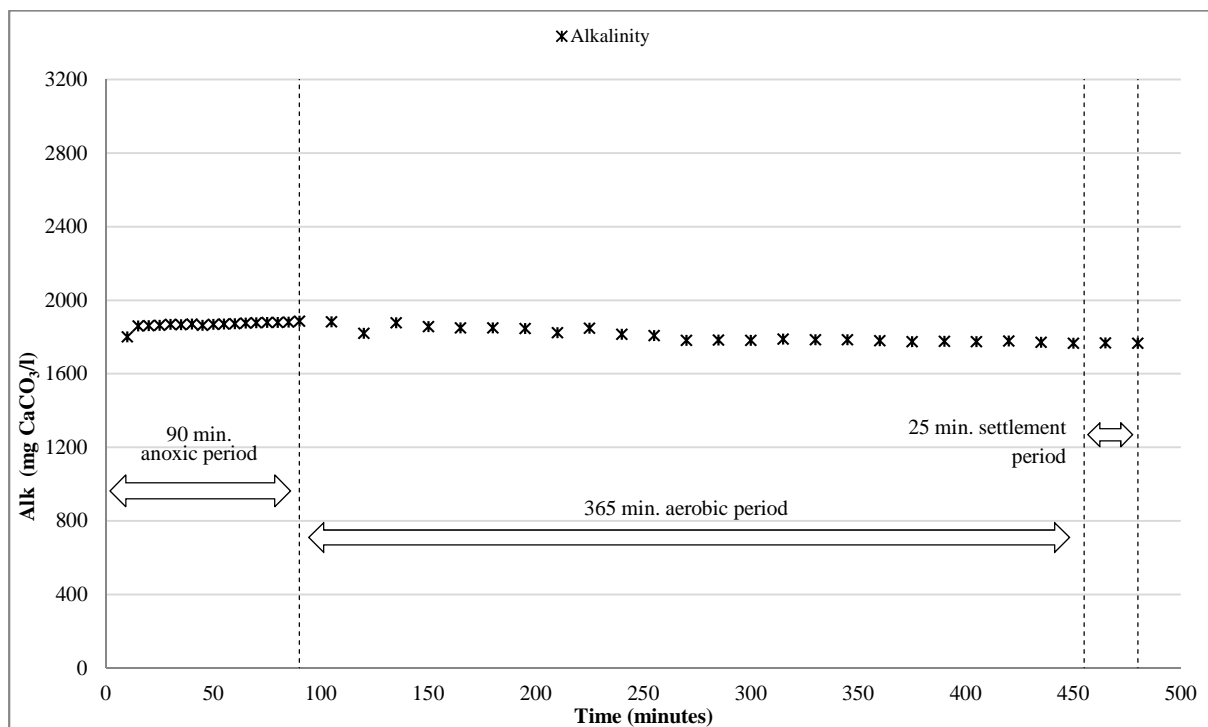
MWIS	N process	Period	Time	Linear trend	R <sup>2</sup>	% D <sub>avg</sub>
			min	mg/l		
MWIS 9	NH <sub>4</sub> reduction	Anoxic	5-90	NH <sub>4</sub> = - 0.022 t + 174.1	0.51	0.6
	NH <sub>4</sub> oxidation	Aerobic	90-450	NH <sub>4</sub> = - 0.041 t + 175.7	0.34	-0.2
	NO <sub>3</sub> reduction	Anoxic	5-90	NO <sub>3</sub> = - 0.648 t + 267.7	0.59	-0.3
	NO <sub>3</sub> increase	Aerobic	90-450	NO <sub>3</sub> = 0.022 t + 220.4	0.10	-0.4
MWIS 10	NH <sub>4</sub> reduction	Anoxic	10-90	NH <sub>4</sub> = - 0.008 t + 152.7	0.01	-0.1
	NH <sub>4</sub> oxidation	Aerobic	90-450	NH <sub>4</sub> = - 0.042 t + 162.4	0.58	0.1
	NO <sub>3</sub> reduction	Anoxic	10-90	NO <sub>3</sub> = - 0.013 t + 216.1	0.003	-0.1
	NO <sub>3</sub> increase	Aerobic	90-450	NO <sub>3</sub> = 0.036 t + 214.1	0.06	-0.7

**Note:** The low R<sup>2</sup> values for the linear relationship of the NO<sub>3</sub> and NH<sub>4</sub> with respect to time in MWIS 9 and MWIS 10 are attributable to the high concentrations of the nutrients and the relatively modest changes observed.

To validate the linear relationship, the difference (D<sub>avg</sub>) between the linear relationship and the actual concentration was calculated, the average of which is presented in Table 6.21.

#### 6.5.4.3 Alkalinity

During the anoxic period of both LLIS 9 and LLIS 10, there was an increase in the alkalinity of the LL due to denitrification, while alkalinity was consumed during the nitrification process. Figure 6.17 presents the alkalinity profile observed during LLIS 9 (typical of that observed during LLIS 10 also).

**Figure 6.17 Alkalinity profile – LLIS 9**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.



Table 6.25 summarises the rate of change of alkalinity during LLIS 9 and LLIS 10, along with the quantity of alkalinity produced per unit NO<sub>3</sub> removed and the rate of alkalinity consumption per unit NH<sub>4</sub> oxidised.

**Table 6.25 Alkalinity production and removal rates – Phase 2c<sup>a</sup>**

	Anoxic period <sup>1,2</sup>		Aerobic <sup>3</sup>	
	mg CaCO <sub>3</sub> produced/ m <sup>2</sup> media/min <sup>b</sup>	mg CaCO <sub>3</sub> produced/ mg NO <sub>3</sub> reduced <sup>b</sup>	mg CaCO <sub>3</sub> consumed/ m <sup>2</sup> media/min <sup>b</sup>	mg CaCO <sub>3</sub> consumed/ mg NH <sub>4</sub> oxidised <sup>b</sup>
LLIS 9	3.52	3.24	1.38	6.89
LLIS 10	2.46	1.99	0.77	6.25

<sup>a</sup> Glucose added at a rate of 20 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l; <sup>b</sup> Timing as per Table 6.20; <sup>1</sup> Post dilution; <sup>2</sup> Surface area loading rate calculated based on surface area in R1 (4.1m<sup>2</sup>); <sup>3</sup> Surface area loading rate based on surface area of R1 and R2 (8.2m<sup>2</sup>).

## 6.6 Excess sludge production

Desludging was not required during this study as visual inspections found no accumulation of excess sludge in the base of either reactor. Estimates were made on the overall excess sludge production, as outlined in Section 3.5.4.2. The excess sludge production in Phase 1, Phase 2a, Phase 2b and 2c is presented in Table 6.26.

**Table 6.26 Excess sludge production**

	V <sub>treated</sub>	ΔSS	ΔCOD <sub>f</sub>	Excess Sludge
	l	g	g	g SS/g COD
Phase 1	128	2673	130	0.05
Phase 2a	11	25	6	0.24
Phase 2b	16	46	18	0.39
Phase 2c	8	9	82	0.94

The excess sludge production observed during Phase 1, Phase 2a and Phase 2b compares favourably with those observed by Klimiuk and Kuliowska (2006), who reported an excess sludge production of between 0.04 and 0.6 mg SS/mg COD depending on the operating configuration of the treatment unit. The excess production observed in Phase 2c is likely due to increased biomass growth as a result of increased influent COD<sub>f</sub> concentrations.

### 6.7 A comparison of the treatment performance of the ASF-BR in Study 1, 2 and 3

Table 6.27 presents a summary of the highest level of treatment performance achieved by the ASF-BR in Phase 2 of Study 1, Phase 2b of Study 2 and Phase 2c of Study 3.

