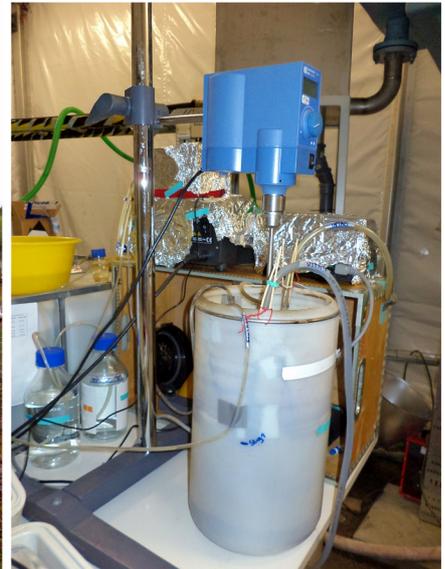




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# **Removal of Nitrogen from Landfill Leachates with MBBR and SBR**

## **A Pilot Study at the Brudaremossen Landfill**

Master's Thesis in the Master's Program Infrastructure and Environmental Engineering

HANNA PORSGAARD  
SOFIA SÖDERSTRÖM



MASTER'S THESIS 2015:65

REMOVAL OF NITROGEN FROM LANDFILL LEACHATES WITH MBBR AND SBR  
A Pilot Study at the Brudaremossen Landfill

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*Division of Water Environment Technology*  
CHALMERS UNIVERSITY OF TECHNOLOGY  
Göteborg, Sweden 2015



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Cover:

Picture of Brudaremossen landfill to the left, Brudaremossen pilot plant 2015 in the middle and one of the lab scale SBR's maintained in this Master's Thesis to the right.

Department of Civil and Environmental Engineering, Göteborg, Sweden, 2015



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### ABSTRACT

Landfill leachates are the effluent produced when rain percolates through waste sustained within a landfill. Leachates are often characterised by high concentrations of toxic metals, carcinogenic substances and ammonium nitrogen  $NH_4-N$ . The landfill Brudaremossen in Gothenburg, Sweden, was operated from 1938 to 1978 and its leachates are currently being channelled to the wastewater treatment plant Gryaab. In 2009, Gryaab acquired a REVAQ certificate, which guarantees a certain quality of the sludge produced in the plant. Due to the negative characteristics of the leachates from Brudaremossen, Gryaab will cease to treat the leachates after year 2018. The consultancy firm Ramböll was assigned to investigate treatment techniques suitable for a permanent treatment facility on site.

The aim of the Master's Thesis was to compare nitrogen removal efficiency for two biological treatment techniques, Moving Bed Biofilm Reactor *MBBR* and Sequencing Batch Reactor *SBR*, and decide which one that is most suitable to apply at the Brudaremossen landfill area. In the beginning of 2015, Ramböll designed a pilot plant on site in which the *MBBR* was analysed. Two lab scale *SBR*'s were constructed in the pilot plant and was operated during 50 days. One *SBR* - Reactor 1 - received leachate water that had been chemically treated and filtered through lamellae, whilst the other - Reactor 2 - received water that had been chemically treated, filtered through lamellae, sand and granulated active carbon filters. This was done in order to assess if toxic compounds in the leachates would inhibit the treatment capacity.

The average removal efficiency of total nitrogen  $N_{tot}$  for the lab scale *SBR*'s were 51 % in Reactor 1 and 76 % in Reactor 2, whereas the *MBBR* had an average removal rate of 97 %. The average removal of  $NH_4-N$  reached 96 % for Reactor 1, 94 % for Reactor 2 and 99.8 % for the *MBBR*. The nitrification rates in Reactor 1 and Reactor 2 were assessed to 14 and 16 mg  $NH_4-N/L \cdot h$  respectively, whilst the denitrification rates were 42 and 39 mg  $NO_3-N/L \cdot h$ . The results show no inhibition due to toxic compounds. The *MBBR* had more even removal rates compared to the lab scale *SBR*'s, which was evaluated to be due to that it had 54 days of acclimatisation before sampling begun, whereas the *SBR*'s had seven days. Based on the results, the *MBBR* was considered most suitable for Brudaremossen. However, an accurate comparison would require a longer trial period to optimise the *SBR* and more advanced laboratory analyses.

Key words: *SBR*; *MBBR*; landfill leachates; ammonium nitrogen; nitrification; denitrification; biological nitrogen removal; Brudaremossen landfill

Kväverening av lakvatten med MBBR och SBR  
En pilotstudie vid avfallsdeponin Brudaremossen

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Engineering* HANNA PORSGAARD

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## SAMMANFATTNING

Lakvatten är ett begrepp på det vatten som produceras då regnvatten filtreras genom avfallsdeponier. Lakvatten karaktäriseras ofta av höga koncentrationer av giftiga metaller, cancerframkallande ämnen och ammoniumkväve  $NH_4-N$ . Avfallsdeponin Brudaremossen i Göteborg var aktiv från 1938 till 1978 och dess lakvatten skickas för tillfället till reningsverket Gryaab. År 2009 ansökte Gryaab om ett REVAQ certifikat som garanterar en viss kvalitet på slammet som produceras på reningsverket. Efter 2018 kommer Gryaab att stoppa mottagning och rening av lakvattnet från Brudaremossen på grund av risken för negativ slampåverkan. Konsultföretaget Ramböll tilldelades uppdraget att undersöka möjliga reningstekniker för ett permanent reningsverk på plats.

Målet med examensarbetet var att jämföra effektivitet av kväverening för två biologiska reningstekniker, rörlig biofilmsreaktor *MBBR* och satsvis biologisk rening *SBR*, och bestämma den mest passande för rening av lakvatten från Brudaremossen. Ramböll installerade en pilotanläggning på plats i början av 2015 inom vilken *MBBR* analyserades. Två *SBR*-tankar i laborationsskala installerades inuti pilotanläggningen och studerades under 50 dagar. En *SBR*-tank – Reaktor 1 – tillfördes lakvatten som hade renats med kemisk fällning, filtrerats genom lameller medan den andra – Reaktor 2 – tilldelades vatten som även hade filtrerats genom sand- och kolfilter. Detta genomfördes för att undersöka om giftiga ämnen i lakvattnet kunde hämma reningseffektiviteten.

Den genomsnittliga reningen för totalt kväve  $N_{tot}$  i de laborationsskaliga *SBR*-tankarna var 51 % i Reaktor 1 och 76 % i Reaktor 2, medan *MBBR*-tanken hade ett motsvarande värde på 97 %. Den genomsnittliga reningen av  $NH_4-N$  uppnådde 96 % i Reaktor 1, 94 % i Reaktor 2 och 99,8 % i *MBBR*-tanken. Nitrifikationshastigheterna i Reaktor 1 och Reaktor 2 utvärderades till 14 respektive 16 mg  $NH_4-N/L \cdot h$  medan denitrifikationshastigheterna var 42 respektive 39 mg  $NO_3-N/L \cdot h$ . Resultatet indikerade på att giftiga ämnen inte hämmade processen. *MBBR*-tanken hade jämnare rening jämfört med *SBR*-tankarna vilket troddes bero på att den acklimatiserades i 54 dagar innan provtagning, jämfört med sju dagar för *SBR*-tankarna. *MBBR*-tekniken ansågs, baserat på resultaten, mest lämplig för Brudaremossen. Dock kräver en mer likvärdigare jämförelse en längre testperiod för optimering av *SBR*-tekniken samt mer avancerade laborationsanalyser.

Nyckelord: *SBR*; *MBBR*; lakvatten; ammoniumkväve, nitrifikation; denitrifikation; biologisk kväverening; Brudaremossen avfallsdeponi

# Table of Contents

ABSTRACT	I
SAMMANFATTNING	II
TABLE OF CONTENTS	III
PREFACE	VII
1 INTRODUCTION	1
1.1 Aim	2
1.2 Research questions	2
1.3 Method	3
1.4 Scope and delimitations	4
2 LANDFILL AND LANDFILL LEACHATES	5
2.1 The characteristics of leachates	5
2.1.1 Nutrients in leachates	5
2.1.2 Organic matter in leachates	6
2.1.3 Additional parameters for leachates	7
2.2 Ageing of landfills	8
2.3 Regulations and maintenance concerning landfills	9
2.3.1 The REVAQ certificate	10
2.3.2 The Natura 2000 network	10
3 BIOLOGICAL PROCESSES FOR REMOVAL OF NITROGEN	11
3.1 The nitrogen cycle	11
3.2 Nitrification and denitrification	12
3.3 Parameters affecting biological processes	13
3.3.1 Parameters affecting nitrification	13
3.3.2 Parameters affecting denitrification	14
3.4 Activated sludge in biological treatment	14
4 BIOLOGICAL TREATMENT TECHNIQUES	17
4.1 Moving Bed Biofilm Reactors <i>MBBR</i>	17
4.2 Sequencing Batch Reactor <i>SBR</i>	18
4.2.1 The SBR cycle phases	20
4.2.2 Cycle analysis of the SBR	21
4.2.3 Operation of the SBR	22
4.3 A comparison between <i>MBBR</i> and <i>SBR</i>	22
5 NATIONAL AND INTERNATIONAL CASES	25
5.1 Summarisation of the literature review	25

5.1.1	SBR removal efficiencies	25
5.1.2	SBR operational parameters	26
5.2	National cases	28
5.2.1	CASE 1 - Filborna landfill, Helsingborg	28
5.2.2	CASE 2 – Spillepeng landfill, Malmö	29
5.2.3	CASE 3 - Norsa landfill, Köping	31
5.3	International cases	32
5.3.1	CASE 4 - Comparison of MBBR and SBR, Munich	32
5.3.2	CASE 5 – Landfill near Wysieka, Poland	33
5.3.3	CASE 6 –Lab scale SBR trial at the University of Florence	33
5.3.4	CASE 7 – Corsa landfill, Spain	34
5.3.5	CASE 8 - Puigpalter landfill, Spain	35
5.4	SBR results applicable for case Brudaremossen	36
6	DESCRIPTION OF CASE STUDY AREA BRUDAREMOSSEN	39
6.1	Leachate quality at Brudaremossen	42
6.2	Potential recipients for the treated leachates	43
6.2.1	The WWTP Gryaab	45
6.2.2	The river Göta älv	45
6.2.3	The stream Säveån	45
6.2.4	Lake Svartjärn & Finngösabäcken	45
6.2.5	The stream Mölndalsån	46
6.2.6	The creek Delsjöbäcken	46
6.3	Pilot plants at Brudaremossen	46
6.3.1	Lamellae sedimentation at Brudaremossen pilot plant 2015	48
6.3.2	MBBR treatment at Brudaremossen pilot plant 2015	49
6.3.3	Sedimentation at Brudaremossen pilot plant 2015	49
6.3.4	Sand and GAC filters at Brudaremossen pilot plant 2015	50
7	EXPERIMENTAL TRIAL WITH LAB SCALE SBR'S	53
7.1	Design of the lab scale SBR's	53
7.1.1	Lab scale SBR construction and operation	54
7.1.2	Stepwise increase of the exchange ratio	55
7.2	Addition of chemicals to the lab scale SBR's	56
7.2.1	Calculation of phosphorus dosage	56
7.2.2	Calculation of methanol dosage	57
7.3	SBR cycle analysis	57
7.4	Lab scale SBR sampling procedure	58
7.4.1	Analysis of the water samples	59
7.4.2	Analysis of the sludge samples	60
7.5	Full scale SBR design	62
8	RESULTS & DISCUSSION	63
8.1	Removal efficiency of the lab scale SBR's and the MBBR	63

8.1.1	Removal of nitrogen	64
8.1.2	Removal of organic matter	75
8.1.3	Removal of metals	79
8.2	Operational parameters for the lab scale SBR's and the MBBR	82
8.2.1	The pH during the experimental trial	82
8.2.2	Dissolved Oxygen during the experimental trial	83
8.2.3	Temperature during the experimental trial	84
8.2.4	Dosage of phosphorus	84
8.2.5	Dosage of methanol	86
8.2.6	Precipitation during the experimental trial	87
8.3	Cycle analysis of the lab scale SBR's	88
8.3.1	Dissolved Oxygen, pH and temperature during the cycle analysis	90
8.3.2	Nitrification and denitrification rates during the cycle analysis	91
8.3.3	Cycle analysis - discussion of the results	92
8.4	Quality of the activated sludge in the lab scale SBR's	94
8.5	Result & Discussion of Full scale SBR design	97
9	CONCLUSION	99
10	REFERENCES	101
11	APPENDICES	109



## Preface

During this Master's Thesis, two lab scale SBR's have been constructed for treating leachates on site at Brudaremassen landfill area. The experimental trial was carried out during the spring in 2015, from the 11<sup>th</sup> of March until the 29<sup>th</sup> of April. The purpose with analysing the lab scale SBR's treatment efficiency was to enable a comparison with an MBBR technique which was evaluated on site by the consultancy firm Ramböll. The project was carried out at the Department of Civil and Environmental Engineering at Chalmers University of Technology, Sweden. All the laboratory analyses were performed at the laboratory of Water Environment Technology at Chalmers.

We would like to start with thanking the water department at Ramböll for letting us be a part of their team, as well as providing us with well needed energy in form of freshly baked bread every morning. A special gratitude is expressed to Måns Lundh and Rita Marques for consulting us during the thesis. Thereafter, we would like to thank our supervisor Britt-Marie Wilén, professor and head of Division Water Environment Technology at Chalmers, for assisting us with her extensive knowledge of biological treatments. We would also like to acknowledge Ann-Margret Strömvall, associate professor at Chalmers, for her interest in our work and for her valuable feedback. The laboratory work would have been far more challenging without the much needed help from the staff at Chalmers; Mona Pålsson, Enikö Barbara Szabo, Soroush Saheb Alam, Jesper Knutsson and Sebastien Rauch.

We would also like to thank our stately opponents, Joachim Jensen and Fredrik Svensson, for helping us with improving our work and for the nice coffee dates. We are furthermore grateful for the pleasant conversations and helpful inputs from our fellow Master's Thesis companions Mathias Andersson and Hanna Sätterskog.

Last but not least, we would like to send our gratitude to Lars-Ove Sörman whose practical knowledge and cheerfulness were very valuable throughout the study.

Gothenburg, June 2015

Hanna Porsgaard & Sofia Söderström

## List of Abbreviations

<i>Anammox</i>	Anaerobic ammonium oxidation bacteria
<i>F/M</i>	Food to Microorganism ratio
<i>GAC</i>	Granular Activated Carbon
<i>IC analysis</i>	Ion Chromatograph analysis
<i>ICP-MS</i>	Inductively Coupled Plasma Mass Spectrometry
<i>MBBR</i>	Moving Bed Biofilm Reactor
<i>RPM</i>	Revolutions Per Minute
<i>SBR</i>	Sequencing Batch Reactor
<i>SWWA</i>	Swedish Water & Wastewater Association
<i>WWTP</i>	Wastewater Treatment Plant

<i>BOD</i>	Biochemical Oxygen Demand
<i>COD</i>	Chemical Oxygen Demand
<i>DO</i>	Dissolved Oxygen
<i>DOC</i>	Dissolved Organic Content
<i>IC</i>	Inorganic carbon
<i>MLSS</i>	Mixed Liquor Suspended Solids
<i>MLVSS</i>	Mixed Liquor Volatile Suspended Solids
<i>mV</i>	Millivolts
<i>N<sub>tot</sub></i>	Total nitrogen
<i>ORP</i>	Oxygen Reduction Potential
<i>PAH</i>	Polycyclic Aromatic Hydrocarbon
<i>PCB</i>	Polychlorinated Biphenyl
<i>PHC</i>	Petroleum Hydrocarbon
<i>P<sub>tot</sub></i>	Total phosphorus
<i>SRT</i>	Sludge Retention Time
<i>SV</i>	Volume of settled solids
<i>SVI</i>	Sludge Volume Index
<i>TKN</i>	Total Kjeldahl Nitrogen
<i>TOC</i>	Total Organic Carbon
<i>TSS</i>	Total Suspended Solids
<i>VSS</i>	Volatile Suspended Solids
<i>HRT</i>	Hydraulic Retention Time

<i>Ag</i>	Silver
<i>Cd</i>	Cadmium
<i>Cl</i>	Chloride
<i>Cu</i>	Copper
<i>Cr</i>	Chromium
<i>Fe</i>	Iron
<i>Hg</i>	Mercury
<i>Ni</i>	Nickel
<i>Pb</i>	Lead
<i>Sn</i>	Tin
<i>V</i>	Vanadium
<i>Zn</i>	Zink

<i>CH<sub>4</sub></i>	Methane
-----------------------	---------

$CO_2$	Carbon dioxide
$H^+$	Hydrogen
$H_2$	Hydrogen gas
$HNO_2$	Nitrous acid
$KH_2PO_4$	Monopotassium phosphate
$N_2$	Nitrogen gas
$NH_3$	Ammonia
$NH_3-N$	Ammoniacal nitrogen
$NH_4^+$	Ammonium ion
$NH_4-N$	Ammonium nitrogen
$NO$	Nitric oxide
$NO_2^-$	Nitrite
$NO_2-N$	Nitrite nitrogen
$NO_3^-$	Nitrate
$NO_3-N$	Nitrate nitrogen
$O_2$	Oxygen
$OH$	Hydroxide ions



# 1 Introduction

Landfilling has been and is still a common method to dispose of waste all around the world (Renou, et al., 2008). In 2009, there were 157 active landfills in Sweden of which 28 allowed disposal of hazardous waste. The number of landfills in Sweden has however decreased since 2001 due to the Landfill Directive 1999/31/E introduced by the European Union *EU* (Swedish EPA, 2014). The aim of the directive is to prevent negative environmental impacts from landfills (European Commission, 2015).

The existing landfills have created a threat to the environment, humans and groundwater in form of contaminated landfill leachates (Swedish EPA, 2008). Leachate is formed when rain percolates through the waste in a landfill (Mor, et al., 2006). This effluent contains numerous metals, high concentrations of ammonium nitrogen  $NH_4-N$ , carcinogenic substances and other organic components that can harm the ecosystems in lakes or streams and contaminate the groundwater (Klinck & Stuart, 1999). Nitrogen is one of the problematic substances that can be found in landfill leachates and generally require some form of biological treatment (Wärjerstam, 2009). It can exist in many different chemical forms but especially ammonium ion  $NH_4^+$  can have a toxic effect on aquatic organisms and cause oxygen depletion if untreated (Swedish EPA, 2008). In order to reduce the risk of contamination, the water from landfills needs to be treated either on site or in local treatment plants before discharged into a recipient. Treatment of leachates demands complex techniques due to its undesirable characteristics, but there are today many different treatment methods that could be applied (Wiszniewski, et al., 2006).

Brudaremassen is a landfill that has been used for municipal waste from 1938 to 1978 in Gothenburg, Sweden (Fersters, 2003). The leachates from this landfill is characterised by high levels of organic material, iron *Fe* and extremely high concentrations of nitrogen in the form of  $NH_4-N$  (Ramböll, 2015 B). According to the consultancy firm Ramböll Sweden, Brudaremassen contains large volumes of organic household waste (Ramböll, 2015 A). Household waste is a source of nitrogen and therefore can explain the high concentrations of  $NH_4-N$  in the deposit. Production of methane gas  $CH_4$  has also been detected at Brudaremassen, which also confirms that large volumes of organic substances are present within the landfill (Ramböll, 2015 B).

Since 1962, the leachates from Brudaremassen have been channelled and further treated at the Wastewater Treatment Plant *WWTP* Gryaab in Gothenburg (Johansson & Johansson, 1973). However, landfill leachates can contain toxic metals and persistent organic pollutants which can affect the quality of the sludge produced in the treatment plant. In 2009, the *WWTP* Gryaab acquired a REVAQ certificate, which implies that the produced sludge at the plant can be used as fertilizers without any risk of spreading hazardous elements (SP Sveriges Tekniska Forskningsinstitut, 2011). Gryaab therefore informed the Gothenburg municipal executive board, City of Gothenburg, that they would not be able to receive and treat the leachates from Brudaremassen after year 2018 if the quality of the leachates does not improve (Moberg, 2013).

Due to these circumstances, the City of Gothenburg were obliged to find a suitable treatment for the leachates at Brudaremassen in order to either; treat the most hazardous elements on site and be able to continue sending it to Gryaab or; treat the leachates completely and discharge it into suitable water recipient without risking the safety of the environment (Moberg, 2013). A division at the City of Gothenburg,

*Kretslopp och vatten*, responsible for the landfill Brudaremossen, hired the consultancy firm Ramböll to investigate the current conditions in possible recipients and to analyse available treatment techniques that would fulfil the requirements for local treatment of the leachates (Ramböll, 2015 A). Ramböll decided to analyse the possible treatment techniques on site in a pilot plant for further evaluation.

Biological treatment in the pilot plant was implemented in order to reduce total nitrogen  $N_{tot}$  in the leachates. The concentrations had to be significantly reduced in order to be discharged into surrounding watercourses (Ramböll, 2015 B). Two different biological treatment techniques were evaluated for this purpose; Moving Bed Biofilm Reactor *MBBR* and Sequencing Batch Reactor *SBR*. These two techniques has shown to be feasible for leachate treatment with high removal rates for nitrogen (SYSAV AB, 2008). Ramböll installed a MBBR on site as a treatment step in the pilot plant, whilst Chalmers University of Technology was asked to evaluate and construct a lab scale SBR as a part of a Master's Thesis (Ramböll, 2015 C). The aim of the Master's Thesis, presented in this report, was to evaluate and compare the MBBR and SBR concerning nitrogen removal efficiency. An additional Master's Thesis was conducted parallell with this thesis concerning the overall treatment removal efficiency for the pilot plant at Brudaremosen, regarding organic pollutants and toxic metals. Since the biological treatment step might be affected by organic pollutants in the leachates - which can inhibit the nitrogen removal process - their results was taken into consideration.

According to Qarani Aziz et al (2013), there is a gap in the litterature concerning the SBR operation parameters such as cycle time, duration of chosen phases, aeration rate etcetera. In order to optimise the SBR treatment performance, it is vital to study operational parameters (Qarani Aziz, et al., 2013). This project can, by studying the pilot plant at Brudaremossen, contribute with further knowledge regarding how different parameters affect the SBR treatment efficiency concerning leachates.

## 1.1 Aim

The aim of this Master's Thesis was to compare nitrogen removal efficiency for two different biological treatment techniques, MBBR and SBR, and decide which one that is most suitable to apply at Brudaremossen landfill area.

## 1.2 Research questions

Three research questions (RQ) were compiled in order to reach the aim.

**RQ1:** *How should the SBR technique be designed, optimised and analysed for the case Brudaremossen?*

**RQ2:** *How will operational parameters affect the treatment techniques removal efficiency?*

**RQ3:** *How should an accurate comparison between the two biological treatment techniques – MBBR and SBR – be performed?*

### 1.3 Method

A flow chart illustrating the chronological order of processes carried out during the Master's Thesis is presented in Figure 1.1. The first step was to formulate the overall aim of the study. Thereafter, three research questions were formed in order to strive towards achieving the aim.

The initial processes carried out during the thesis are summarised in the marked area defined as A. in the flow chart of processes, see Figure 1.1. During January and February a literature study was carried out regarding the case study area Brudaremissen. A field visit took place in January to get an overview of the current situation and the location of the new pilot plant designed by Ramböll. During the same period, theoretical background needed for the project was gathered. Also, a literature review was performed in order to assemble knowledge from earlier similar national and international cases with focus on lab scale SBR's.

The gained knowledge from the initial processes was used to plan and construct a lab scale SBR processes in collaboration with Chalmers University of Technology. After a trial run at Chalmers, the treatment was installed at Brudaremissen in the beginning of March. A cycle analysis was performed once the lab scale SBR's had been adapted to the conditions on site. Samples were taken from the lab scale SBR process two times a week during March and April and subsequently analysed at the laboratory at Chalmers. Results from the sample analysis of the MBBR treatment were provided by the consultancy firm Ramböll. This second stage in the study is summarised in the marked area defined as B. in the flow chart of processes, see Figure 1.1.

The collected results from the two biological treatment techniques were compared and evaluated. Also, a full scale SBR treatment facility was designed. Key critical operational parameters for the two techniques were determined and taken into consideration when the final recommendations were concluded. The recommendations were presented to the City of Gothenburg regarding whether it would be possible to remove enough nitrogen and which biological treatment that was considered the most suitable for the case Brudaremissen. These final steps, which resulted in a licensed Master's Thesis, are presented in the lower part of the flow chart of processes, see Figure 1.1.

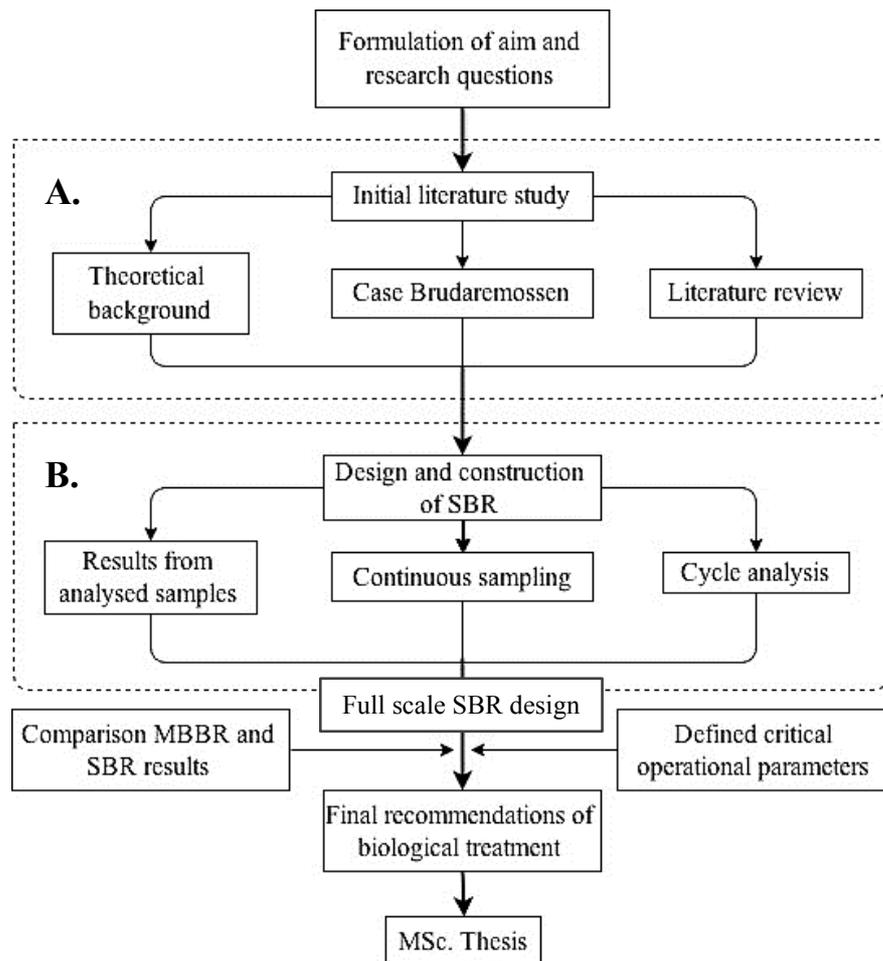


Figure 1.1 Flow chart of processes carried out in order to achieve the aim of the thesis.

## 1.4 Scope and delimitations

The project mainly focused on the biological treatment step for removal of  $N_{\text{tot}}$  and  $NH_4\text{-N}$  in the pilot plant at Brudaremassen. In the beginning of the project it was decided to only focus on the two biological treatment techniques; MBBR and SBR. This delimitation was based on a previous study, performed by Ramböll, which analysed suitable treatment techniques at Brudaremassen (Ramböll, 2015 C).

The cases included in the literature review were chosen based on their similarities to the study at Brudaremassen. Thus, the focus was on finding cases which involved a lab scale SBR that treated an effluent with similar characteristics as the leachates at Brudaremassen. Afterwards, the number of cases was narrowed down to eight, three national and five international.

Sampling and analysis of the lab scale SBR's treatment capacity were only carried out during winter and spring in 2015, due to a limited time frame. In order to minimise the number of samples, some results concerning the quality of the incoming leachates to the lab scale SBR's were received from the analysis performed on the leachates in the pilot plant by Ramböll. The analysis performed in the laboratory at Chalmers University of Technology, on water and sludge samples, were chosen based on available equipment and if the results would provide information about the lab scale SBR's treatment capacity.

## 2 Landfill and landfill leachates

Landfills are a result of an old waste disposal technique in society where mostly solid waste is disposed of in quantity, forming piles of waste, on specified sites (Schmoll, et al., 2006; Abbas, et al., 2009). The categories of waste that can end up in landfills are everything from household or construction waste to non-hazardous or hazardous industrial waste. A modern landfill technique, known as sanitary landfilling, is applied at carefully selected sites where the area is designed to provide isolation of the waste from the surrounding environment (UNEP, 2005). This technique was developed in order to prohibit the negative effects that landfills can have on the surrounding environment, mainly caused by landfill leachates.

According to S. Renou in the journal article *Landfill leachate treatment: Review and opportunity* (2008) landfill leachates can be defined as “the aqueous effluent generated as a consequence of rainwater percolation through wastes, biochemical processes in waste’s cells and the inherent water content of wastes themselves.”

The characteristics of leachates are affected by the waste in the current landfill and often contain a large amount of organic matter, ammonium nitrogen  $NH_4-N$ , toxic metals, chlorinated organic and inorganic salts (Renou, et al., 2008). Other parameters used to characterise leachates are pH, electrical conductivity, alkalinity and content of solids (Swedish EPA, 2008). Some of the mentioned parameters are described further in Section 2.1. Uncontrolled leaching from landfills can cause pollution of surrounding surface water or groundwater (Kjeldsen, et al., 2002). Many factors influence the volumetric amount of leachate that is leaching from a landfill every year. The amount of precipitation or evaporation will decide how much water that enters the landfill. More compacted waste will reduce the permeability of the landfill. This will impact the residence time of leachates in deposits, which has shown to vary from a few days to several years (Schmoll, et al., 2006; Abbas, et al., 2009). The age of the landfill will affect both element concentrations in the leachates and the landfills possibility to absorb percolated water.

### 2.1 The characteristics of leachates

Physical, chemical and biological elements in an effluent can provide information about its characteristics and quality. Key parameters used for leachate treatment are presented in this section.

#### 2.1.1 Nutrients in leachates

Nitrogen and phosphorous are the two main nutrients vital for the growth of microorganisms, plants and animals (Tchobanoglous, et al., 2003). Nevertheless, if too high concentrations of these substances enter a hydraulic environment it can cause eutrophication, excess growth of algae, disruption in the current ecosystem and finally oxygen depletion (SwAM, 2014).

#### NITROGEN

Nitrogen is a cornerstone in the synthesis of protein and therefore essential for life. Sources of nitrogen are mainly atmospheric nitrogen, sodium nitrate and the nitrogenous compounds in living organisms. Nitrogen can exist in many forms and its chemistry is complex due to its many levels of oxidation state. The most important nitrogenous fractions concerned in water treatment are organic nitrogen, ammonia

$NH_3$ , ammonium  $NH_4^+$ , nitrite  $NO_2^-$  and nitrate  $NO_3^-$ . Total nitrogen  $N_{tot}$  is the sum of the prior mentioned nitrogenous fractions, expressed as their nitrogen content. Total Kjeldahl Nitrogen  $TKN$  is also a common parameter used in this field, which is the sum of the amount of organic nitrogen, ammoniacal nitrogen  $NH_3-N$  and ammonium nitrogen  $NH_4-N$ . The organic nitrogen consists of amino acids, amino sugars and proteins (Tchobanoglous, et al., 2003).

When analysing results of nitrogen concentrations, it is important to be aware of the different methods of referring to various types of concentrations. For example, if the results are presented as  $NH_4^+$ , it refers to the concentration of ammonium molecules that is present in the sample. If the results are presented as  $NH_4-N$  instead, it refers to the concentration of nitrogen within the ammonium molecule. The latter example is often used in scientific reports to enable an easy comparison between nitrogen concentrations bound to different molecules (California Environmental Protection Agency, 2015).

## PHOSPHORUS

Phosphorous is also an important nutrient for biological organisms and can be found in high concentrations in the soil (Börling, et al., 1999). It is less abundant compared to carbon and nitrogen; hence it is often the limiting factor for growth in lakes. However, increased levels of phosphorous from domestic and industrial discharge can cause extensive growth of algae in lakes and watercourses. Phosphorus exists in both organic and inorganic form which can be either dissolved or suspended in water. Total phosphorus  $P_{tot}$  includes all forms of phosphorus in a water sample (U.S. EPA, 2012).

### 2.1.2 Organic matter in leachates

Leachates contain high concentration of organic matter which can have a negative impact on ecosystems and the environment if the leachates reach nearby watercourses. This is since the decomposition of organic matter is very oxygen-consuming, causing oxygen depletion in aquatic environments (Swedish EPA, 2008). Measurement of organic matter is often divided into two different groups depending on concentration of organics. Large concentrations, greater than 1 mg/L, can be analysed by measuring the Biochemical Oxygen Demand  $BOD$ , Chemical Oxygen Demand  $COD$  and Total Organic Carbon  $TOC$  in the sample (Tchobanoglous, et al., 2003).

#### BIOCHEMICAL OXYGEN DEMAND $BOD$

The  $BOD$  is, according to Tchobanoglous et al. (2003), the most commonly used parameter for measuring organic pollution in an effluent. The parameter implies the amount of Dissolved Oxygen  $DO$  that is consumed by microorganisms in the process during which organic matter is biologically oxidised. Hence, it is a measure of how much degradable organic matter that is available in an effluent (Swedish EPA, 2008). It can be analysed by measuring the oxygen  $O_2$  before and after letting a sample rest in darkness and at a constant temperature of 21°C for a number of days. Five or seven days are commonly used, resulting in a concentration of  $BOD_5$  or  $BOD_7$ . The results can thereafter be used to determine the concentration of  $O_2$  that is needed to stabilise the organic matter, regulate the design of the treatment process and analyse the treatment removal rate. It can also be used when deciding a suitable recipient and discharge point after treatment of the water (Tchobanoglous, et al., 2003).

## CHEMICAL OXYGEN DEMAND *COD*

The COD is a measure of the O<sub>2</sub> required to chemically oxidise organic matter in an effluent, which is a useful parameter when monitoring and designing a treatment plant. The COD is often higher than the BOD, since addition of chemicals makes it possible to oxidise more organic substances compared to biologically (Tchobanoglous, et al., 2003). The measuring procedure for COD does however include manufacturing of hazardous wastes such as mercury *Hg* and hexavalent chromium. In a study by Dubber & Gray (2010), 11 Waste Water Treatment Plants *WWTP*'s were studied in order to find a relationship between TOC and COD. It was found that TOC can be a reliable replacement for COD in both influent and effluent wastewaters according to equation (1) and equation (2) respectively (Dubber & Gray, 2010).

$$COD(\text{influent}) = 49.2 + 3 \cdot TOC \quad (1)$$

$$COD(\text{effluent}) = 7.25 + 2.99 \cdot TOC \quad (2)$$

According to Tchobanoglous et al, (2003) a COD concentrations in wastewater below 250 mg/L is considered to be low, whilst concentrations above 800 mg/L is considered high.

## TOTAL ORGANIC CARBON *TOC*

The TOC is the amount of both dissolved and undissolved organic matter in a water or sediment sample (Swedish EPA, 2008). According to Tchobanoglous et al, (2003), a TOC concentration in wastewater below 80 mg/L can be considered as low, whilst a concentration above 260 mg/L is considered high. The TOC concentration can be analysed by performing a TOC test in which heat, O<sub>2</sub>, ultraviolet radiation and chemical oxidants are used to convert organic carbon into carbon dioxide CO<sub>2</sub> which thereafter can be measured. Similar to the BOD/COD ratio, there is a ratio between BOD and TOC which can be used to characterise an untreated effluent.

## DISSOLVED ORGANIC CARBON *DOC*

A concentration of DOC represents the dissolved organic carbon in a sample. DOC can be analysed by filtration of the sample through a filter with the filter size 0.45 µm (Swedish EPA, 2008).

### 2.1.3 Additional parameters for leachates

An effluents electrical conductivity, alkalinity and contents of solids can in addition to pH and temperature provide necessary information about the analysed leachates. These parameters can vary a lot for leachates, depending on the characteristics of the landfill. According to a study performed on 11 Swedish deposits it was found that the electrical conductivity had a median value of 913 mS/m, the content of solids varied from 9 to 210 mg/L and the pH ranged from 6.4 to 8.5 (Swedish EPA, 2008).

## ELECTRICAL CONDUCTIVITY

Electrical conductivity indicates the amount of dissolved ions in the effluent and thereby the ability to conduct electricity. The conductivity for leachates based on a study performed on 11 Swedish deposits, ranged from 430 to 2 700 mS/m (Swedish EPA, 2008). If the concentration of ion increases in an effluent, this will increase the

electrical conductivity as well since the electrical current is transported by ions (Tchobanoglous, et al., 2003).

#### ALKALINITY

Alkalinity defines a waters ability to resist acidification and should be higher than 0.10 mekv/L to indicate a good buffering capacity. The concentration of bicarbonate and carbonate ions in a sample will determine the alkalinity since they react with hydrogen  $H^+$  and hydroxide ions  $OH^-$  which neutralises the water. It is important to be aware of the alkalinity in waters since it represents the capability to resist sudden changes in pH which will protect the aquatic life (Swedish EPA, 2008).

#### CONTENT OF SOLIDS

Content of solids is an important physical parameter to analyse in an effluent. If too high concentrations of suspended solids are released into watercourses, it can cause sludge deposits and anaerobic conditions (Tchobanoglous, et al., 2003). Suspended solids are also potential distributors of bacteria and substances that can bio accumulate to the solid matter (Swedish EPA, 2008).

## 2.2 Ageing of landfills

The constituents of leachates have shown to be affected by the age of the waste. This is due to the altered rate of decomposition that changes depending on the number of years that a substance has been placed on a landfill (Kang, et al., 2002). Scientists argue that the landfill will undergo at least four stages of decomposition; aerobic, acid, initial methanogenic and stable methanogenic, see Figure 2.1 (Kjeldsen, et al., 2002). The aerobic stage can also be called the initial stage, where the organic substance in the waste is decomposed (Hoyer & Persson, 2007). In the acid stage, pH-levels drops and concentrations of e.g. toxic metals and chloride  $Cl^-$  rise due to precipitation. The decomposition causes  $O_2$  levels to drop and the conditions will become anaerobic. Also, the concentrations of hydrogen gas  $H_2$  and  $CO_2$  will increase during the acid stage. Thereafter the initial and stable methanogenic stages will occur (Kjeldsen, et al., 2002). Due to  $CO_2$  and methane  $CH_4$  produced in this stage, pH will increase to 6.8–8 and the ratio between BOD and COD will become low (Hoyer & Persson, 2007).