**Table 6.27 Comparison of treatment performance in Study 1, 2 and 3**

	<b>COD<sub>f</sub></b>	<b>COD<sub>f</sub></b>	<b>%</b>	<b>NH<sub>4</sub></b>	<b>NH<sub>4</sub></b>	<b>%</b>
	<b>loading rate</b>	<b>removal rate</b>	<b>Removal</b>	<b>loading rate</b>	<b>removal rate</b>	<b>Removal</b>
	mg/m <sup>2</sup> media/d	mg/m <sup>2</sup> media/d		mg/m <sup>2</sup> media/d	mg/m <sup>2</sup> media/d	
Study 1	2.49	2.42	97	0.21	0.20	99
Study 2	4.72	4.19	89	0.14	0.13	93
Study 3	0.36	0.12	33	0.11	0.08	73

From Table 6.24, it can be seen that the performance of the ASF-BR declined as the complexity of the wastewater increased. Phase 2 of Study 1 yielded the highest level of overall treatment performance at the highest NH<sub>4</sub> loading rate. This is the result of the use of synthetic wastewater, which ensured a consistent source of nutrients ideal for bacterial growth and development, free of inhibitory compounds which can be found in municipal wastewater and landfill leachate. The high COD<sub>f</sub> loading rate in Phase 2b was due to amendment of the influent municipal wastewater with an external source of carbon (as previously discussed). Study 3 had the lowest loading rates together with the lowest removal efficiencies. Landfill leachate is regarded as one of the most difficult wastewaters to treat and in this study, the majority of the carbon was found to be non-biodegradable. There was also evidence from the ISs that a high portion of the nitrogen was recalcitrant in nature (Cortez *et al.*, 2010; Loukidou and Zouboulis, 2001).

### 6.8 Performance comparison to alternative treatment process

Table 6.28 and Table 6.29 present the treatment performance of alternative biofilm treatment processes used in the treatment of landfill leachate found in the literature. All studies referred to the difficulties associated with the biological treatment of landfill leachate.

**Table 6.28 Comparison of COD removal performance with alternative biofilm treatment systems**

Study	Treatment Process	Organic loading rate	Removal rate	Comments
			0.13 g COD/m <sup>2</sup> media/d <sup>1</sup> ; 65% removal efficiency <sup>1</sup>	
	ASF-BR – Phase 1	0.20 g COD/m <sup>2</sup> media/d <sup>1</sup> 0.20 g COD <sub>f</sub> /m <sup>2</sup> media/d <sup>1</sup>	0.12 g COD <sub>f</sub> /m <sup>2</sup> media/d <sup>1</sup> ; 60% removal efficiency <sup>1</sup>	
	ASF-BR – Phase 2a	0.23 g COD/m <sup>2</sup> media/d 0.21 g COD <sub>f</sub> /m <sup>2</sup> media/d	0.01 g COD/m <sup>2</sup> media/d; 4% removal efficiency 0.01g COD <sub>f</sub> /m <sup>2</sup> media/d; 5% removal efficiency	Unit operated at 11 °C; Excellent removal of both NH <sub>4</sub> and COD; Potential to capture of all gases produced
	ASF-BR – Phase 2b	0.26 g COD/m <sup>2</sup> media/d 0.24 g COD <sub>f</sub> /m <sup>2</sup> media/d	0.01 g COD/m <sup>2</sup> media/d; 4% removal efficiency 0.01 g COD <sub>f</sub> /m <sup>2</sup> media/d; 4% removal efficiency	
	ASF-BR – Phase 2c	0.42 g COD/m <sup>2</sup> media/d 0.36 g COD <sub>f</sub> /m <sup>2</sup> media/d	0.14 g COD/m <sup>2</sup> media/d; 33% removal efficiency 0.12 g COD <sub>f</sub> /m <sup>2</sup> media/d; 33% removal efficiency	
Chen <i>et al.</i> , 2008	MBBR – Anaerobic aerobic reactors	0.33 g COD/m <sup>2</sup> media/d	95% removal efficiency; 0.31 g COD/m <sup>2</sup> media/d	HRT - 24 h; Unit operated at 35 °C; LF operational for 4 years
Castillo <i>et al.</i> , 2007	RBC	24.7 g COD/m <sup>2</sup> media/d	13.1 g COD/m <sup>2</sup> media/d; 53% COD removal efficiency	HRT – 24 h; No details on nitrogen removal
Loukidou <i>et al.</i> , 2001	MBBR	Influent 5000 mg COD/l <sup>2</sup> ; polyurethane cubes used as media	81% COD removal efficiency;	Biomass detachment from the media and began to hydrolyse in the bulk fluid; Molasses used as a carbon source to improved denitrification

Nomenclature: <sup>1</sup> Average value; <sup>2</sup> Surface area of media not specified; RBC- Rotating biological contactor; MWW – Municipal wastewater; SMWW – Synthetic municipal wastewater; MBBR – Moving bed biofilm reactor; APSBBR – Alternating pumped sequencing batch biofilm reactor; LF – Landfill.

**Table 6.29 Comparison of ammonium removal performance with alternative biofilm treatment systems**

Study	Treatment Process	Organic loading rate	Removal rate	Comments
	ASF-BR – Phase 1a	0.14 g NH <sub>4</sub> /m <sup>2</sup> media/d <sup>1</sup>	0.12 g NH <sub>4</sub> /m <sup>2</sup> media/d; 86% removal efficiency	Unit operated at 11 °C; Excellent removal of both NH <sub>4</sub> and COD; Potential to capture of all gases produced
	ASF-BR – Phase 2a	0.10 g NH <sub>4</sub> /m <sup>2</sup> media/d <sup>1</sup>	0.08 g NH <sub>4</sub> /m <sup>2</sup> media/d; 80% removal efficiency	
	ASF-BR – Phase 2b	0.09 g NH <sub>4</sub> /m <sup>2</sup> media/d <sup>1</sup>	0.07 g NH <sub>4</sub> /m <sup>2</sup> media/d; 78% removal efficiency	
	ASF-BR – Phase 2c	0.11 g NH <sub>4</sub> /m <sup>2</sup> media/d <sup>1</sup>	0.08 g NH <sub>4</sub> /m <sup>2</sup> media/d; 73% removal efficiency	
Kulikowska <i>et al.</i> , 2010	RBC	3.58 g NH <sub>4</sub> /m <sup>2</sup> media/d	99% NH <sub>4</sub> removal efficiency	Unit operated at 20 °C; LF operational for 12 years
Chen <i>et al.</i> , 2008	MBBR	0.20 g NH <sub>4</sub> /m <sup>2</sup> media/d	0.19 g NH <sub>4</sub> /m <sup>2</sup> media/d; 97% NH <sub>4</sub> removal efficiency	HRT 24 h; Unit operated at 35 °C
Tengrui <i>et al.</i> , 2007	SBBR <sup>2</sup>	0.51 kg NH <sub>4</sub> /m <sup>3</sup> /d	89% NH <sub>4</sub> removal efficiency	Unit operated at 30-33 °C Biomass was observed to detach from the media and begin to hydrolyse in the bulk fluid; Molasses used as a carbon source to improved denitrification; Poor denitrification noted
Loukidou <i>et al.</i> , 2001	MBBR – polyurethane cubes used as media <sup>2</sup>	Influent 1800 mg NH <sub>4</sub> /l	85% removal efficiency	

Nomenclature: <sup>1</sup> Average values; <sup>2</sup> Surface area of media not specified; SBBGR – Sequencing batch biofilm granular reactor; MBBR – Moving bed biofilm reactor; LF – Landfill.

## 6.9 Discussion

### 6.9.1 Future research

The findings in this study are supported by the views of Bohdziewicz *et al.* (2001), Loukidou and Zouboulis (2001) and Ushikoshi *et al.* (2002), who state that LL is one of the most difficult wastewaters to treat. Future studies investigating the treatment of LL using the ASF-BR could employ an advanced oxidation pre-treatment process (AOP) to enhance the biodegradability of the LL. De Morais and Zamora (2005) and Lin and Kiang (2003) both used AOP methods (photochemical and Fenton methods respectively) to successfully convert non-biodegradable compounds into biodegradable intermediaries. By pre-treating the LL through an AOP unit, the biodegradable organic content of the LL can be increased.

## 6.10 Conclusion

The performance of a pilot scale ASF-BR treating LL was investigated over a 329-day period. The objective of this study was to evaluate the efficacy of the ASF-BR in removing nitrogen and carbon from LL. The study consisted of two phases: Phase 1 (254 days) and Phase 2 (75 days). During Phase 1, the LL was treated using biological nitrification over an 8.25 h cycle. The average removals efficiencies in Phase 1 were 57% COD, 63% COD<sub>f</sub> and 86% NH<sub>4</sub>.

During Phase 2, a 90-minute anoxic period was added to facilitate the biological denitrification of the LL. Batch studies revealed that a high portion of the carbon content of the LL was non-biodegradable. Thus during Phase 2b, an external carbon source was added to the influent LL. The mass added was increased incrementally over time. At the highest carbon loading rate, 20 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l was added to the influent wastewater. The average COD and COD<sub>f</sub> removal were 34% and 33% respectively. Biological denitrification resulted in an average removal of 13% in the influent TN, with an average TN<sub>f</sub> removal of 7% and an 80% reduction in the average influent NH<sub>4</sub>. Despite positive results from the influent and effluent sampling, this study further confirms that LL is one of the most difficult wastewaters to treat. Other researchers have employed a pre-treatment process to increase the biodegradability of the LL. Future investigations into the treatment of LL using the ASF-BR will require pre-treatment to increase its biodegradability.

Studies 1, 2 and 3 have identified a number of issues which could be addressed in future site units:

- In designing future ASF-BR units, an alternative method of monitoring the wastewater levels within the reactor should be investigated. The float level switches employed were found to be susceptible to biofilm accumulation, and required regular maintenance. A possible alternative could be pressure transducers.
- In the pilot scale unit used in these studies, the lids of the reactors were bolted to the reactor with a rubber gasket to maintain an air tight seal. Thus opening the reactors to monitor the biofilm accumulation on the media was a time consuming, laborious process. Future ASF-BRs should encompass a quick and simple way of viewing the media without compromising the operation of the unit.
- The use of vacuum pumps would need to be further investigated at a larger scale to ensure they are cost-effective, robust and reliable.

# **CHAPTER 7**

## **An analysis of N<sub>2</sub>O emissions**

## 7.1 Introduction

Biological treatment of wastewater results in the production of a variety of gases, the most prevalent of which are carbon dioxide (CO<sub>2</sub>), methane (CH<sub>4</sub>) and nitrous oxide (N<sub>2</sub>O). The emission of N<sub>2</sub>O is of particular concern due to its global warming potential (GWP) of 300 relative to CO<sub>2</sub>. Legislative requirements are becoming increasingly stringent with regard to the permissible effluent nitrogen limits from wastewater treatment plants. To comply with such limits, an increase in the utilisation of nitrification and denitrification processes will be required. However, this could potentially result in an increase in the production of N<sub>2</sub>O. Thus environmentally sustainable wastewater treatment systems of the future must encompass the ability to control gases produced. The ASF-BR offers the potential to control gaseous emissions arising from the biological treatment of wastewater. This chapter presents the results of an analysis of nitrous oxide (N<sub>2</sub>O) emissions during the treatment of municipal wastewater (Phase 2, Study 2) and landfill leachate (Phase 2, Study 3). Both investigations were carried using the sampling and analysis procedures outlined in Section 3.5.2

## 7.2 N<sub>2</sub>O emissions arising from the treatment of municipal wastewater

### 7.2.1 Study configuration

The apparatus, operational configuration and influent wastewater characteristics of the MWW used in this study were as outlined in Sections 3.5.2, Section 5.6.1 and Section 5.6.2 respectively.

### 7.2.2 Overall performance – Study 2 – Phase 2

As outlined in Section 5.6, during Phase 2a, the COD<sub>f</sub>/TN<sub>f</sub> (measured as biodegradable carbon) ratio was likely limiting for optimal denitrification performance, which is generally the process responsible of the production of N<sub>2</sub>O. Thus during Phase 2b, an external carbon source in the form of glucose (C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>) was added to the influent municipal wastewater at a rate of 0.8 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l to provide the required source of readily biodegradable organic carbon, resulting in an increase in the COD<sub>f</sub>/TN<sub>f</sub> ratio to 30.9. A summary of the system performance from Study 2 is outlined in Table 7.1.



**Table 7.1 Average municipal wastewater characteristics before and after glucose addition**

	Phase 2a <sup>1</sup>			Phase 2b <sup>2</sup>		
	Influent conc.	Loading rate <sup>a, c</sup>	Removal %	Influent conc.	Loading rate <sup>b, c</sup>	Removal %
	mg/l	g/m <sup>2</sup> media/d		mg/l	g/m <sup>2</sup> media/d	
COD <sup>**</sup>	359 (82)	3.07	75	870 (106)	7.41	75
COD <sub>f</sub>	79 (29)	0.67	50	554 (121)	4.72	89
TN <sup>**</sup>	23.2 (5.1)	0.20	53	16.5 (3.6)	0.14	73
TN <sub>f</sub>	21.5 (4.5)	0.18	57	15.9 (3.6)	0.14	75
NH <sub>4</sub>	21.9 (5.2)	0.19	95	16.0 (3.7)	0.14	93
TON	0.9 (1.3)	0.01	-	0.3 (0.2)	0	-
NO <sub>2</sub>	0.1 (0.2)	0	-	0.1 (0.17)	0	-
NO <sub>3</sub>	0.8 (1.3)	0.01	-	0.1 (0.12)	0	-
Alk	219 (71.0)	1.87	-	269 (25)	-	-

All parameters marked \*\* were measured unfiltered, all other samples were filtered; <sup>1</sup> Raw MWW – COD<sub>f</sub>/TN<sub>f</sub> = 1.9; <sup>2</sup> carbon added at a rate of 0.8 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l – COD<sub>f</sub>/TN<sub>f</sub> = 30.9; <sup>a</sup> average influent volume 35.1 l/d; <sup>b</sup> Average influent volume 34.9 l/d; <sup>c</sup> surface area of 4.1 m<sup>2</sup> media; standard deviations shown thus ( ).

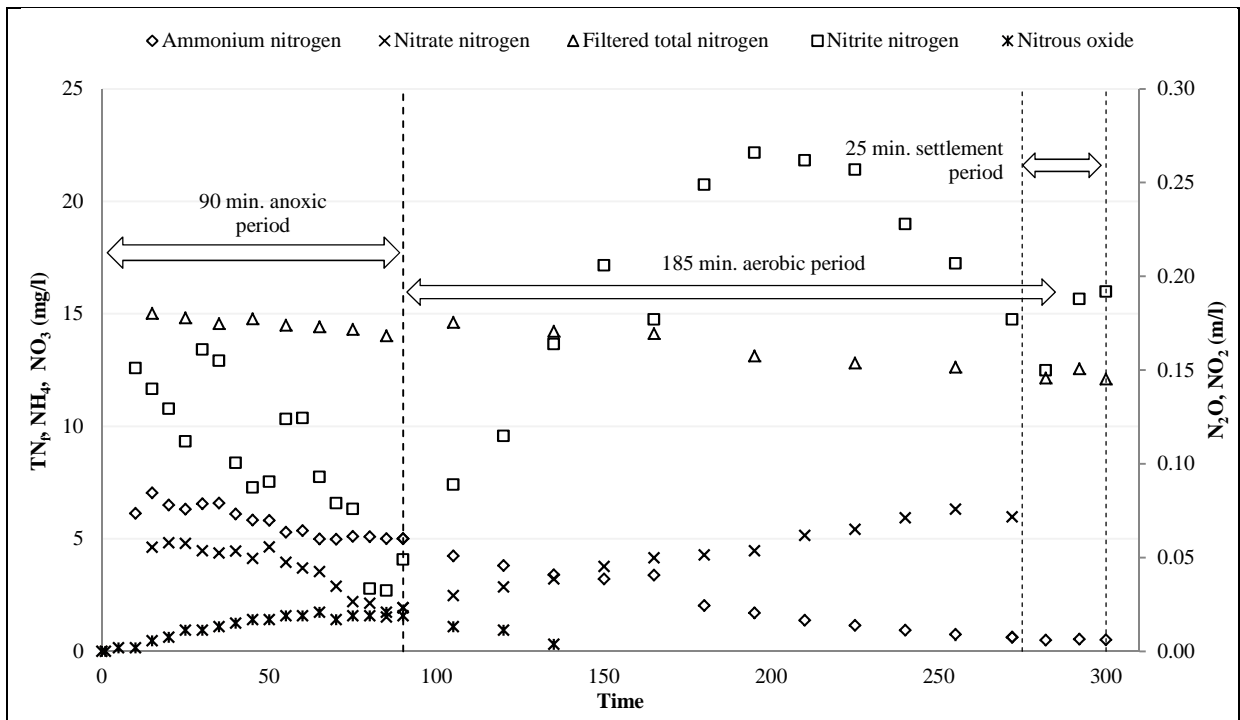
### 7.2.3 Intensive study analysis

The N<sub>2</sub>O emitted was monitored, as outlined in Section 3.5.2, during the four of the five ISs conducted, as outlined in Table 7.2.