It is uncertain what will happen to a landfill after these stages, since decomposition of waste can still occur after hundreds of years and well-monitored landfills are often less than 30 years old. However, scientists believe that the later stages of the landfill will become aerobic due to minor rates of decomposition e.g. microbial oxygen depletion that will be lower than the  $O_2$  diffusion into the landfill. They also believe that the amount of nitrogen gas  $N_2$  will increase with age (Kjeldsen, et al., 2002). It is important to understand that several stages can occur at the same time within a landfill (Renou, et al., 2008).

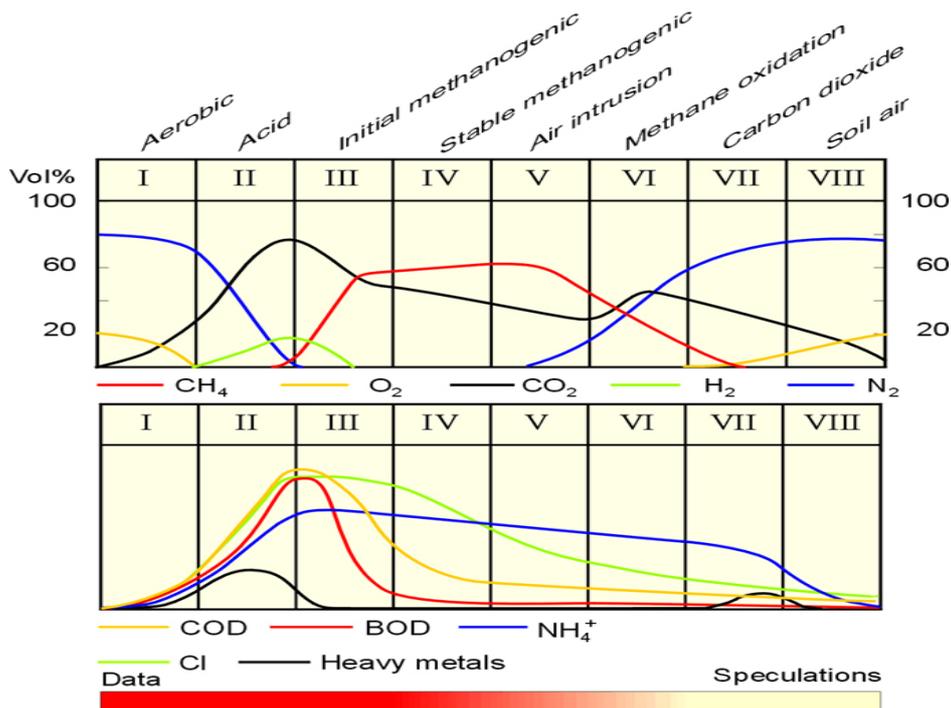
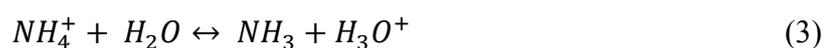


Figure 2.1 Phases of decomposition that is assumed for a landfill (Kjeldsen, et al., 2002) [permitted by author].

The concentrations of different forms of nitrogen in leachates are also affected by the age of the landfill. High levels of nitrogen in the leachates are often due to large amount of household waste that is contained within the landfill. The waste consists of biodegradable substrate with nitrogenous fractions which are converted into  $\text{NH}_4^+$  during hydrolysis and fermentation (Kulikowska, 2012). A higher pH shifts the equilibrium, see equation (3), between  $\text{NH}_4^+$  and  $\text{NH}_3$ , towards  $\text{NH}_3$ . As can be seen in Figure 2.1, the level of  $\text{NH}_4^+$  is decreasing when a landfill ages, confirming the previous stated change in equilibrium.



At pH 8, it can be assumed that the amount of  $\text{NH}_3$  is approximately ten percent of the amount of  $\text{NH}_4^+$  (IVL Svenska Miljöinstitutet AB, 2000).

### 2.3 Regulations and maintenance concerning landfills

Many improvements regarding deposition and recycling of waste have been implemented in Sweden during the last decade, due to sharper introduced regulations. The European Union *EU* has implemented stricter demands concerning deposition of waste and regulations have been introduced regarding goals for enhanced recycling, taxes on still operating deposits and prohibition for disposal of certain wastes (Swedish EPA, 2008).

In 2008, leachates were distributed to municipal WWTP's from approximately 100 landfill facilities in Sweden. About 140 were at the same time applying some form of local treatment of the leachates before discharge to nearby watercourse or the municipal WWTP (Swedish EPA, 2008). The REVAQ certificate was implemented during 2008 in Sweden, which can be issued to a WWTP if it fulfils certain

requirements. The aim with the certificate is to improve quality of the sludge produced in WWTPs, and in that way be able to bring back valuable nutrients in the sludge to the agricultural landscape (Svenskt Vatten, 2015). Consequently, many WWTPs have chosen to disconnect local landfills from their system, since the leachates reduces the quality of the produced sludge (Svenskt Vatten, 2013). The REVAQ certificate is further described in Section 2.3.1.

In many cases when REVAQ has been implemented this implies the need for local treatment of the leachates. The negative characteristics of leachates can however complicate how the treatment process should be designed and often limits the discharge possibilities. One option can be to continue to channel the water to a municipal WWTP after sufficient treatment of leachates. Another alternative is to discharge the water into a nearby recipient, which demands a higher level of treatment (Swedish EPA, 2008). If the second option is chosen, local environmental conditions and regulations such as Natura 2000 sites and protected areas must be taken into consideration and be investigated. The Natura 2000 network is further explained in Section 2.3.2.

### **2.3.1 The REVAQ certificate**

Phosphorus is an essential nutrient source for agricultural industries and it is usually extracted from the non-renewable resource, phosphate rock. Scientists believe that the future reserve of phosphate rock on earth, considering economical and technical aspects, is uncertain with the current production rates and uncontrolled management (Schröder, et al., 2010; White & Cordell, 2011). In order to solve this global problem, a more sustainable approach must be driven. One approach is to recycle the phosphorus in the sludge from WWTPs. Phosphorus in wastewater originates mostly from households, such as from water used for personal hygiene or flushing the toilet (Lusk, et al., 2011)

The reason for the upcoming of the REVAQ certificate is to enable sludge to be used as fertilizers in agriculture in order to recycle phosphorus (Svenskt Vatten, 2015). Sludge from WWTPs that have acquired a REVAQ certificate can be used as fertilizers in agriculture since the certificate ensures that there are limited hazardous elements in the sludge. The company that authorises applications of REVAQ certificates is the SP Technical Research Institute of Sweden according to the rules of certification decided by The Swedish Water & Wastewater Association *SWWA* (SP Sveriges Tekniska Forskningsinstitut, 2015).

### **2.3.2 The Natura 2000 network**

The Natura 2000 is a network of protected areas created by the EU with the aim to protect and preserve valuable habitat and endangered species in a long-term perspective (European Commission, 2015). Each state member in EU is obliged to strive towards creating these protected areas according to two directives: the Habitats Directive and the Birds Directive. The two directives state a selection of species and types of habitats - based on scientific facts - which presence should result in a protected area. Once a protected area is assigned the EU state should implement measures for preservation and maintenance of that area and also prevent human activities which could endanger the site. The Natura 2000 network was implemented the first of July in 2001 in the Swedish law and thereby became a national interest (Swedish EPA, 2009).

### 3 Biological processes for removal of nitrogen

As mentioned in Chapter 2, Section 2.1.1, nitrogen is vital for biological growth. But some forms of nitrogen, such as the ammonium ion  $NH_4^+$ , can be harmful to aquatic life in too high concentrations (SwAM, 2014; Tchobanoglous, et al., 2003). It is therefore desirable to reduce the concentration of nitrogen in an effluent before it is discharged into a recipient, which is often performed by biological treatment techniques.

This chapter provides an overview of the different chemical forms of nitrogen and the naturally occurring processes of nitrification, denitrification and Anammox. It also describes how the biological processes can be implemented by using activated sludge in water treatment and related characteristic parameters. The theory in this chapter aims to give a basic understanding of the mechanisms involved in the biological treatment techniques presented in Chapter 4.

#### 3.1 The nitrogen cycle

Nitrogen is transmitted from the atmosphere to the biosphere in a continuous cycle, called the nitrogen cycle, see Figure 3.1. Nitrogen gas  $N_2$  is the main component in the atmosphere; however its chemical form makes it unavailable for plants and organisms. During the process of fixation,  $N_2$  is biologically fixed by soil bacteria which transform  $N_2$  to ammonia  $NH_3$ , which can be absorbed by the roots of plants and form organic components. This is also the case for inorganic nitrate  $NO_3^-$  in the soil (Allaby & Park, 2013). When the organisms and plants are decomposed, the organic nitrogen is converted into the ammonium ion  $NH_4^+$  by certain bacteria during the process of ammonification. Thereafter  $NH_4^+$  is transformed into nitrite  $NO_2^-$  and later  $NO_3^-$  during the biological process of nitrification. Both  $NO_3^-$  and  $NH_4^+$  can be utilised by plants, whilst animals have to digest plants in order to provide the necessary amount of nitrogen. The cycle is completed by a certain type of bacteria, called denitrifying bacteria, which converts  $NO_3^-$  to  $N_2$  and to some extent nitrous oxide  $N_2O$  which are released to the atmosphere (Bolin, et al., 2015).

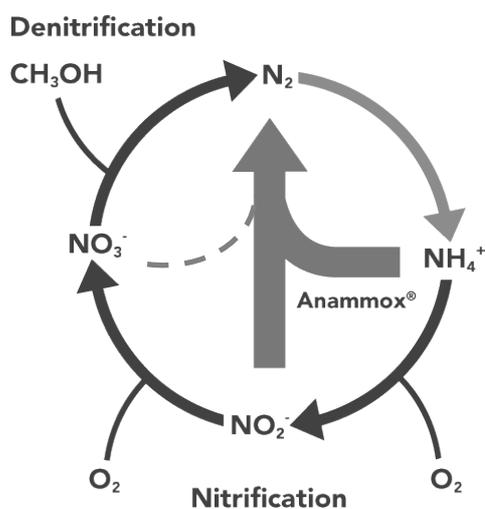


Figure 3.1 Illustration of the nitrogen cycle including the Anammox process (Paques, 2015)[permitted by author].

The nitrogen cycle illustrates the processes involved in conventional biological treatment, where  $\text{NH}_4^+$  is removed by naturally converting it to  $\text{N}_2$ . However, during recent decades anaerobic ammonium oxidation bacteria *Anammox* have been discovered. These specific bacteria have an ability to use  $\text{NH}_4^+$ ,  $\text{NO}_2^-$  and  $\text{NO}_3^-$  to some extent to form  $\text{N}_2$ , see Figure 3.1 and equation (4) (Gijs, 2008).

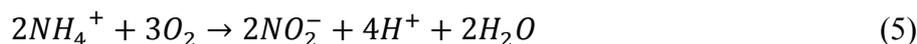


This shortcut in the nitrogen cycle is called the Anammox process (Gijs, 2008). It has been observed that 50 % of the nitrogen conversion in the oceans is due to the Anammox process. The Anammox bacteria use carbon dioxide  $\text{CO}_2$  as carbon source; hence they do not require organic carbon, which is the case for nitrifying bacteria. The advantages with the Anammox process compared to conventional biological treatment are that there is no need to create an oxygen  $\text{O}_2$  rich environment, hence energy consumption decreases, and no external carbon source is required. However, there are still some drawbacks with the Anammox process which needs to be solved. The process of Anammox bacteria is most optimal at high temperatures, approximately  $35^\circ\text{C}$ , which limits the process's applicability (Kartal, et al., 2010).

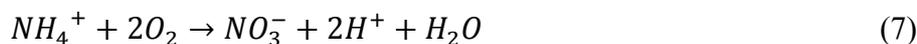
## 3.2 Nitrification and denitrification

The naturally occurring biological processes called nitrification and denitrification are both a part of the nitrogen cycle, as explained in previous section. These processes lead to the conversions between nitrogen fractions performed by different bacteria (Tchobanoglous, et al., 2003).

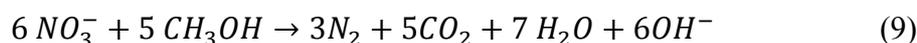
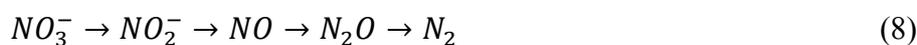
Nitrification is a two-step process by autotrophic bacteria where  $\text{NH}_4^+$  is oxidised to  $\text{NO}_2^-$ , see equation (5), and  $\text{NO}_2^-$  is further on oxidised to  $\text{NO}_3^-$ , see equation (6). The two oxidation steps are performed by different bacteria, such as *Nitrosomonas* and *Nitrobacter*. Addition of  $\text{O}_2$ , or an oxic environment, is needed in order for the bacteria to thrive and nitrification to occur (Tchobanoglous, et al., 2003).



Total oxidation reaction for nitrification is shown in equation (7).



Denitrification is a process by heterotrophic bacteria, such as *Pseudomonas* which are the most common denitrifying bacteria. During the denitrification process,  $\text{NO}_3^-$  is reduced by several steps to  $\text{N}_2$ , see equation (8). Denitrification can occur in anoxic, e.g. oxygen absent, environments since the process only relies on addition of an external carbon source and presence of  $\text{NO}_3^-$ . Methanol  $\text{CH}_3\text{OH}$  is a commonly used carbon source for denitrification, see equation (9) (Tchobanoglous, et al., 2003).



When a carbon source, such as methanol, is added to an anoxic environment Chemical Oxygen Demand *COD* is formed, which is necessary to achieve denitrification. Theoretically, 2.86 g *COD* is required during the denitrification process in order to convert one gram of nitrate nitrogen  $NO_3-N$  into  $N_2$  (Leslie Grady, et al., 2011). The *COD* will however also be utilised in order to oxidise the  $O_2$  that enters the denitrification process, since a zero level of  $O_2$  during the beginning of denitrification is difficult to achieve. A larger required amount of *COD* is therefore almost always assumed, often an amount between 3.5–5 g *COD*/ g N is used instead (Ekenberg, 2007).

The nitrifying and denitrifying organisms are together known as the active part of the biomass in a biological water treatment process since they actively take part in treating the water. Biomass also consists of slowly-degradable products that are formed during biomass decay, often called inert residue (Von Sperling, 2007). Biomass in the field of wastewater treatment can be considered to have a molecular form of  $C_{60}H_{87}O_{23}N_{12}P$  (Leslie Grady, et al., 2011). This illustrates that biomass consists of a small amount of phosphorus, P, which also implies that phosphorus is needed in order for biomass to form. The required dosage is however quite small, 7–18 grams of phosphorus per kilogram *COD*. A more common way of describing biomass in molecular form is instead as  $C_5H_7O_2N$  (Henze, et al., 2002).

### 3.3 Parameters affecting biological processes

The bacteria performing nitrification and denitrification are sensitive and their ability to grow will affect the performance of the processes during treatment (Henze, et al., 2002). Operational parameters that can affect the efficiency of biological wastewater techniques are explained in following sections.

#### 3.3.1 Parameters affecting nitrification

The nitrifying bacteria are sensitive to toxic compounds and metals such as silver *Ag*, mercury *Hg*, nickel *Ni*, chromium *Cr*, copper *Cu* and zinc *Zn* in the effluent. The concentration of toxic metals should be below 5 mg/L in order to achieve optimum nitrification (Bitton, 2005). The nitrification process is also sensitive to high concentrations of *COD*, since this will cause growth of other bacteria that can transcend the nitrifying ones (Morling, 2010). The concentration of dissolved oxygen *DO* should be high in order for nitrification to occur effectively, not less than 1 mg/L, and it should be evenly distributed in the tank (Bitton, 2005). An optimum nitrification requires a *DO* concentration from two to 3 mg/L (Gerardi, 2002).

The temperature affect the growth of nitrifying bacteria and should not be below 8°C or above 30°C in order to achieve an ideal growing rate. Studies performed in 2006 regarding the effect of ammoniacal nitrogen  $NH_3-N$  on nitrification showed that a high concentration of free  $NH_3$  will affect nitrifying bacteria negatively – a concentration of 0.7 mg/L resulted in a decreasing effectivity of 50 % (Kim, et al., 2006). The optimum pH for most nitrifiers is studied to be from 7.5 to 8.5. Since the process of nitrification will produce hydrogen  $H^+$ , see equation (7) in Section 3.2, which causes a lowered pH, the alkalinity in the water should be adequate to balance the produced acidity (Rusten, et al., 2006).

### 3.3.2 Parameters affecting denitrification

The denitrification process is not as sensitive to parameters in the effluent as nitrification. It is however dependent on the presence of methanol, or other carbon based molecules, for the denitrification process to function (Bitton, 2005). The concentration of DO should be low in order for denitrification to occur, since if there is DO available the denitrifiers will use that for the denitrification instead of the bound  $O_2$  in  $NO_3^-$  (Eriksson, 2011). Temperature affect denitrification - similar to nitrification - and the optimal temperatures has shown to be from  $5^\circ C$  to  $30^\circ C$ . The pH will also have an impact on the process of denitrification, and the optimal pH has shown to be from 7 to 9. A pH below 7 can cause production of nitric oxide  $NO$  which is a toxic gas (Rusten, et al., 2006; Henze, et al., 2002).

### 3.4 Activated sludge in biological treatment

Activated sludge is a by-product formed during the process of wastewater treatment. The term *activated* originates from that the sludge contains active microorganisms that will use the incoming wastewater as nourishment, which contributes to a cleaner effluent (NSFC, 2003). During the processes of nitrification or denitrification microorganisms degrade some of the constituents in the incoming wastewater - such as soluble organic substances - in order to obtain carbon and energy (Gerardi, 2003). This creates the biomass that forms new sludge (UNEP, 2014). The activated sludge will also adsorb metals that can be present in the leachates (Carlsson & Kanerot, 2006). These qualities of activated sludge have resulted in biological treatment techniques based on the biological processes performed by the microorganisms within the sludge (Swedish EPA, 2008), see Chapter 4.

According to Tchobanoglous et al (2003), it is important to determine and optimise the physical characteristics of the activated sludge in a biological treatment process in order to achieve an efficient treatment. An analysis will evaluate if the microorganisms are flourishing or if something is affecting their growth. A thriving activated sludge mass is important in biological wastewater treatment, since it will reduce constituents from the incoming wastewater more effectively (Carlsson & Kanerot, 2006). Commonly used parameters for analysis of sludge are presented in the following paragraphs.

The Sludge Volume Index *SVI* is an index which sets the character of the flocculation and settling of the sludge in a tank. It implies the reciprocal concentration of the sludge after a settling period of 30 minutes (Henze, et al., 2002). Equation (10) below explains how the SVI is estimated, where  $X_{0.5}$  is the sludge concentration after 30 minutes. A SVI below 150 mL/g is considered to be an indication of a good sludge settling according to wastewater treatment theory and a SVI below 100 mL/g indicates excellent sludge settling rate (Spellman, 2004). A too rapidly settling sludge with an SVI below 70 mL/g is however not always preferable since it has shown to result in a turbid effluent due to weakly structured flocs (Janczukowicz, et al., 2001).

$$SVI = \frac{1}{X_{0.5}} \quad [ml/g] \quad (10)$$

Volatile Suspended Solids *VSS* is a fraction of the suspended solids in a sample with activated sludge and constitutes of the organic content. The measured amount of organic content is in many cases equal to the content of biomass in the sample. VSS

can be measured by first filtering a known volume of a sample, then heating the filter in an oven at 550°C. Before this is performed, the filter must be dried in an oven at 105 °C. The difference between the original filter weight and weight after 105°C oven corresponds to the Total Suspended Solids *TSS*. The difference in weight before and after the 550 °C oven corresponds to the *VSS*, or organic content (Spagni & Marsili-Libelli, 2009).

Mixed Liquor Suspended Solids *MLSS* stands for the concentration of TSS in a mixed sample as mg/L and is analysed with the same technique as TSS, except that the sample used contains mixed liquor. A high value of MLSS is undesirable since this may lead to an overload of the activated sludge system and a too low value may imply that the processes are not working correctly. In wastewater treatment, the concentration of MLSS should be between two to five g/L in order for the process to work (Partech Electronics Ltd., 2014; WasteWater System, 2013). Mixed Liquor Volatile Suspended Solids *MLVSS* is used to estimate the amount of active microorganisms in the mixed liquor and is analysed with the same technique as VSS, except that the sample used contains mixed liquor (Larsen, 2010). The amount of MLVSS represents the microbial suspension, hence the organic content, and should at least be 1,500 to 2,000 mg/L for achieving nitrification (Shammas, et al., 2009; Tchobanoglous, et al., 2003).

Sludge Volume *SV* is the volume of settled solids in a one litre cylindrical tank after 30 minutes of settling, and can be estimated as described by equation (11) (WasteWater System, 2013).

$$SV = SVI \cdot MLSS \quad [ml/L] \quad (11)$$

The Hydraulic Retention Time *HRT* is the number of hours it takes for the water to pass through a tank during a process (Gerardi, 2003). The age of the sludge is defined as how long the biomass is available in the system in number of days. This is defined as Sludge Retention Time *SRT* which can be estimated as shown in equation (12). The SRT has shown to result in better efficiency of treatment if it is longer when treating water with a lower temperature (Grady, et al., 2011).

$$SRT = \frac{\text{Biomass in aerated reactor [kg]}}{\text{Biomass removal rate [kg/day]}} \quad [days] \quad (12)$$

The Food to Microorganism *F/M* ratio defines the amount of available food - or organic matter measured as Biochemical Oxygen Demand *BOD*, COD or TOC - in balance with the available microorganisms in the treatment. A high F/M ratio indicates that there is a high concentration of food relative to concentration of microorganisms, which is beneficial for the bacterial growth. A too high F/M ratio will however affect the settling rate of the sludge to become insufficient, which will result in a turbulent effluent. The optimal ratio is hard to predict, since every treatment process has different optimal F/M ratios (Gerardi, 2003). During an SBR optimisation trial in 2004, two SBR treatments were tested with different F/M ratios, 0.2 and 0.4 kg COD/kg MLSS·day. The results showed that both experienced similar treatment efficiencies. However, it was evaluated that an lower F/M ratio, for example 0.14 kg COD/kg MLSS·day, would result in a more stable treatment (Schwarz Pharma, 2004). Equation (13) illustrates how to calculate the F/M ratio with incoming food as COD and sludge amount as MLSS.

$$F/M = \frac{COD \cdot Q_{incoming}}{MLSS \cdot V_{aeration\ tank}} \quad (13)$$

The concentration of toxic metals is generally high in sludge due to physical-chemical processes that leads to accumulation of the metals in the sludge (Stylianou, et al., 2007). Analysis of the concentrations can be performed in order to validate the metal removal efficiency of the treatment, since heavy metals existing in the sludge must originate from the incoming water. By knowing the metal concentration in the sludge, it is also possible to decide how the sludge should be disposed of. Sludge can either be further treated, put on a landfill or reused for purposes such as fertilizers for agriculture.

## 4 Biological treatment techniques

Biological treatment techniques are commonly used for removal of nitrogen from leachates through the processes of nitrification and denitrification. The techniques have also shown efficient for removal of organic matter, since they are based on microorganisms which biodegrades the organic material. Some commonly used biological treatment techniques are aerated lagoons, activated sludge processes such as Sequencing Batch Reactor *SBR*, attached-growth biomass systems such as Moving Bed Biofilm Reactor *MBBR*, anaerobic filters, plant or root systems and constructed wetlands (Renou, et al., 2008; Swedish EPA, 2008). The *SBR* and *MBBR* will be further explained in Section 4.1 and 4.2 respectively.

When using the *MBBR* and *SBR* technique for treating landfill leachates, it is mostly removal of total nitrogen  $N_{tot}$ , or the ammonium ion  $NH_4^+$ , that are the main objective with the treatment (Abbas, et al., 2009). These concentrations can be high in leachates from older landfills – a study showed that the  $NH_4^+$  concentration can be as high as 400 mg/L for a landfill older than five years (Alvarez-Vazquez, et al., 2004). The  $N_{tot}$ , hence the  $NH_4^+$ , threshold value for discharge into watercourses in the Gothenburg region was 1.25 mg/L in 2013 (Ramböll, 2015 B).

### 4.1 Moving Bed Biofilm Reactors *MBBR*

The *MBBR* is a biological technique based on attached biomass which, compared with many activated sludge systems, enables a high volumetric load, stability and more compact reactors (Comett-Ambriz, et al., 2003). The technique was developed during the years 1980-1990 in Norway (Rusten, et al., 2006). Earlier developed biological treatment techniques, such as trickling filters and granular media biofilters, often had problems with occupying large areas or demanded continuous cleaning by backwashing. The *MBBR* technique was developed in order to minimise these problems and became popular for various treatment plants. It needs less area to function and there is no need for recycling sludge (Renou, et al., 2008). Nitrogen removal rates for the *MBBR* process are generally high. According to Butkovskiy (2009), the *MBBR* has shown  $N_{tot}$  removal rates from 85–90 % whilst the reduction of Chemical Oxygen Demand *COD* in previous cases has not exceeded 20 %.

One advantage with *MBBR* is its low sensitivity to toxic compounds, since it is based on a principle of multi-layered bio systems within plastic porous polymeric carriers which protects the biomass to some extent (Headworks BIO, 2015). Earlier studies have also shown that the *MBBR* treatment have more stable treatment results during different loading rates and concentrations in wastewater compared to conventional treatment techniques (Camper & Bott, 2006).

The volume of the tanks can be optimised by small suspended porous polymeric carriers that flow freely in the tank, enabling a large residence area for the bacteria to grow on (Rusten, et al., 2006). These bio carriers are designed to give the bacteria a calm environment for growing and at the same time to provide good transfer of substrate or oxygen  $O_2$  (Ayub & Alam, 2011). The *MBBR* treatment technique can be designed for different purposes since choice of order and presence of oxic or anoxic tanks can be made. This enables a choice of sequence in which the denitrification and nitrification will occur (Rusten, et al., 2006). Part of the sludge can be returned to the beginning of the oxic tank in order for the incoming leachate to come in direct contact

with active microorganisms (Ayub & Alam, 2011). See Figure 4.1 for an illustration of the MBBR process.

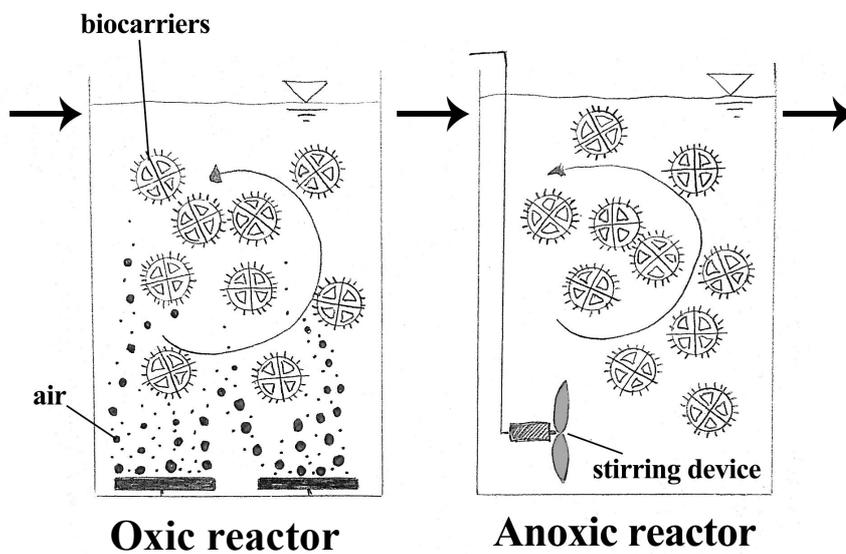


Figure 4.1 Typical design of a MBBR with an oxic reactor followed by an anoxic reactor.

## 4.2 Sequencing Batch Reactor *SBR*

The SBR consists of one single volume tank where phases of aerobic and anaerobic processes take place, see Figure 4.2. Each process occurs in a sequence with intervals of normally four to eight hours (Cerne, et al., 2007). The process is built on a fill- and draw scheme and activated sludge. The SBR became popular and started to be further developed during the 1950's and 60's but similar fill and draw systems was operated already in 1914 (U.S. EPA, 1999).

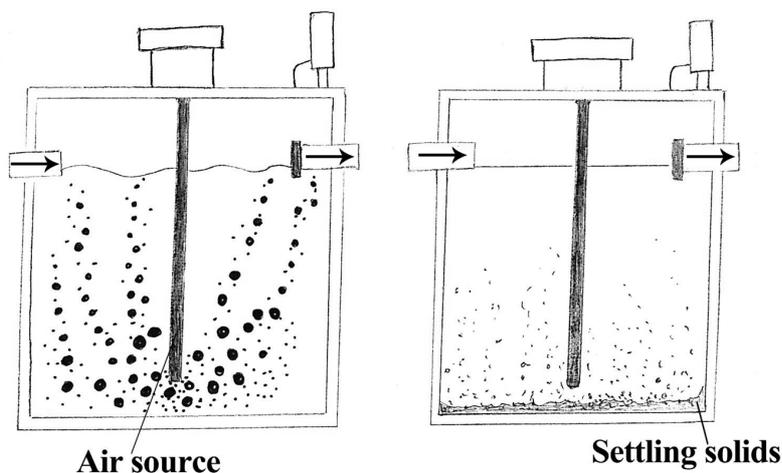


Figure 4.2 An SBR reactor where different processes occur in the same tank at different sequences.

The purpose with the SBR is mainly to biologically remove nitrogen from leachates (Spagni & Marsili- Libelli, 2009). Earlier studies performed regarding the treatment capacity of the SBR have shown high removal efficiency for the parameters COD,  $N_{\text{tot}}$ , Biochemical Oxygen Demand *BOD*, Total Kjeldahl Nitrogen *TKN* and Total Suspended Solids *TSS* (Mahvi, 2008). The removal rate of COD for example has reached a rate of 75 % during earlier studies (Renou, et al., 2008). In a project where different advanced landfill leachate treatments were tested during low temperature, the SBR results showed an ammonium nitrogen  $NH_4-N$  removal rate of 99.7 % at temperatures between 10.9°C –14.9°C (Sun, et al., 2010).

Oxic conditions are generated in the tank by aeration equipment placed at the bottom of the tank, which aerates the effluent. Autotrophic bacteria need  $O_2$  to be able to decompose the BOD and for the nitrification process to start (Cerne, et al., 2007). During the nitrification process,  $NH_3$  is oxidised into nitrate  $NO_3^-$  in two oxidation steps (Spagni & Marsili- Libelli, 2009). A precipitant can be added to the water at the end of the oxic phase which causes reduction of phosphorus if required. In the anoxic stage, it is often necessary to add an external carbon source to achieve a complete denitrification (Cerne, et al., 2007). The external source performs as an electron donor when  $NO_3^-$  is reduced to nitrogen gas  $N_2$ , which is the final stage in the treatment process (Spagni & Marsili- Libelli, 2009).

Organic material is the main component decomposed during the SBR process. Metals and persistent organic compounds are absorbed to the sludge to some extent but the separation seems to be limited for the case of leachates compared to wastewater (Cerne, et al., 2007).

Depending on the level of activated sludge and retention time, the treatment capacity can vary (Cerne, et al., 2007). This technique is most suitable to use for leachates with a low or intermediate flow (U.S. EPA, 1999). One way to optimise the system is by constructing several tanks which are operating according to a specific sequence (U.S. EPA, 1999). Another alternative for optimising the technique can be by applying the so called step-feed SBR. This version of the SBR is widely used in the United States for removal of nitrogen and BOD (Sahlstedt, et al., 2012). To achieve nitrogen removal, anoxic and oxic phases are alternated in a sequence and the effluent is added in the beginning of each anoxic phase.

The advantages with the SBR technique is that the level of activated organisms are very high compared to for example aerated ponds, due to the high level of bacteria fixed to the sludge (Cerne, et al., 2007). The SBR treatment enables flexibility in the form of controlling the time step for each process in the tank. There is also no need for return of activated sludge since all the processes occur in the same tank. However, the microorganisms will create excess sludge that needs to be removed in order to achieve the proper amount of sludge in the tank (Comett-Ambriz, et al., 2003). Furthermore, it is easily operated, has a minimal footprint and requires a limited space. Some disadvantages are that the technique demands high levels of maintenance and qualified personnel since it involves advanced systems and controls. Also, depending on the type of SBR, there might be a risk of clogging the aeration membranes (U.S. EPA, 1999).

## 4.2.1 The SBR cycle phases

Each operation cycle in the SBR process generally consists of five different phases: fill, react, settle, draw and idle. The length of each phase is determined based on the characteristics of the treated effluent (Wang & Shammas, 2009). The total cycle time is expressed as the sum of the five phases (Tomaszek, 2005). Figure 4.3 illustrates a common sequence of the SBR cycle phases. Each phase in the process is described further in the following paragraphs.

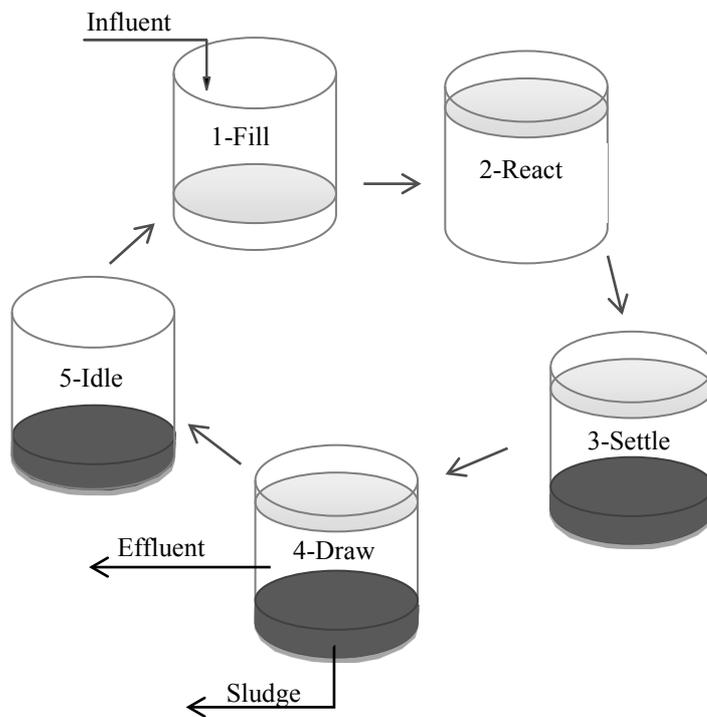


Figure 4.3 Cycle phases in an SBR process [based on figure from (Wang & Shammas, 2009)].

### 1 – FILL PHASE

During the fill phase, new untreated effluent is added to the treatment tank statically. At the same time constant mixing or aeration can be performed. Static filling will create a high concentration of food to microorganisms  $F/M$  ratio once the reaction phase begins. This condition favours some specific organisms which results in an easily settled sludge and good conditions for biological removal of phosphorus. Fill during mixing creates direct biological reactions where organic material is degraded by bacteria. This can create an anoxic environment where denitrification can take place. Aeration during the fill phase works in the opposite way by creating an environment rich of  $O_2$  which enables aerobic reactions to initiate, which are completed during the reaction phase (U.S. EPA, 1999).

### 2 – REACTION PHASE

During the reaction phase, nitrification or denitrification occurs depending on which conditions; oxic, anoxic or both during different times, that has been decided. The choice of method in the fill phase will affect the conditions and biological reactions in this phase (U.S. EPA, 1999).

### 3 – SETTLING PHASE

After the reaction phase the mixing and aeration equipment are turned off to create a calm environment in the tank. This condition is called the settling phase, which allows the solids in the effluent to settle at the bottom of the tank as sludge, hence creating a clear water phase at the top (Butkovskiy, 2009). Slow mixing during the initial part of this phase can sometimes achieve a clearer effluent but it often takes place without any stirring.

### 4 – DRAW PHASE

During the draw phase, clarified effluent is discharged from the tank (U.S. EPA, 1999). This can be achieved by using a floating or fixed weir, which is the most common technique to use. The decanting procedure and its capacity to remove treated effluent is one of the major operational and mechanical limitations of the SBR. This is due to the possibility of discharging sludge along with the treated effluent if the decanting equipment is not correctly operated (Wang & Shammas, 2009). The volume of the clarified effluent drawn from this phase, in relation to the total reactor working volume defines the exchange ratio of the SBR process (Liu, 2006).

### 5 – IDLE PHASE

Finally, the tank is kept inoperative during the idle phase. Thereafter, new untreated effluent is added to the settled activated sludge and a new cycle begins (Butkovskiy, 2009). During the idle phase it is possible to achieve equalisation and perform regulation depending on flow rates and operation strategies (U.S. EPA, 1999).

## 4.2.2 Cycle analysis of the SBR

An optimisation of the SBR can be performed by completing a cycle analysis. The cycle analysis is accomplished by taking samples continuously during one cycle and thereafter analysing the nitrogenous fractions in each sample to see how their concentrations change during each phase. The results will show if the length of each SBR phase is long enough for the processes to function properly (Butkovskiy, 2009).

According to Tchobanoglous et al (2003) if assuming that there is a full conversion from  $\text{NH}_4^+$  to  $\text{NO}_3^-$  during nitrification - see equation (7) in Chapter 3, Section 3.2 - then the concentration of  $\text{NO}_3^-$  should be the same as for  $\text{NH}_4^+$  after the reaction phase in the oxic tank, e.g. after 1.75 hours in the presented case, see Figure 4.4. For the anoxic tank, where denitrification occurs, the concentration of  $\text{N}_2$  should be the same as nitrite  $\text{NO}_2^-$  after the reaction phase. However,  $\text{N}_2$  is usually not measured since it is difficult to analyse it compared to other nitrogenous fractions.

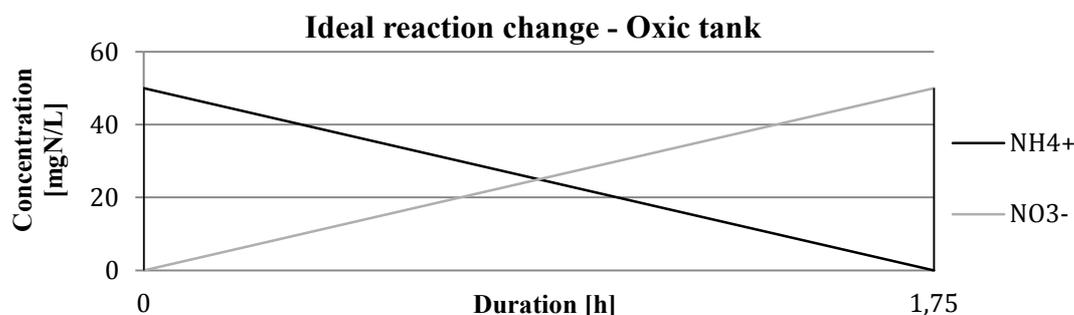


Figure 4.4 Ideal reaction change in the oxic reaction tank during nitrification

### 4.2.3 Operation of the SBR

The fact that the SBR only involves one tank is beneficial from an operational point of view, since it reduces the number of equipment that needs to be maintained, compared to conventional biological treatment techniques. The SBR can replace the need for separate primary and secondary clarifiers as well as related mixers, pumps, aeration equipment and anoxic or oxic basins which are needed in a conventional nutrient removal processes (Wang & Shamma, 2009).