**Table 7.2 IS conducted during Phase 2 – Study 2**

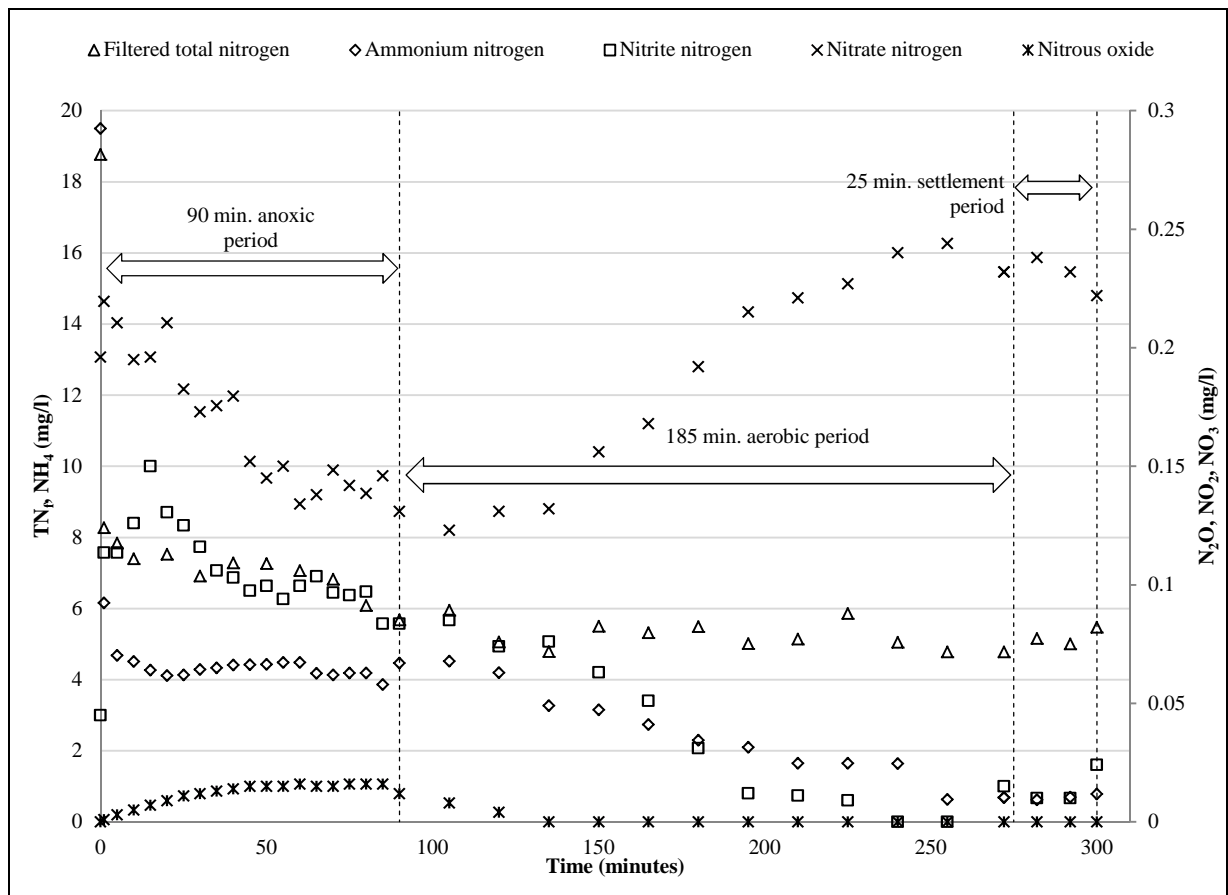
Phase	Additional carbon	IS	Day
Phase 2a	No glucose	MWIS 5	40
		MWIS 6	48
Phase 2b	0.8 g C <sub>6</sub> H <sub>12</sub> O <sub>6</sub> /l	MWIS 7	104
		MWIS 8	117

Figure 7.1 (a) presents the nitrogen and N<sub>2</sub>O profiles observed during MWIS 5, and Figure 7.1 (b) depicts the N<sub>2</sub>O and nitrogen profiles from MWIS 8. Both of these profiles are typical of those obtained in Phases 2a and 2b respectively. The N<sub>2</sub>O emissions from all ISs conducted during Phases 2a and 2b are available in Appendix G.



**Figure 7.1 (a) Nitrogen profiles – MWIS 5**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.



**Figure 7.1 (b) Nitrogen profiles – MWIS 8**

From the profiles presented in Figure 7.1 (a) and Figure 7.1 (b), it can be seen that in both Phases 2a and 2b, on commencement of the aerobic period, emissions of N<sub>2</sub>O were observed to decrease. After approximately 20 to 30 minutes, there was no further production of N<sub>2</sub>O. The average N<sub>2</sub>O production as a percentage of the observed changes in the NH<sub>4</sub>, TN and TN<sub>f</sub> from the ISs conducted in Phase 2a and Phase 2b are presented in Table 7.3.

**Table 7.3 Average N<sub>2</sub>O production during the anoxic periods – Study 2**

	$\Delta$ N <sub>2</sub> O	$\Delta$ TN <sub>f</sub>	COD <sub>f</sub> /TN <sub>f</sub>	N <sub>2</sub> O emission <sup>a</sup>		
	mg N <sub>2</sub> O/l/min	mg TN <sub>f</sub> /l/min		% NH <sub>4</sub>	% TN	% TN <sub>f</sub>
<b>Phase 2a</b> <sup>1</sup>	0.00019	0.015	1.9	0.71	0.70	0.68
<b>Phase 2b</b> <sup>2</sup>	0.00013	0.023	30.9	0.38	0.43	0.57

<sup>a</sup> N<sub>2</sub>O emitted as a % of nitrogen species removed; <sup>1</sup> No glucose added; <sup>2</sup> 0.8g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l added to influent wastewater.

The addition of the glucose resulted in an improvement in the denitrification performance of the ASF-BR. A lower average percentage of the influent TN was emitted as N<sub>2</sub>O when the COD<sub>f</sub>/TN<sub>f</sub> was increased to 30.9 through the addition of 0.8g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l. This was a result of an increase in the TN and TN<sub>f</sub> removal efficiency from 13% and 7% (Figure 5.19) respectively over the ISs in Phase 2a to an average of 39% and 36% in Phase 2b (Figure 5.24). N<sub>2</sub>O emissions during Phase 2b were less than those observed during Phase 2a, which suggests that the N<sub>2</sub>O was reduced to nitrogen gas in the final step of the denitrification process.

#### **7.2.4 N<sub>2</sub>O produced under optimal denitrification conditions – Phase 2b**

NO<sub>3</sub> concentrations were reduced, on average, by 32% during the anoxic periods of MWIS 7 and MWIS 8. The average rate of N<sub>2</sub>O production was 0.18 mg N<sub>2</sub>O/mg NO<sub>3</sub>, removed during MWIS 7 and MWIS 8. In Phase 2a, 0.71%, 0.70% and 0.68% of the influent NH<sub>4</sub>, TN and TN<sub>f</sub> respectively were converted to N<sub>2</sub>O following dilution. In Phase 2b an improvement in the denitrification performance resulted in a reduction in the N<sub>2</sub>O emission rate, and the percentage of the influent NH<sub>4</sub>, TN and TN<sub>f</sub> emitted as N<sub>2</sub>O, following dilution, was also reduced to 0.38%, 0.43% and 0.57% respectively. These results compare favourably with previous studies (Table 2.7).

Thus based on a wastewater production rate of 150 l/population equivalent (PE)/d (Tchobanoglous *et al.*, 2004) with a concentration of 16 mg NH<sub>4</sub>/l, as was observed in Phase 2b of this study, and N<sub>2</sub>O production rate of 0.07 % of the influent NH<sub>4</sub>, the ASF-BR would

produce 61.3 mg N<sub>2</sub>O/PE/yr, equivalent to 18,390 mg CO<sub>2</sub>/PE/yr, which could be controlled using the ASF-BR.

### **7.3 N<sub>2</sub>O emissions arising from the treatment of landfill leachate (LL)**

#### **7.3.1 Study configuration**

The apparatus, operational configuration and influent wastewater characteristics are outlined in Sections 3.5.3, Section 6.5.1 and Section 6.5.2 respectively.

#### **7.3.2 Overall performance – Study 3 – Phase 2**

As detailed in Section 6.5.2, Phase 2 of Study 3 consisted of three sub periods: Phase 2a (10 days), when raw LL was being treated; Phase 2b (32 days), when an external carbon in the form of glucose was added at a rate of 2.4 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l to the influent LL to increase the availability of biodegradable carbon required for the denitrification process; and Phase 2c (25 days) when the mass of carbon added was increased to 20 g C C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l. The hydraulic loading rate in this study averaged 1 l/d, with 3 g CaCO<sub>3</sub>/l being added to the influent leachate to ensure the alkalinity of the LL was sufficient for nitrification. Limited nitrification and denitrification were achieved when treating the LL due to the recalcitrant nature of the nitrogen contained in the leachate. Table 7.4 presents a summary of the average treatment performance in Phase 2. As outlined in Section 6.5, the availability of biodegradable carbon was insufficient, which limited the denitrification. Thus the results presented here refer to Phase 2b and Phase 2c, when the addition of an external source of carbon provide the conditions necessary for denitrification to occur (Section 2.5.2.5.2).

**Table 7.4 Average landfill leachate characteristics before and after glucose addition**

	Phase 2b <sup>1</sup>			Phase 2c <sup>2</sup>		
	Influent conc. mg/l	Loading rate <sup>a, c</sup> g/m <sup>2</sup> media/d	Removal %	Influent conc. mg/l	Loading rate <sup>b, c</sup> g/m <sup>2</sup> media/d	Removal %
COD <sup>**</sup>	2107 (59)	0.26	4	3440 (169)	0.42	34
COD <sub>f</sub>	1938 (42)	0.24	3	2973 (229)	0.36	33
TN <sup>**</sup>	1175 (21.5)	0.14	3	1229 (12.3)	0.16	13
TN <sub>f</sub>	1085 (26.9)	0.13	2	1080 (36.7)	0.14	7
NH <sub>4</sub>	710 (59.2)	0.09	78	879 (85.9)	0.11	73
TON	49.7 (16.5)	0.01	-	51.1 (24.2)	0.003	-
NO <sub>2</sub>	27.9 (7.9)	0.003	-	40.7 (19.4)	0.002	-
NO <sub>3</sub>	21.2 (14.7)	0.003	-	10.3 (5.7)	0.001	-
Alk	5197 (1282)	0.63	95	4557 (138)	0.58	71

All parameters marked <sup>\*\*</sup> were measured unfiltered, all other samples were filtered; <sup>1</sup> carbon added at a rate of 2.4 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l – COD<sub>f</sub>/TN<sub>f</sub> = 2.4; <sup>2</sup> carbon added at a rate of 20 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l – COD<sub>f</sub>/TN<sub>f</sub> = 8.2; <sup>a</sup> based on an average influent volume of 1 l/d; <sup>b</sup> based on an average influent volume of 1.05 l/d; <sup>c</sup> a surface area of 8.2m<sup>2</sup>; standard deviation shown thus ( ).

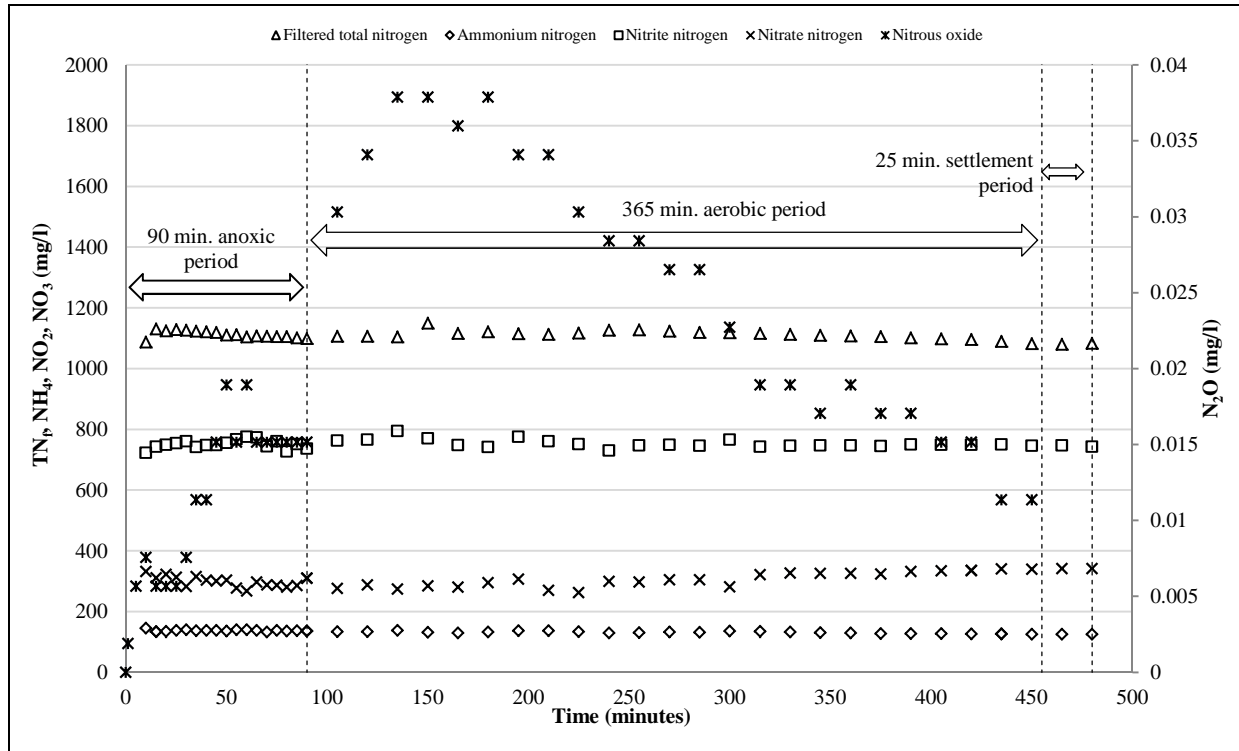
### 7.3.3 Intensive phase studies

Over Phase 2b and Phase 2c, a total of 4 ISs were conducted, as outlined in Table 7.5.