On the other hand, since the SBR is based on a more advanced and automatic system, this makes the process more vulnerable to failure. Therefore, higher demands on maintenance of the control system, valves, panels, switches etcetera are required compared to a conventional system. But the SBR is still a very flexible system since it is easy to change cycle times per day, and it can be adapted to resemble any other conventional activated sludge process (Wang & Shamma, 2009). It is important to continuously monitor certain parameters in the treated effluent, such as temperature, Dissolved Oxygen *DO*, pH and alkalinity, since these affect the biological processes in the SBR (NEIWPCC, 2005).

## 4.3 A comparison between MBBR and SBR

In Table 4.1, a comparison between the MBBR and SBR technique is presented regarding several operational parameters. The comparison of the two techniques is based on studies conducted at Spillepeng landfill in Malmö (Ramböll, 2015 C).

*Table 4.1 Comparison between the MBBR and SBR regarding certain operational parameter [based on data compiled by (Ramböll, 2015 C)].*

<b>Operational parameters</b>	<b>MBBR</b>	<b>SBR</b>
Robustness of the process	+	
Need for machinery	+	
Need for personnel	+	
Required space	+	
Energy demand		+
Treatment efficiency		+
Need for chemicals	equal	

According to these studies, the MBBR has shown to be more robust compared to the SBR (Ramböll, 2015 C). As mentioned in Section 4.1, the polymeric plastic biocarriers creates a protected environment for the microorganisms which makes the process more resilient towards toxic compounds and changes in loading rates compared to the SBR. The need for machinery and personnel is also less for the MBBR than for the SBR, which is economically favourable. As mention in Section 4.2, the SBR requires a high level of maintenance and skilled operators due to the advanced control system, whereas the MBBR technique involves less maintenance.

Furthermore, the MBBR can be operated within a smaller space than the SBR when having the same volumetric loads. However, it must be taken into consideration that the SBR might be more flexible than the MBBR, since all the processes are operated in one single reactor. Also, the energy demand will be less for an SBR treatment since the requirement for aeration is lower. This is mainly because of the plastic polymeric

biocarriers that can be used for the MBBR, which will demand more O<sub>2</sub> in order for it to reach the bacteria inside the protected environment.

The studies by Ramböll showed that the advantages with the SBR technique are foremost its treatment capacity and efficiency regarding removal of nitrogen, which was shown to be higher than for the MBBR (Ramböll, 2015 C). Regarding COD, the MBBR technique does not generally exceed 20 % removal efficiency whereas the SBR has reached efficiencies of 75 % in earlier studies. When comparing the need for dosage of chemicals it can be seen that it is equal for the two techniques.

The comparison of the MBBR and SBR presented in Table 4.1 indicates that the MBBR is the most favourable technique to choose for wastewater treatment. However, this could be misleading since the required quality of the treated effluent, local conditions, economic assets etcetera differs from case to case. For example, if it is required to reach a high removal efficiency of nitrogen and organic matter the SBR could be the more suitable option. Therefore, the mentioned operational parameters must be evaluated for each case in relation to current circumstances.



## 5 National and International Cases

A literature review was performed in order to gain knowledge of both national and international cases of earlier performed lab scale Sequencing Batch Reactor *SBR* trials for treatment of landfill leachates. In total, eight cases were reviewed, including three national and five international cases. Even though the main focus in this chapter has been on previous *SBR*'s, CASE 2 and CASE 4 also includes trials performed on Moving Bed Biofilm Reactors *MBBR*'s. The knowledge and data gained during the review was considered and applied during the construction and operation of the lab scale *SBR*'s at Brudaremossen. The most significant data from the review are summarised in Section 5.1. Thereafter, Section 5.2 and 5.3 will provide short descriptions of the cases. Finally, results and recommendations from the cases, applicable for case Brudaremossen, are presented in Section 5.4.

### 5.1 Summarisation of the literature review

In this section, the most significant data from the literature are summarised. It includes a comparison of resulted total nitrogen  $N_{tot}$ , ammonium nitrogen  $NH_4-N$  and Chemical Oxygen Demand *COD* removal efficiencies and some important design parameters applied in the design of the *SBR* treatments.

#### 5.1.1 *SBR* removal efficiencies

Table 5.1 presents the concentrations of  $N_{tot}$ ,  $NH_4-N$  and *COD* in the influent to the *SBR* treatments, as well as related removal efficiencies. Overall, the *SBR*'s studied in the cases showed high removal efficiencies of  $N_{tot}$  and  $NH_4-N$ , except for CASE 7, see Table 5.1. The removal efficiency of *COD* ranges from 25 % to 55 % and the average removal efficiencies of  $N_{tot}$ ,  $NH_4-N$  and *COD* for the cases were 79 %, 87 % and 38 % respectively. The lower removal efficiencies presented for CASE 7 could be related to the high concentrations of  $N_{tot}$ ,  $NH_4-N$  and *COD* in the incoming leachates to the *SBR*, which are considerably higher compared to in the other cases.

Table 5.1 Summary of removal efficiencies for the eight cases

Case	Source	Concentrations in the influent [mg/L]			Removal efficiency [%]		
		$N_{tot}$	$NH_4-N$	<i>COD</i>	$N_{tot}$	$NH_4-N$	<i>COD</i>
1	(Butkovskyi, 2009)	163	131	578	93	99	< 40
2	(Ekenberg, 2007)	325 – 344	265 – 273	655–718	-	97	30
3	(Junestedt, et al., 2003)	235	95	250	93	99.7	36
4	(Comett-Ambriz, et al., 2003)	263	232	3.16	99	99	55
5	(Kulikowska & Klimiuk, 2004)	397	362	757	-	99	49–51
6	(Spagni, et al., 2008)	-	1199	2055	90	-	30–40
7	(Ganigué, et al., 2009)	3774 ± 960	3772 ± 956	4357 ± 692	15–20	10–36	25–30
8	(Monclús, et al., 2009)	-	1268	1674	80	98	29

## 5.1.2 SBR operational parameters

Table 5.2 provides a summarisation of the length of the study period, temperature, Sludge Retention Time *SRT*, exchange ratio, cycle time and the duration of the phases for each case.

Table 5.2 Summarisation of operational parameters for the lab scale SBR's

Case	Period [d]	Temp. [°C]	SRT [days]	Ex. ratio [%]	Cycle time [h]	Fill	Oxic	Anoxic	Settle	Draw
1	160	20	15.60	80	8	0.80	1 <sup>st</sup> : 2–5 2 <sup>nd</sup> : 0.33	1.33–4.33	1	0.33–0.50
2	~730	21	25	93	8	-	4.25	2.08	0.50	0.33
3	~180	-	-	26	8	0.08	4	1.25	1.67	1
4	79	-	-	8	8	0.08	6.75	-	1	0.17
5	~365	20	-	33–50	12/ 24	0.25	20.50	10	1.5/3	0.25
6	1,095	20	-	38	24	-	2	3.75	1	0.05
7	450	36	25–30	50	24	0.25	0.25	1.42	0.33	0.33
8	170	16–19	29	7–10	8/12	-	1.17–1.67	1.67–2.33	0.75	0.50

The length of the study periods ranges from just about 80 days up to three years, with an average duration of 400 days. In many cases, the study has been divided into two or more stages. The first stage focuses on adapting the technique to current conditions and to start up the biological processes. Thereafter the main focus in many studies has been to optimize or atomise the SBR technique. Since the average study period exceeds one year, it indicates that a longer time period is relevant to consider when performing a lab scale SBR trial in order to achieve valid results and to be able to optimise the process.

Table 5.2 illustrates that most of the SBR's were operated in room temperature, around 20 °C, which is preferable for achieving nitrification and denitrification. For CASE 6, 7 and 8, which were applying the step-feed method - hence strived for achieving the Anaerobic ammonium oxidation bacteria *Anammox* process - a temperature close to 35 °C would have been desirable. The *Anammox* bacteria require a high temperature in order for the processes to function. Nevertheless, only CASE 7 reached a temperature close to 35 °C whilst CASE 6 and 8 was operated in temperatures close to or below 20 °C. Still, these cases reached higher removal rates compared to CASE 7. It could be that other operational parameters had a negative impact on the removal rates for CASE 7 and that the results for CASE 6 and 8 could have been improved with an increased operational temperature.

SBR treatments commonly design for SRT between 10 to 30 days according to Tchobanoglous et al. (2003). When comparing the SRT's from the cases they ranged from 15.60 up to 30 days.

The exchange ratio varied significantly from case to case, seven percent for CASE 8 up to 93 % for CASE 2, with an average value of 40 %. The cycle times were more similar for all the cases ranging from eight to 24 hours. Eight was the most commonly

used cycle time, applied in five cases. Concerning the phase times, it can be seen that the fill and draw phase were the shortest phases, generally lasting not longer than half an hour. The average settling time was approximately one hour and 20 minutes. For CASE 1 to 5 the oxic phase was longer than the anoxic phase whereas it was the opposite for CASE 6 to 8. The first five cases were based on more conventional operation of the SBR technique, whilst the last three were based on the step-feed method.

In Table 5.3 it can be seen that methanol was the most commonly used external carbon source, used in three cases, CASE 3, 5 and 8. Ethanol was applied in CASE 1 and 2, whilst bicarbonate was used in the two international Spanish cases, CASE 7 and 8. The last two mentioned applied the steep-feed method. In CASE 7 the addition of bicarbonate caused declined denitrification efficiency and problems related to the dosage apparatus (Ganigué, et al., 2009). For CASE 8 bicarbonate was added in order to improve nitrification since it consumes alkalinity. It was noted that the cases during which bicarbonate was applied achieved the lowest removal rates whilst the cases that applied methanol and ethanol seems to reach rather equivalent results. Phosphoric acid or Monopotassium phosphate  $KH_2PO_4$  has been added for the cases where the incoming concentration of phosphorus has been too low to achieve biomass growth.

Table 5.3 Summarisation of added chemicals and dosages to the SBR's

Case	Chemicals Oxic phase	Dosage [mg/L]	Chemicals Anoxic phase	Dosage [mg/L]	COD/N ratio
1	-	-	Ethanol	1.37	4.43
2	Phosphoric acid	0.73	Ethanol	62–164	2.96-3.89
3	Phosphorus	-	Methanol	-	-
4	-	-	-	-	-
5	-	-	Methanol	790 – 1,190	5.36-7.02
6	$KH_2PO_4$	0.50–1.00*	Sodium acetate trihydrate	20,000	9.51
7	-	-	Bicarbonate	10,200–20,400	-
8	-	-	Bicarbonate Methanol	B: 0.78 M: 2,930**	4.79

\*Maintained that level of phosphate in the SBR

\*\*Based on the density of Methanol; 791 mg/ml

Concerning the dosage of phosphorus and carbon source it can be seen that it varies significantly from case to case. Since the dosage of phosphorus only was obtained from two cases, the data was insufficient for making any conclusions. As for the dosage of an external carbon source, the dosages range from 0.78 mg/L up to 20,400 mg/L. Since the required dosages of chemicals vary with the concentrations of ammonium nitrogen in the incoming leachates, the dosages for the cases was expected to be varying due to that every leachate has different characteristics. The COD/N ratio illustrates the incoming COD concentrations together with the dosage of carbon source expressed as COD divided by the incoming concentration of ammonium nitrogen. The different ratios are therefore easier to compare when having varying leachate characteristics, and it can be noted that the methanol generally requires a higher COD/N ratio than ethanol throughout the cases.

## 5.2 National cases

The national cases include three substantial lab scale SBR studies performed in the city of Helsingborg, Malmö and Köping. The location of the three cities and during which years the studies were performed are presented in Figure 5.1.

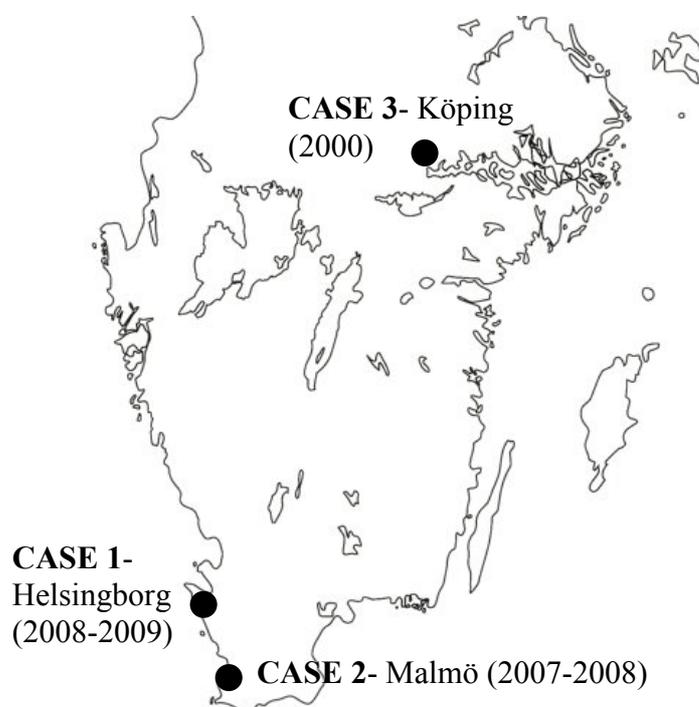


Figure 5.1 Location of the three national Swedish cases and during which years they were performed.

### 5.2.1 CASE 1 - Filborna landfill, Helsingborg

The possibility to use SBR for onsite treatment of leachate from the landfill Filborna, southern Sweden, was evaluated during a Master's Thesis at the University of Lund in 2009 (Butkovskiy, 2009). The leachates from the landfill were at the time collected and locally treated in treatment ponds and thereafter channelled to the Wastewater Treatment Plant *WWTP* in the municipality of Helsingborg. Critical parameters in the leachates from Filborna were Biochemical Oxygen Demand *BOD*, Total Organic Carbon *TOC*, total phosphorus  $P_{tot}$  and  $N_{tot}$  which all exceeded the threshold for discharge into the strait Öresund. Both a lab scale and industrial scale SBR for removal of nitrogen were studied and evaluated from November 2008 until April 2009, which provided design parameters used to calculate a full scale SBR facility.

The lab scale SBR at Filborna Landfill was operated during 160 days from the 5<sup>th</sup> of November in 2008 until the 15<sup>th</sup> of April in 2009 (Butkovskiy, 2009). The installation included a reactor tank with a total volume of five litres and a working volume of four litres, attached mixer and aerator, vessels for adding an external carbon source, incoming and outgoing effluent, connected pumps and timers, see Figure 5.2.

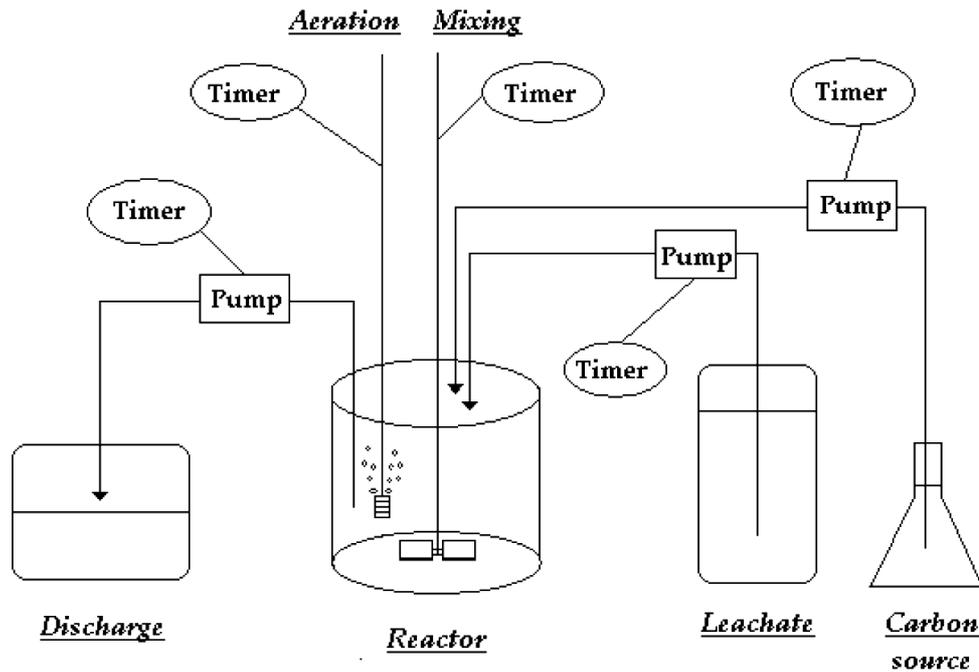


Figure 5.2 Illustration of the components included in the lab scale SBR at Filborna (Butkovskiy, 2009) [permitted by author].

The total cycle time was eight hours; including two oxic phases to be able to oxidize easily biodegradable organic substances left after the anoxic phase (Butkovskiy, 2009). The maximal hydraulic load of the process in the course of one cycle was 4.6 litres of leachates added to three litres of sludge. Ethanol was used as external carbon source whilst addition of phosphorus was not necessary in this case due to high concentrations in the raw leachates.

The lab scale SBR reached a 93 % removal rate of  $N_{\text{tot}}$  and 99 % of  $\text{NH}_4\text{-N}$  which implied concentrations lower than the requirements for discharge into Öresund (Butkovskiy, 2009). Reduction of COD and  $P_{\text{tot}}$  measured lower than 40 % and 20 % respectively. Results from the analysis showed that the nitrification process was the limiting step in the process, when comparing a nitrification rate of 23.0-23.6  $\text{NH}_4\text{-N/L}\cdot\text{h}$  to a denitrification rate of 178.8-184.5 Nitrate nitrogen  $\text{NO}_3\text{-N/L}\cdot\text{h}$ . Critical operational parameters were examined and it was concluded that the temperature, pH and concentration of Dissolved Oxygen  $DO$  highly affect the nitrification rate. Further evaluation of techniques for reduction of COD and  $P_{\text{tot}}$  was recommended after this study. Some technical problems, such as clogging and breaking of the energy supply decreased the treatment efficiency during some short periods of the study.

### 5.2.2 CASE 2 – Spillepeng landfill, Malmö

The Spillepeng landfill in Malmö is maintained by a waste management company in southern Sweden entitled SYSAV. It is still an active landfill area, permitted to be used for waste disposal until 2017. Thereafter, the strategy from the City of Malmö was to reconstitute the closed down parts of the site to act as a park for recreational purpose (Jonsson & Andersson, 2011). The leachates from Spillepeng landfill has previously been discharged to the local municipal WWTP Sjölanda, but during 2007 new regulations demanded that the leachates should be treated on site (Heander,

2007). The leachate at the landfill was during 2007 characterized by high concentrations of  $\text{NH}_4\text{-N}$  and COD. The leachates also had a low  $\text{BOD}_7$  concentration, indicating a low  $\text{BOD}/\text{COD}$  ratio (Ekenberg, 2007).

Two Master's Thesis studies were carried out at Spillepeng, one during 2007 by Heander and one during 2008 by Gørfelt, in order to analyse the pilot plant SBR treatment that was installed near the landfill in 2007 (Ramböll, 2015 C). A lab scale MBBR was also constructed during 2007 by the company AnoxKaldnes, on behalf of SYSAV. When comparing the results from the study of the lab scale SBR and the MBBR respectively, it was showed that the SBR had a higher removal rate of  $\text{N}_{\text{tot}}$  and BOD.

The aim with the SBR pilot plant treatment was to reach effluent concentrations of  $\text{N}_{\text{tot}}$  below 15 mg/L and COD below 500 mg/L. The SBR pilot plant had an effective volume of 2,800 litres and a total cycle time of eight hours. The phases included in the cycle are presented in Figure 5.3 (Gørfelt, 2008). The Hydraulic Retention Time  $HRT$  was 1.7 days and both phosphorus and ethanol was added to the process. The mean temperature in the SBR tank was 21°C and the mean pH was 7.5 (Heander, 2007).

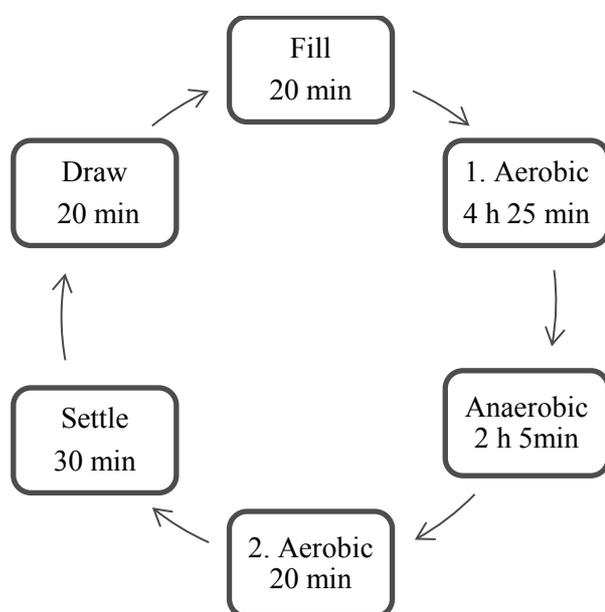


Figure 5.3 Schematic illustration of one cycle in the pilot plant SBR at Spillepeng

#### MASTER'S THESIS BY HEANDER (2007)

The Master's Thesis by Heander was carried out during the 166 first operating days of the pilot plant SBR (Heander, 2007). After six operation days, the nitrification process was successfully removing 97 % of incoming  $\text{NH}_4\text{-N}$ , reaching an effluent concentration of approximately 2 mg/L. The denitrification was started up after 98 operation days, and the optimal ethanol dosage was calculated to be 5-5.6 grams COD, as ethanol, per gram  $\text{NO}_3\text{-N}$ . The start of the ethanol dosage begun 10 minutes after the oxidation process was finished, in order to make sure that the ethanol was not used for oxic digestion of the organic matter. Accumulation of nitrite  $\text{NO}_2^-$  occurred four times and was evaluated to be caused by increasing influent treatment volume or oxygen  $\text{O}_2$  concentration dropping below two milligrams  $\text{O}_2$  per litre. The removal efficiency for organic matter, measured as total COD, was approximately 30 % at

best, which resulted in insufficient treatment of organic matter. Some problems with loss of sludge, foaming and overflows were observed.

#### MASTER'S THESIS BY GÖRFELT (2008)

The aim of the Master's Thesis by Görfelt was to further evaluate and make the SBR pilot plant more effective (Görfelt, 2008). Results showed that the treated effluent from the pilot plant had high concentrations of phosphate phosphorus  $PO_4-P$  during 2007, approximately 1.6 mg/L, whilst the threshold value for  $P_{tot}$  discharge into surrounding recipients was 0.5 mg/L. The study showed that 11 mg P/g N was enough to fulfil the demand for biological processes in the SBR, instead of the first approximated dosage of 20 mg P/g N. This resulted in a treated effluent concentration for  $P_{tot}$  below 0.5 mg/L. The nitrification and denitrification rates were also analysed during the study, resulting in rates between 7.6-18 mg  $NH_4-N/L\cdot h$  and 11-180 mg  $NO_3-N/L\cdot h$  respectively, whereas the latter results were analysed at the end of the study when the aeration functioned properly.

#### LAB SCALE MBBR - AnoxKaldnes (2007)

During 2007, a study was performed at the Spillepeng landfill by Ekenberg from the company AnoxKaldnes (Ekenberg, 2007). By orders from SYSAV, the study aimed to evaluate the possibility of treating the leachates at Spillepeng with a MBBR process. A lab scale MBBR was constructed and analysed at room temperature, 22–24 °C, with a nitrification tank volume of one litre followed by a denitrification tank with a volume of 0.5–1.0 litre.

The HRT was decreased from 30 hours to 15 hours when the leachates showed to have negligible toxic effects on the nitrifying bacteria. However, a HRT of approximately 25 hours showed to give the optimal nitrification rate (Ekenberg, 2007). The denitrification process was started up 41 days after the nitrification with a hydraulic retention time of 10-15 hours, which later was increased to 30 hours since the denitrification rate was low in the beginning of the study.

The results from the lab scale MBBR study by Ekenberg showed that the leachates at Spillepeng can be treated with biologically without the leachate having an obstructive effect on the nitrification or denitrification (Ekenberg, 2007). Some problems with the denitrifying step were evaluated to originate from a too low dosage of ethanol during certain days and also a noticeable concentration of  $O_2$  that leached into the denitrification tank at certain times. The dissolved COD was reduced with 37 % and the  $N_{tot}$  concentration was reduced to 15 mg/L in the treated effluent. Roughly 3 mg/L of the  $N_{tot}$  was assessed to consist of inorganic fractions such as  $NH_4-N$ ,  $NO_3-N$  or nitrite nitrogen  $NO_2-N$  and the rest, 12 mg/L, was organic fractions of nitrogen. The organic fractions of nitrogen were evaluated to be difficult to biodegrade during the biological processes.

### 5.2.3 CASE 3 - Norsa landfill, Köping

Junestedt et al (2003) at the IVL Swedish Environmental Research Institute Ltd performed a study where an SBR treatment plant at Norsa landfill close to Köping, near Stockholm in Sweden, was analysed and discussed. The SBR treatment plant was installed during spring and summer in 2000. The landfill leachates were at the time of sampling, in 2001, characterized by generally increased concentration of critical parameters. This was conducted to be because of a continual rain period before the previous sampling occasion in 2000, which lead to dilution of the leachates.

The SBR tank had a total volume of 300,000 litres, where the exchange volume was approximately 80,000 litres (Junestedt, et al., 2003). The total cycle time was eight hours during which methanol and phosphorus was added to the process in order to ensure efficient treatment.

The SBR showed to have a virtuous treatment of  $\text{NH}_4\text{-N}$  with a removal efficiency of 99.7 %, with values from 190 mg/L in influent to 0.56 mg/L in the treated effluent (Junestedt, et al., 2003). The concentrations of  $\text{N}_{\text{tot}}$  were 220 mg/L in the influent and 15 mg/L in the treated effluent. Results showed that the reduction of metal concentrations were minor, except for lead *Pb*, Tin *Sn* and Vanadium *V* of which some difference were noted. Observations showed that sludge was accidentally transported with the treated effluent at several occasions during the treatment process.

### 5.3 International cases

The international cases include five substantial lab scale SBR studies performed in German, Poland, Italy and Spain. The location of the cases and during which years they were performed are presented in Figure 5.4.

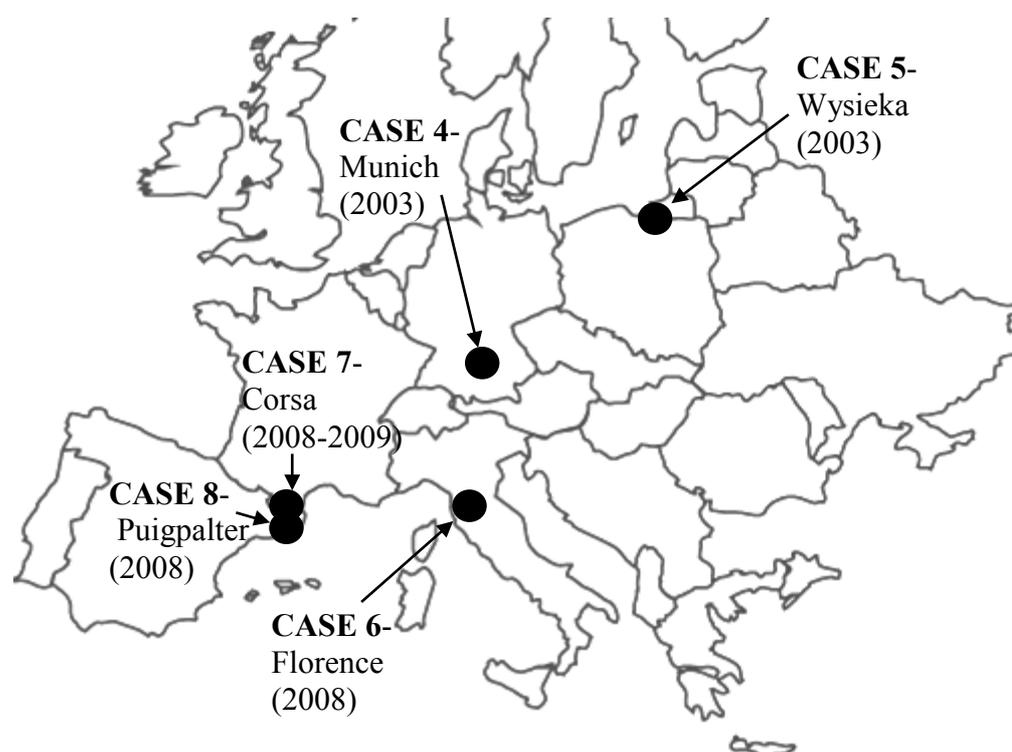


Figure 5.4 Location of the international studied cases in Europe

#### 5.3.1 CASE 4 - Comparison of MBBR and SBR, Munich

A comparison between MBBR and SBR was carried out by Comett-Ambriz et al (2003) for effluents from a biowaste fermentation plant in Munich. The effluent had a high  $\text{NH}_4^+$  concentration, an average of 232 mg/L, and a high COD/BOD<sub>5</sub> ratio, an average of 3.5 mg/L.

The aim of the study was to evaluate the removal efficiency of organic carbon and  $\text{N}_{\text{tot}}$  during the nitrification process of MBBR and SBR (Comett-Ambriz, et al., 2003).

This was performed by constructing a lab scale tank where both the MBBR and SBR treatment process were achieved with the same operating conditions, except the settling time. The first phase, lasting 70 days, simulated the MBBR process and included bio carriers. The test was carried out with two different settling times; 15 minutes and one hour. Thereafter, the second phase lasted 79 days and simulated the SBR process with an exchange volume of three litres and a settling time of one hour.

Results from the study showed that the MBBR and SBR had very similar removal capacities of nitrogen, both having 99 % removal efficiency, but that the removal of total COD was higher in average for the SBR, 53 % compared to 40 % for the MBBR (Comett-Ambriz, et al., 2003). This difference was evaluated to originate from a higher biomass concentration created during the SBR process, ranging from 1.07 to 4.4 kg Total Suspended Solids  $TSS/m^3$ , in comparison with a concentration from 0.8 to 2.4 kg  $TSS/m^3$  for the MBBR. The two different tested settling times for the MBBR was proven to influence the removal of COD to become lower with a shorter settling time.

### **5.3.2 CASE 5 – Landfill near Wysieka, Poland**

A SBR treatment study of landfill leachates was performed by Kulikowska & Klimiuk in 2003 from a municipal landfill in Wysieka, Poland. The landfill has been in operation since 1996 and the leachates were at the time characterized by a  $BOD_5$  concentration of 105 mg/L and a COD concentration of 757 mg/L, which implies a low  $BOD_5/COD$  ratio. The concentration of  $NH_4-N$  was high, 362 mg/L.

The lab scale SBR treatment was designed with a two-stage nitrification and denitrification system operated in room temperature conditions (Kulikowska & Klimiuk, 2004). Two identical nitrification tanks were installed, one with HRT of three days and one with HRT of two days, with a reactor volume of six litres each and a total cycle time of 24 hours. Thereafter, the treated water was channelled into four different denitrification tanks, with varying dosages of methanol ranging between 1.8-7.3 mg COD/mg N. The HRT in all the denitrification tanks was one day and the total cycle time was 12 hours.

The study showed that the difference in HRT between the two nitrification tanks, resulted in an organic COD removal efficiency of 51 % and 49 % respectively (Kulikowska & Klimiuk, 2004). A  $NH_4-N$  removal rate of approximately 20 mg/L·h was evaluated to occur in both tanks, but the tank with lower HRT was observed to have an accumulation of  $NO_2^-$  during the nitrification process. A low  $BOD_5/COD$  ratio in combination with low removal efficiency of organics was evaluated to be due to slowly or non-biodegradable organics present in the leachates.

The study also showed that the different methanol dosages added to the different denitrification tanks affected the process (Kulikowska & Klimiuk, 2004). The optimal methanol dosage was evaluated to be 3.6 mg COD/mg N. The concentration of  $NH_4-N$  in the treated effluent steadily reached 0.08 mg/L. It was concluded that the nitrogen/COD ratio in the landfill leachates affect the treatment process and that a higher ratio would enable a higher  $NH_4-N$  removal rate. A preferable nitrogen/COD ratio for an optimal nitrification process was evaluated to be 1.08.

### **5.3.3 CASE 6 –Lab scale SBR trial at the University of Florence**

The aim with the study was to evaluate a real-time control system for a SBR, treating leachates from an old landfill (Spagni, et al., 2008). Focus was on improving the

nitrification and denitrification process by adjusting the length of the anoxic and oxic phases as well as the amount of added external carbon source. The goal was also to increase nitrogen removal by nitrate, reduce the addition of external carbon source and energy used for oxidizing ammonia  $NH_3$  (Spagni & Marsili- Libelli, 2009). Leachates were received from a municipal landfill in the northern part of Italy. The leachates contained high concentrations of nitrogen, low concentrations of COD and volatile fatty acids. The BOD/COD ratio was considered low and metals were detected in the range normal for many municipal landfills.

A lab scale SBR was constructed with a working volume of 24 litres and was operated during approximately three years in room temperature,  $20 \pm 0.5 \text{ C}^\circ$  (Spagni & Marsili- Libelli, 2009). Conventional treatment via nitrification and denitrification was optimised and monitored during the first year. During the second year, the SBR was operated in order to build up  $NO_2^-$  and to analyse if this method would result in a higher nitrogen removal. Throughout the final year, the process was operated both locally and by using a computer based fuzzy supervisory control system. The SBR had a total cycle time of 24 hours which was divided into four sub-cycles and an additional hour for settling, see Figure 5.5.

One Cycle, 24 h				
<b>Sub-cycle 1</b> Anoxic 2 h Oxic 3.75 h	<b>Sub-cycle 2</b> Anoxic 2 h Oxic 3.75 h	<b>Sub-cycle 3</b> Anoxic 2 h Oxic 3.75	<b>Sub-cycle 4</b> Anoxic 2 h Oxic 3.75 h	Settle 1h

Figure 5.5 SBR cycle, including four sub-cycles, applied during the first 80 days of the study (Spagni & Marsili- Libelli, 2009).

During the anoxic phases, sodium acetate trihydrate was added in order to increase the amount of COD needed for denitrification (Spagni & Marsili- Libelli, 2009). Addition of new leachates can affect the Oxygen Reduction Potential  $ORP$  and the pH, causing measurement errors in the control system. Therefore, acetate was added in this initial phase to avoid misreading. Phosphorus was also added in this phase to maintain a  $P_{tot}$  concentration of 0.5–1.0 mg/L in the reactor tank.

Removal of nitrogen reached approximately 95 % and COD was reduced by 20–30 % (Spagni & Marsili- Libelli, 2009). During the first 80 days of the experiment  $NO_3^-$  was the main fraction of the  $N_{tot}$  whilst  $NO_2^-$  was detected in concentrations lower than 0.5 mg/L. When the control system was applied,  $NO_2^-$  increased at once and  $NO_3^-$  decreased to concentrations below 1.0 mg/L. Even so, the nitrification efficiency reached 98 % during the study. The results indicated that it is possible to build up  $NO_2^-$  by adjusting the length of the oxic phase. The nitrite path made it possible to optimise the nitrogen removal process by reducing the amount of external COD added to complete the denitrification by almost 35 %.

### 5.3.4 CASE 7 – Corsa landfill, Spain

Ganiqué et al (2009) performed a study of the possibility to treat leachates with high concentrations of  $NH_4^+$  during 2009 at the University of Girona in Spain. The study pointed out that, since the characteristics of leachates often contain high  $NH_4^+$  concentrations and low organic content of biodegradable substance, a conventional treatment of nitrification and denitrification will become expensive. The aim was to

analyse the SBR technique as a pre-treatment before treatment in order to minimise the costs of additional carbon source and high O<sub>2</sub> demand.

The leachates studied during this project originated from Corsa landfill in Spain with high mean NH<sub>4</sub><sup>+</sup> concentration of 3772 mg/L (Ganigué, et al., 2009). The mean COD and BOD concentration were 4357 and 810 mg/L respectively, indicating that the leachates had a low BOD/COD ratio of 0.18.

The SBR pilot plant was designed to encourage partial nitrification, which implies that approximately half of the incoming NH<sub>4</sub><sup>+</sup> is oxidised to NO<sub>2</sub><sup>-</sup> (Ganigué, et al., 2009). The pilot plant SBR had a tank volume of 250 litres with an exchange ratio of approximately 0.5 and the temperature of the incoming leachates was maintained at 36°C in order to achieve NO<sub>2</sub><sup>-</sup> build-up, or inhibit growth of nitrifying bacteria's. The DO concentration was minimised to 2 mg/L during aeration and a maximum of pH 8 was adjusted by addition of hydrochloric acid. Since the SBR was designed according to the step-feed method - where the total cycle of the SBR involves multiple anoxic and oxic phases - it had a total cycle time of 24 hours with 14 sub-cycles á 100 minutes.

During the first period of the SBR process, from day one to day 59, no additional carbon source was added to the denitrification process (Ganigué, et al., 2009). This resulted in an insufficient oxidation of NH<sub>4</sub><sup>+</sup>. During the second period, from day 59 to day 220, bicarbonate was added directly to the tank. There was however some problems with the dosage apparatus, resulting in a third period, from day 220 to day 450, where the bicarbonate was added as solids into the pre-treatment tank instead. The denitrification efficiency was however affected negatively when solid bicarbonate was added, resulting in a declining performance.

The results from a DNA isolation and screening showed that the SBR was successful in forming the proper conditions for the Anammox bacteria to thrive (Ganigué, et al., 2009). However, it also showed that some nitrifying bacteria's still were able to grow in the reactor tanks. The average N<sub>tot</sub> removal of the denitrification ranged between 15–20 % of the incoming amount, which declined to approximately 5 % during the third period of the process. The study showed that the anoxic-oxic step-feed cycle design of the SBR reduced the N<sub>tot</sub> concentration, but that the denitrification process still needed further improvement.

### **5.3.5 CASE 8 - Puigpalter landfill, Spain**

A SBR pilot plant was studied in Spain during 170 days in order to analyse the techniques reliability to biologically remove nitrogen from landfill leachates (Monclús, et al., 2009). Focus was also addressed on optimising the operational process and to reduce the addition of carbon and bicarbonate. The leachates were received from the urban landfill site, Puigpalter, located in Banyoles, Girona, in the north-east region of Spain. It was characterized by very high concentrations of ammonium NH<sub>4</sub><sup>+</sup> and very low concentrations of biodegradable organic matter.

The lab scale SBR was constructed on site and consisted of a stainless steel square reactor with a volume of 1,000 litres and a capacity to treat from 220 to 300 litres of leachates per day (Monclús, et al., 2009). An automatic controller measured the level of DO and kept it at 2.0 mg O<sub>2</sub>/L during the oxic phase. Also, the performance of the SBR was continuously monitored by a control system which measured the pH, ORP, DO and temperature. Electromagnetic pumps added bicarbonate in order to increase the alkalinity and methanol to enable the biological processes to occur.

The experimental period was divided into three different periods which made it possible to optimise the operation of the SBR (Monclús, et al., 2009). Period 1 *PI* lasted for 36 days and had a cycle time of eight hours. The aim of this period was to adapt the process to the present conditions on site. Period 2 *PII* lasted for 92 days with an unchanged cycle time of eight hours and aimed to improve the efficiency of the biological processes. The third and last period *PIII* lasted for 47 days and focused on stabilizing the removal rate of nitrogen. During *PI* two anoxic feeding steps were implemented to optimise the usage of the limited amount of raw biodegradable organic matter in the leachates. In *PII*, the duration time of the two oxic phases were prolonged by almost 50 % which made the nitrification process more efficient. Due to low values of carbon and alkalinity, methanol and bicarbonate were added in this period as well. During *PIII*, two additional feeding steps were implemented and the cycle time was extended to 12 hours. The different periods and related cycle times and phases are illustrated in Figure 5.6.

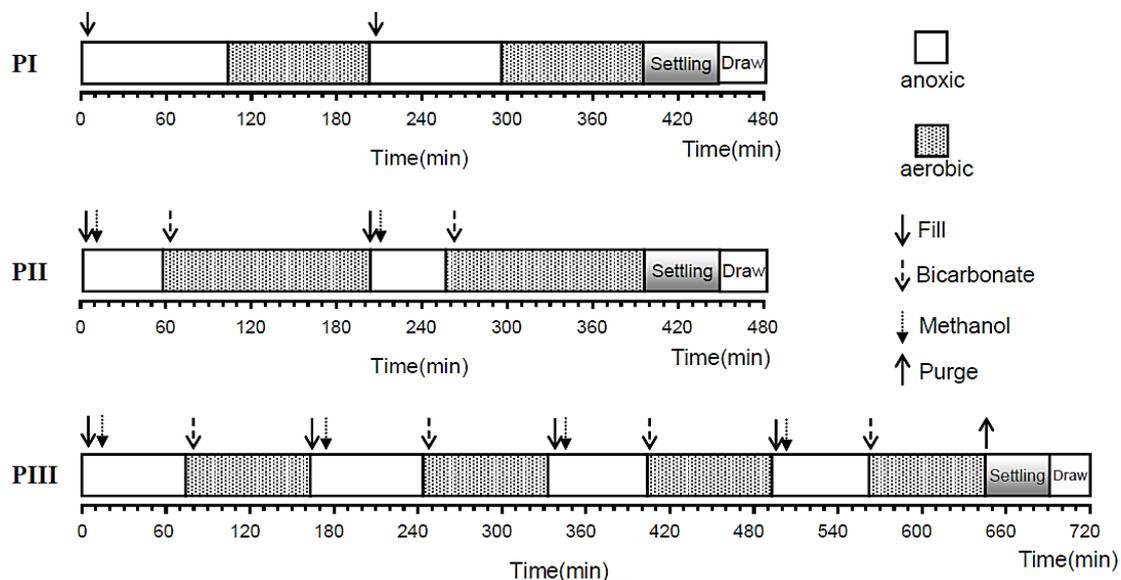


Figure 5.6 The SBR periods and related cycles (Monclús, et al., 2009)[Licensed by author].

Nitrification and denitrification efficiency reached 53 % respectively 10 % during *PI* and increased to above 90 % and 70 % at the end of *PII* (Monclús, et al., 2009). Due to the addition of two extra feed steps in *PIII*, the nitrification and denitrification efficiencies were increased to 95 % and 98 % respectively. The study concluded that it is possible to minimize the usage of external alkalinity and carbon sources by applying the method of step-feed and alteration of the anoxic and oxic phases. The study also showed that it required some time to adapt the system to addition of methanol.

## 5.4 SBR results applicable for case Brudaremsen

In this section, valuable recommendations and practical knowledge regarding the operation, design and performance of the studied SBR's are presented. Gained information was considered during the experimental trial with lab scale SBR's at Brudaremsen.

Throughout the cases, emphasis was on optimising the SBR process including altering the length and number of phases as well as changing the dosages of chemicals. These operational parameters are the most commonly altered and optimised throughout the study periods. For example, during CASE 1 and 2 excess carbon in the SBR was reduced by adding a 2<sup>nd</sup> oxic phase. In CASE 6 the nitrogen removal rate was increased by adjusting the length of the oxic phase, which made it possible to build up  $\text{NO}_2^-$  and thereby create a nitrite path. By altering the oxic and anoxic phases - called the step-feed method - which was used in CASE 7 and 8, it was possible to increase the removal rate of  $\text{N}_{\text{tot}}$  and the nitrification and denitrification rate reached 95 % and 98 % respectively. By optimizing the various phases the studies have shown that it is possible to reduce the amount of added chemicals to the SBR process.

There are many aspects concerning the dosage of an external carbon source to the SBR which should be taken into consideration. For example, as presented in CASE 2 it is advantageous to start the dosage of external carbon ten minutes after the oxic phase has stopped, since this will avoid that the carbon source becomes digested by the aerobic microorganisms (Ganigué, et al., 2009). Also, according to Monclús et al. (2009), CASE 8, it might take some time to adapt the system to the addition of carbon. Kulikowska & Klimiuk (2004), CASE 5, states that the nitrogen/COD ratio in the landfill leachates affect the treatment process and that a higher ratio will result in a higher  $\text{NH}_4\text{-N}$  removal rate. In order to optimise the nitrification process and to find the optimal nitrogen/COD ratio one method is to test and analyse different dosages of external carbon. During the study performed by Kulikowska & Klimiuk (2004), the optimal nitrogen/COD ratio was found to be 1.08.

The loading of leachates to the SBR tank and the HRT can also be optimised in order to achieve sufficient treatment. If the influent volume is rapidly increased it might cause accumulation of  $\text{NO}_2^-$  in the SBR, due to a drop in  $\text{O}_2$  concentration during nitrification. This might also occur if the  $\text{O}_2$  concentration falls below two milligrams  $\text{O}_2$  per litre (Ganigué, et al., 2009). Due to this, it might be advantageous to increase the influent volume stepwise during several stages. As presented in CASE 5, the HRT in the SBR was shown to influence the COD removal efficiency. But, as for the dosage of external carbon source it is of interest to perform trials and analyse the result of different HRT. According to Kulikowska & Klimiuk (2004), a too low HRT should be avoided since it can cause accumulation of  $\text{NO}_2^-$ . According to Comett-Ambriz et al. (2003), CASE 4, the length of the settle phase can also affect the COD removal rate. A shorter settling time resulted in a lower COD removal rate.

During analysis of the SBR treatment efficiency, it is important to take into account that the amount of precipitation before and during a study, can affect quality and quantity of the incoming leachates. This was observed in CASE 3 during which it was conducted that due to a continual rain period before sampling, the leachates had been diluted considerably (Junestedt, et al., 2003). Finally, it is important to consider that the data, results and recommendations from the cases are not directly applicable for the case Brudaremsøen. Characteristic parameters in the leachates differ significantly for some cases if compared to the leachates from Brudaremsøen landfill. Also, the climate, average operational temperature, age of the studied landfill etcetera varies throughout the cases.



## 6 Description of case study area Brudaremossen

Brudaremossen is a landfill area situated close to the city of Gothenburg on the Swedish western coast. The landfill is located in the Delsjö area, five kilometres east of the city centre, and is surrounded by a natural reserve, including several forest lakes and a recreational area, see Figure 6.1 (Ramböll, 2015 A).

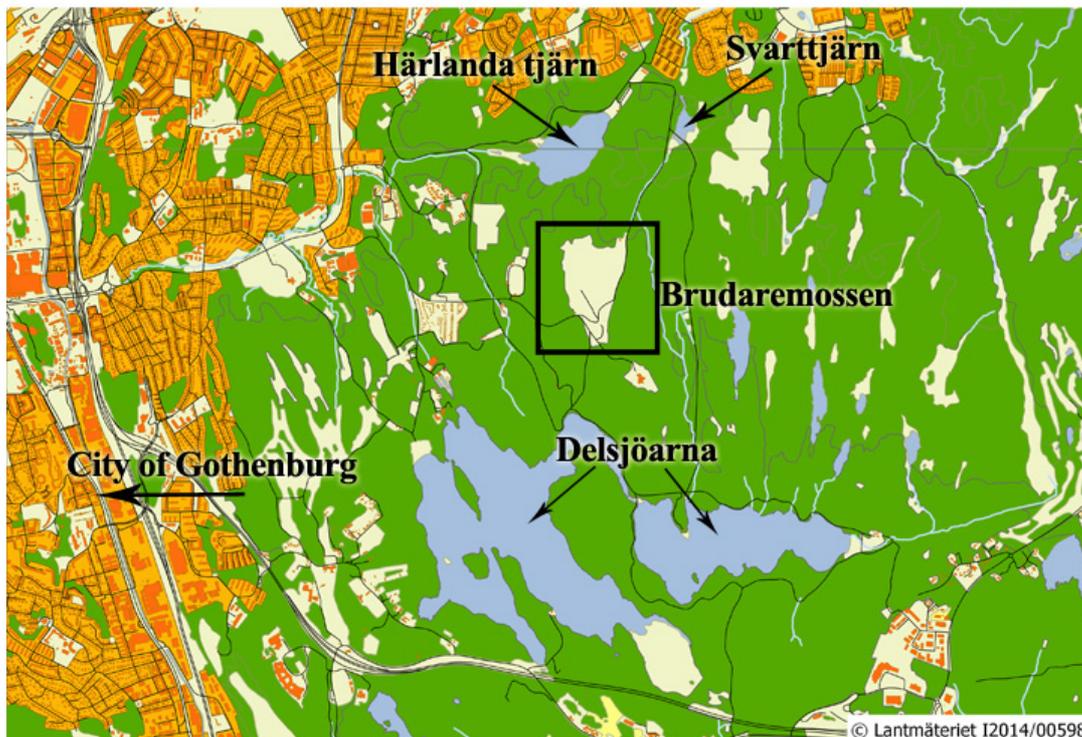


Figure 6.1 The area of landfill Brudaremossen, outlines of the area is highlighted with rectangular shape [background map from (Lantmäteriet, 2014)].

The deposit was operative from 1938 to 1978 and was until 1972 the main waste disposal site for the municipality of Gothenburg. During that period, numerous different wastes were deposited at Brudaremossen but the exact volumes unclear since the documentations from that period are insufficient. However, some records show that waste from industries, households, construction sites as well as sludge, oil, various chemicals, excavated material and carcasses have been disposed at Brudaremossen (Fersters, 2003; Ramböll, 2015 B). The incautious deposition of chemicals and oil caused several fires on site and pollution of the nearby creek Finngösabäcken during the 1950's and 60's. To prevent additional accidents, specific lagoons and pits were built where oil, sludge containing toxic metals and chemicals were collected; see Figure 6.2 (Ramböll, 2015 B).



Figure 6.2 Oil lagoon on the crest of Brudaremossen- aerial photograph in 1977 (Ramböll, 2015 B)[permitted by author].

Today the deposit covers an area of approximately 25 hectares and forms the shape of a hill 130 meters above sea level (Krewer & Moona, 2012). A clay layer has been applied to cover the hill and the vegetation consists of grass and some lower bushes. The waste masses can roughly be estimated to four million cubic meters and the maximum depth of the waste is approximately 40 meters (Jansson, et al., 2012; Ramböll, 2015 C). A large part of the landfill is placed directly above an old moss area but some parts rest directly above gneissic rock. The substrate can be assumed to be relatively impermeable with the only concern being the sections of gneissic rock. It has been observed that the rock contains a lot of slate which might lead to fractions; hence a risk of leakage of leachates through the rock (Fersters, 2003). Due to the age of Brudaremossen landfill, it is likely that it has reached the more stabilised methanogenic phase (Krewer & Moona, 2012).

Situated to the south of the landfill are the two lakes Delsjöarna, see Figure 6.1, that serves as the main water reserve for the municipality of Gothenburg (Fersters, 2003). Due to the negative characteristics of the leachates that percolate from the landfill, there have been concerns regarding the risk of pollution of the lakes. However, according to Jansson et al in (2012) there is a low risk for pollution of Stora Delsjön. This risk assumption was based on measured concentrations of pollutants in the groundwater and leachates surrounding Brudaremossen. The results did however show an indication of groundwater flow from the deposit towards north, which suggested a risk for leachates affecting Lake Svartjärn and Lake Härlanda tjärn.

Numerous of improvements have been applied to the landfill Brudaremossen in order to minimise the risk of leachates affecting the surrounding environment. For example, during the 40's, a concrete screen was constructed in the soil to the south of the landfill, acting as a wall to protect the groundwater and main water reserve from pollution (Fersters, 2003; Johansson & Johansson, 1973). During the 60's, a drainage system was installed in order to enable collection of leachate and runoff. Problems with excessive landfill leachate from Brudaremossen lead to a decision in 1991 to prevent too high inflows to the landfill during precipitation. This was solved by applying a clay layer on top of the landfill (Fersters, 2003). However, only 50 % of the total landfill's surface area was concealed with clay at that time and further clay layers was sporadically placed on top of Brudaremossen until 1995. In order to analyse the constituents in the leachates frequently, further automatic measurement devices were installed from 2003 to 2008. During that time, more ditches as well as pipelines were created to prevent intrusion of surface runoff into surrounding lakes. Also, an oil and sludge separator was installed near the intake of the pipeline leading to the Wastewater Treatment Plant *WWTP* Gryaab (Krewer & Moona, 2012).

Figure 6.3 illustrates the existing treatment on site for Brudaremossen landfill leachates, starting with separation of oil and sludge. Subsequently, the water enters a leachate pond where the flow is static. Water is then discharged into a well in the northern part of the pond and further channelled through the municipal sewage system to the *WWTP* Gryaab. An existing wetland lies further north of the leachate pond, see Figure 6.3. During periods with extensive precipitation, it is possible to discharge some of the leachates from the pond into the wetland to avoid flooding (Ramböll, 2015 B).



Figure 6.3 The existing treatment of landfill leachates; oil and sludge separator followed by a leachate pond and wetland area situated to the north of Brudaremossen [background map from (Lantmäteriet, 2014)].

## 6.1 Leachate quality at Brudaremossen

Measurements and analysis of the incoming leachates after the oil and sludge separator were performed by Ramböll, four times each year, from 2008 until 2013. Organic pollutants, such as Petroleum Hydrocarbons *PHC*, Polycyclic Aromatic Hydrocarbons *PAH*'s and Polychlorinated Biphenyls *PCB*'s, were detected during the measurements in concentrations higher than the threshold values recommended by the City of Gothenburg (Ramböll, 2015 C). High concentrations of PAH and PCB in water environments has shown to cause negative effects, such as carcinogenic or hormonal disrupting, on aquatic organisms (Grolander, 2010; U.S.EPA, 2013).

According to the measurements by Ramböll from 2008 to 2013 the untreated leachates at Brudaremossen have a neutral pH, high concentrations of iron *Fe* and organic substances in comparison with unaffected surface water. Very high values of total nitrogen  $N_{tot}$  were detected, and the main fraction was ammonium nitrogen  $NH_4-N$  (Ramböll, 2015 C). Such high concentrations, which can be found in the leachates at Brudaremossen, can cause eutrophication in the nearby environment which eventually can lead to depletion of oxygen  $O_2$  (Butkovskyi, 2009).

Other leachate parameters that exceeded the threshold values demanded by the City of Gothenburg or the European Directive 2013/39/EU for surface water or salmon fishing were; Total Organic Carbon *TOC* and Total Phosphorus  $P_{tot}$  (Ramböll, 2015 C). The yearly average of mentioned parameters in untreated leachate exceeded the threshold values more than twice and was therefore considered high. In Table 6.1, the average concentrations of parameters measured by Ramböll from 2008 to 2013 are presented (Ramböll, 2015 A). Values exceeding a certain threshold value are marked and referred accordingly.

Table 6.1 Average concentrations from 2008 to 2013 measured at Brudaremossen in the leachates after the oil and sludge separator (Ramböll, 2015 A).

Parameter	Average value	Unit	
pH	7.50	-	Below guideline value for the city of Gothenburg
Temp	8.27	°C	
Alkalinity	1082	mg/l	
COD Mn	25.20	mg/l	
TOC	58.93	mg/l	Above guideline value for the city of Gothenburg
$O_2$	7.61	mg/l	
$P_{tot}$	0.14	mg/l	Above guideline value for the city of Gothenburg
$NH_4-N$	74.01	mg/l	Exceeding threshold value SPS 2001:554
$NH_3-N$	0.28	mg/l	
$N_{tot}$	78.47	mg/l	Above guideline value for the city of Gothenburg
$PCB_{tot}$	0.26	µg/l	Value above guideline value for the city of Gothenburg
$PAH_{sum}$ carcinogenic	3.24	µg/l	
Benzo(a)pyrene	0.02	µg/l	Exceeding threshold value Directive 2013/39/EU
Fluoranthene	0.17	µg/l	
Oil index	0.14	mg/l	Below guideline value for the city of Gothenburg

Measurements at Brudaremossen show that the quantity of outflowing leachates from the deposit is rather constant. In 2012, the leachate runoff was approximately 500 litres per square meter. During heavy rains the flow increases rapidly during a shorter period, which indicates that the system is easily affected by storm water (Ramböll, 2015 B). Therefore, it is likely that the current precipitation affect both the quantity and quality of the outflowing leachates from the deposit.

## 6.2 Potential recipients for the treated leachates

During the beginning of 2015, a pilot plant was constructed at Brudaremossen, in this report entitled the Brudaremossen pilot plant 2015. The initiator for the construction was the City of Gothenburg. Due to the REVAQ certification implemented on the sludge produced at the WWTP Gryaab, the board of directors at Gryaab decided that the plant would not receive leachates from Brudaremossen after year 2018 (Moberg, 2013). This implied that the City of Gothenburg became responsible for treating the leachates after this date. The municipal division *Kretslopp och vatten* commissioned the consultancy firm Ramböll to investigate possible techniques for local treatment of the leachates from the landfill and to analyse potential recipients, including current status and possible discharge locations (Ramböll, 2015 B). The investigation aimed to provide necessary data in order to be able to construct and operate a permanent treatment facility on site.

Six possible discharge points, involving four different recipients, for the treated leachates from Brudaremossen were identified by Ramböll on behalf of the City of Gothenburg, see Figure 6.4 (Ramböll, 2015 C). The available options were:

- The WWTP Gryaab
- The river Göta älv
- The stream Säveån
- Lake Svarttjärn & Finngösabäcken- discharge into Säveån
- The stream Mölndalsån
- The creek Delsjöbäcken- discharge into Mölndalsån

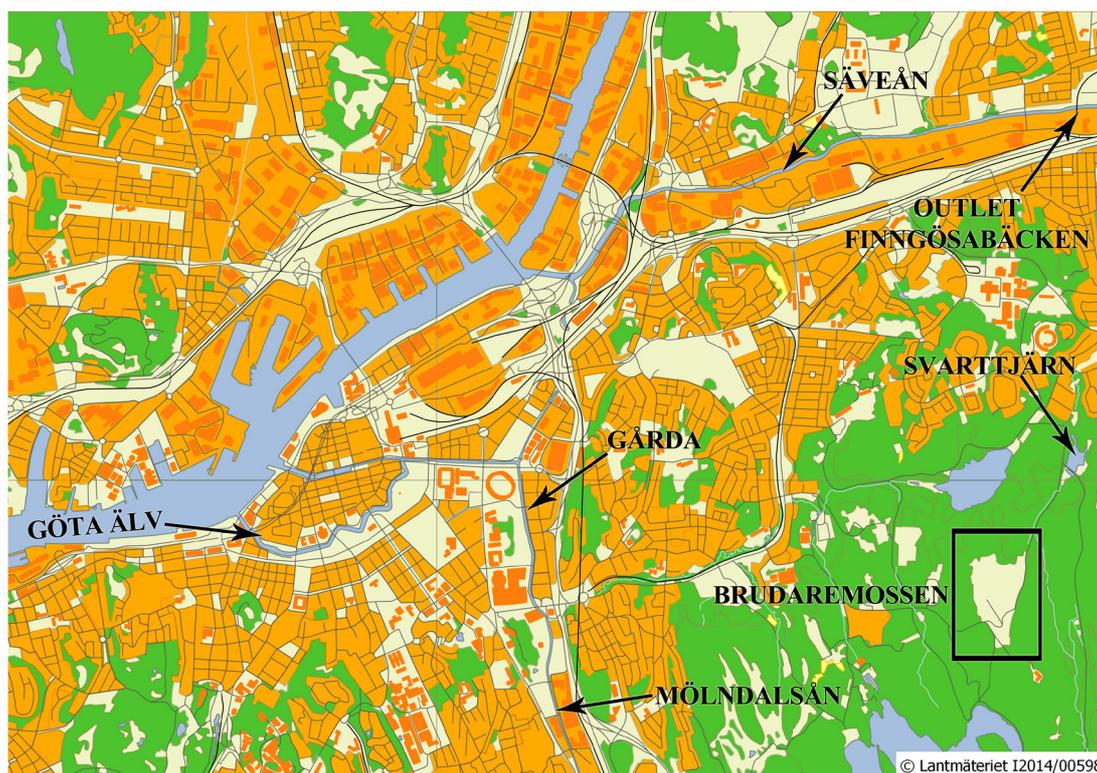


Figure 6.4 Map with possible recipients for the treated leachates from Brudaremossen [private illustration (background map from (Lantmäteriet, 2014))].

Each recipient applicable for the case Brudaremossen requires a certain quality of the treated leachate before discharged. The guideline values for each recipient, as well as Finngösabäcken and Delsjöbäcken, are summarised in Table 6.2. All the possible recipients and related requirements are further explained in the paragraphs below.

Table 6.2 Average concentrations between 2008 and 2013 in the leachates after the oil and sludge separator together with threshold values for recipients (Ramböll, 2015 A; Ramböll, 2015 C).

Parameter	Unit	Average value	WWTP	Göta älv	Säveån	Finngösa bäcken	Delsjö bäcken	Möldals ån
N <sub>tot</sub>	mg/l	78.47	-	30	30	30	-	-
NH <sub>4</sub> -N	mg/l	74.01	-	10	3.0	3-10	3	3-10
P <sub>tot</sub>	mg/l	0.14	-	0.50	0.50	0.50	0.3	0.30
BOD <sub>7</sub>	mg/l		25	15	15	15	15	15
TOC	mg/l	58.93		-	50	50	50	50
MLSS (supernatant)	mg/l	29.53	-	25-30	25-30	25-30	25-50	25-50
PCB <sub>tot</sub>	µg/l	0.26	-	0.05	0.05	0.05	0.05	0.05
PCB <sub>7</sub>	µg/l	0.05	-	0.01	0.01	0.01	0.01	0.01
Benzo(a)pyrene	µg/l	0.02	-	0.02	0.02	0.02	0.02	0.02
Fluoranthene	µg/l	0.17	-	0.10	0.10	0.10	0.10	0.10
Oil index	mg/l	0.14	-	1.00	1.00	1.00	1.00	1.00

### 6.2.1 The WWTP Gryaab

If it is decided to distribute the locally treated leachates to the WWTP Gryaab, the characteristic values of the treated water must fulfil the requirements according to the REVAQ certificate. This implies a reduction of organic pollutants as well as toxic metals.

### 6.2.2 The river Göta älv

The river Göta älv runs through the central part of the City of Gothenburg where the nearby urbanisation has influenced the river significantly, see Figure 6.4 (Ramböll, 2015 B). In 2014, the ecological and chemical status of the river was defined as unsatisfied respectively not achieved (VISS, 2014). Göta älv has been classified as a priority class one according to a water protection plan implemented by the City of Gothenburg. The classification is a part of the water plan *Vattenplan* conducted by the City of Gothenburg with the aim to protect and implement measurements for stormwater recipients. It consists of a scale from one to four where one indicates the highest priority class (Ramböll, 2015 A).

If decided to release the treated leachates into Göta älv, the oil index and the organic substances has to be reduced to acceptable levels (Ramböll, 2015 C). Also, Table 6.2 show that there are regulations concerning the level of  $N_{\text{tot}}$  but not for  $NH_4\text{-N}$ .

### 6.2.3 The stream Säveån

Discharge of treated leachates to the stream Säveån would most likely take place in the lower part of the stream along the stretch Finngösa, in the municipality of Partille east of the landfill, see Figure 6.4 (Ramböll, 2015 B). This lower part of Säveån and its surroundings has been defined as a Natura 2000 site since it has a significant value for the reproduction of an Atlantic salmon stock and as a nesting spot for the Kingfisher (Länsstyrelsen Västra Götalands Län, 2005). Also, Säveån has been classified as a priority class one according to *Vattenplan*, implemented in the city of Gothenburg (Ramböll, 2015 A).

If the stream Säveån would be chosen as recipient there are specific demands for the concentrations of the parameters released into the river due to the Natura 2000 regulations. The acceptable levels are presented in Table 6.2, and are based on guideline and threshold values summarised by Ramböll (Ramböll, 2015 C).

### 6.2.4 Lake Svarttjärn & Finngösabäcken

Lake Svarttjärn has been observed to endorse a rich variety of birds and has a great recreational value (Tamulénas, 1983). From Lake Svarttjärn towards the stream Säveån runs the brook Finngösabäcken. The brook is a part of the watershed for the stream Säveån, a water protected Natura 2000 site, hence certain regulations must be taken into consideration for discharge into Lake Svarttjärn (Ramböll, 2015 B).

Finngösabäcken has been classified as priority class four based on its sensitivity and load of stormwater according to *Vattenplan*. Even though Finngösabäcken has a low priority class the point of discharge lies within the municipality of Partille and the level of pollution in the creek is unknown, hence further investigations could become necessary (Ramböll, 2015 A).

### 6.2.5 The stream Mölndalsån

Mölndalsån runs through the central part of Gothenburg and has been highly affected, both chemically and physical, by the surrounding urban landscape. Implemented measures have steadily improved the conditions in the stream and problems with eutrophication and level of pollutions has been reduced. It has a priority class two according to *Vattenplan* and should therefore be protected against contamination (Ramböll, 2015 A).

### 6.2.6 The creek Delsjöbäcken

Delsjöbäcken is one tributary flow to the stream Mölndalsån. The possible point of discharge is situated in a brook ravine close to an allotment area. Current levels of critical parameters in the creek are at the moment unknown, but it is likely that a discharge of treated leachates from Brudaremosen will increase the concentration of nutrients (Ramböll, 2015 A). Delsjöbäcken has been assigned priority class two according to *Vattenplan*, since its aquatic fauna and species has been considered rich and valuable from a preservation point of view (Ramböll, 2015 B).

## 6.3 Pilot plants at Brudaremosen

During a Master's Thesis by Krewer & Moona (2012), a small pilot plant was constructed in order to test treatment of the leachates from Brudaremosen with two series of sorption columns. The sorbing materials chosen for the study was Sphagnum peat moss and Granular Activated Carbon *GAC*. The lastly mentioned is a very porous material made out of materials which have a high content of carbon, for example coal and wood. Its high porosity makes it very efficient to adsorb various components from an effluent since it provides a large surface area compared to other filters. It is common to place the *GAC* filter after a conventional filtration process, such as a flocculation/sedimentation step followed by some form of filtration (U.S. EPA, 2015).

In the pilot plant, one series consisted of two columns with *GAC* and the other of a peat moss column followed by a column with *GAC*. The aim was to sorb and reduce the organic content in the leachates adequate enough to enable further discharge of leachates to the WWTP Gryaab. (Krewer & Moona, 2012). Both filter series was similar in treatment and could be used to treat the leachate. The results indicated sufficient reduce in oil, PHC and PAH concentrations. The *GAC* filter removal efficiency regarding Dissolved Organic Carbon *DOC* and TOC was evaluated to be higher than for the peat filter. High concentrations of Fe in the leachates did however clog the filters and recommendations from the study were to solve this before installation of a permanent treatment facility at Brudaremosen.

In the beginning of 2015, a new pilot plant was constructed at Brudaremosen. The design of the pilot plant was based on a pre-study conducted by Ramböll in 2014, concerning suitable treatment techniques. The aim with the pilot plant was to study if the selected techniques were suitable to apply at Brudaremosen or if modification with the suggested processes had to be made (Ramböll, 2014). During the pre-study, two different designs of treatment processes were developed based on two possible recipients; the river Göta älv or the WWTP Gryaab. The two different process designs and involved treatment techniques are illustrated in Figure 6.5. The treatment process based on Göta älv as recipient is in this report defined as the *biological process* (I) and the one based on treatment at the WWTP Gryaab is defined as the *chemical process* (II).

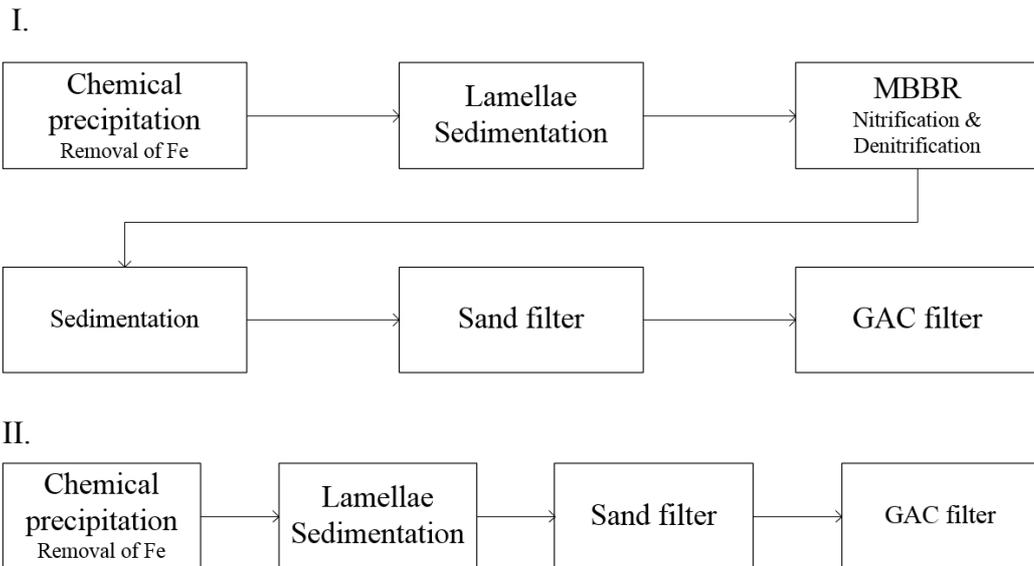


Figure 6.5 I. Suggested design for the biological process, river Göta älv as recipient.  
 II. Suggested design for the chemical process, WWTP Gryaab as recipient (Ramböll, 2015 A).

The pilot plant was therefore constructed so that it was possible to run both process designs simultaneously, for the Göta älv alternative and the WWTP Gryaab alternative respectively. See Figure 6.6 for an illustration of the Brudaremossen pilot plant 2015, including sampling points A-G'. The two process designs were achieved by installing a distribution gear, marked as point B in Figure 6.6, which discharged the flow after the sedimentation through lamellae in three different directions. One part of the effluent was lead directly to the sewer system, one to the biological treatment step in the Moving Bed Biofilm Reactor *MBBR* and the last amount of effluent was distributed to the final sedimentation step through the sand and GAC filters (Ramböll , 2014).

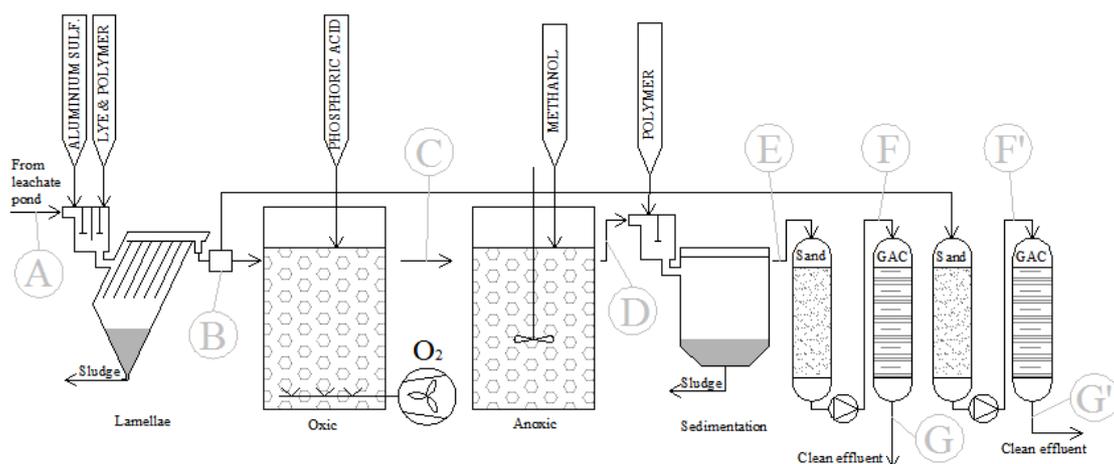


Figure 6.6 Illustration of the Brudaremossen pilot plant 2015 [based on figure from (Ramböll , 2014)].

On site, the pilot plant was constructed in between the leachate pond and the existing wetland – see Figure 6.3 in Chapter 6. The already installed oil and sludge separators

for reduction of oil fractions and organic pollutants in the leachates were utilised as pre-treatment for the pilot plant. Sludge produced in the plant was continuously discharged from the treatment process through time regulated control valves and pumps. The sludge was stored in a tank with a volume of ten cubic meters, equipped with level monitors to avoid too large volumes of sludge. Every two weeks the sludge was collected by a sludge truck (Ramböll , 2014). The treatment techniques used in the pilot plant are described further in the following paragraphs.

### 6.3.1 Lamellae sedimentation at Brudaremossen pilot plant 2015

Chemical precipitation and sedimentation through lamellae were the first treatment steps in the pilot plant, illustrated in Figure 6.6 between sampling points A and B. In this step, the aim was mainly to precipitate Fe from the leachates and thereby reduce the risk of coating or clogging of the subsequent equipment (Ramböll , 2014). The leachates were pumped from the leachate pond into the pilot plant by a recently installed pumping station, and thereafter lead to a tank for chemical precipitation. Aluminium sulphate was used as precipitant, sodium hydroxide (lye) for decreasing the pH and polymers in order to form flocs (Ramböll, 2015 C). Aluminium sulphate is a commonly used coagulant for water treatment and forms flocs which agglomerate into larger flocs able to settle (American Water Works Association, 2010). Large amounts of the flocs were thereafter removed in lamellae sedimentation filters, see Figure 6.7. The flocs formed sludge at the bottom of the lamella tank which was regularly removed through a time- regulated valve (Ramböll , 2014). In addition to Fe removal, the intention with this step was also to reduce the concentrations of suspended material, metals and organic particle bound substances (Ramböll, 2015 C).

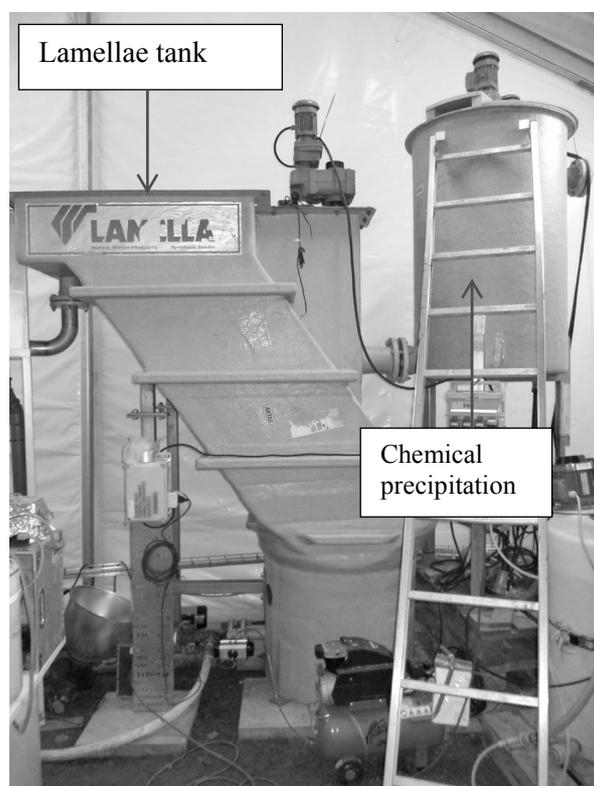
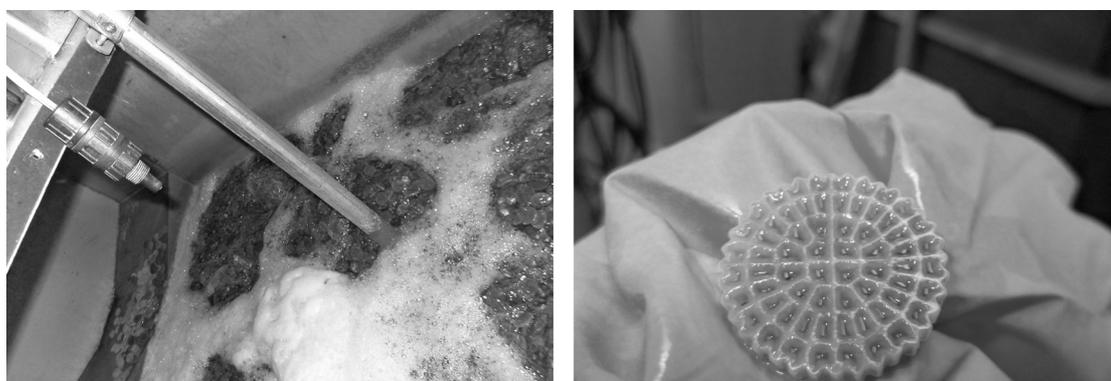


Figure 6.7 The lamellae tank at Brudaremossen pilot plant 2015

### 6.3.2 MBBR treatment at Brudaremossen pilot plant 2015

Biological treatment with MBBR in two tanks with moving bio carriers followed the chemical precipitation and sedimentation step, illustrated in Figure 6.6 between sampling points B and D. The MBBR was hired from the company AnoxKaldnes with a capacity to receive a maximum of 5,900 litres of leachates per day (Ramböll , 2014). The water was first lead into an aerated aerobic zone where nitrification occurred, followed by an anoxic zone with denitrification. It was necessary to add both phosphorus and methanol to the MBBR in order to ensure sufficient nitrogen removal (Ramböll, 2015 C). Plastic polymeric carriers were used both in the nitrification and denitrification tanks in order to provide the bacteria with a sufficient surface area to grow on. See Figure 6.8 for a picture inside the MBBR oxidic tank and a plastic polymeric carrier on site at the pilot plant. For more detailed description about the mechanisms involved in the MBBR technique, see Chapter, 4, Section 4.1.



*Figure 6.8 The MBBR tank at Brudaremossen pilot plant 2015, oxidic tank to the left and a plastic polymeric carrier to the right.*

### 6.3.3 Sedimentation at Brudaremossen pilot plant 2015

Sedimentation was implemented after the MBBR treatment as a pre-step before the final filtration steps, illustrated in Figure 6.6 between sampling points D and E. This took place in a tank equipped with a stirrer and a pump. Sludge assembled at the bottom of the tank was continuously discharged (Ramböll , 2014). See Figure 6.9 for a picture of the sedimentation tank on site at the pilot plant. The aim with the sedimentation was to remove the biological sludge produced during the MBBR treatment, which otherwise could cause clogging of the sand filters (Ramböll, 2015 C).