**Table 7.5 IS conducted during Phase 2 – Study 3**

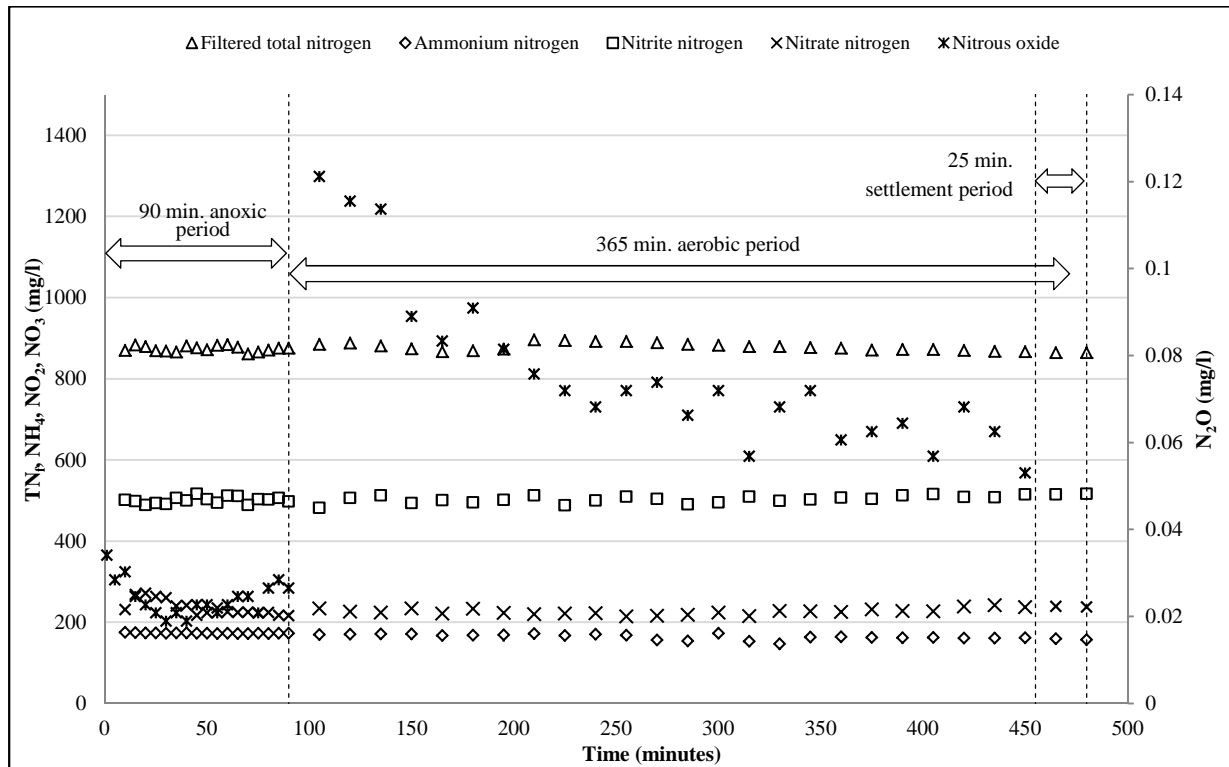
Phase	Additional carbon	IS	Day
Phase 2b	2.4 g C <sub>6</sub> H <sub>12</sub> O <sub>6</sub> /l	LLIS 7	31
		LLIS 8	46
Phase 2c	20 g C <sub>6</sub> H <sub>12</sub> O <sub>6</sub> /l	LLIS 9	57
		LLIS 10	74

The ISs from Phase 2b and Phase 2c have been presented in Section 6.5.3 and 6.5.4 in the context of the performance of the ASF-BR in the treatment of the landfill leachate. Figure 7.2 presents the results of LLIS 7 (Figure 7.2 (a)) and LLIS 10 (Figure 7.2 (b)), with the emitted N<sub>2</sub>O included. Full details of the N<sub>2</sub>O produced during the ISs in Phase 2b and 2c are available in Appendix G.



**Figure 7.2 (a) Nitrogen profiles – LLIS 7**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.



**Figure 7.2 (b) Nitrogen profiles – LLIS 10**

Note: Initial concentration is the theoretical post dilution concentration; Influent concentrations omitted.

From Figure 7.2 (a) and Figure 7.2 (b), it can be seen that the emission of N<sub>2</sub>O increased immediately in both Phase 2b and Phase 2c on the commencement of the aerobic period and subsequently decreased at a steady rate throughout the aerobic period. This may be due a number of factors, including: (i) release of dissolved N<sub>2</sub>O generated in the leachate during the anoxic period due to turbulence, (ii) simultaneous nitrification and denitrification (SND) during the aerobic period, and (iii) nitrifier denitrification. The reduction in N<sub>2</sub>O emissions throughout the aerobic period would tend to support the theory of a release of N<sub>2</sub>O which had been dissolved in the leachate during the anoxic period. O'Reilly (2011) attributed an increase in the dissolved N<sub>2</sub>O during an aerobic period to hydroxylamine oxidation as the NH<sub>4</sub> was oxidised to NO<sub>3</sub> when treating wastewater using a pumped flow biofilm reactor. It is possible that hydroxylamine oxidation may also have contributed to the observed increase in the N<sub>2</sub>O emission during the aerobic period in this study.

Table 7.6 presents the average N<sub>2</sub>O production during the anoxic periods of the ISs conducted during Phase 2b and Phase 2c.

**Table 7.6 Summary of average N<sub>2</sub>O production during the anoxic periods – Study 3**

	$\Delta$ N <sub>2</sub> O	$\Delta$ TN <sub>f</sub>	COD <sub>f</sub> /TN <sub>f</sub>	N <sub>2</sub> O emission <sup>a</sup>		
	mg N <sub>2</sub> O/l/min	mg TN <sub>f</sub> /l/min		% NH <sub>4</sub>	% TN <sub>t</sub>	% TN <sub>f</sub>
Phase 2b <sup>1</sup>	0.0003	0.38	0.06	0.18	0.03	0.05
Phase 2c <sup>2</sup>	0.0003	0.55	8.6	0.51	0.22	0.10

<sup>a</sup> N<sub>2</sub>O emitted as a % of nitrogen species removed; <sup>1</sup> 2.4 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l added to influent leachate; <sup>2</sup> 20 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub>/l added to influent leachate.

#### 7.3.4 N<sub>2</sub>O production highest denitrification performance – Phase 2c

In Phase 2c, the NO<sub>3</sub> removal rate increased from an average of 0.08 mg NO<sub>3</sub>/l/min in LLIS 7 and LLIS 8 to an average removal rate of 0.10 mg NO<sub>3</sub>/l/min in LLIS 9 and LLIS 10. The increase in the glucose added to the influent LL resulted in an increase in the TN and TN<sub>f</sub> reduction during the anoxic period, from an average of 3% and 5% respectively in Phase 2b to average reductions of 10% TN and 11% TN<sub>f</sub> respectively in Phase 2c, an N<sub>2</sub>O production rate of 2.6x10<sup>-5</sup> mg NO<sub>2</sub>/mg NO<sub>3</sub> removed. In Phase 2c, the increased level of denitrification resulted in an increase in the percentage of the influent NH<sub>4</sub>, TN and TN<sub>f</sub> emitted as N<sub>2</sub>O, which indicates that the unit was failing to complete the denitrification process. A further increase in the mass of glucose added may have resulted in an improvement in performance; however, such an approach would not be viable from a financial or logistical perspective in practical applications of the ASF-BR.

## 7.4 Summary

A pilot scale ASF-BR was monitored to determine the nitrous oxide generation and emission while operating under steady state conditions when treating municipal wastewater and landfill leachate. The N<sub>2</sub>O was measured in gaseous form emitted during the treatment cycle.

When treating municipal wastewater, it was found that under optimal denitrification conditions, 0.57% of the influent TN<sub>f</sub> was emitted as N<sub>2</sub>O. A reduction in the N<sub>2</sub>O emission following the addition of glucose indicated the addition of the carbon enabled more of the N<sub>2</sub>O to be converted to N<sub>2</sub> in the final step of the denitrification process. An increase in the carbon content of the effluent wastewater meant that the quantity of carbon exceeded the requirements of denitrification, resulting in carbon breakthrough.

When treating landfill leachate, limited denitrification resulted in a 0.10% of the influent TN<sub>f</sub> being emitted as N<sub>2</sub>O. On commencement of the aerobic period, the N<sub>2</sub>O emissions initially increased and subsequently decreased, indicating that N<sub>2</sub>O may have been dissolved within the LL and was released as the LL was alternated between the reactors during the treatment cycle. Calculations to determine the required carbon found that such an increase would not be logistically or financially feasible.