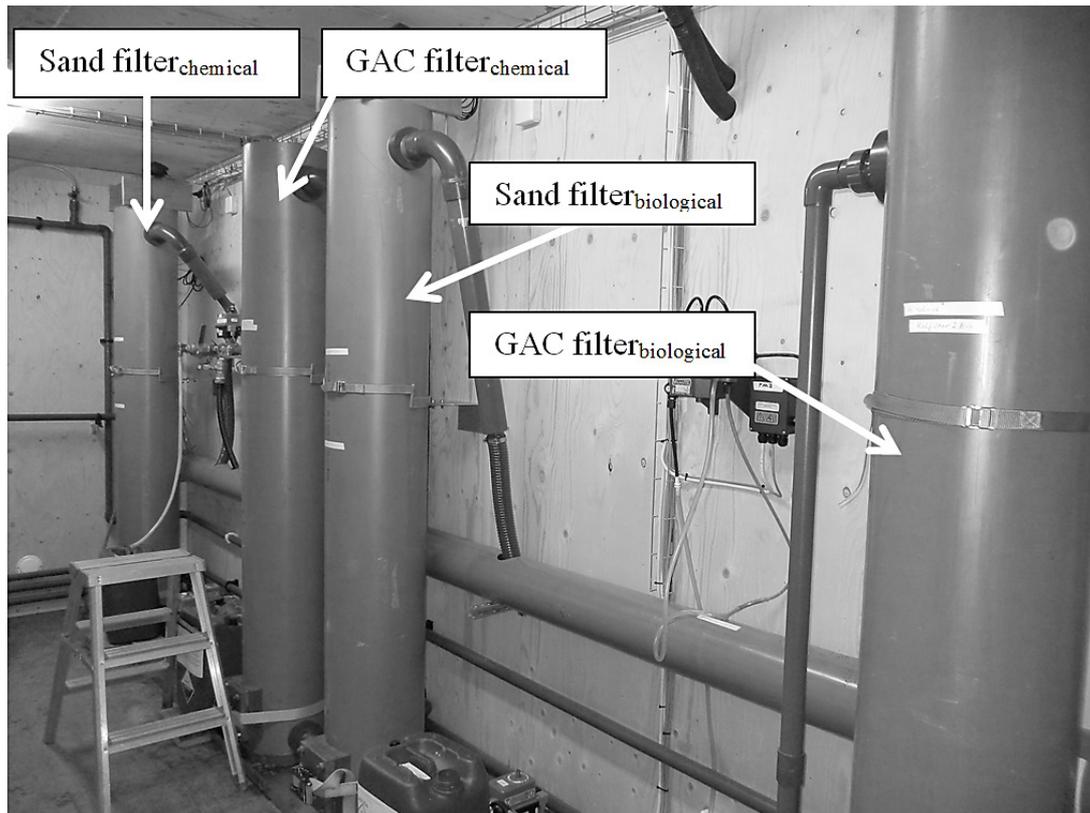


*Figure 6.9 The sedimentation tank at Brudaremassen pilot plant 2015*

#### **6.3.4 Sand and GAC filters at Brudaremassen pilot plant 2015**

Filtration through sand filters was performed with the purpose to reduce the concentration of particles and dissolved organic substances in order to prolong the life length of the subsequent GAC filters (Ramböll, 2015 C). Two sand filters were installed in the facility. The one connected to the biological process received water from the sedimentation step - sampling point E in Figure 6.6 - and the one connected to the chemical process received water from the lamellae step - sampling point B in Figure 6.6. Filtration through GAC filters was the final process step in the plant for removal of dissolved organic substances, see Figure 6.6. Two filters with GAC were installed in the plant following the preceding lines with sand filters (Ramböll, 2015 C).

The two filter types were constructed within plastic tubes with a height of 2.5 meters, a diameter of 30 centimetres and a filter height of 1.5 meters (Ramböll , 2014). Two pumps were installed to enable backwashing of both sand and GAC filters. After treatment through the filters steps the treated leachates were discharged into the already existing sewer system and channelled to the WWTP Gryaab. See Figure 6.10 for a picture of the filters on site at the Brudaremassen pilot plant 2015.



*Figure 6.10 The sand and GAC filters at Brudaremsen pilot plant 2015*



## 7 Experimental trial with lab scale SBR's

An experimental trial with two lab scale Sequencing Batch Reactors *SBR's* was conducted during 50 days, from the 11<sup>th</sup> of March until the 29<sup>th</sup> of April. This chapter will describe the methods applied during the trial including the design of the lab scale *SBR's*, cycle times, construction and the continuous operational work. It will also provide the approach for calculating the required chemical dosages. Thereafter, the procedures and standard methods used for sampling and analysing of the water and sludge samples will be explained.

Finally, the process used for calculating and estimating the design of a full scale *SBR*, based on the result from the experimental trial of the lab scale *SBR's*, will be explained.

### 7.1 Design of the lab scale *SBR's*

In order to investigate how much toxic metals or organic pollutants affect or might inhibit the biological processes in the *SBR*, two duplicates of a lab scale *SBR* were constructed defined as *Reactor 1* and *Reactor 2*. *Reactor 1* was fed with leachate from point B, after the lamellae in the Brudaremissen pilot plant 2015 - see Figure 6.6 in Chapter 6, Section 6.3. *Reactor 2* was fed with leachate from point G', after the Granulated Activated Carbon *GAC* filter - see Figure 6.6 in Chapter 6, Section 6.3.

The design of the two lab scale *SBR's* was planned during February at Chalmers University of Technology together with the supervisor for the project. The initial design and operation - cycle time, number and length of the chosen phases - of the lab scale *SBR's* were decided based on previous knowledge of wastewater treatment *SBR* trials at Chalmers and data gathered from the literature study on previous lab scale *SBR* trials. Figure 7.1 show the process design of one lab scale *SBR* including the five determined phases for one cycle at exchange ratio 0.50. The phases were; fill (1), react oxic (2), react anoxic (3), settle (4) and draw (5).

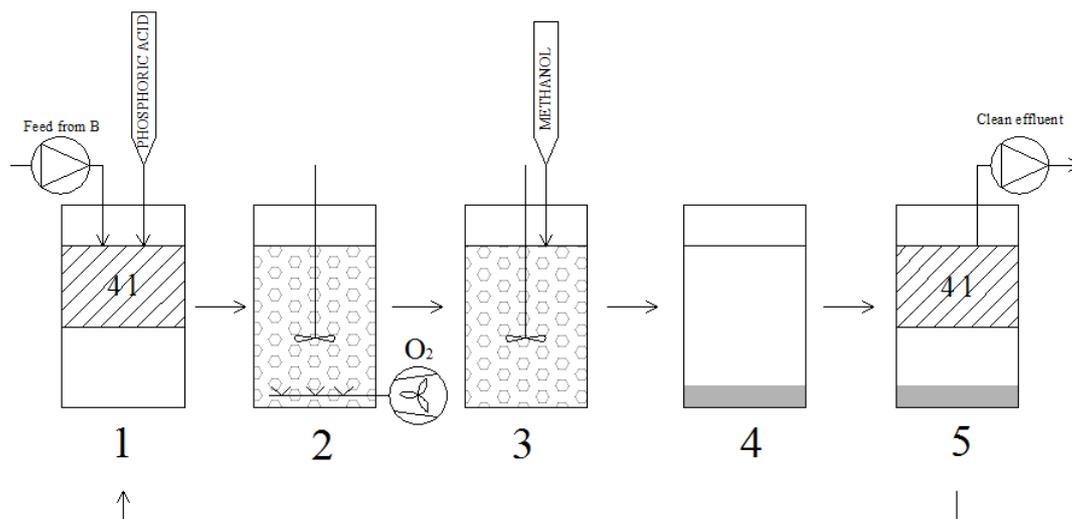


Figure 7.1 Phases included in the process design of the lab scale *SBR* treatment

The total cycle time was four hours, which resulted in six cycles performed per day. Maximum flow of incoming leachate to each lab scale SBR was 168 litres per week, hence 24 litres each day. Table 7.1 presents the duration of the five phases and gives a description of the addition of chemicals or mechanisms taking place during each phase.

*Table 7.1 Phases for one cycle of the lab scale SBR treatment, what to add during the phases, the amount of dosage and duration of the phases*

Phase order	Phase	Addition of	Duration [h]
1	Fill	Leachate	0.50
		Phosphorus	0.50
2	React Oxidic	Air	1.75
		Stirring	1.75
3	React Anoxic	Stirring	1.00
		Methanol	2 [minutes]
4	Settle	-	0.50
5	Draw	Extraction of treated water	0.25

### 7.1.1 Lab scale SBR construction and operation

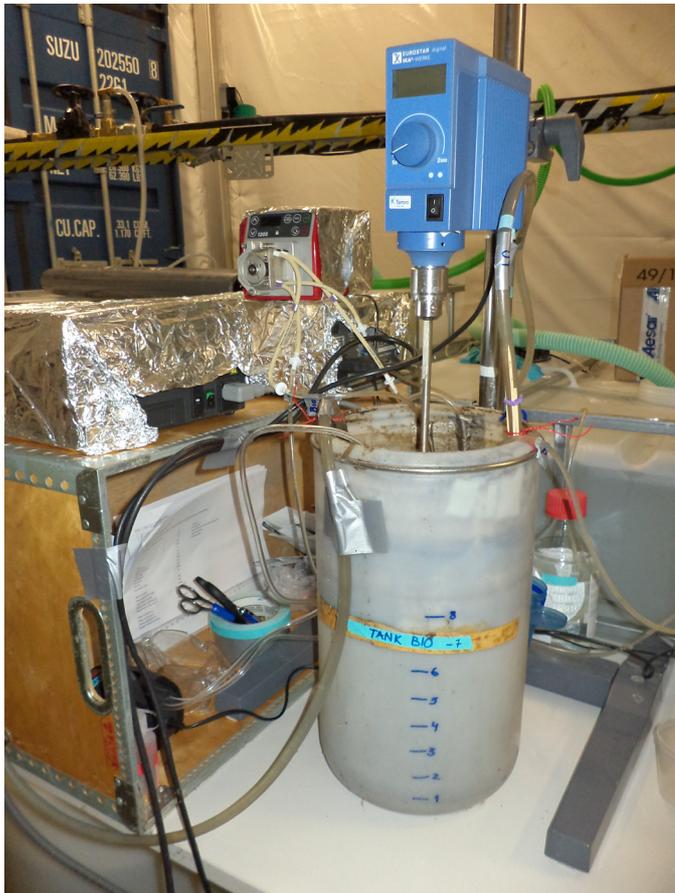
The lab scale SBR's were installed the 11<sup>th</sup> of March at Brudaremsen and were operated during two months, until the 29<sup>th</sup> of April. Activated sludge used for the experimental trial was retrieved during day one from the Waste Water Treatment Plant *WWTP* in Hammargård, Kungsbacka. Before installing the SBR's on site, the capacities of all the pumps were tested. This was done in order to obtain the correct Revolutions per Minute *RPM* corresponding to the right velocities for inflow and outflow of leachates and chemicals. Table 7.2 describes the materials that were needed in order to construct the two lab scale SBR's.

*Table 7.2 Material for constructing the two lab scale SBR's*

Material	No. of	Size / Properties
Feed tank	2	100 L
Reactor tank	2	10 L
Stirring paddles	2	
Pumps for outgoing leachate	2	
Pumps for incoming leachate	2	
Dosage pump methanol	1	
Dosage pump phosphorus	1	
Air pumps	2	
Glass bottles for chemicals	4	
Timers	5	With possibility to program an on/off time schedule

All equipment was connected to timers which were programmed to run automatically. Maintenance of the treatment during the study included continuous, manual sludge withdrawal, refill of the feed tanks and the enclosed glass bottles containing chemicals. Concerning the Sludge Retention Time *SRT*, according to theory it is

preferable to have a longer SRT when operating at lower temperatures. Since the lab scale SBR's operated in a rather cold climate - 8 °C to 13 °C – it was decided to have a SRT of 32 days. Based on the desire to achieve a SRT of 32 days and a Mixed Liquor Suspended Solids *MLSS* concentration of 3.5 mg/L, it was calculated that 1.75 litres of mixed sludge should be removed each week. To avoid clogging of the tubes and aeration equipment, these were regularly controlled and cleansed if needed. See Figure 7.2 for one of the lab scale SBR's after installation at Brudaremassen.



*Figure 7.2 One of the lab scale SBR's successfully installed at Brudaremassen*

### **7.1.2 Stepwise increase of the exchange ratio**

In order to adapt the bacteria in the sludge retrieved from the WWTP in Hammargård, Kungsbacka, to the conditions on site at Brudaremassen, the exchange ratio was increased stepwise. During the first 29 days, the pumps were set on a RPM related to an exchange ratio of 0.17. From Day 29 to Day 36 the RPM for each pump was increased based on an exchange ratio of 0.33. Finally, the exchange ratio was increased to 0.50 which was kept during the remaining 14 days. The three different exchange ratios and related RPM's for each pump are presented in Table 7.3. How the exchange ratios were calculated for each exchange ratio is described in Appendix A.

Table 7.3 The various exchange ratios and related RPM settings for each pump.

Exchange ratio	RPM-inflow	RPM-outflow	RPM-Phosphorus	RPM-Methanol
0.17	10-13	29-30	1	4
0.33	23-25	52	2	7
0.50	36-37	75-76	3	11

## 7.2 Addition of chemicals to the lab scale SBR's

To be able to compare the results from the Moving Bed Biofilm Reactor *MBBR* with the results from the lab scale SBR's, it was decided to add the same kind of chemicals, phosphoric acid and methanol. The chemical dosages that needed to be added to the lab scale SBR's were calculated before installation at Brudaremossen. They were calculated according to basic wastewater design theory; to add the amount that is required to remove unwanted substance by stoichiometric assumptions (Henze, et al., 2002).

The amount of incoming nitrogen can be used in order to calculate the required dosage of phosphorus and methanol. For these calculations, a concentration of incoming total nitrogen  $N_{tot}$  was assessed as 100 mg/l which was an average measured value of  $N_{tot}$  in the leachates after the oil and sludge separator during the analysis by Ramböll from 2008 to 2013 (Ramböll, 2015 A). Since the incoming flow into the SBR's was assumed to be 24 litres per day, an  $N_{tot}$  load per day could be calculated, see equation (14). By knowing that six cycles would be performed each day, an  $N_{tot}$  load per cycle could be assessed.

$$N_{tot} = (24 \cdot 100)/1000 = 2.4 [g/d] = 400 [mg/cycle] \quad (14)$$

### 7.2.1 Calculation of phosphorus dosage

The calculation of phosphorus dosage was based on that the ratio of Chemical Oxygen Demand *COD*:N:P should be 200:5:1 for optimal conditions during wastewater treatment (Henze, et al., 2002). With the assumed  $N_{tot}$  load of 400 mg/cycle, the required phosphorus was calculated according to the known ratio 200:5:1. As previously mentioned, three different exchange ratios – 0.17, 0.33 and 0.50 – were applied to the lab scale SBR's during the experimental trial. Therefore, three various dosages of phosphorus were calculated. Firstly, the required amount of phosphorus during one cycle for the various exchange ratios were calculated, see equation (15) to equation (17).

$$P_{add 1} = N_{tot (133 mg)}/5 = 27 [mg/cycle] \quad (15)$$

$$P_{add 2} = N_{tot (267 mg)}/5 = 53 [mg/cycle] \quad (16)$$

$$P_{add 3} = N_{tot (400 mg)}/5 = 80 [mg/cycle] \quad (17)$$

The phosphorus was added as 75 % phosphoric acid with a density of 367 g/L. It was diluted 100 times in order to get a viable amount to add per cycle. When knowing that the phosphorus should be added during 30 minutes of the feed phase, phosphorus dosages in ml/min during 30 minutes was assessed for each exchange ratio, see equation (18) to equation (20).

$$P_{dose\ 1} = ((P_{add\ 1}/367) \cdot 100)/30 = 0.242 \text{ [ml/min]} \quad (18)$$

$$P_{dose\ 2} = ((P_{add\ 2}/367) \cdot 100)/30 = 0.484 \text{ [ml/min]} \quad (19)$$

$$P_{dose\ 3} = ((P_{add\ 3}/367) \cdot 100)/30 = 0.727 \text{ [ml/min]} \quad (20)$$

## 7.2.2 Calculation of methanol dosage

The calculation of methanol dosage was based on that the ratio of COD: Nitrate nitrogen  $NO_3-N$  should be approximately (3.5–5):1 for wastewater treatment (Ekenberg, 2007). The highest ratio was used for this calculation since oxygen  $O_2$  left from previous nitrification steps can oxidise some of the methanol. The assumed load of 100 mg  $N_{tot}$  /L was used in the calculations since it was assumed that it would be converted into  $NO_3-N$  during nitrification. As for the calculated dosages of phosphorus, three different methanol dosages for the exchange ratios – 0.17, 0.33 and 0.50 – were calculated. Firstly, the required COD amounts for the various exchange ratios were determined; see equation (21) to equation (23).

$$COD_{Me\ 1} = N_{tot(133\ mg)} \cdot 5 = 667 \text{ [m g/cycle]} \quad (21)$$

$$COD_{Me\ 2} = N_{tot(267\ mg)} \cdot 5 = 1333 \text{ [m g/cycle]} \quad (22)$$

$$COD_{Me\ 3} = N_{tot(400\ mg)} \cdot 5 = 2000 \text{ [m g/cycle]} \quad (23)$$

A known assumption is that 1.5 kg COD corresponds to 1 kg of methanol (Leslie Grady, et al., 2011). Methanol with a density of 0.792 g/m<sup>3</sup> was diluted three times and was thereafter added to the lab scale SBR during two minutes every cycle, which generated the required dose of methanol for each exchange ratio as explained in equation (24) to equation (26).

$$Me_{dose\ 1} = (((COD_{Me\ 1}/1.5)/0.792) \cdot 3)/2 = 0.84 \text{ [ml/min]} \quad (24)$$

$$Me_{dose\ 2} = (((COD_{Me\ 2}/1.5)/0.792) \cdot 3)/2 = 1.69 \text{ [ml/min]} \quad (25)$$

$$Me_{dose\ 3} = (((COD_{Me\ 3}/1.5)/0.792) \cdot 3)/2 = 2.53 \text{ [ml/min]} \quad (26)$$

## 7.3 SBR cycle analysis

In order to investigate critical operational parameters during the SBR process, a cycle analysis was performed the 23<sup>rd</sup> of April, Day 44 of the experimental trial. The results from the cycle analysis showed if the duration of the phases, based only on previous knowledge and literature study, was enough for the biological processes to function

completely. The sequence of phases during one cycle of the lab scale SBR's as well as related durations were presented in Table 7.1.

Samples were taken from the two reactor tanks during an entire cycle, which resulted in 17 samples from each reactor. The first sample was taken in the beginning of the fill phase and thereafter the samples were taken with a time interval of 15 minutes. For each sample, the collected effluent first was placed into a smaller vessel during five minutes, which allowed sludge in the sample to settle, see Figure 7.3. Thereafter, five mL of the water phase of the sample was pipetted and immediately filtered through a 0.45 µm membrane filter in order to avoid further biological processes to occur. The samples were collected in ten mL plastic tubes and directly placed into a cooler, see Figure 7.3.



*Figure 7.3 Cycle analysis performed the 23<sup>rd</sup> of April; settled sample in vessel to the left and five mL of filtered sample to the right.*

The concentrations of  $\text{NH}_4\text{-N}$ ,  $\text{NO}_3\text{-N}$  and nitrite nitrogen  $\text{NO}_2\text{-N}$  were later analysed at the laboratory at Chalmers for each sample in order to perform a cycle analysis. By analysing the conversion between the different nitrogen fractions throughout the cycle, it was possible to determine if the processes of nitrification and denitrification were working properly in the SBR's.

## **7.4 Lab scale SBR sampling procedure**

During the study, both water and sludge samples were taken on site from the installed lab scale SBR's. Water samples were taken two times each week - every Wednesday and Friday - from the feed tanks and after treatment in the reactor tanks during the draw phase. During Wednesdays, the samples were taken simultaneously with Ramböll, who analysed the overall treatment capacity of the Brudaremossen pilot plant 2015. Results from analysis of the incoming leachates during the first four Wednesdays were obtained by Ramböll in order to minimise the number of samples. Sludge samples were taken once a week from the reactor tanks during the react anoxic phase, since that resulted in least impact on the biological process and in a more homogenous sample. During each sampling occasion the current water temperature, pH, electrical conductivity, Dissolved Oxygen  $\text{DO}$ , volume of the inflow and climate conditions were measured or noted on site.

All the water samples were placed into a cooler and transported directly to the laboratory at Chalmers. Thereafter, both filtered and unfiltered samples, before and after treatment in the lab scale SBR's, were stored in a freezer. All the sludge samples

were placed into a cooler and transported to the laboratory at Chalmers, where they were analysed directly.

### 7.4.1 Analysis of the water samples

The parameters and related analytical methods that were used for analysing the water samples for the lab scale SBR's are presented in Table 7.4.

Table 7.4 Parameters analysed in the water samples from the lab scale SBR's.

Parameter	Unit	Analysis method
pH	-	MultiLine P4
Temperature	°C	MultiLine P4
DO	mg/L	MultiLine P4
Alkalinity	mmol/L	EN ISO 9963-1:1995 (Titration)
NH <sub>4</sub> -N	mmol/L	EN ISO 10304-2-1996 (Ion Chromatograph)
NO <sub>3</sub> -N	mmol/L	EN ISO 10304-2-1996
NO <sub>2</sub> -N	mmol/L	EN ISO 10304-2-1996
N <sub>tot</sub>	mg/l	EN 1484-1997
TC	mg/L	EN 1484-1997
IC	mg/L	EN 1484-1997
TOC	mg/L	EN 1484-1997
DOC	mg/L	EN 1484-1997
BOD <sub>7</sub>	mg/L	BOD <sub>7</sub> -day test
P <sub>tot</sub> Metals (Ag, Al, Co, Cd, Cu, Cr, Fe, Hg, Mn, Ni, Pb, Pt, Sb, Ti, V and Zn)	mg/l µg/l	ISO 15587-2 (ICP-MS)

A portable MultiLine P4 meter was used out in field to measure the pH, temperature, electrical conductivity and the DO concentration. Remaining parameters were analysed in the laboratory at Chalmers University of Technology according to the standard methods presented in Table 7.4.

The alkalinity was measured according to the standard method EN ISO 9963-1:1995. Total alkalinity was thereafter calculated according to equation (27).

$$Tot\ Alk = \frac{2 \cdot V_{pH\ 4.5} \cdot Normality\ of\ acid\ \left[\frac{eq}{L}\right] \cdot 50,000\ [mg_{CaCO_3}/eq]}{V_{sample}} \quad [mg/L] \quad (27)$$

Where  $V_{pH\ 4.5}$  – Volume of added acid  
 $V_{sample}$  – Volume of the tested sample

Inductively Coupled Plasma Mass Spectrometry *ICP-MS* analysis was used for detecting the metals and total phosphorus  $P_{tot}$  in the water samples. The technique is based on a high temperature Inductive Coupled Plasma *ICP* which enables a transformation of the element atoms into ions, which thereafter are separated. The mass spectrometer can thereafter detect the separated ions based on their mass-to-

charge ratio. By creating an electrostatic filter in the mass spectrometer, only ions of a certain mass-to-charge ratio are able to pass through rods in the detector, during a certain time period. The ICP-MS analysis was run with both filtered and unfiltered samples - in order to obtain the dissolved concentrations of the ions - and was performed according to the standard method ISO 15587-2.

Ion Chromatograph *IC* analysis was used in order to detect anions and cations in the samples. The technique is based on ion exchange chromatography, eluent suppression and conductimetric detection (Cheremisinoff, 2001). The process of ion exchange takes place in a column with a flow of an eluent, to which the sample is added. Different anions and cations present in the sample adhere to the column in different ways, hence travels through the column with the eluent in different speeds. Subsequently the sample passes through a detector which measures the conductivity, which can provide information about the concentration of anions and cations present in the sample (Otto, 2015; Cheremisinoff, 2001). The analysis was performed according to the standard method EN ISO 10304-2-1996.

Determination of Total Organic carbon *TOC*, Dissolved Organic Carbon *DOC*, Inorganic Carbon *IC* and  $N_{\text{tot}}$  in the water samples was performed according to the European Standard method; EN 1484:1997 (SIS, 1997). The test is based on oxidation of organic carbon in water into carbon dioxide  $CO_2$ , which can be achieved by adding an oxidant or by applying UV radiation. Inorganic carbon can be analysed separately or by acidifying the sample.

The Biochemical Oxygen Demand *BOD* for the samples was measured at the laboratory at Chalmers. Organic material is biochemically oxidised by microorganism and the level of DO needed for that process was determined in a  $BOD_7$ -day test (Tchobanoglous, et al., 2003). The samples were taken in glass BOD bottles and incubated for seven days in  $20^\circ\text{C}$ . The levels of DO were measured before and after the incubation and the difference in mg/L represented the BOD of the sample, see equation (28).

$$BOD_7 = DO_1 - DO_2 \quad [\text{mg/L}] \quad (28)$$

Where  $DO_1$  – Concentration of DO during sampling [mg/L]  
 $DO_2$  – Concentration of DO after 7 days [mg/L]

#### 7.4.2 Analysis of the sludge samples

The sludge samples were taken as a mixed liquor sample during the react phase in the lab scale SBR's. The parameters and related analytical methods that were used for analysing the sludge samples from the SBR's are presented in Table 7.5. All parameters analysed for the sludge samples are described in Chapter 3, Section 3.4.

Table 7.5 Parameters analysed in the sludge samples from the lab scale SBR's (Rice, et al., 2012; Gerardi, 2003) .

Parameter	Unit	Analysis method
SV	ml/L	Method 2710 C
MLSS	mg/L	Heating to 105 °C according to Method 2540 G
MLVSS	mg/L	Method 2540 G
SVI	ml/g SS	Method 2710 D, Equation (10)
F/M ratio	kg COD/kg MLSS *day	Calculation equation (13)

The settled Sludge Volume *SV* of the sludge samples was tested according to Method 2710 C *Settled Sludge Volume* in Standard Methods for the Examination of Water and Wastewater (Rice, et al., 2012). Figure 7.4 illustrates the sludge volume test after 30 minutes of settling.



Figure 7.4 Sludge settled in a 1000 mL cylindrical tube after 30 minutes

In order to measure the MLSS and the Mixed Liquid Volatile Suspended Solids *MLVSS* of the sludge samples, analysis were performed according to Method 2540 G *Total, Fixed and Volatile Solids in Solid and Semisolid Samples* (Rice, et al., 2012). The same analysis was also performed with Milli-Q water due to the possibility of filter loss during high temperature burn of the glass fibres in the filters. MLSS and MLVSS were thereafter calculated according to equation (29) and equation (30).

$$MLSS = \frac{w_{105} - w_{filter} + w_{filter\ loss\ 105}}{V_{filtered\ sample}} \quad (29)$$

$$MLVSS = \frac{w_{550} - w_{105} + w_{filter\ loss\ 550}}{V_{filtered\ sample}} \quad (30)$$

Where

- $w_{105}$  –weight after heated in oven at 105 °C
- $w_{filter}$  –weight of filter before heating
- $V_{filtered\ sample}$  –volume of the filtered sample
- $w_{550}$  –weight after heated in oven at 550 °C

## 7.5 Full scale SBR design

A full scale SBR was designed based on evaluation of the experimental trial with the lab scale SBR's together with a design calculation approach presented in *Wastewater Engineering: Treatment and Reuse* p. 724-733 by Tchobanoglous et al. (2003).

Some of the design parameters are based on data from the report *Lakvattenhantering vid Brudaremossen* - Delrapport 2 by Ramböll, see Table 7.6.

Table 7.6 Design parameters based on data from the report *Lakvattenhantering vid Brudaremossen*- Delrapport 2 by Ramböll

Design parameter	Unit	Value
Average flow	m <sup>3</sup> /d /reactor	40·24 = <b>960</b>
Average BOD load	kg/d	7·960,000/1,000,000 = <b>6.72</b>
Average NH <sub>4</sub> -N load	kg/day	55·960,000/1,000,000 = <b>52.8</b>

It was decided that the already installed treatments, such as the oil and sludge separator and the leachate pond, could continue be used to reduce the oil fractions as well as organic pollutants. As in the Brudaremossen pilot plant 2015, chemical precipitation with addition of aluminium sulphate, sodium hydroxide and polymers would be used. Thereafter, lamellae sedimentation with a flow limit of 80 m<sup>3</sup>/h would be installed as pre-treatment before the full scale SBR.

While assessing a full scale MBBR process for Brudaremossen, it was evaluated by Ramböll that a leachate flow from 30 m<sup>3</sup>/h to 50 m<sup>3</sup>/h could pass through a full scale treatment. Flows above the limit would be discharged into a levelling magazine with a volume of 500 m<sup>3</sup>. Also, a flow before the GAC filter of 30 m<sup>3</sup>/h would have to be discharged into the levelling magazine. Therefore, an average flow into a full scale SBR technique was estimated to be 40 m<sup>3</sup>/h and a levelling magazine was decided to be necessary for the SBR treatment as well.

Some of the design parameters were assessed from the measured data provided by Ramböll. The average BOD load in point B – see Figure 6.6 in Chapter 6, Section 6.3 - was approximately 7 mg/L and the average NH<sub>4</sub>-N load was approximately 55 mg/L from 2015-01-22 to 2015-04-29. As previously stated, the resulted BOD<sub>7</sub> concentrations for the lab scale SBR's were uncertain which explains why the BOD load for the MBBR was applied.

After the two full scale SBR reactors, it is suggested that the decanted water will be channelled to sand and GAC filters as a final treatment step before discharge in to a nearby recipient. The report by Ramböll (Ramböll, 2015 C) evaluated that three sand filters with a total area of 12 m<sup>2</sup> and two GAC filters with a total area of four m<sup>2</sup> would be necessary for a full scale treatment.

## 8 Results & discussion

The lab scale Sequencing Batch Reactors *SBR*'s were operated and analysed during 50 days, from 2015-03-11 to 2015-04-29. The results from the samples taken in Reactor 1 and Reactor 2 during the trial were compared with results from sample analysis of the Moving Bed Biofilm Reactor *MBBR* process in the Brudaremossen pilot plant 2015. The data for the *MBBR* was provided by Ramböll and covered the period from 2015-02-11 to 2015-05-06.

In this chapter, the treatment capacity of lab scale *SBR*'s and the *MBBR* concerning removal of nitrogen, organic compounds and metals will be presented and discussed. Critical operational parameters during the experimental trial will also be illustrated and argued in relation to the treatment techniques performance. Thereafter, the lab scale *SBR*'s treatment capacity will be further presented and discussed based on the results from the cycle analysis performed on Day 44, the 23<sup>rd</sup> of April, and the analysed activated sludge parameters. Finally, the calculated design of a full scale *SBR* will be provided and argued.

### 8.1 Removal efficiency of the lab scale *SBR*'s and the *MBBR*

Table 8.1 presents a summary of the average concentrations of the most important parameters measured in the influent and effluent, at the highest exchange ratio of 0.50, for the lab scale *SBR*'s, Reactor 1 and Reactor 2, and of the *MBBR* process in the Brudaremossen pilot plant 2015. The standard deviations are included in the table as well, in order to get an overview regarding how the concentrations varied at the end of the trial. Similar summaries of average concentrations during the periods with the exchange ratios 0.17 and 0.33 are presented in Table 1 and Table 2 in Appendix B. It should be noted that only two sampling occasions were performed when the exchange ratio was set to 0.33 and 0.50, which makes the results for those periods somewhat uncertain.

The influent samples from the lab scale *SBR*'s were taken from each reactor's feed tank and the effluent samples were taken during the draw phase in Reactor 1 and Reactor 2 separately. The influent samples of the *MBBR* process was taken from data in sampling point B and the effluent samples from data in sampling point F - see Figure 6.6 in Chapter 6, Section 6.3. On Day 50, a technical error related to the aeration equipment was detected which caused disruption of the biological processes in the lab scale *SBR*'s. The resulted concentrations of parameters on that day were therefore not included when the average concentrations, presented in Table 8.1, were calculated.

All measured parameters in the water and sludge samples for each sampling occasion during the experimental trial are presented in Appendix C.

Table 8.1 Average concentrations and standard deviations in the influent and effluent for the lab scale SBR's, with an exchange ratio of 0.50, and of the MBBR process in the Brudaremossen pilot plant 2015.

Exchange ratio 0.50	Parameter	Unit	Lab scale SBR REACTOR 1		Lab scale SBR REACTOR 2		Brudaremossen pilot plant 2015 MBBR	
			Influent	Effluent	Influent	Effluent	Influent	Effluent
	pH	-	7.3 ± 0.03	8.0 ± 0.1	7.4 ± 0.01	8.2 ± 0.1	n.a.	7.9 ± 0.1
	Temperature	°C	11.3 ± 1.5	12.5 ± 2.2	11.0 ± 1.2	12.6 ± 2.2	8.0 ± 0.9	11.9 ± 1.5
	NH <sub>4</sub> -N	mg/L	27.1 ± 5.3	3.5 ± 2.4	27.8 ± 3.3	2.9 ± 1.8	52.1 ± 10.6	0.1 ± 0.2
	NO <sub>2</sub> -N	mg/L	1.3 ± 0.1	1.1 ± 0.0	1.3 ± 0.1	1.1 ± 0.0	0.01 ± 0.004	2.2 ± 0.9
	NO <sub>3</sub> -N	mg/L	6.4 ± 4.0	5.7 ± 3.6	10.3 ± 5.9	3.2 ± 0.7	0.2 ± 0.2	2.2 ± 0.1
	N <sub>tot</sub>	mg/L	52.6 ± 1.9	11.4 ± 10.1	51.3 ± 1.6	6.5 ± 5.1	58.4 ± 11.6	1.5 ± 0.3
	P <sub>tot</sub>	mg/L	0.03 ± 0.004	12.5 ± 1.6	0.04 ± 0.004	14.6 ± 0.5	0.04 ± 0.03	2.5 ± 2.4
	BOD <sub>7</sub>	mg/L	0.4 ± 0.3	0.1 ± 0.005	0.9 ± 0.3	0.03 ± 0.02	7.4 ± 1.3	< 3
	TOC	mg/L	13.8 ± 0.2	33.9 ± 3.6	11.6 ± 2.4	36.9 ± 2.2	31.0 ± 2.5	50.7 ± 20.8
	DOC	mg/L	12.0 ± 0.5	33.8 ± 1.4	8.3 ± 1.6	38.5 ± 0.9	28.7 ± 3.0	62.0 ± 26.1
	Fe	mg/L	0.6 ± 0.2	0.3 ± 0.1	0.2 ± 0.004	0.1 ± 0.003	5.1 ± 2.1	0.13 ± 0.14
	Fe filtered	mg/L	0.1 ± 0.004	0.1 ± 0.0	0.1 ± 0.004	0.04 ± 0.002	0.1 ± 0.01	0.07 ± 0.03
	Mn	mg/L	0.4 ± 0.04	0.4 ± 0.03	0.4 ± 0.02	0.4 ± 0.03	n.a.	0.06 ± 0.03

n.a. –not analysed

\*From sampling point E due to lack of measured data in sampling point F

### 8.1.1 Removal of nitrogen

The following paragraphs will present and discuss the minimum, maximum and average removal rates of ammonium nitrogen  $NH_4-N$  and total nitrogen  $N_{tot}$  in relation to the various exchange ratios.

#### REMOVAL OF AMMONIUM NITROGEN

The minimum, maximum and average removal rates of  $NH_4-N$  were analysed for the lab scale SBR's. Each concentration was calculated separately for the three periods with the exchange ratios; 0.17, 0.33 and 0.50 see Table 8.2. The same parameters were also analysed for the MBBR process in the Brudaremossen pilot plant. The resulted minimum, maximum and average removal rates of  $NH_4-N$  for the MBBR were 98.11 %, 99.98 % and 99.68 % respectively.

Table 8.2 Minimum, maximum and average  $\text{NH}_4\text{-N}$  removal rates for the lab scale SBR's in relation to the exchanges ratios 0.17, 0.33 and 0.50.

Exchange ratio	$\text{NH}_4\text{-N}$ removal	Unit	Lab scale SBR REACTOR 1	Lab scale SBR REACTOR 2
0.17	Min	%	95.51	88.71
	Max	%	98.31	99.59
	Average	%	97.07	94.63
0.33	Min	%	97.59	97.82
	Max	%	99.48	97.84
	Average	%	98.53	97.83
0.50	Min	%	81.82	80.86
	Max	%	94.86	96.40
	Average	%	88.34	88.63

Generally, the removal rate of  $\text{NH}_4\text{-N}$  during the experimental trial for the lab scale SBR's was high, which indicated that the nitrifying bacteria thrived at once and acclimated fast in the reactors. In Table 8.2 it can be seen that the  $\text{NH}_4\text{-N}$  removal in Reactor 1 was slightly higher than in Reactor 2 during period one and two and thereafter rather similar throughout the last period. The only parameter that differs somewhat between the two reactors is the minimum  $\text{NH}_4\text{-N}$  removal when the exchange ratio was set to 0.17, which were 95.51 % and 88.71 % in Reactor 1 and Reactor 2 respectively. On the other hand, the maximum removal rate for the same period was measured to be higher in Reactor 2 than in Reactor 1, which makes it difficult to draw any conclusions.

The removal rate of  $\text{NH}_4\text{-N}$  was highest when the exchange ratio was 0.33 and lowest when the exchange ratio was set to 0.50. During the first and second period the average removal rate reached close to 98 % in the two reactors, whilst it only reached approximately 88 % during the last period. This can be explained by that less data was obtained for the last period and since it was observed that the lab scale SBR's performed less efficient during one day, it had a significant impact on the results.

Throughout the experimental trial there were some concentrations that were analysed to be under the detection limit of approximately 1 mg/L, set in the Ion Chromatograph IC. These concentrations are therefore somewhat uncertain and could in fact be lower than 1 mg/L, which would have resulted in higher removal rates of  $\text{NH}_4\text{-N}$  than the results provided in Table 8.2.

It can be seen that the removal rate of  $\text{NH}_4\text{-N}$  was higher for the MBBR process when comparing minimum, maximum and average removal rates with the lab scale SBR's. The maximum removal rates presented for the various exchange ratios and the minimum removals during the second period in the lab scale SBR's are similar to the removal rates of the MBBR. Regarding the average removal in the lab scale SBR's, it can be seen that it is foremost the removal rates at the exchange ratio of 0.50 that are significantly lower than the value presented for the MBBR. Even if the lab scale SBR's reached high  $\text{NH}_4\text{-N}$  removal rates, the result implies that the MBBR had a more stable removal of  $\text{NH}_4\text{-N}$  during the sampling period, which is preferable when treating leachates. However, the operation of the MBBR was started week 51 in 2014, in order to enable acclimatisation and build-up of the biomass, and the first samples

were taken approximately 54 days later. This allowed the MBBR to reach a more stable result and removal rate than the lab scale SBR's, which were installed week 11 in 2015 and the first samples was taken only seven days after start-up.

When comparing the resulted removal rates for the lab scale SBR's at Brudaremsen with the removal rates presented for the eight reviewed cases, see Table 5.1 in Chapter 5, Section 5.1.1, it can be seen that they are within the same range. Four cases had reported a  $\text{NH}_4\text{-N}$  removal rate of 99 % or higher and the average removal rate for all the cases combined were 87 %. This shows that the lab scale SBR's at Brudaremsen removed  $\text{NH}_4\text{-N}$  to a rate that could be expected for such a technique. It is likely that the resulted  $\text{NH}_4\text{-N}$  removal rate would have been more stable and similar to the results for the MBBR if the technique had been optimised and operated during a longer period.

The average removal of  $\text{NH}_4\text{-N}$  for the entire experimental trial were also analysed and was 96 % for Reactor 1 and 94 % for Reactor 2.

#### REMOVAL OF TOTAL NITROGEN

The minimum, maximum and average removal rates of  $\text{N}_{\text{tot}}$  during the three periods with different exchange ratios were also analysed for the lab scale SBR's.  $\text{N}_{\text{tot}}$  removal for the MBBR was taken from provided data during the same period as the pilot plant  $\text{NH}_4\text{-N}$  removal analysis. The minimum, maximum and average  $\text{N}_{\text{tot}}$  removal rates for the MBBR were 97.02 %, 97.71 % and 97.38 % respectively. See Table 8.3 for the removal rates of  $\text{N}_{\text{tot}}$  in relation to the various exchange ratios; 0.17, 0.33 and 0.50.

*Table 8.3 Minimum, maximum and average  $\text{N}_{\text{tot}}$  removal rates for the lab scale SBR's in relation to the exchanges ratios 0.17, 0.33 and 0.50.*

<b>Exchange ratio</b>	<b><math>\text{N}_{\text{tot}}</math> removal</b>	<b>Unit</b>	<b>Lab scale SBR REACTOR 1</b>	<b>Lab scale SBR REACTOR 2</b>
0.17	Min	%	0.00	59.44
	Max	%	65.04	95.37
	Average	%	39.46	83.93
0.33	Min	%	46.86	75.07
	Max	%	46.86	75.72
	Average	%	46.86	75.39
0.50	Min	%	60.70	78.24
	Max	%	97.52	97.14
	Average	%	79.11	87.69

As can be seen in Table 8.3, the minimum, maximum and average  $\text{N}_{\text{tot}}$  concentrations for Reactor 1 are the same for the period with exchange ratio 0.33. This is due to that some samples for that period were wrongly handled during preparation and therefore only one sample was analysed for that period.

The results show that Reactor 2 had a higher  $\text{N}_{\text{tot}}$  removal rate than Reactor 1, both regarding minimum and average removal rates. The zero minimum removal for Reactor 1 during the first period was since  $\text{N}_{\text{tot}}$  was measured in a higher

concentration in the effluent compared to the influent, which resulted in a negative removal rate.

Concerning the maximum  $N_{\text{tot}}$  removal rate it can be seen that it was significantly higher in Reactor 2, except for during the last period during which the rates were close to 97 % in both reactors. However, the result implies that the removal of  $N_{\text{tot}}$  was more even in Reactor 2 and that the biological processes worked better compared to in Reactor 1. Generally, it can be seen in Table 8.3 that the minimum, maximum and average removal rates improved when the exchange ratio increased. This is probably since the biological processes in the lab scale SBR needed some time in order to acclimatise to the current situation at Brudaremossen.

Similar to the results of  $\text{NH}_4\text{-N}$  removal rates, the  $N_{\text{tot}}$  removal was higher for the MBBR than the lab scale SBR's. It can be noted in Table 8.3 that the maximum  $N_{\text{tot}}$  removal rates at exchange ratio 0.50 were very similar to the maximum  $N_{\text{tot}}$  removal for the MBBR. The higher  $N_{\text{tot}}$  removal rates for the MBBR is a result of that the lab scale SBR's had a shorter acclimatisation time than the MBBR before the sampling period begun.

When looking at the  $N_{\text{tot}}$  removal rates presented for the eight reviewed cases, see Table 5.1 in Chapter 5, Section 5.1.1, it can be seen that it was lower than the  $\text{NH}_4\text{-N}$  removal rates. The average  $N_{\text{tot}}$  removal rate for the cases were 79 % which correlates well with the average  $N_{\text{tot}}$  removal rates presented for Reactor 2 and the average removal in Reactor 1 during the exchange ratio of 0.50.

The average removal of  $N_{\text{tot}}$  for the entire experimental trial were also analysed and was 51 % for Reactor 1 and 76 % for Reactor 2.

#### NITROGENOUS FRACTIONS IN REACTOR 1

The various nitrogenous fractions were also analysed during the experimental trial in order to assess the efficiency of the nitrification and denitrification processes. The outgoing concentration of  $\text{NH}_4\text{-N}$ , nitrite nitrogen  $\text{NO}_2\text{-N}$ , nitrate nitrogen  $\text{NO}_3\text{-N}$  and  $N_{\text{tot}}$  for the lab scale SBR Reactor 1 are presented in Figure 8.1. The exchange ratio was increased twice during the experimental trial and the duration of the three stages with different loading are marked in the figure. The concentration of  $\text{NH}_4\text{-N}$  and  $\text{NO}_2\text{-N}$  were low in all samples from the effluent of Reactor 1 during the sampling period. However, the  $\text{NO}_3\text{-N}$  concentrations increased in the beginning of the study until Day 22. Thereafter, it decreased, reaching the lowest measured concentrations at Day 44. During Day 50, the aeration timer was observed to be malfunctioning which caused a breakdown of the nitrification processes, explaining the sudden increase of  $\text{NH}_4\text{-N}$ .

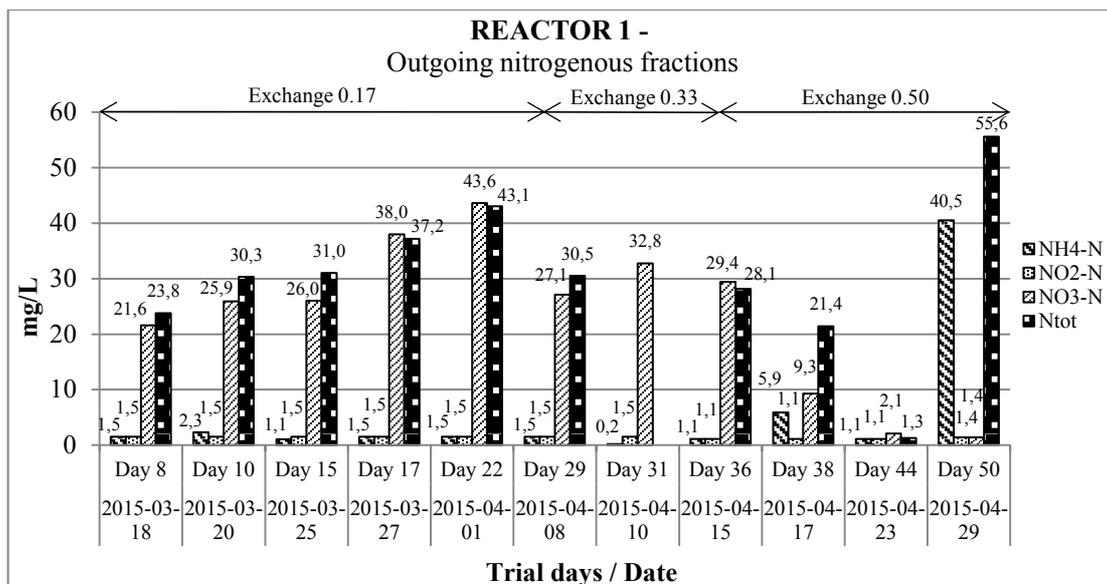


Figure 8.1 Outgoing nitrogenous fractions during the operation of the lab scale SBR Reactor 1.

The high NO<sub>3</sub>-N concentration in the samples during the beginning of the sampling period indicated that there was no active denitrification process in Reactor 1 throughout that time. The high NO<sub>3</sub>-N concentration resulted in a high average outgoing concentration of N<sub>tot</sub>, which explains the low N<sub>tot</sub> removal rates for Reactor 1 earlier presented in Table 8.3. During the lab scale SBR trial performed by Heander (2007) – CASE 2 – the denitrification was started up after 98 days, indicating that the denitrifying bacteria might need time to adapt to certain conditions.

The nitrification process in Reactor 1 appears to have functioned effectively during the entire operational period, although it showed a slightly higher concentration of NH<sub>4</sub>-N during Day 38. Moreover, the outgoing concentrations of NH<sub>4</sub>-N from Reactor 1 were lower than 3 mg/L - except for on Day 38 - which was the lowest threshold value for discharge of the treated leachates to the recipients presented by Ramböll, see Table 6.1 in Chapter 6, Section 6.2. At the end of the experimental trial, the outgoing concentrations of N<sub>tot</sub> were also lower than the threshold value of 30 mg/L for discharge to the recipients.

#### NITROGENOUS FRACTIONS IN REACTOR 2

The outgoing concentrations of NH<sub>4</sub>-N, NO<sub>2</sub>-N, NO<sub>3</sub>-N and N<sub>tot</sub> were also analysed for the lab scale SBR Reactor 2, see Figure 8.2. As for Reactor 1, the exchange ratio was increased twice during the experimental trial and the duration of the three stages with different loading are marked in the figure. Both nitrification and denitrification took place in the reactor throughout the experimental trial. Especially the nitrification seemed to have functioned well since the outgoing concentrations of NH<sub>4</sub>-N and NO<sub>2</sub>-N generally were low. On Day 8, it is likely the nitrification process had not started properly since there were higher concentrations of NH<sub>4</sub>-N and NO<sub>2</sub>-N on that day. Concerning the denitrification, the NO<sub>3</sub>-N concentration decreased until day 17 and then increased until Day 31. Thereafter, during the period with the exchange ratio 0.50, the denitrification improved and NO<sub>3</sub>-N was measured in the lowest concentrations on Day 44. As mentioned before, there was a malfunction of the

aeration equipment during Day 50, explaining the sudden increase of nitrogenous fractions on that day.

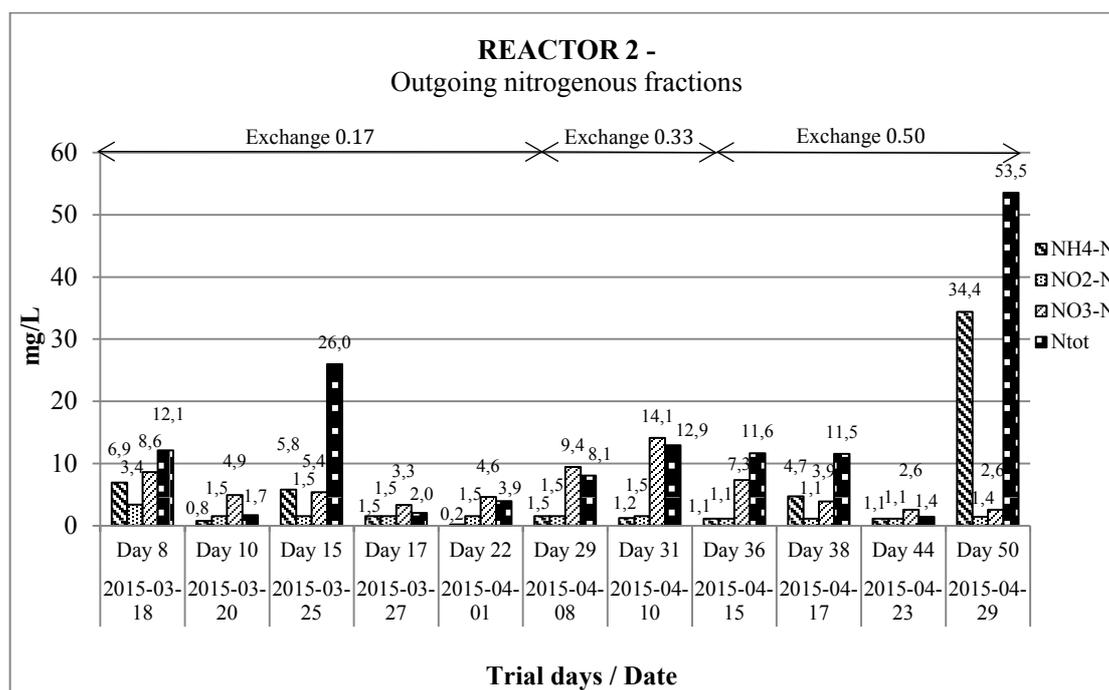


Figure 8.2 Outgoing nitrogenous fractions during the operation of the lab scale SBR Reactor 2.

The analysis shows that Reactor 2 experienced a more efficient denitrification during the first period of the operation compared to Reactor 1. The nitrification seemed to result in a similar removal of NO<sub>2</sub>-N in both reactors, except for on Day 8 where the outgoing concentration was slightly higher for Reactor 2 than Reactor 1. Similar to Reactor 1, the outgoing concentration of NH<sub>4</sub>-N was generally lower than 3 mg/L which was the lowest threshold value for discharge of the treated leachates into a nearby recipient; see Table 6.1 in Chapter 6, Section 6.2. The outgoing concentrations of N<sub>tot</sub> remained below the threshold value of 30 mg/L for discharge to the recipients.

As previously stated, the exchange ratio was increased twice during the experimental trial which might have affected the biological processes in the reactors. In Figure 8.1 and Figure 8.2 it can be seen that the NO<sub>3</sub>-N concentrations increased on Day 31 and that a high NH<sub>4</sub>-N concentration in both reactors was measured on Day 38. For both cases, the exchange ratio was increased during the preceding sampling occasion, which could have affected the biological processes. Furthermore, when observing the diagrams in Figure 8.1 and Figure 8.2 it seems like that the concentrations of outgoing nitrogenous fractions in the two reactors decreased when the exchange ratio was increased to 0.50. It could be that the microorganisms in the activated sludge required a higher input of nitrogen fractions during each cycle in order to thrive and create more biomass.

When comparing the outgoing concentrations of nitrogenous fractions for lab scale SBR Reactor 1 and Reactor 2, it can be seen that the denitrification for Reactor 1 was malfunctioning during the first operational period. The nitrification process did however function for both reactors. At first, it was believed that the denitrification process in Reactor 1 might had been affected by higher concentrations of toxic

compounds in the leachates, since Reactor 1 received the least treated water directly after the lamellae sedimentation in the Brudaremossen pilot plant 2015. However, denitrification has shown to have a lower sensibility to toxic compounds than nitrification and the nitrification process in Reactor 1 performed properly. It could be that the denitrifying bacteria are sensitive to other components, which has not been analysed during this thesis.

One other influencing factor for the denitrification process is the available concentration of methanol. Since the same amount of methanol was added to both reactors, using the same dosage pump, the methanol dosage should have been enough in Reactor 1 to enable denitrification. It is however possible that the dosage was provided unevenly to the reactors since the methanol dosage pump added the methanol solution during two minutes only. Problems with the dosage, such as uneven pumping or clogging of the narrow distribution tubes, would during two minutes be devastating for the distribution. It is possible that this might have caused a smaller amount of methanol distribution to Reactor 1 during the first operational period. This could have been solved by diluting the methanol solution even further and adding it during a longer time period instead, which would have minimised the risk of uneven methanol addition.

The high concentrations of nitrogenous fractions in Reactor 1 and Reactor 2 during Day 50 are an example of how a failure of the aeration equipment will affect the SBR treatment capacity. If a full scale SBR facility is to be built, then it will require proper timer functions and system controls in order to prohibit such system failures.

The Sludge Retention Time *SRT* in the lab scale SBR's was 32 days and according to the diagrams of the outgoing nitrogenous fractions presented in Figure 8.1 and Figure 8.2, the concentrations were decreasing after approximately 32 days. After 32 days, the sludge that was added when the lab scale SBR's were installed should have been removed from the system. At the same time, new bacteria were added to the reactors and biomass was built-up. It is likely that the new biomass was better acclimatised to the conditions at Brudaremossen, which contributed to the improved nitrogen removal.

#### NITROGENOUS FRACTIONS IN MBBR

The outgoing concentrations of  $\text{NH}_4\text{-N}$ ,  $\text{NO}_2\text{-N}$ ,  $\text{NO}_3\text{-N}$  and  $\text{N}_{\text{tot}}$  were also analysed for the MBBR process in the Brudaremossen pilot plant 2015, see Figure 8.3. The data was acquired from analysed samples in sampling point F - see Figure 6.6 in Chapter 6, Section 6.3 - provided by Ramböll. During the experimental trial the volumetric load of incoming leachates to the MBBR was increased twice, see Figure 8.3. The results showed that the MBBR process had an overall efficient treatment of all nitrogenous fractions during the studied period. The average outgoing concentration of  $\text{NH}_4\text{-N}$  was 0.11 mg/L which is well below the threshold value of 3 mg/L presented by Ramböll, see Table 6.1 in Chapter 6, Section 6.2.

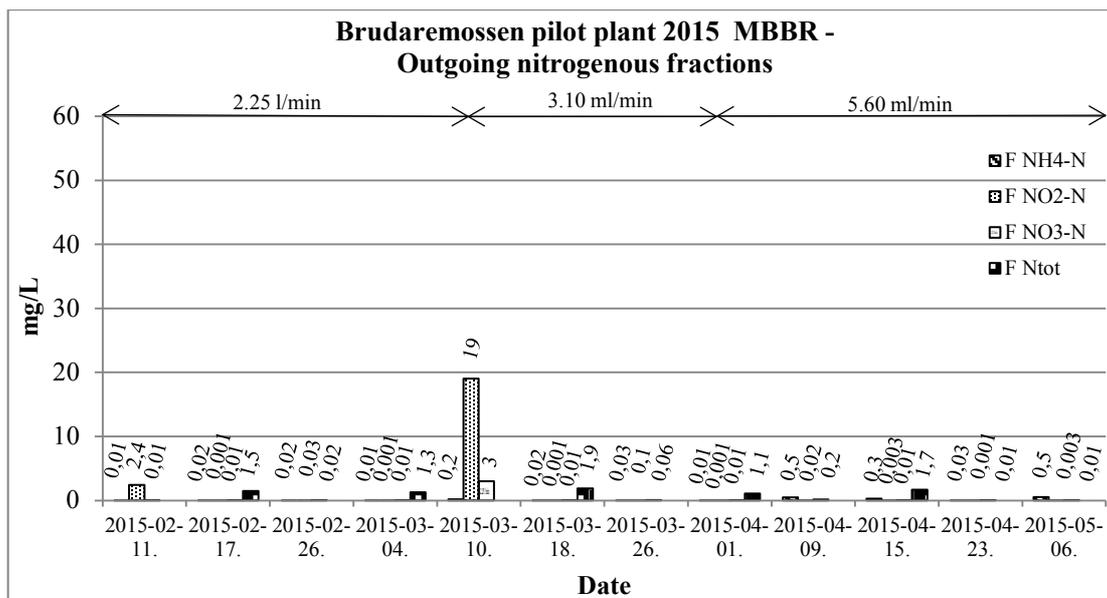


Figure 8.3 The outgoing nitrogenous fractions from the MBBR process in the Brudaremossen pilot plant 2015.

The high concentration of NO<sub>2</sub>-N measured on 2015-03-10 was by Ramböll evaluated to be due to a too low dosage of methanol solution during the first operational period, 90 mL/h, which was solved by subsequently adding more methanol resulting in a final methanol solution addition of 259 mL/h. The methanol solution consisted of methanol concentrate with a density of 0.792 g/m<sup>3</sup> that was diluted three times.

Due to lack of analysed parameters after the MBBR process, data had to be taken from sampling point F in the Brudaremossen pilot plant, which was located after the sand filter. This might have affected the results and in order to achieve a more accurate comparison between the MBBR and the lab scale SBR's, all parameters should have been analysed directly after the MBBR instead.

The outgoing nitrogenous fractions show that the MBBR was more efficient than the lab scale SBR's. However, as mentioned before, the MBBR had 54 days of acclimatisation time before sampling was begun compared to the seven days of acclimatisation for the lab scale SBR's. The most efficient nitrogen removal by the lab scale SBR's was achieved during Day 44 for both reactors. It is therefore believed that the lab scale SBR's would have shown similar nitrogen removal capacities as the MBBR if it was allowed the same amount of acclimatisation time.

#### NITROGENOUS FRACTIONS IN FEED TANKS

The distribution of nitrogenous fractions in the feed tanks into the lab scale SBR's were analysed in order to understand the variations of incoming NH<sub>4</sub>-N in relation to N<sub>tot</sub>. The result from the analysis will be presented and analysed in subsequent paragraphs and a more general discussion of the variations in both reactors will thereafter follow.

Figure 8.4 presents the nitrogenous fractions in the feed tank to Reactor 1 measured during the experimental trial. It can be seen that the concentrations varied throughout the measured period, with accumulation of NO<sub>2</sub>-N and NO<sub>3</sub>-N during Day 10 and of NO<sub>3</sub>-N especially during Day 44.

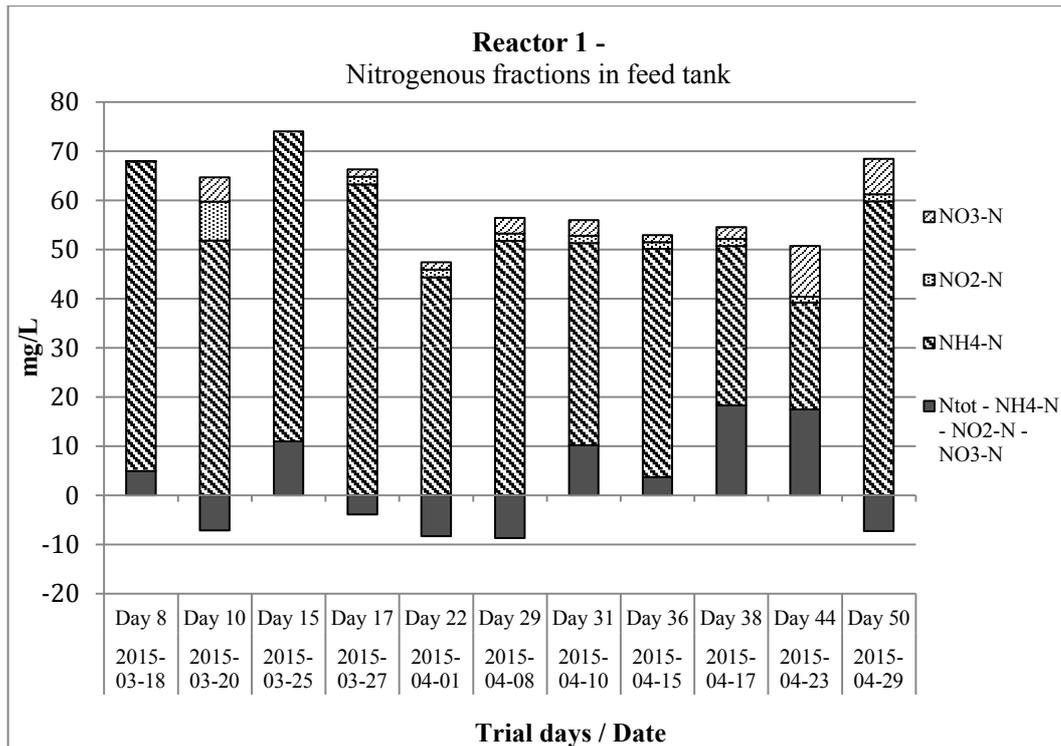


Figure 8.4 Nitrogenous fractions in feed tank to Reactor 1 during the experimental trial.

Figure 8.5 presents the nitrogenous fractions in the feed tank to Reactor 2 measured during the experimental trial. It can, as for the feed tank to Reactor 1, be seen that the concentrations varied throughout the measured period, with accumulation of  $\text{NO}_3\text{-N}$  during Day 38.

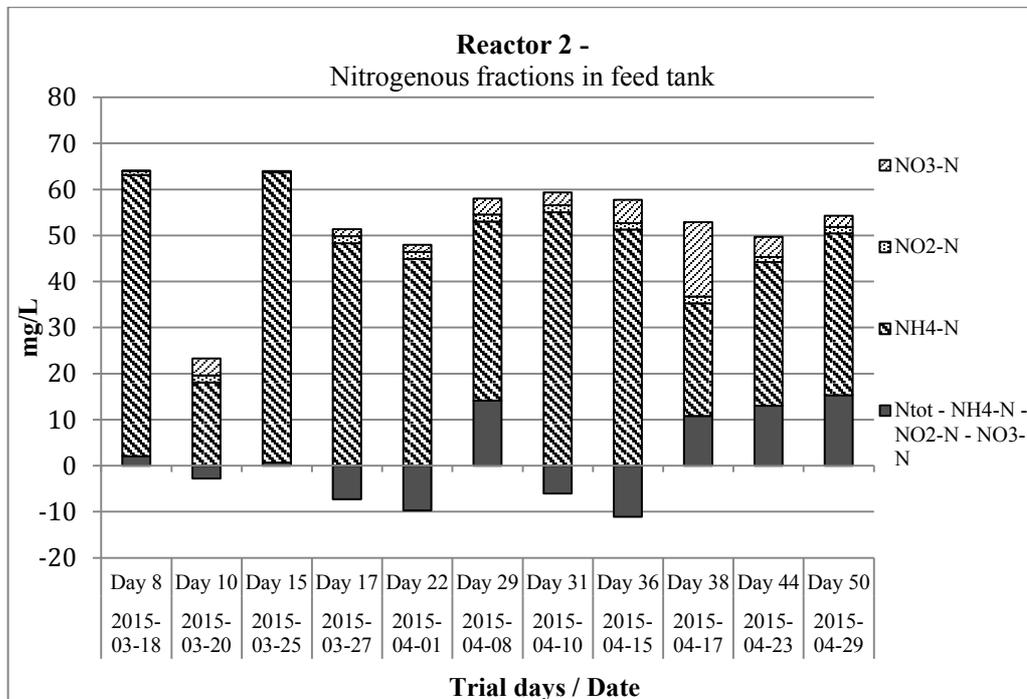


Figure 8.5 Nitrogenous fractions in feed tank to Reactor 2 during the experimental trial.

The varying nitrogenous fractions in the feed tanks may indicate that nitrification occurred in the tanks. The temperature in the feed tanks was relatively high during the trial which could have enabled biological reactors to occur. This would explain the uneven incoming concentrations of  $\text{NH}_4\text{-N}$  into both reactors. The nitrogenous fractions could have been further studied in order to be aware of that different incoming concentrations of  $\text{NH}_4\text{-N}$  to the reactors might occur and that regulations should be performed accordingly.

Since the  $\text{N}_{\text{tot}}$  concentration represents the sum of organic nitrogen, ammonia nitrogen  $\text{NH}_3\text{-N}$ ,  $\text{NH}_4\text{-N}$ ,  $\text{NO}_2\text{-N}$  and  $\text{NO}_3\text{-N}$  – see Chapter 2, Section 2.1.1 –  $\text{N}_{\text{tot}}$  minus all known nitrogenous fractions must represent the amount of organic nitrogen and  $\text{NH}_3\text{-N}$  in a sample. This was considered when analysing the fractions in the feed tanks since it was observed that the  $\text{N}_{\text{tot}}$  concentration was considerably larger than the sum of analysed nitrogenous fractions for a couple of samples. However, for some samples the  $\text{N}_{\text{tot}}$  was lower than the sum of nitrogenous fractions, symbolised by the negative concentrations in Figure 8.4 and Figure 8.5. This indicated that the methods used for analysis of the samples might have contributed to inaccurate concentrations, which should be considered when analysing the results.

As previously stated, it was believed that the lab scale SBR's nitrogen removal rate was the most efficient on Day 44. However, when observing the various incoming concentrations of nitrogenous fractions it could be seen that the  $\text{NH}_4\text{-N}$  concentration was low on Day 44, which could have contributed to the high removal rates on that day.

#### LOADING AND REMOVAL OF NITROGEN IN LAB SCALE SBR'S

The loading and removal of  $\text{NH}_4\text{-N}$  and  $\text{N}_{\text{tot}}$  in both reactors are presented and discussed in subsequent paragraphs. This inquiry was made in order to evaluate how the loading changed during the experimental trial and how this might have affected the removal of nitrogen. The incoming loading of milligrams  $\text{NH}_4\text{-N}$  or  $\text{N}_{\text{tot}}$  per litres and hours to the reactors was estimated by the concentration in each reactor's feed tank together with the feed phase time. The removal rate in milligrams  $\text{NH}_4\text{-N}$  or  $\text{N}_{\text{tot}}$  per litres and hours in the reactors was estimated by the difference between incoming and outgoing concentrations together with the oxic phase time for  $\text{NH}_4\text{-N}$  removal or oxic plus anoxic phase times for  $\text{N}_{\text{tot}}$  removal.

Figure 8.6 presents the loading and removal of  $\text{NH}_4\text{-N}$  and  $\text{N}_{\text{tot}}$  per litres and hours for Reactor 1 during the experimental trial. It can be seen that the loading as well as removal increases from the beginning of the trial period until Day 44. On Day 50, the loading increased whilst the removal remained low due to observed malfunction of the aeration tanks.

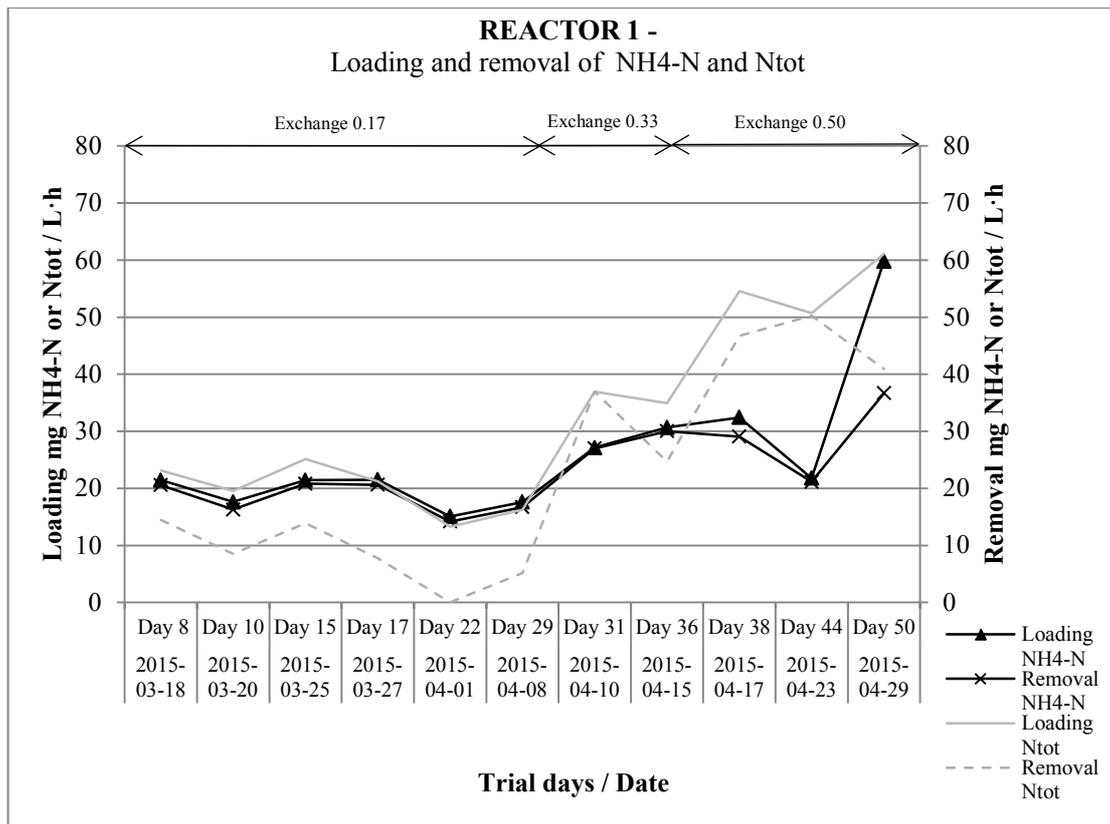


Figure 8.6 Loading and removal of NH<sub>4</sub>-N and N<sub>tot</sub> during the experimental trial

The loading of NH<sub>4</sub>-N and N<sub>tot</sub> per litres and hours varied throughout the experimental trial. It was assumed that the loading would increase when increasing the exchange ratio, as can be seen in Figure 8.6, it generally was the case. The decreased loading of NH<sub>4</sub>-N during Day 10 and Day 44 into Reactor 1 was believed to be a result of that more concentrations of other nitrogenous fractions was found during those days when analysing samples from the feed tank, see Figure 8.4. The fact that the loading of N<sub>tot</sub> was higher than the loading of NH<sub>4</sub>-N during those days strengthens that belief.

Figure 8.7 presents the loading and removal of NH<sub>4</sub>-N and N<sub>tot</sub> per litres and hours for Reactor 2 during the experimental trial. It can be seen that the loading as well as removal decreased significantly during Day 10. There was also a general increase of loading and removal from the beginning of the experimental trial until Day 38. On Day 50, the loading increased whilst the removal remained low due to observed malfunction of the aeration tanks.

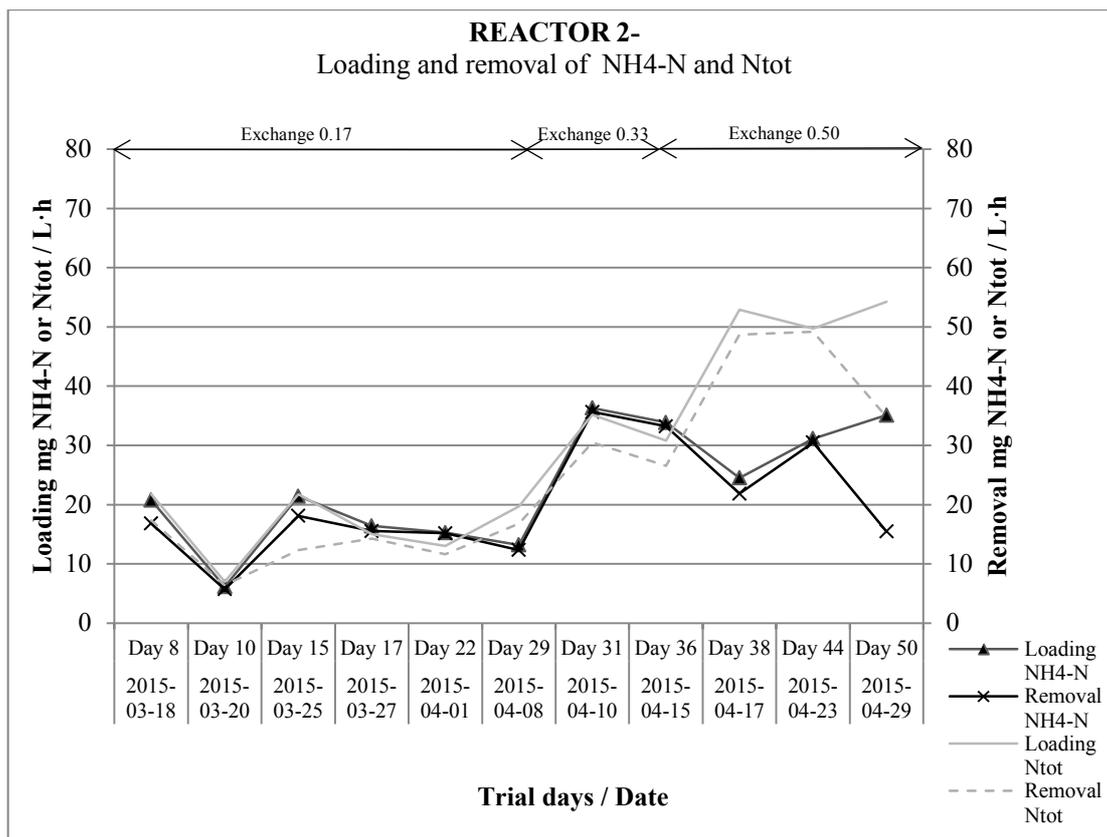


Figure 8.7 Loading and removal of NH<sub>4</sub>-N and N<sub>tot</sub> during the experimental trial

During Day 10, a significant decrease of loading can be observed. This is due to low concentrations of nitrogenous fractions in the feed tank during that day, see Figure 8.5. What caused the low concentrations in the feed tank is unknown, but since the concentrations in the feed tank to Reactor 1 were unaffected it might have been a result of the sand and Granulated Activated Carbon *GAC* filters. The decrease of NH<sub>4</sub>-N loading on Day 29 and during Day 38 to Day 50 were, as for Reactor 1, believed to be due to presence of other nitrogenous fractions in the feed tank to Reactor 2, see Figure 8.5. This seems reasonable since the N<sub>tot</sub> loading remained high during these days.

The variations in loading for both reactors was also believed to be due to a lower incoming concentration of NH<sub>4</sub>-N into the Brudaremossen pilot plant 2015, see Appendix D, at the end of the experimental trial since this would have affected the concentrations in the feed tanks and subsequently the loading. The results show that the standard deviations for NH<sub>4</sub>-N and NO<sub>3</sub>-N at the end of the trial were rather high, see Table 8.1, which also indicates that the incoming concentrations of those nitrogenous fractions varied significantly. Furthermore, it can be seen that the concentrations of loading and removal for NH<sub>4</sub>-N and N<sub>tot</sub> respectively is related, as a higher loading generally gives a higher removal.

### 8.1.2 Removal of organic matter

Table 8.4 presents the average concentrations of Total Organic Carbon *TOC*, Dissolved Organic Carbon *DOC*, Chemical Oxygen Demand *COD* and Biochemical Oxygen Demand *BOD* in the influent and effluent for Reactor 1, Reactor 2 and the MBBR during the experimental trial. The presented concentrations are for the period with the exchange ratio 0.50. It should be noted that the effluent concentrations for the

MBBR were based on concentrations in sampling point F and E, which is likely to have influenced the results.

*Table 8.4 Average TOC, DOC, COD, COD removal rate and BOD<sub>7</sub> in the influent and effluent to the lab scale SBR's, at exchange ratio 0.50, and the MBBR at Brudaremossen pilot plant 2015.*

<b>Exchange ratio 0.50</b>		<b>Lab scale SBR REACTOR 1</b>		<b>Lab scale SBR REACTOR 2</b>		<b>Brudaremossen pilot plant 2015 MBBR</b>	
<i>Parameter</i>	<i>Unit</i>	<i>Influent</i>	<i>Effluent</i>	<i>Influent</i>	<i>Effluent</i>	<i>Influent</i>	<i>Effluent</i>
TOC	mg/L	13.78	33.92	11.60	36.85	31.00	50.70
DOC	mg/L	12.01	33.78	8.31	38.53	28.67	62.00
COD	mg/L	90.53	108.66	84.01	117.42	85.33	210.67
BOD <sub>7</sub>	mg/L	0.41	0.09	0.93	0.03	7.40	< 3.00

In Table 8.4, it can be seen that the average concentrations of TOC, DOC and COD were higher in the effluent compared to the influent. This is probably since the dosages of metanol to the lab scale SBR's and the MBBR were too high or that biomass did not settle completely after the biological treatment, which therefore could have been retrieved when sampling. During the experimental trial, no measures were performed in order to optimise the dosage of methanol, which if performed would have reduced the concentrations of organic matter in the effluent.