# **CHAPTER 8**

## **Conclusions**

## 8.1 Introduction

During this research, the efficacy of an innovative wastewater treatment technology, the Air Suction Flow Biofilm Reactor (ASF-BR), developed at NUI Galway, was investigated. The performance of the pilot scale ASF-BR was evaluated across four studies:

- (i) Study 1 – Treatment of high strength synthetic wastewater,
- (ii) Study 2 – Treatment of municipal wastewater,
- (iii) Study 3 – Treatment of landfill leachate,
- (iv) Study 4 – An analysis of N<sub>2</sub>O emissions from the ASF-BR.

The conclusions drawn from these studies are as follows:

## 8.2 Treatment of high strength synthetic wastewater

The performance of the ASF-BR in treating high strength synthetic wastewater was investigated over a period of 430 days and two phases (Phase 1 and Phase 2). Synthetic wastewater was used to ensure reasonable consistency of the influent wastewater.

- The following conclusions can be drawn from Phase 1.
  - The ASF-BR was capable of achieving average COD<sub>f</sub>, NH<sub>4</sub> and TN<sub>f</sub> removal efficiencies in Phase 1 of 98%, 97% and 86% respectively, with loading rates of 3.00 g COD<sub>f</sub>/m<sup>2</sup> media/d, 0.24 g NH<sub>4</sub>/m<sup>2</sup> media/d and 0.28 g TN<sub>f</sub>/m<sup>2</sup> media/d.
  - Heavy biofilm accumulation was observed on the plastic media. The dense thick biofilm is likely to have provided conditions conducive to simultaneous nitrification and denitrification (SND).
  - DO profiling conducted during this phase indicated that the treatment cycle could be configured in a more efficient manner by reducing the number of pumping cycles between the reactors.
- In Phase 2, the energy efficiency of the unit was increased by prolonging the time that wastewater was held in each reactor. This resulted in a 66% reduction in the energy requirements of the system with limited impact of the performance of the unit. The following conclusions were drawn from Phase 2:
  - The ASF-BR was operated at a loading rate of 2.49 g COD<sub>f</sub>/m<sup>2</sup> media/d, 0.21 g NH<sub>4</sub>/m<sup>2</sup> media/d and 0.24 g TN<sub>f</sub>/m<sup>2</sup> media/d, with removal efficiencies of 97%, 99% and 88% achieved respectively.
  - Minimum DO concentrations between pumping phases were lower than in Phase 1. However, this did not appear to inhibit reactor performance.

- As in Phase 1, heavy biofilm accumulation was observed on the media, which may have allowed SND to occur.
- Samples of biofilm were taken periodically throughout the study. Ammonia oxidising bacteria (AOB) populations were relatively constant throughout. This indicated the use of a partial vacuum within the ASF-BR did not inhibit bacterial growth. Furthermore the reduced oxygen concentrations in Phase 2 did not significantly affect AOB populations.
- AOB densities (per g dry cell mass) were significantly higher on the plastic media than in suspended biomass.
- The float level switches used to control the movement of the wastewater were subject to biofilm accumulation, which effected the operation of the unit. There were no issues with any of the other mechanical components during Study 1.

### **8.3 Treatment of municipal wastewater**

In Study 2, the performance of a pilot scale ASF-BR was monitored when treating municipal wastewater over a 329-day period. This study consisted of two phases: (i) Phase 1, which focused on organic carbon removal and nitrification, and (ii) Phase 2, in which the wastewater was treated using both nitrification and denitrification.

- The following conclusions were drawn from Phase 1:
  - Operating at average loading rates of 1.43 g COD<sub>f</sub>/m<sup>2</sup> media/d, 0.32 g NH<sub>4</sub>/m<sup>2</sup> media/d and 0.35 g TN<sub>f</sub>/m<sup>2</sup> media/d in Phase 1, the ASF-BR achieved average removal efficiencies of 68% COD<sub>f</sub>, 89% NH<sub>4</sub> and 31% TN<sub>f</sub>.
  - The wastewater used in Phase 1 was sourced from a municipal wastewater treatment facility which accepted landfill leachate from a local municipal landfill site. Contaminants in the leachate possibly inhibited biological activity during the treatment process. This was exacerbated in the latter stage of the phase due to a revised regime for adding the landfill leachate at the wastewater treatment plant.
- In Phase 2, the wastewater was sourced from an alternative wastewater treatment plant and an anoxic period was incorporated into the cycle to facilitate denitrification. The main conclusions of Phase 2 are:
  - The carbon content of the municipal wastewater used in Phase 2 probably limited biological denitrification of the wastewater. Thus an external source of carbon in the form of glucose was added to the wastewater.

- When glucose was added to the influent wastewater at a rate of 0.8 g C<sub>6</sub>H<sub>12</sub>O<sub>6</sub> /l, average removal efficiencies of 89% COD<sub>f</sub>, 93% NH<sub>4</sub> and 75% TN<sub>f</sub> were achieved (the average loading rates were 4.72 g COD<sub>f</sub>/m<sup>2</sup> media/d, 0.14 g NH<sub>4</sub>/m<sup>2</sup> media/d and 0.14 g TN<sub>f</sub>/m<sup>2</sup> media/d).
- The addition of external carbon requires careful control to avoid increases in effluent organic carbon.
- The performance of the ASF-BR in Study 2 was found to be comparable to alternative treatment processes used in the treatment of municipal wastewater.

#### **8.4 Treatment of landfill leachate**

In Study 3, the efficacy of the ASF-BR in treating landfill leachate was investigated over a 329-day period. Landfill leachate is acknowledged as being one of the most difficult wastewaters as it can contain a diverse range of toxins and recalcitrant compounds. Study 3 consisted of 2 phases: (i) Phase 1, which focused on organic carbon removal and nitrification, and (ii) Phase 2, which focused on nitrification and denitrification.

- The following conclusions were drawn from Phase 1 (Study 3):
  - Over the first 134 days of Phase 1, the ASF-BR achieved average removal efficiencies of 73% COD<sub>f</sub>, 89% NH<sub>4</sub> and 50% TN<sub>f</sub> at average loading rates of 0.20 COD<sub>f</sub>/m<sup>2</sup> media/d, 0.14 g NH<sub>4</sub>/m<sup>2</sup> media/d and 0.14 g TN<sub>f</sub>/m<sup>2</sup> media/d. Effluent TN and TN<sub>f</sub> concentrations increased at a constant rate throughout this period.
  - After 160 days, the treatment performance of the ASF-BR began to decline, which was found to be the result of a deterioration of the biofilm. The unit was subsequently reseeded, resulting in a limited improvement in performance.
  - During the initial 193 days of the phase, the availability of alkalinity may have limited the nitrification process. Following the reseeded on Day 193, alkalinity was added to the influent leachate to prevent any possible limitation.
- In Phase 2, an anoxic period was scheduled at the beginning of the treatment cycle to provide conditions conducive to denitrification.
  - Batch tests conducted on the leachate, found that a high portion of the carbon contained in the leachate was non-biodegradable. This is often characteristic of leachate originating from older landfills and thus an external carbon source was added to the influent wastewater.

- In Phase 2c, when the mass of external carbon added was increased to 20 g  $C_6H_{12}O_6/l$ , the average loading rates were 0.42 g  $COD_f/m^2$  media/d, 0.11 g  $NH_4/m^2$  media/d and 0.14 g  $TN_f/m^2$  media/d, with the ASF-BR achieving average removal efficiencies of 33%  $COD_f$ , 73%  $NH_4$  and 7%  $TN_f$ .
- The addition of an external source of carbon was shown to improve the denitrification performance of the ASF-BR when treating landfill during Phase 2.
- The relatively low removal rates and efficiencies of both the carbon and nitrogen may be partly attributable to the recalcitrant nature of the carbon and nitrogen contained in the landfill.

### **8.5 Analysis of $N_2O$ emissions**

$N_2O$  production during the treatment of municipal wastewater (Study 2 – Phase 2) and landfill leachate (Study 3 – Phase 2) were investigated during Study 4.