Based on the theory concerning TOC in Chapter 2, Section 2.1.2, the presented effluent concentrations in Table 8.4 are still considered to be low. The average concentrations of TOC in the effluent from the lab scale SBR's are below the threshold value of 50 mg/L, required for discharge into potential recipients. Regarding the MBBR, the average TOC concentration was 50.70 mg/L, which is slightly above the threshold value. Nevertheless, considering the fact that the effluent from the MBBR passed through both sand and GAC filters subsequently, it is likely that the TOC will be further reduced before discharge.

Table 8.4 illustrates that the average concentrations of DOC in the effluent were similar to the amount of TOC. This indicates that the organic carbon in the leachates mainly existed in its dissolved form. This result is reasonable since it is likely that a part of the undissolved organic matter in the incoming leachates was removed during the preceding chemical precipitation and sedimentation.

The average concentrations of COD presented in Table 8.4 are considered to be low according to the theory presented in Chapter 2, Section 2.1.2. However, it should be noted that the presented concentrations for the lab scale SBR's were calculated by using the TOC concentrations according to equations (1) and (2), see Chapter 2, Section 2.1.2. These correlation equations were evaluated in a study of wastewater treatment, which makes the COD results somewhat uncertain. Nevertheless, they can give an indication of the COD concentration in the influent and effluent to the lab scale SBR's.

As previously stated, the COD concentrations in the effluent from the lab scale SBR's and the MBBR were higher than in the influent, which resulted in negative COD removal rates. The SBR treatments studied in the literature review had an average COD removal rate of 38 % - see Table 5.1 in Chapter 5, Section 5.1.1 - and earlier

studies has shown that SBR can reach a COD removal rate up to 75 %. If the dosage of external carbon source had been optimised, it is likely that the COD removal rate would have reached similar values. There was no threshold value presented for the recipients regarding COD.

The low TOC and COD concentrations in the incoming leachates correlates to the fact that the Brudaremossen landfill is assumed to have reached the more stable methanogenic phase, during which the amount of organic matter – hence TOC and COD – is reduced.

The average BOD<sub>7</sub> concentrations in the influent and effluent for the two reactors are presented in Table 8.4. There was a reduction of BOD<sub>7</sub> after treatment, which indicated that there was degradable organic matter available for the microorganisms in the incoming leachates. Also, the results show that the two biological treatment techniques fulfils the required threshold concentration of 15 mg BOD<sub>7</sub> per litre for all six potential recipients, see Table 6.2 in Chapter 6, Section 6.2.

When comparing the BOD<sub>7</sub> in the influent and effluent to the two biological processes in Table 8.4, it is noticeable that the results differs considerably. For the lab scale SBR's the highest concentration in the influent was calculated to be 2.10, meanwhile the average BOD<sub>7</sub> in the influent to the MBBR reached 7.40. One explanation to the lower concentrations in the lab scale SBR's could be that some of the organic matter in the leachates were oxidised in the feed tanks, before the leachates were fed into the reactors. Still, it was considered unlikely that the BOD<sub>7</sub> would differ to such extent between the SBR's and the MBBR. The BOD<sub>7</sub> result was based on the measured Dissolved Oxygen *DO* concentrations, which were considered to be unreliable and therefore contributed to a uncertain BOD<sub>7</sub> result.

#### ORGANIC MATTER IN THE LAB SCALE SBR'S

To get an overview of how the amount of organic material changed in the effluent from the lab scale SBR's during the experimental trial, the concentrations of TOC, DOC and COD in the effluent from Reactor 2 are illustrated in Figure 8.8. The concentrations of organic matter varied in a similar manner for Reactor 1, see Figure 1 in Appendix E.

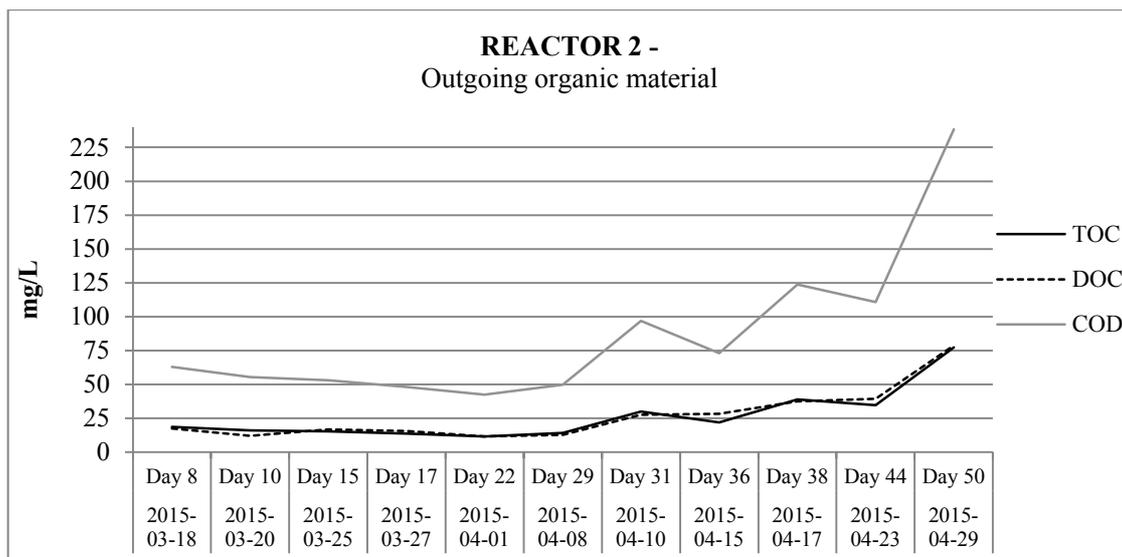


Figure 8.8 Concentrations of TOC, DOC and COD in the effluent from Reactor 2 throughout the experimental trial.

It can be seen in Figure 8.8 that all three parameters increased gradually during the experimental trial. The concentrations peaks on Day 31 and Day 38, which could be since the exchange ratio - hence the amount of methanol - in the lab scale SBR's were increased on Day 29 and Day 36. Since the aeration equipment did not work properly the days before and on Day 50, this could explain the high concentrations on that day. As previously stated, it can be observed in Figure 8.8 that the concentrations of TOC and DOC correlate well throughout the experimental trial.

#### ORGANIC MATTER IN THE MBBR

The concentrations of TOC and DOC in the effluent from the MBBR remained more consistent during the period for the experimental trial compared to the lab scale SBR's, see Figure 8.9. The COD was only measured three times during the experimental trial, which makes it difficult to evaluate.

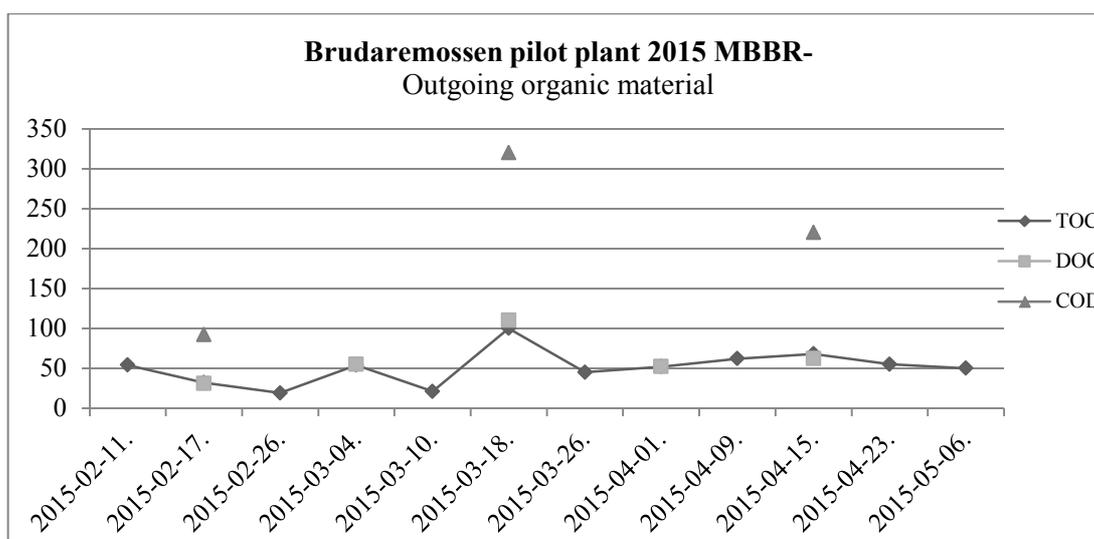


Figure 8.9 Concentrations of TOC, DOC and COD in the effluent from the MBBR during the experimental trial.

It can be seen in Figure 8.9, that the TOC, DOC and COD peaked in 2015-03-18 and that increased concentrations were noted the 2015-04-15. On the 2015-03-18 the methanol dosage was increased, which could have caused the peak in concentration on that day.

### 8.1.3 Removal of metals

Table 8.5 presents average concentrations of metals during last period of the experimental trial –when the exchange ratio was 0.50 - in the influent and effluent to Reactor 1 and Reactor 2. It also illustrates the average metal concentrations for the MBBR measured from 2015-02-11 to 2015-05-06. Metals that were detected in very low concentrations are not presented in this section. The result, including all the measured metals, is provided in Appendix C. The influent samples of the MBBR process was taken from data in sampling point B and the effluent samples from data in sampling point F - see Figure 6.6 in Chapter 6, Section 6.3.

*Table 8.5 Average concentrations of metals in the influent and effluent to the lab scale SBR's, at exchange ratio 0.50, and the MBBR at Brudaremsen pilot plant 2015.*

Exchange ratio 0.50		Lab scale SBR REACTOR 1		Lab scale SBR REACTOR 2		Brudaremsen pilot plant 2015 MBBR	
Parameter	Unit	Influent	Effluent	Influent	Effluent	Influent	Effluent
Fe	mg/L	0.58	0.26	0.17	0.10	5.10	0.13
Fe filtered	mg/L	0.08	0.05	0.08	0.04	0.05	0.07
Mn	mg/L	0.41	0.40	0.41	0.38	n.a.	0.06
Al	mg/L	0.23	0.12	0.10	0.10	0.17	n.a.
Toxic metals							
Ti	mg/L	0.39	0.41	0.41	0.39	n.a.	n.a.
Pb	µg/L	1.75	1.98	1.89	2.21	0.20	9.80
Cr	µg/L	0.00	0.00	0.00	0.00	3.35	n.a.
Co*	µg/L	3.09	3.37	2.13	2.29	n.a.	n.a.
Ni	µg/L	3.16	3.31	3.22	3.69	7.95	6.80
Li	µg/L	0.02	0.02	0.02	0.02	n.a.	n.a.
Cu *	µg/L	11.23	18.50	9.40	15.00	4.90	2.30
Zn	µg/L	9.11	8.12	7.32	10.70	21.50	8.10
Ag*	µg/L	<0.10	<0.10	<0.10	<0.10	<0.10	<0.10
Cd	µg/L	1.71	1.79	1.75	1.75	0.02	<0.02
Hg	µg/L	0.01	0.01	0.01	0.01	<0.01	<0.01

n.a. – not analysed

\* Based on measured concentrations at the exchange ratio 0.17

Metals in the influent and effluent were detected in very low concentrations for the two biological treatment techniques, see Table 8.5. In general the incoming concentrations were a bit less for Reactor 2, hence after the GAC filter. All toxic metals were detected in concentrations below 5 mg/L in the influent to the lab scale SBR's and the MBBR, which is favourable in order to avoid inhibition of the nitrification process. Some concentrations of the toxic metals were higher in the

effluent compared to the influent for the lab scale SBR's and the MBBR. This was believed to be since some of the metals accumulate in the biomass and therefore remained in the reactors. Since the water samples for the lab scale SBR's were taken manually, biomass - and thereby a certain concentration of metals - was inevitably discharged with some of the samples.

Since the incoming leachates to Reactor 2 had passed through more treatment steps within the Brudaremsøen pilot plant 2015 compared to the incoming leachates to Reactor 1, it was expected that the concentrations of metals should have been lower in the influent for Reactor 2. However, the results presented in Table 8.5 show that the average concentrations of metals in the influent to the reactors were very similar. The result indicates that the sand and GAC filters in the pilot plant did not remove toxic metals extensively. However, as previously stated, the metals concentrations in the incoming leachates to the pilot plant were low and many concentrations were measured to be below the detection limit.

When comparing the metal concentrations in the influent to the lab scale SBR's with the concentrations in the incoming leachates to the pilot plant, it can be seen that lead *Pb* and cadmium *Cd* were approximately ten times higher in the incoming leachates. This was also the case when comparing the *Pb* and *Cd* concentrations in the influent to the lab scale SBR's with the concentrations in sampling point B and G' in the pilot plant. Based on the results, it appears that *Pb* and *Cd* accumulated within the feed tanks. However, if that was the case it does not explain why the concentrations were high in the water samples, since it is more likely that the metals would have accumulated in the sediments at the bottom of the tanks. The high concentrations could also be a result of an inaccurate analysis or that the samples became contaminated at some point during preparation. To get an overview of how the metal concentrations varied during the experimental trial, some that deviated from the average values are further discussed in the following paragraphs.

The concentrations of metals remained relatively stable in Reactor 2 during the experimental trial. However, some metals in Reactor 1 deviated from the average concentrations. Figure 8.10 presents the concentrations of manganese *Mn*, iron *Fe* and Titanium *Ti* in the influent to Reactor 1. The same metals for Reactor 2 are illustrated in Figure 2, Appendix E.

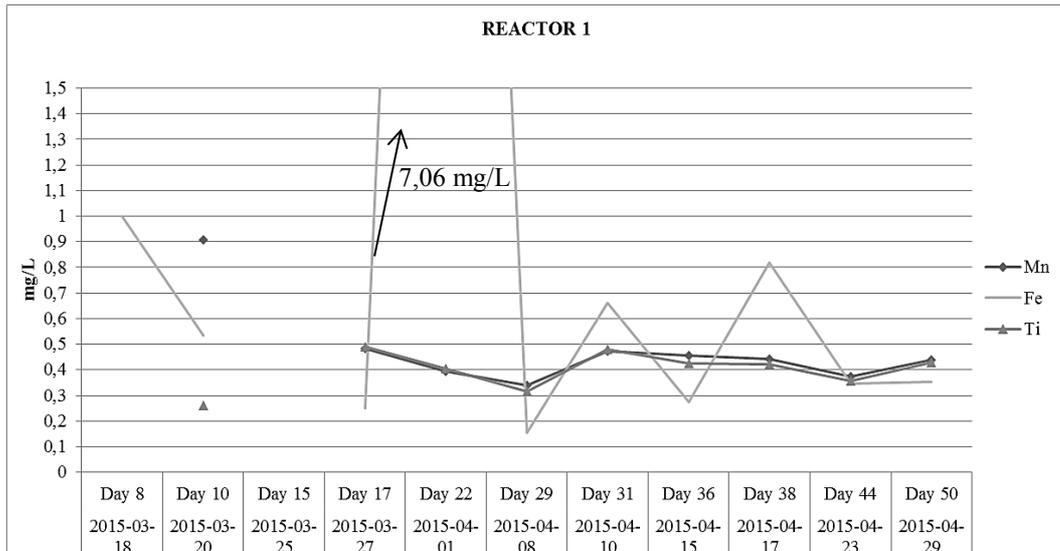


Figure 8.10 Concentrations of Mn, Fe and Ti in the influent to Reactor 1 throughout the experimental trial.

On Day 22 the concentration of Fe peaked at 7.06 mg/L, which was approximately six times higher than the average Fe concentration presented in Table 8.5. During the sampling occasion on Day 22, the feed tanks to the lab scale SBR were mistakenly refilled before sampling. It was also noted that sediments had been accumulated at the bottom of the feed tanks. The sediment in the feed tank for Reactor 1 was believed to consist of unsettled flocs, distributed from the lamellae sedimentation. The brownish red colour of the sediment indicated high concentration of Fe in the settled flocs. When new leachates were added to the feed tank, it stirred up the sediment which could explain the increased concentrations on that particular day. At all the other sampling occasions, the samples for the incoming leachates were taken before the feed tanks were refilled.

For Reactor 1, the concentration of zink Zn also peaked once during the experimental trial. The concentrations of Pb, chromium Cr, Zn and copper Cu in the influent to Reactor 1 are presented in Figure 8.11. See Figure 3 in Appendix E for the same metal concentrations in Reactor 2 during the experimental trial.

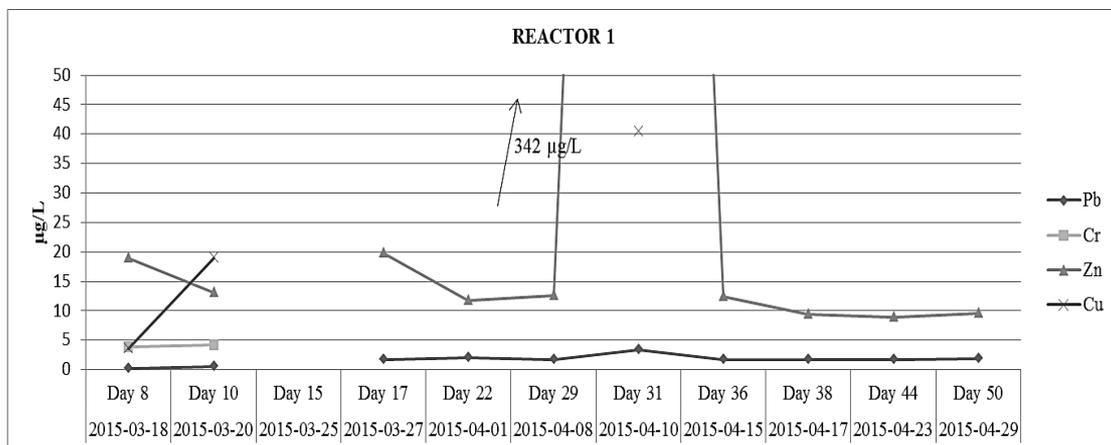


Figure 8.11 Concentrations of Pb, Cr, Zn and Cu in the influent to Reactor 1 throughout the experimental trial.

For Day 31 the concentration of Zn in the influent for Reactor 1 was measured to be 342 µg/L, which is more than seven times higher than the average value presented in Table 8.5. No operational errors were observed during Day 31, but as can be seen in Figure 8.10, the concentrations of Mn, Fe and Ti also increased slightly during that day. The extreme peak could also be due to errors performed during analysis.

## 8.2 Operational parameters for the lab scale SBR's and the MBBR

In this section, critical operational parameters for the lab scale SBR's and the MBBR analysed during the experimental trial are presented and discussed. It is also evaluated if the treatment techniques efficiency were affected by the parameters.

### 8.2.1 The pH during the experimental trial

The pH during the experimental trial in the lab scale SBR's are presented in Figure 8.12. In the beginning of the experimental trial the pH was measured to be close to eight in both reactors. Thereafter the pH decreased down to 7.07 on Day 22 which was the lowest measured value during the experimental trial, see marked area in Figure 8.12. The pH in the incoming leachates to the MBBR was not analysed by Ramböll, but the average pH during nitrification and denitrification were 7.49 and 7.86 respectively.

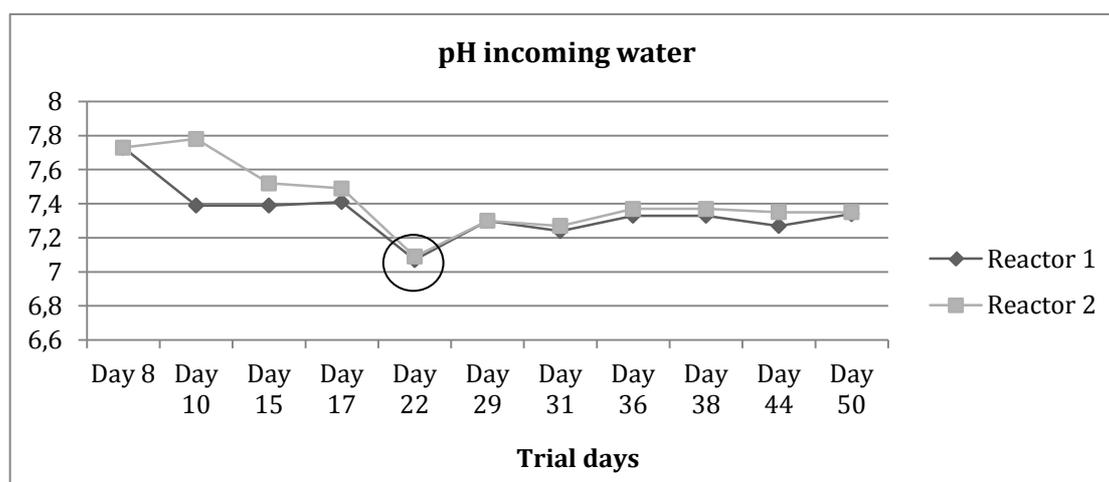


Figure 8.12 The pH in the incoming leachates to the lab scale SBR's during the experimental trial.

It can be evaluated that during the later operational period of the trial, the pH in the incoming leachates to both reactors stabilised at approximately 7.30, see Table 8.1. The optimal pH for nitrifying and denitrifying bacteria is from 7.50 to 8.50 and 7.00 to 9.00 respectively. Therefore, increased levels of NH<sub>4</sub>-N and NO<sub>2</sub>-N were expected for the effluent samples in both reactors for Day 22. However, the decrease in pH did not affect the process of nitrification significantly in any of the reactors - see Figure 8.1 and Figure 8.2. The concentration of NO<sub>3</sub>-N and N<sub>tot</sub> in the effluent from Reactor 1 did however peak on Day 22 which could be related to the drop in pH.

The pH in the MBBR during the biological processes was evaluated to be adequate for achieving efficient treatment.

## 8.2.2 Dissolved Oxygen during the experimental trial

The DO concentrations in the lab scale SBR's were measured continuously during one cycle at two occasions – Day 13 the 23<sup>rd</sup> of March and during the cycle analysis on Day 44 the 23<sup>rd</sup> of April. Measured DO concentrations during the oxic and anoxic phase for the two reactors are presented in Table 8.6. The concentrations represent the lowest and highest DO concentrations during the oxic and anoxic phase. The average DO concentration during the oxic phase in the MBBR was analysed to be 10.30 mg/L, whereas the DO during the anoxic phase was not measured but was considered to be low.

*Table 8.6 DO concentrations during the oxic and anoxic phase in both reactors on Day 13 and Day 44.*

Parameter	Unit	Day 13		Day 44	
		Oxic	Anoxic	Oxic	Anoxic
DO Reactor 1	mg/L	1.60-4.35	3.50-0.30	0.61-0.70	0.28
DO Reactor 2	mg/L	1.27-4.50	2.76-0.29	0.51-2.60	0.29

When comparing the DO concentrations for the two days it can be seen that the oxygen  $O_2$  supply to the reactors worked better on Day 13 compared to Day 44. During the oxic phase on Day 13 the  $O_2$  levels reached more than 4 mg/L, which is higher than the lower recommended limit of two to 3 mg/L for achieving nitrification. The measured concentrations one month later showed a significantly decrease of DO in the reactors. In Reactor 2, the DO concentration still exceeded 2 mg/L during the oxic phase whilst the DO concentration in Reactor 1 was measured to be lower than 1 mg/L, which is insufficient for achieving nitrification. However, as can be seen in Section 8.1.1, Figure 8.1 and Figure 8.2, the process of nitrification still seemed to occur in both reactors and no significant increase of  $NH_4-N$  or  $NO_2-N$  were detected. The DO concentrations at the end of the anoxic phase did not change between Day 13 and Day 44 and were approximately 0.30 mg/L.

The reduction of DO was believed to be caused due to clogging of the aeration equipment. Since the equipment used was designed to aerate aquariums, it is likely that the amount of sludge in the reactors caused deterioration of its capacity. To achieve a more continuous level of DO for the lab scale SBR's, more robust aeration equipment would have been required. Furthermore, it must be taken into consideration that Multiline P4 is a measuring device with limited accuracy. The parameter  $O_2$  is difficult to measure and in order to obtain more certain results, a more advanced measurement apparatus would have been required. One option could have been to install an automatic control system on site, which probably would have provided more continuous and reliable DO results.

The DO concentration during the oxic phase in the MBBR was significantly higher than for the lab scale SBR's. This was due to that the  $O_2$  supply was regulated automatically for the MBBR and with a more well-suited aeration device for this purpose.

### 8.2.3 Temperature during the experimental trial

As can be seen in Figure 8.13, the temperature in the incoming leachates to the two reactors steadily increased throughout the experimental trial, from approximately 7.8 °C to 13 °C. Since the feed tanks for the incoming leachates were placed in the same room at the Brudaremossen pilot plant 2015, the measured temperatures were similar. The average temperature in the influent and effluent for the lab scale SBR's were slightly higher compared to the MBBR. As can be seen in Table 8.1, Section 8.1, the temperature in the reactors increased approximately 2 °C after treatment. During the biochemical processes of nitrification and denitrification, energy was produced which caused the temperature to rise in the reactors.

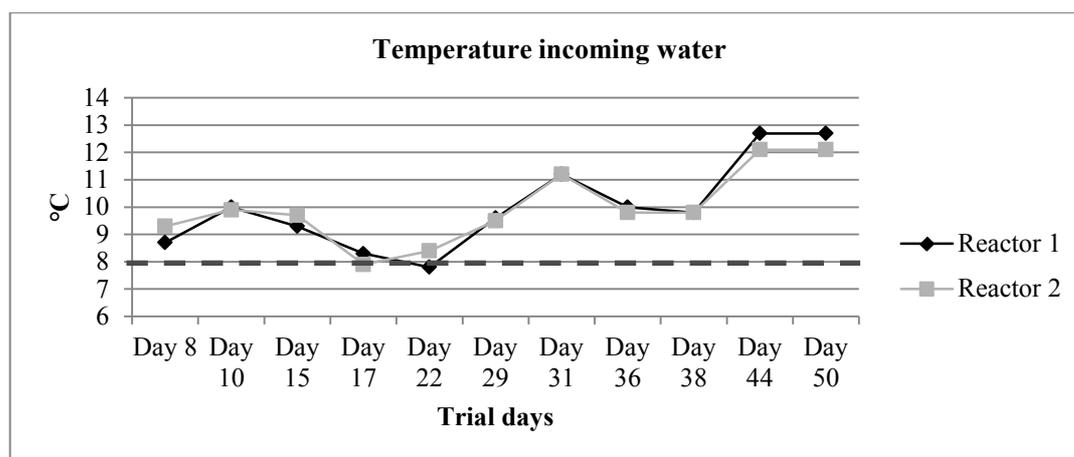


Figure 8.13 Temperature in the incoming leachates to the lab scale SBR's during the experimental trial

During Day 17 and Day 22, the temperature dropped beneath 8 °C in Reactor 2 and Reactor 1 respectively, see marked line in Figure 8.13. Temperatures below 8 °C can have a negative impact on the growth of nitrifying bacteria whilst the denitrifying bacteria are less sensitive and can survive down to a temperature of 5 °C. When comparing with the result regarding nitrogen fractions in Reactor 1, see Figure 8.1 in Section 8.1.1, it can be observed that the nitrification worked properly during these days whilst the denitrification was not working. Since the denitrification should be able to take place at lower temperatures, the drop in temperature is not believed to be related to the malfunction of the denitrification process in Reactor 1. In Reactor 2, the biological processes seemed to be unaffected by the drop in temperature.

When comparing the operational temperature during the experimental trial with the temperatures in the cases studied during the literature review it can be seen that the latter mentioned were significantly higher – see Table 5.2 in Chapter 5, Section 5.1. During the cases the SBR's were operated at room temperatures – close to 20 °C or higher – which is favourable for achieving an efficient biological treatment. Therefore, it would have been interesting to analyse the lab scale SBR's treatment capacity during summer in order to determine if the nitrification and denitrification rates improved.

### 8.2.4 Dosage of phosphorus

The measured concentrations of total phosphorus  $P_{tot}$  after treatment in Reactor 1 and Reactor 2 are presented in Figure 8.14. At exchange ratio 0.50 the average  $P_{tot}$  in the

influent to the two reactors were 0.03 – 0.04 mg/L and approximately 13–15 mg/L in the effluent. According to Ramböll in *Lakvattenhantering vid Brudaremossen - Delrapport 2* (Ramböll, 2015 C), the threshold value regarding  $P_{tot}$  for discharge into the river Göta älv, the stream Sävån and the creek Finngösabäcken is 0.50 mg/L and 0.30 mg/L for the creek Delsjöbäcken and the stream Mölndalsån. The resulted  $P_{tot}$  concentrations in the effluent from the lab scale SBR's are all above 0.50 mg/L, see Figure 8.14, hence does not fulfil the threshold values for neither of the potential recipients.

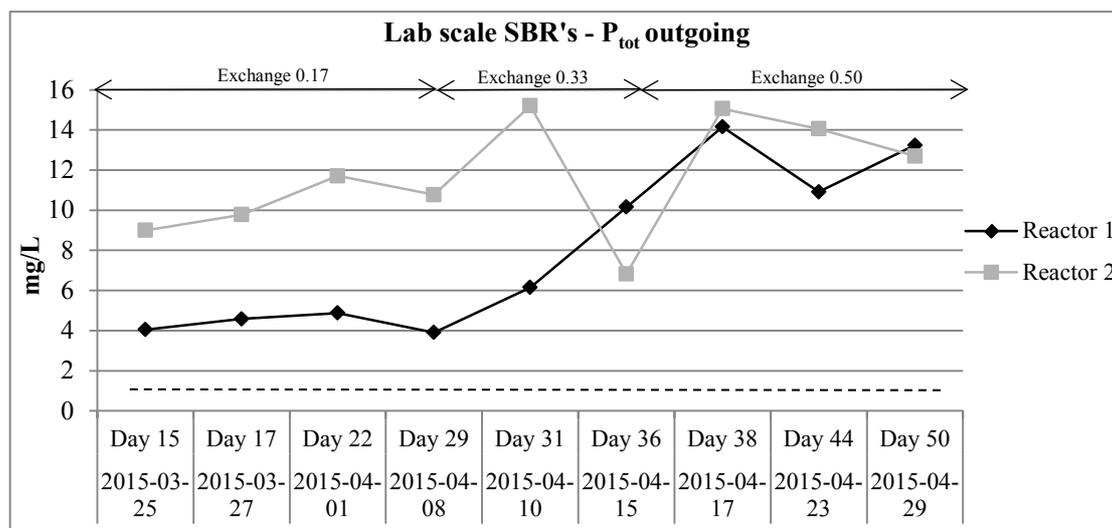


Figure 8.14 Concentrations of  $P_{tot}$  after treatment in the lab scale SBR's throughout the experimental trial.

When calculating the required dosages of phosphoric acid to the lab scale SBR's, see Chapter 7, Section 7.2.1, an assumed load of 100 mg  $NH_4-N/L$  – 400 mg  $NH_4-N/cycle$  – was used. This value was based on the average and maximum  $N_{tot}$  concentrations in untreated leachates from Brudaremossen during the period from 2008 to 2012 (Ramböll, 2015 B). However, the results from the analysed samples taken from the feed tanks show that the average concentrations of  $NH_4-N$  in the incoming leachates were 49 mg/L and 43 mg/L for Reactor 1 and Reactor 2 respectively, see Table 8.1 in Section 8.1. When applying the measured average concentrations of  $NH_4-N$ , it resulted in the reduced dosages of phosphoric acid; 0.11, 0.22 and 0.33 ml/min for the three different exchange ratios. The new concentrations are approximately half of the dosages that were added to the lab scale SBR's. To obtain the optimal dosage of phosphoric acid to the reactors, further analysis would have been required.

Reactor 2 generally had higher concentrations of  $P_{tot}$  compared to Reactor 1, see Figure 8.14. During the experimental trial, the amount of phosphoric acid in the glass bottles was noted. The two bottles were continuously filled with the same amount of phosphoric acid. However, it was observed that the amount of acid in the bottle distributing phosphorous to Reactor 2 was reduced faster compared to Reactor 1. It is uncertain why the dosage of phosphorous was uneven, but as seen in Figure 8.14 the levels of  $P_{tot}$  in the two reactors became more even at the end of the study. This may indicate that the pump needed time to adjust to the programmed Revolutions Per Minute *RPM*.

When comparing the dosage of phosphorus for the lab scale SBR's with the added dosages for the studied SBR's in the cases studied during the literature review, see Table 5.2 in Chapter 5, it can be seen that they were lower than 1 mg/L whereas seven to 20 mg/L was added to the reactors at Brudaremossen. On the other hand, the dosage of phosphorus was only presented in two of the cases which make the comparison uncertain. Also, since the concentration of  $P_{tot}$  was very low in the incoming leachates from Brudaremossen, higher dosages of phosphorus to the lab scale SBR's were required in order for the nitrification to function. It could be that the leachates applied in the studied cases had a higher  $P_{tot}$  concentration, which could explain the lower phosphorus dosages to the SBR's.

The concentration of  $P_{tot}$  in the effluent from the MBBR process in the Brudaremossen pilot plant 2015 during the experimental trial is presented in Figure 8.15. Compared to the results for the lab scale SBR's, the concentration of  $P_{tot}$  decreased in the MBBR and during April the concentrations were measured to be below the established threshold values for discharge.

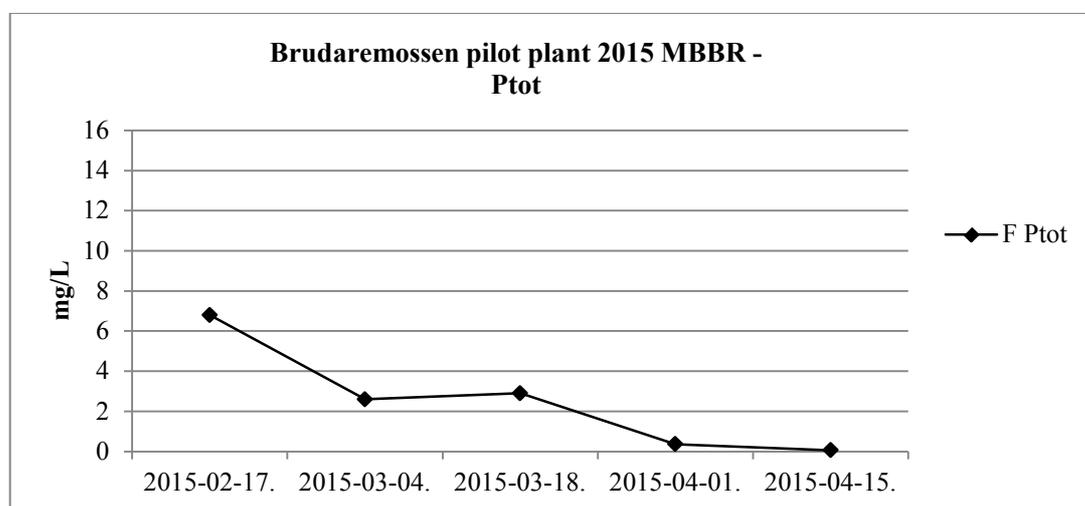


Figure 8.15 Concentrations of  $P_{tot}$  (measured in sampling point F) in the leachates after treatment in the MBBR throughout the experimental trial.

### 8.2.5 Dosage of methanol

Similar to the calculated dosage of phosphorus, the required dose of methanol added to the lab scale SBR's was based on an assumed load of 100 mg  $NH_4-N/L$ . As presented in Section 8.1.2 the levels of TOC, DOC and COD were higher in the effluent compared to in the influent for the lab scale SBR's and the MBBR. The increased amount of carbon in the effluent is likely to have been caused by overdosage of methanol during the anoxic phase or that biomass did not settle properly in the lab scale SBR's and in the settling tank after the MBBR.

No optimisation of methanol for the two techniques was performed during the experimental trial, which could have reduced the usage of methanol and the outgoing concentrations of organic compounds if performed. Based on the assumed load of 100 mg  $NH_4-N/L$  the required methanol dosage was calculated to be 0.84, 1.69 and 2.52 mL/min for each exchange ratio. However, when applying the measured average concentration of  $NH_4-N/L$  in the influent to the reactors, the methanol dosages were reduced to approximately 0.46, 0.93 and 1.24 mL/min.

The final dosage of methanol to the MBBR was 4.32 mL/min, which is significantly higher compared to the lab scale SBR's. As presented in Section 8.1.2, the outgoing concentrations of organic compounds were higher for the MBBR which could be due to the higher methanol dosage.

When comparing the methanol dosages applied in the reactors and the dosages applied in the studied cases in the literature review, see Table 5.3 in Chapter 5, they differed significantly. When the applied methanol dosages for the lab scale SBR's were converted into milligrams per litre, it could be seen that they ranged from nearly four to 11 mg/L. Much greater or significantly lower dosages of external carbon sources were applied to the SBR's in the cases, which makes it difficult to compare and to draw any conclusions. The various required amount of carbon is likely to be related to the quality of the leachates and how high the COD concentrations were in the influent to the SBR's. Therefore, it is easier to compare the COD/N ratio for the cases and the lab scale SBR's. The average COD/N ratio for Reactor 1 and Reactor 2 was evaluated to be 44.85 and 33.88 respectively. Compared to the cases, where the COD/N ratio varied from 2.96 to 9.51, the ratio for the lab scale SBR's were much higher. According to theory – see Chapter 7, Section 7.2.2 – the ratio of COD:NO<sub>3</sub>-N should be approximately (3.5–5):1 for wastewater treatment which corresponds more to the COD/N ratio evaluated in the literature review for the cases. This shows that there were an excess of COD in the lab scale SBR's which implies an overdosage of methanol.

Furthermore, in CASE 5 – Chapter 5, Section 5.3.2 – it was concluded that the optimal nitrogen/COD ratio was 1.08. The nitrogen/COD ratios for the lab scale SBR's were calculated to range from 0.02 to 0.03 which were very low. If applying the theoretical relation between nitrogen and COD of (3.5–5):1, which results in a nitrogen/COD ratio of 0.20, the resulted ratio for the lab scale SBR's confirms that the addition of external carbon source to the reactor was too high.

In Reactor 1, the denitrification did not start to work properly until Day 38. According to Monclús et al. (2009), it might take some time to adapt the system to the addition of carbon, which could be the case for the processes in Reactor 1.

### **8.2.6 Precipitation during the experimental trial**

Data showing the level of precipitation from a nearby monitoring station at Lake Tvärnsjön during the period 2015-02-28 to 2015-04-29 is presented in Figure 8.16. It can be observed that the precipitation decreased by approximately 75 % during the studied period.