- The following conclusions were drawn from the evaluation of  $N_2O$  emissions during Study 2:
  - Under optimal denitrification conditions,  $N_2O$  was emitted at a rate of 0.00014 mg  $N_2O/l/min$ .
  - The percentage of influent  $TN_f$  emitted as  $N_2O$  was 0.57%, which compares favourably with the findings of previous literature.
- The following conclusions can be drawn from the evaluation of the  $N_2O$  emission when treating landfill leachate using the ASF-BR:
  - When the highest mass of glucose was added to the influent leachate,  $N_2O$  emissions were 0.003 mg  $N_2O/l/min$ , which was equivalent to 0.10% of the influent  $TN_f$ .
  - On commencement of the aerobic period,  $N_2O$  emissions increased immediately, and then decreased throughout the aerobic period. This could indicate that there was a significant quantity of  $N_2O$  dissolved within the leachate, which was released as the leachate was agitated during the aerobic period.

### **8.6 Potential of the ASF-BR**

The ASF-BR may offer an innovative alternative to conventional wastewater treatment systems. Based on this study, the system could offer a number of key advantages when compared to conventional technologies.

- The ASF-BR employs a passive aeration mechanism, where a vacuum is used to alternate the wastewater between two reactors. As the density of air is lower than that of water, the costs associated with creating a partial vacuum required can be less than those associated with pumping wastewater. Exposing the biofilm to air when the wastewater is in the alternate reactor eliminates the need for mechanical aeration. Thus the system could significantly reduce the costs associated with aeration.
- The treatment of wastewater in the ASF-BR is carried out within two enclosed reactors, which allow all gases produced during the treatment of the wastewater to be controlled. Using the ASF-BR, these gases can be captured and controlled and subsequently treated, eliminating the emission of greenhouse gases and potentially reducing public objections to wastewater treatment facilities.
- The ASF-BR operates using a series of valves and a vacuum pump, all of which are commercially available. Over the course of this research, mechanical equipment proved to be robust, with limited maintenance required.
- The ASF-BR provides a high degree of versatility in the operation of the unit. The duration of each stage of the treatment process can be defined using a PLC. Thus the cycle can be configured to achieve biological carbon, nitrogen and phosphorus removal. The duration of the aerobic and anoxic periods can be extended depending on the strength of the wastewater, and the volume of wastewater treated can be increased or reduced to ensure optimisation of the treatment process.

### **8.7 Recommendations for future research**

A number of areas of further research have been identified based on this study.

- A dynamic mathematical model of the ASF-BR should be developed to aid future design work on the ASF-BR. An effective model would allow: (i) process designers to develop the ASF-BR for a range of influent types, and (ii) designers to determine the effects of varying the influent characteristics and operational conditions on the performance of the ASF-BR.
- A field scale unit should be developed to investigate the viability of the ASF-BR on a larger scale. Such a study would be an important aspect of the continued development and commercialisation of the ASF-BR.

- A detailed study should focus on comparing oxygen transfer rates and aeration costs with those achieved by existing technologies. The pumping cycles could be further investigated to ensure oxygen concentrations are optimised.
- In this study, the ASF-BR was configured to achieve nitrification or a combination of nitrification and denitrification. Future studies should investigate the efficacy of the ASF-BR in achieving biological phosphorus removal.
- The recalcitrant nature of the contaminants in the landfill leachate inhibited the biological treatment of the landfill leachate. Pre-treatment processes that increase the biodegradability of the landfill leachate should be considered.
- Real time monitoring of DO, OPR and pH could be utilised in future studies to increase the efficiency of the ASF-BR by the indicating the end of nitrification or denitrification processes.
- The use of fixed media could be investigated as alternative media to the bioballs. This would eliminate sloughing of the biofilm due to collisions between the moving media. However, when using fixed media, regular inspections would be required to ensure clogging or excess accumulation of biofilm does not occur.

## **8.8 Summary**

This study investigated the performance of an innovative technology – the ASF-BR – when treating three distinctly different wastewaters. The study has shown that the unit could offer a novel, effective and efficient alternative to conventional technologies. The design of the system offers a high degree of operational flexibility, and allows control of gases produced during the biological treatment of wastewater. The system utilises a passive aeration regime that could significantly reduce operational costs when compared to mechanically aerated systems. Further work at a field scale is necessary to demonstrate the technology as a potential commercial offering.

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**GLOSSARY OF TERMS**

<b>AOB</b>	Ammonium oxidising bacteria	<b>NH<sub>4</sub>-N</b>	Ammonium nitrogen
<b>ASP</b>	Activated sludge process	<b>NH<sub>3</sub>-N</b>	Ammonia nitrogen
<b>ASF-BR</b>	Air suction flow biofilm reactor	<b>NO</b>	Nitric oxide
<b>APHA</b>	American Public Health Association	<b>NOB</b>	Nitrite oxidising bacteria
<b>NH<sub>4</sub>-N</b>	Ammonium nitrogen	<b>NO<sub>3</sub>-N</b>	Nitrate nitrogen
<b>BOD<sub>5</sub></b>	5-day biochemical oxygen demand	<b>NO<sub>2</sub>-N</b>	Nitrite nitrogen
<b>C<sub>i</sub></b>	Influent parameter concentration	<b>N<sub>2</sub>O</b>	Nitrous oxide
<b>C<sub>e</sub></b>	Effluent parameter concentration	<b>OLR</b>	Organic loading rate
<b>C<sub>6</sub>H<sub>12</sub>O<sub>6</sub></b>	Glucose	<b>PO<sub>4</sub>-P</b>	Ortho phosphorus
<b>CO<sub>2</sub></b>	Carbon dioxide	<b>PAO</b>	Phosphorus accumulating organism
<b>COD</b>	Chemical oxygen demand	<b>PE</b>	Population equivalent
<b>COD<sub>f</sub></b>	Filtered chemical oxygen demand	<b>PHA</b>	Polyhydroxyalkanoates
<b>DO</b>	Dissolved oxygen	<b>PBA</b>	Poly-β-hydroxybutrates
<b>EC</b>	European Commission	<b>PHB</b>	Poly-β-hydroxybutrates
<b>EEC</b>	European Economic Community	<b>PHV</b>	Poly-β-hydroxyvalents
<b>Eff</b>	Effluent	<b>PLC</b>	Programmable logic controller
<b>EPA</b>	Irish Environmental Protection Agency	<b>RBD</b>	River basin district
<b>EU</b>	European Union	<b>SBR</b>	Sequencing batch reactor
<b>GWP</b>	Global warming potential	<b>SND</b>	Simultaneous nitrification and denitrification
<b>GLY</b>	Glycogen	<b>SS</b>	Suspended solids
<b>HFBR</b>	Horizontal flow biofilm reactor	<b>TIC</b>	Total inorganic carbon
<b>HSWW</b>	High strength synthetic wastewater	<b>TKN</b>	Total Kjeldahl nitrogen
<b>HSWIS</b>	High strength synthetic wastewater intensive study	<b>TN</b>	Total nitrogen
<b>HLR</b>	Hydraulic loading rate	<b>TOC</b>	Total organic carbon
<b>Inf</b>	Influent	<b>TON</b>	Total oxidised nitrogen
<b>IS</b>	Intensive study	<b>TP</b>	Total phosphorus
<b>IPCC</b>	Intergovernmental Panel on Climate Change	<b>TSS</b>	Total suspended solids
<b>LL</b>	Landfill leachate	<b>US EPA</b>	United States Environmental Protection Agency
<b>SCADA</b>	Supervisory control and data acquisition	<b>UWWTD</b>	Urban Wastewater Treatment Directive
<b>LLIS</b>	Landfill leachate intensive study	<b>WFD</b>	Water Framework Directive
<b>MWW</b>	Municipal wastewater		
<b>MWWIS</b>	Municipal wastewater intensive study		
<b>MWTP</b>	Municipal wastewater treatment plant		
<b>MIWTP</b>	Mutton Island wastewater treatment plant		
<b>N<sub>2</sub></b>	Nitrogen gas		