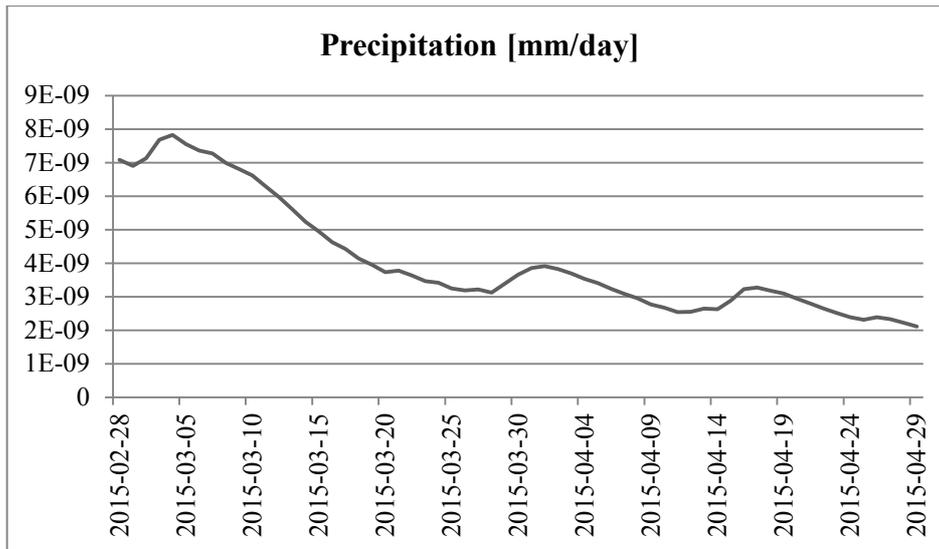


Figure 8.16 Precipitation at a monitoring station at Lake Tvärnsjön located close to Brudaremassen (SMHI Vattenwebb, 2015).

As stated in Chapter 6, Section 6.1, it is likely that the level of precipitation affects the quality and quantity of the leachates from Brudaremassen. However, the results from the experimental trial show little indication of correlation between the decreased level of precipitation and the measured concentrations of parameters in the leachates. Also, when comparing the concentration of parameters in the incoming leachates to the Brudaremassen pilot plant 2015 between February and April, no significantly increased concentrations were noted. It could be that more frequent sampling would have been required in order to determine if there were correlations.

### 8.3 Cycle analysis of the lab scale SBR's

In this section, the results from the cycle analysis performed on Day 44, the 23<sup>rd</sup> of April are presented and thereafter discussed. Initially, the measured concentrations of nitrogenous fractions in the reactors during the cycle analysis are illustrated. Section 8.3.1 and Section 8.3.2 will provide the results concerning operational parameters and nitrification and denitrification rates during the analysis. Finally, the results from the cycle analysis will be discussed in Section 8.3.3.

The results from the cycle analysis performed on Day 44, the 23<sup>rd</sup> of April, showed that the nitrification and denitrification functioned properly in the lab scale SBR's. In Reactor 1, the concentration of  $\text{NH}_4\text{-N}$  increased steady during the fill phase and was thereafter reduced during the oxic phase to approximately 3 mg/L, see Figure 8.17. The concentration of  $\text{NO}_3\text{-N}$  increased as the  $\text{NH}_4\text{-N}$  decreased, but was thereafter significantly reduced during the anoxic phase to approximately 1 mg/L. There was no visible increase of  $\text{NO}_2\text{-N}$ , even though it theoretically should occur during the nitrification process before the formation of  $\text{NO}_3\text{-N}$ .

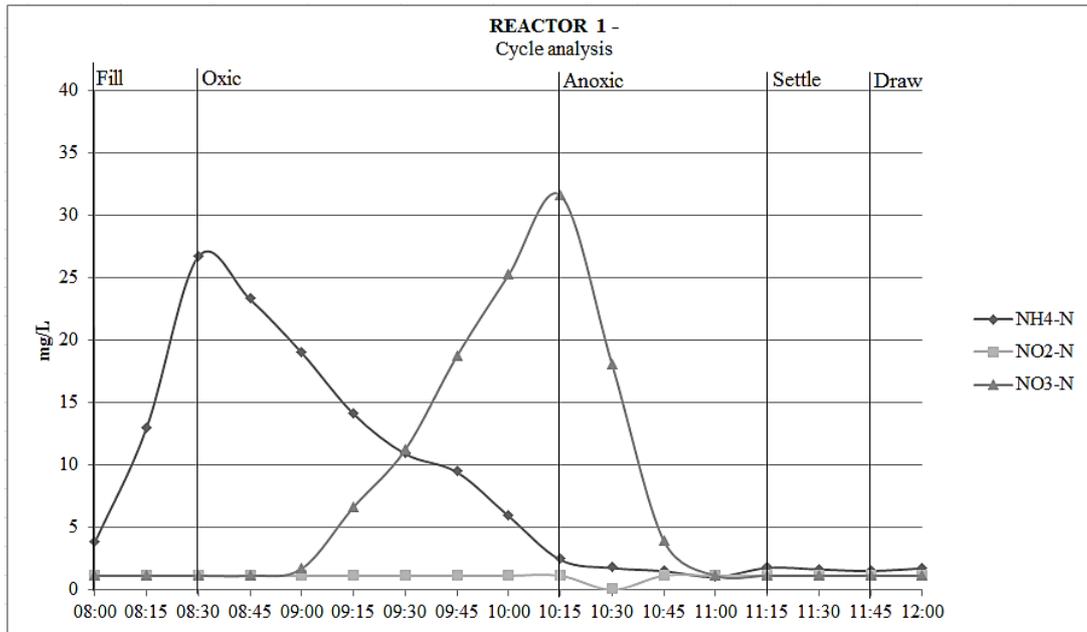


Figure 8.17 Concentrations of nitrogenous fractions during one cycle on Day 44, 2015-04-23, for Reactor 1.

The results from the cycle analysis also showed that the nitrification and denitrification functioned properly during Day 44 in Reactor 2, see Figure 8.18. The concentration of NH<sub>4</sub>-N increased to only 6 mg/L during the fill phase but the next sample, taken at 08:45 in the oxidic phase, showed significantly higher concentrations of NH<sub>4</sub>-N. It was thereafter reduced to approximately 1 mg/L at the end of the oxidic phase. The concentration of NO<sub>3</sub>-N increased as the NH<sub>4</sub>-N decreased, but was thereafter significantly reduced during the anoxic phase to approximately 1 mg/L. Similar to Reactor 1, there was no visible increase of NO<sub>2</sub>-N in Reactor 2 during the cycle analysis.

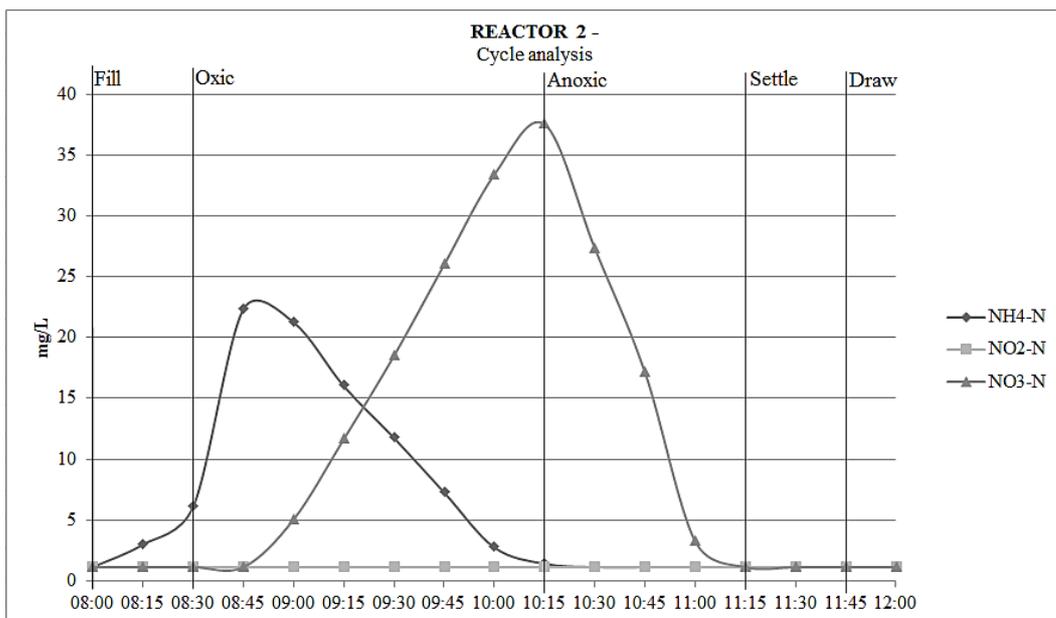


Figure 8.18 Concentrations of nitrogenous fractions during one cycle on Day 44, 2015-04-23, for Reactor 2.

### 8.3.1 Dissolved Oxygen, pH and temperature during the cycle analysis

The concentrations of DO in both Reactor 1 and Reactor 2 were measured during the cycle analysis on Day 44, see Figure 8.19. The results show that the DO concentrations were generally similar for both reactors in all phases except during the oxic phase where Reactor 2 had a considerably higher concentration.

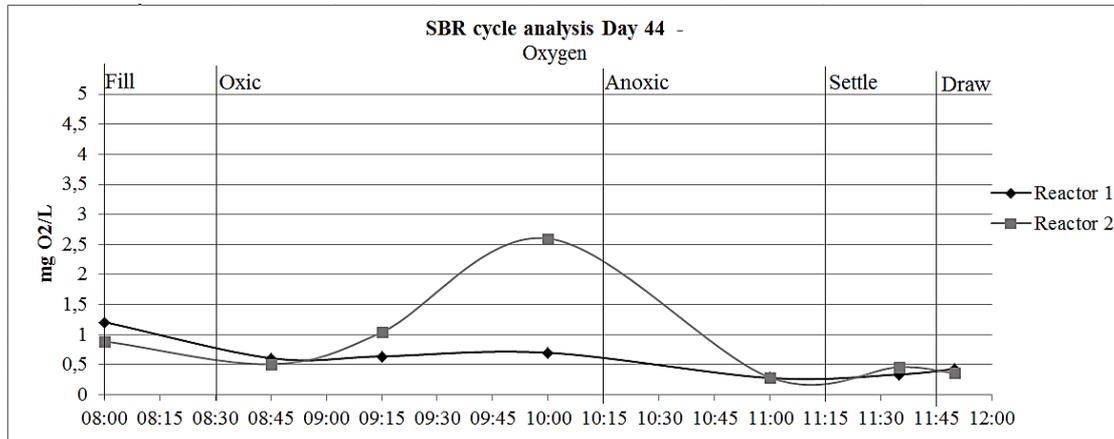


Figure 8.19 DO concentrations during the cycle analysis on Day 44, 2015-04-23, for Reactor 1 and Reactor 2.

As presented in Section 8.2.2, the DO concentration in the reactors was also analysed on the Day 13 the 23<sup>rd</sup> of March, which showed that the aeration equipment operated better in the beginning of the trial. The DO concentrations during one cycle on Day 13 in the reactors are illustrated in Figure 1 in Appendix F.

The pH was also measured for Reactor 1 and Reactor 2 during the cycle analysis on Day 44, see Figure 8.20. Both Reactor 1 and Reactor 2 experiences similar changes in pH during one cycle, with a decrease in pH during the oxic phase and an increase in pH during the anoxic phase. The pH during one cycle on Day 13 in the reactors is illustrated in Figure 2 in Appendix F.

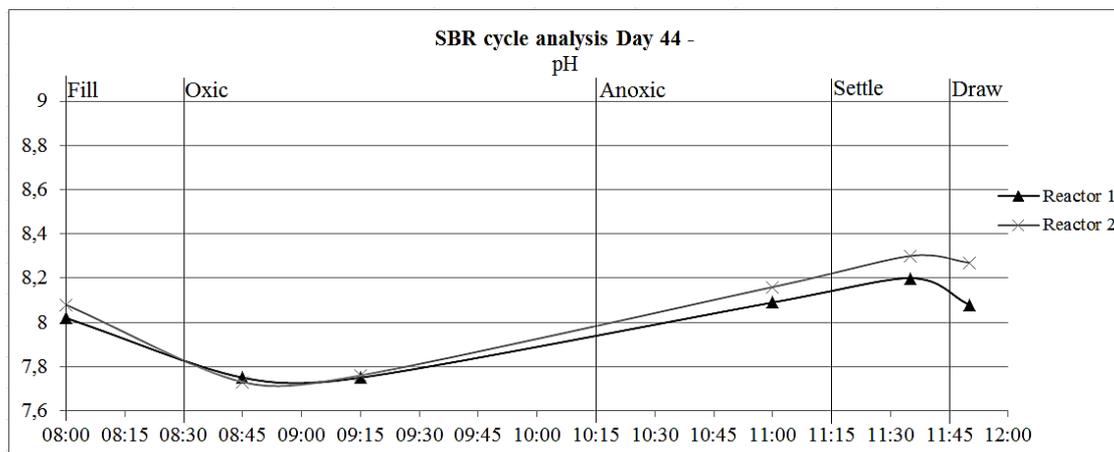


Figure 8.20 The pH measured during the cycle analysis on Day 44, 2015-04-23, for Reactor 1 and Reactor 2.

During the oxic phase, hydrogen ions  $H^+$  are created during the nitrification process - see equation (7), Chapter 3 in Section 3.2 - which causes the pH to drop during that phase. When the denitrifying bacteria reduce  $NO_3^-$  into  $N_2$  hydroxide  $OH^-$  is produced - see equation (9), Chapter 3 in Section 3.2 - causing the pH to increase in the anoxic phase.

Throughout the cycle analysis, the temperature was continuously measured in Reactor 1 and Reactor 2, see Figure 8.21. The temperatures in the reactors were rather alike during the analysis, except from at the beginning of the cycle. It can be observed in Figure 8.21 that the temperatures dropped in the reactors during the fill phase and during the beginning of the oxic phase. Thereafter the temperatures increased similar during the greater part of the oxic phase until the end of the anoxic phase.

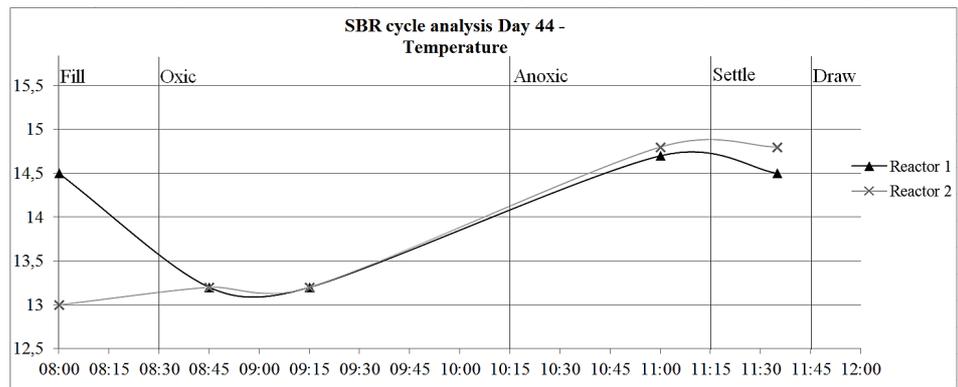


Figure 8.21 Temperatures measured during the cycle analysis on Day 44, 2015-04-23, in Reactor 1 and Reactor 2.

### 8.3.2 Nitrification and denitrification rates during the cycle analysis

The nitrification and denitrification rates were calculated for Day 44, by creating trend lines for the declining concentration of  $NH_4-N$  during nitrification and  $NO_3-N$  during denitrification. The slopes of the trend lines are then considered as the nitrification and denitrification rates in milligrams of removed nitrogen per hour. See Figure 8.22 for an example of the trend lines added for Reactor 1. The calculated removal rates are marked in the figure. Trend lines for Reactor 2 are illustrated in Figure 1 in Appendix G.

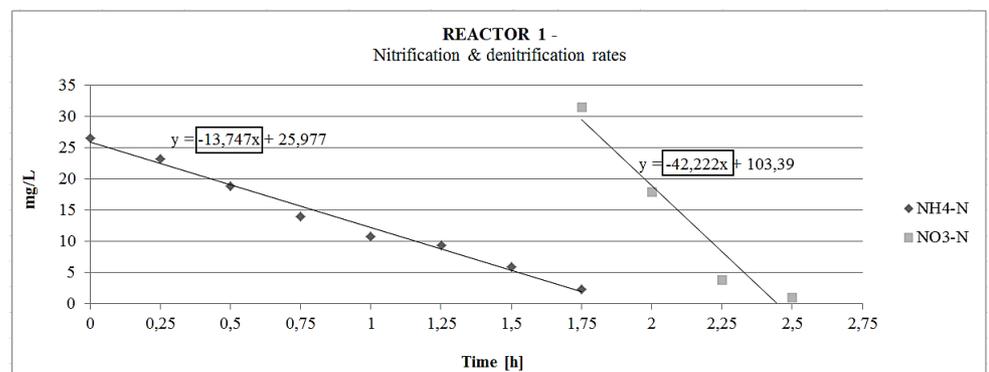


Figure 8.22 Nitrification and denitrification rates during the cycle analysis on Day 44, 2015-04-23, for Reactor 1.

The nitrification and denitrification rates were also calculated in milligrams of removed nitrogen per grams of Mixed Liquor Volatile Suspended Solids *MLVSS*, which was analysed for the activated sludge in both reactors for the sludge samples taken during the cycle analysis. See Table 8.7 for a summary of the nitrification and denitrification rates in Reactor 1 and Reactor 2, presented both in removed nitrogen per hour and grams of *MLVSS*.

*Table 8.7 Nitrification and denitrification rates for Reactor 1 and Reactor 2 in both mg NH<sub>4</sub>-N or NO<sub>3</sub>-N / L·h and mg NH<sub>4</sub>-N or NO<sub>3</sub>-N / g MLVSS·h.*

<b>Day 44</b>	<b>REACTOR 1</b>	<b>REACTOR 2</b>	<b>Unit</b>
Nitrification rates	13.75	15.51	mg NH <sub>4</sub> -N / L·h
	3.80	4.43	mg NH <sub>4</sub> -N / g MLVSS·h
Denitrification rates	42.22	38.77	mg NO <sub>3</sub> -N / L·h
	11.66	11.08	mg NO <sub>3</sub> -N / g MLVSS·h
MLVSS	3,620	3,500	mg/L

### 8.3.3 Cycle analysis - discussion of the results

The absence of increased NO<sub>2</sub>-N concentrations during the cycles in Reactor 1 and Reactor 2 can be explained by that the NO<sub>2</sub>-N was directly oxidised into NO<sub>3</sub>-N. The rapid oxidation of NO<sub>2</sub>-N to NO<sub>3</sub>-N indicates that the level of O<sub>2</sub> available in the reactors were sufficient, hence that the aeration equipment fulfilled the requirements on that day.

The nitrogenous fractions in each sample were analysed using IC analysis, which has a lower detection limit of 0.01 millimoles per litre. Due to a high conductivity in the samples, they had to be diluted from eight to ten times before the analysis in order to minimise risk of damaging the IC machinery. The dilution combined with the lower detection limit lead to an incapability of detecting concentrations lower than 1 mg/L in all samples. Therefore, the impression might be that the lab scale SBR's only decreases the nitrogenous fractions to 1 mg/L, when they in reality were reduced to below 1 mg/L. However, since the concentrations of N<sub>tot</sub> were measured for both tanks using the European Standard method; EN 1484:1997 for the effluent samples from Day 44, it is possible to be certain that the combined outgoing concentrations of all nitrogenous fractions were 1.259 mg/L for Reactor 1 and 1.42 mg/L for Reactor 2.

The incoming concentration of NH<sub>4</sub>-N should theoretically be equal to the concentration of NO<sub>3</sub>-N after nitrification for both Reactor 1 and Reactor 2. The reason to why the results from the IC analysis indicate differently can be due to that reactions might have occurred in the samples after they had been taken, the samples may have been contaminated or due to that measurement errors might have occurred. However, when looking at the incoming concentrations of NO<sub>3</sub>-N and NO<sub>4</sub>-N to the reactors on that day it was noted that they were similar to the highest measured concentrations of NO<sub>3</sub>-N during the cycle analysis. It is therefore the concentrations of NO<sub>4</sub>-N during the oxic phase that are uncertain since they theoretically should have been higher, especially for Reactor 2. In Reactor 2 the NO<sub>4</sub>-N was 6 mg/L at 08.30, which was incomprehensible, since new leachates with NO<sub>4</sub>-N concentrations close to 31 mg/L were added during the fill phase. In order to obtain more reliable results the samples could have been filtered or centrifuged directly on site. This would have

reduced the amount of sludge in the samples and therefore minimising possible reactions or errors.

The results regarding the DO concentrations in both tanks showed that it were considerably higher in Reactor 2 during the oxic phase, see Figure 8.19. According to Gerardi (2002), the DO should be between 2 to 3 mg/L during the oxic phase in order to supply the nitrifying bacteria with enough O<sub>2</sub> for the nitrification process to become optimal. The DO concentration for Reactor 2 reaches 2.5 mg/L during the oxic phase but was less than 1 mg/L for Reactor 1 during the entire cycle. Figure 8.17 does however show that the nitrification in Reactor 1 functioned properly even though the DO was lower than 1 mg/L during the oxic phase.

The resulted pH concentrations confirmed that biological processes were taking place in the reactors at the end of the experimental trial. The decreased and increased pH during the oxic and anoxic phase indicated that H<sup>+</sup> and OH<sup>-</sup> were produced in the reactors which in turn are by-products from nitrification and denitrification. Furthermore, it can be seen in Figure 8.20 that the pH fluctuated from 7.7 up to 8.3 which are within the preferable limits for achieving nitrification and denitrification. As for the DO, the pH was also measured in the reactors on Day 13; see Figure 2 in Appendix F. When comparing the pH throughout one cycle in the beginning and at the end of the experimental trial, it could be seen that it improved and stabilised. On Day 13 only a minor decrease in pH was noted during the oxic phase - which indicates that the nitrification rate was low - and the pH in the anoxic phase decreased, implying that denitrification did not function properly in the reactors on that day.

Concerning the temperature throughout the cycle analysis, the results showed that it initially decreased during the fill phase and thereafter increased until at the end of the anoxic phase. The microorganisms that perform nitrification and denitrification produce energy which explains the increased temperature in the reactors during the oxic and anoxic phase. In Figure 8.21, see Section 8.3.1, it can be noted that the temperature did not start to increase at once during the oxic phase, which could be since it took some time for the nitrifying bacteria to start oxidising the NH<sub>4</sub>-N in the new leachates. The temperature curves are evened out at the end of the anoxic phase, which could indicate that there were less NO<sub>3</sub>-N available in the leachates for the nitrifying bacteria to convert into N<sub>2</sub> at that time.

When comparing the nitrification rates during the cycle, it can be seen that the nitrification rate was higher for Reactor 2 than for Reactor 1, see Table 8.7 in Section 8.3.2. This might be due to the insufficient DO concentration during the oxic phase in Reactor 1, which could have caused a less efficient nitrification process. The aeration equipment used for the experimental SBR trial was originally intended for smaller aquariums and was unable to provide an evenly distributed air flow throughout the reactors. Improved aeration equipment would enable a more even aeration of the reactors which would lead to a higher nitrification rate and more efficient treatment. The denitrification rate was shown to be slightly higher in Reactor 1 during the cycle analysis. This might also be related to the DO concentration, since it can be seen that it was slightly higher in Reactor 2 during the anoxic phase. It is preferable to achieve a low level of DO during the anoxic phase since this will force the denitrifying bacteria to use the bound O<sub>2</sub> in nitrate NO<sub>3</sub><sup>-</sup> during denitrification instead of DO.

The nitrification and denitrification rates achieved during this study were compared to other cases studied in the literature review. CASE 1 at the Filborna landfill achieved

nitrification rates of 23-23.6 mg NH<sub>4</sub>-N/L·h and denitrification rates of 178.8-184.5 mg NO<sub>3</sub>-N/L·h. In CASE 2 at the Spillepeng landfill, nitrification rates of 7.6-32 mg NH<sub>4</sub>-N/L·h and denitrification rates of 11-180 mg NO<sub>3</sub>-N/L·h were achieved. The results from the experimental trial compared to the cases show that both the nitrification and denitrification rates are lower for the lab scale SBR's. However, the first result from the Spillepeng landfill study showed lower rates than for the lab scale SBR's. It was for CASE 2 concluded that it was a result of a low DO concentration during the oxic cycle, but improvement of the aeration resulted in higher rates later on during the study.

Further cycle studies should have been performed during the experimental trial period in order to compare efficiencies and to have a more reliable result of nitrification and denitrification rates. It should be considered that full cycle studies are time consuming and requires many samples to be analysed. However, the results from such a study can be very useful when analysing possible problems or when optimising the SBR treatment. Additional studies of the DO concentrations during one cycle could also improve the reliability of the results and provide further understanding of such an important operational parameter as the aeration during the SBR treatment.

## 8.4 Quality of the activated sludge in the lab scale SBR's

The average concentrations of the analysed sludge parameters for the lab scale SBR's are presented in Table 8.8. The quality of the activated sludge was analysed on fewer occasions compared to the parameters in the influent and effluent for the lab scale SBR's. Therefore, it was determined to calculate one average concentration for each parameter and not divide the results according to the various exchange ratios.

*Table 8.8 The average concentration of SV, MLSS, MLVSS, SVI and F/M ratio for the lab scale SBR's during the experimental trial.*

Parameter	Unit	REACTOR 1	REACTOR 2
MLSS	mg/L	3,147	3,235
MLVSS	mg/L	3,083	2,915
SVI	mL/g	107	71
F/M ratio	kg COD/kg MLSS·day	0.27	0.26

The average concentrations of Mixed Liquor Suspended Solids *MLSS* in Reactor 1 and Reactor 2 were within the interval of two to five g/L, which is required for the biological processes to work properly. The average *MLVSS* in the two reactors exceeded 2,000 mg/L, which has been presented to be a safe design parameter for achieving nitrification. In Table 8.8, it can be seen that the average Sludge Volume Index *SVI* for the two reactors were within the preferable interval of 70 –150 mg/L. The average Food to Microorganism *F/M* ratio was similar for the two reactors. The characteristics of the activated sludge are further discussed in the following paragraphs.

Figure 8.23 presents the *MLSS* concentrations in the lab scale SBR's during the experimental trial. A technical error concerning the aeration equipment caused the drop in *MLSS* on Day 50. Measured values from that day were therefore not included when the average sludge parameters in Table 8.8 were calculated.

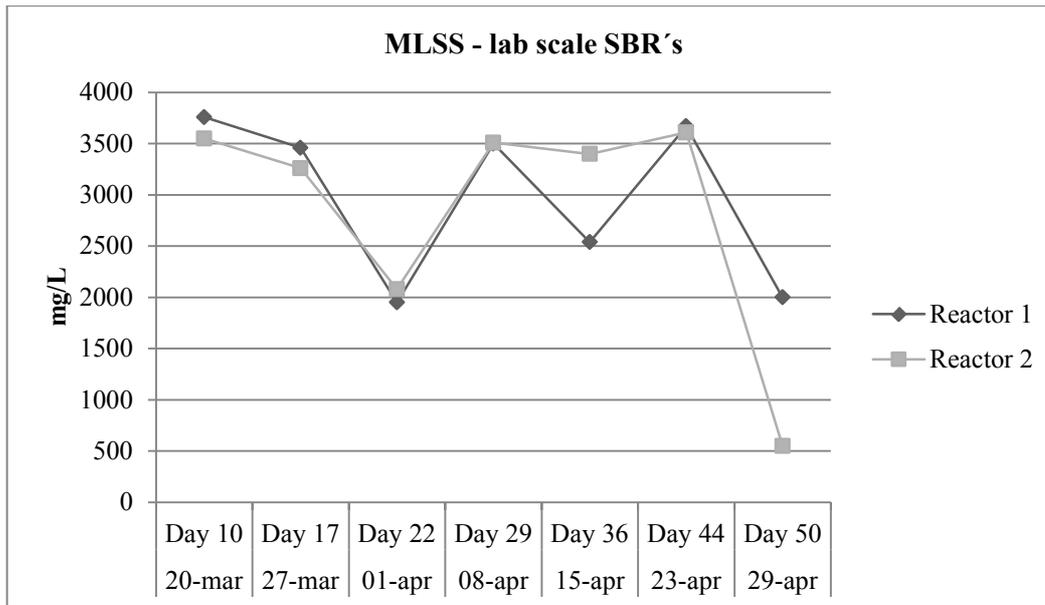


Figure 8.23 MLSS in Reactor 1 and Reactor 2 during the experimental trial

The MLSS remained rather similar in the lab scale SBR's during the experimental trial, except from on Day 36 when a decreased concentration in Reactor 2 was noted. No correlation with any other analysed parameter was found that could have explained the drop. On Day 22, it was noted that the amount of sludge in the reactors had decreased and that the MLSS in both lab scale SBR's dropped significantly, see Figure 8.23. During that day the pH and temperature in the feed tanks decreased, see Section 8.2.1 and Section 8.2.3, which could have contributed to the decrease of MLSS. Despite some decreased values, the MLSS remained within the ideal interval during the experimental trial.

A large part of the MLSS consists of MLVSS which can be said to be equal to the amount of microbial suspensions in the activated sludge. In Figure 8.24, the amount of average fixed solids and organic solids - hence the microbial suspension - are illustrated for the lab scale SBR's.

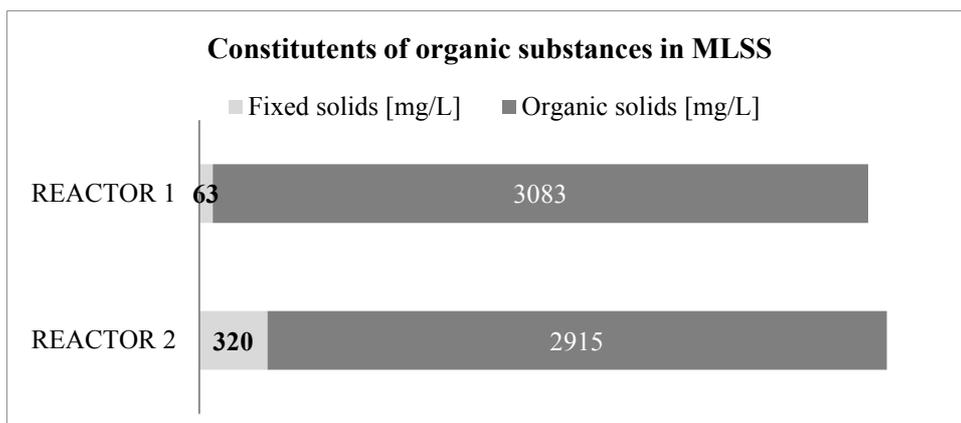


Figure 8.24 Illustration of how large amount of the average MLSS that constitutes of organic substances (MLVSS).

It can be seen that the MLVSS represents 98 % and 89 % of the total MLSS in Reactor 1 and Reactor 2 respectively. According to Andersson & Säterskog (2015), a

part of the fixed solids in the leachates from Brudaremossen - for example iron - was removed in the chemical treatment step with sedimentation through lamellae. This explains why the MLVSS forms such a large part of the MLSS in the reactors. However, it was expected to find a higher amount of fixed solids in Reactor 1 since it receives more Fe.

The SVI in the lab scale SBR's throughout the experimental trial is illustrated in Figure 8.25. The lower and upper recommended SVI concentrations of 70 mg/L and 150 mg/L respectively are marked in the figure with dashed lines. Concerning Reactor 1, it can be seen that the SVI was above the upper SVI limit on Day 22, which indicated that the sludge settled too slow in the reactor. The insufficient sludge characteristics could be since the denitrification process in Reactor 1 did not work in the beginning of the experimental trial, see Figure 8.1, Section 8.1.1, where it can be seen that the concentration of  $\text{NO}_3\text{-N}$  peaked on Day 22. The decreased MLSS on Day 22 correlates with the increased SVI for both reactors on that particular day. When observing the SVI curve for Reactor 1 in Figure 8.25, it appears that the SVI stabilized close to a value of 70 mg/L at the latter part of the experimental trial, during which the denitrification process improved gradually.

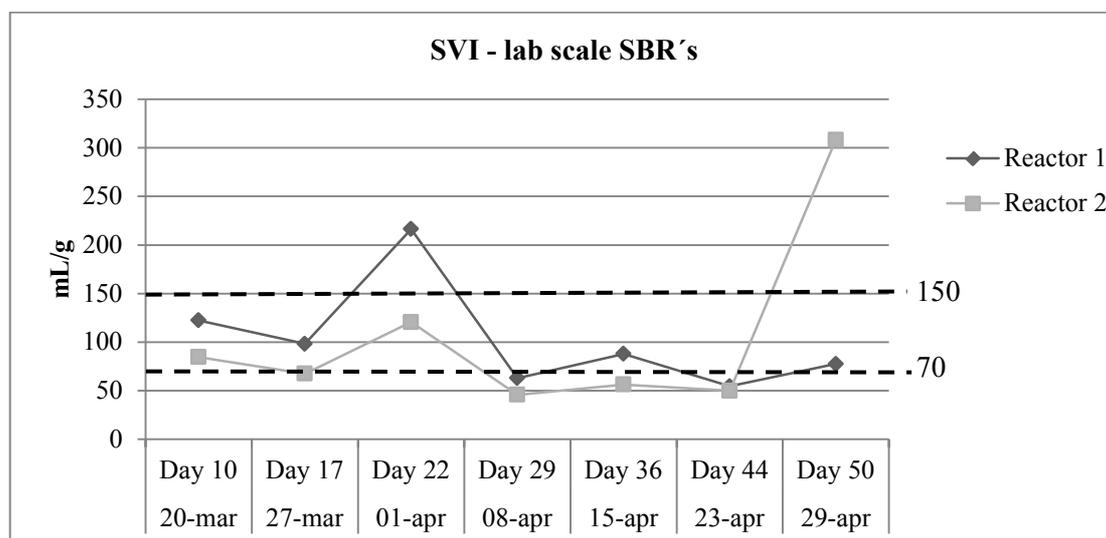


Figure 8.25 The SVI in Reactor 1 and Reactor 2 during the experimental trial

In Figure 8.25, it can be seen that some SVI values for Reactor 2 were measured to be lower than the recommended 70 mg/L, which indicated that the sludge in the reactor settled too rapidly. A low SVI could increase the turbidity of the treated effluent, but since the turbidity was not measured in the reactors during the experimental trial this was not confirmed. Nevertheless, the low SVI did not influence the lab scale SBR's treatment capacity.

As presented in Table 8.8, the average F/M ratio in Reactor 1 and Reactor 2 were 0.27 and 0.26 respectively. The ratios for both reactors were similar to ratios applied in earlier studies - see Chapter 3, Section 3.4.1 - and were within preferable limits. However, a lower ratio would have improved the stability of the treatment capacity. To achieve a more valid result concerning the quality of the activated sludge in the lab scale SBR's, more frequent sampling and analysis would have been required. Also, it should be taken into consideration that the F/M ratios were calculated based on the

COD, which was calculated according to theory that might not be applicable for the leachates.

The SRT applied in the experimental trial was 32 days, which was higher than the SRT's applied in the studied cases, see Table 5.2 in Chapter 5, Section 5.1. Since the results showed that the biological processes improved after the period equal to one SRT, it might have been preferable to use a shorter STR due to the limited timeframe of the project.

## 8.5 Result & Discussion of Full scale SBR design

The results for the lab scale SBR's were used to design a full scale SBR treatment for a permanent facility at Brudaremossen. Table 8.9 below illustrates the recommended design parameters for a full scale SBR treatment of the leachates at Brudaremossen. The result from the experimental trial with the lab scale SBR's indicated no inhibition of the processes due to toxic compounds in the leachates. It was therefore decided that the full scale SBR would be able to receive water directly after the lamellae sedimentation, similar to Reactor 1 during the experimental trial.

*Table 8.9 Design parameters for a full scale SBR treatment of the leachates at Brudaremossen*

Design parameter	Unit	Value
Number of reactors	Number	2
Fill time	h	4.00
Total aeration time	h	4.00*
React oxic time	h	2.00
React anoxic time	h	1.25
Settle time	h	0.50
Draw time	h	0.25
Total cycle time	h	8.00
Total SRT	d	32.00
Reactor volume	m <sup>3</sup>	457
Reactor depth	m	6.00
Reactor area	m <sup>2</sup>	76.20
Fill volume / cycle	m <sup>3</sup>	320
Exchange ratio	-	0.70
Draw depth	m	4.20
MLSS	g/m <sup>3</sup>	3,670
MLVSS	g/m <sup>3</sup>	3,620
Sludge waste (mixed)	L/day/reactor	14,281
Sludge production	kg/day/reactor	52.40
Methanol dosage	L/min*during 4 minutes/cycle	0.56
Phosphorus dosage	L/min*during 4 hours/cycle	4.00
Average oxygen transfer rate	kg O <sub>2</sub> /h/reactor	0.48
Fill pumping rate	L/min	1,333
Draw pumping rate	L/min	21,333

\*2 h during fill, 2 h during react oxic

When deciding the length of the phases for the full scale SBR, the earlier evaluated nitrification and denitrification rates for Reactor 1 was used, see Section 8.1.1. With a

known incoming  $\text{NH}_4\text{-N}$  concentration of 55 mg/L together with the evaluated nitrification rate of 13.75 mg  $\text{NH}_4\text{-N/L}\cdot\text{h}$  and denitrification rate of 42.22 mg  $\text{NH}_4\text{-N/L}\cdot\text{h}$ , the required lengths of the oxic and anoxic phases in order to reduce all incoming nitrogen could be estimated. The calculations can be seen in equation (3) and equation (4) respectively in Appendix H.

The evaluations showed that the total aeration time for the full scale SBR ought to be four hours and that the anoxic phase have to be approximately one hour. In order to manage having four hours of total aeration time, it was decided that there would be aeration during two hours in the fill phase, resulting in a react anoxic phase time of two hours. However, it should be considered that the nitrification and denitrification rates were evaluated only once, during the cycle analysis on Day 44. In order to validate this result, further cycle analyses should be performed. Also, the nitrification rate in Reactor 1 could have been affected by the low DO concentrations during the oxic phase, which also should be considered.

It was decided that a full scale SBR technique would benefit by having two reactors so that different phases can be in progress at the same time. When one reactor is in the fill period, all other phases should be able to occur in the other reactor. Therefore, the fill phase of 0.5 hours assumed during the experimental trial was too short and a fill phase of four hours was decided for the full scale SBR instead, see equation (5) in Appendix H. This resulted in a recommended total cycle time of eight hours, see equation (6) in Appendix H. From the literature review it was found the eight hours was the most common cycle time, which align with the calculated cycle time.

The analysed exchange ratio of 0.50 for the lab scale SBR's proved to be a valid exchange ratio, resulting in an efficient nitrogen removal. However, it was discovered that a higher loading would provide with at higher removal efficiency. Therefore, it was decided that a higher exchange ratio than 0.50 was preferable for the full scale SBR. Equation (7) to equation (11) in Appendix H illustrates that an exchange ratio of 0.70 is possible.

The overall Hydraulic Retention Time *HRT* for the full scale SBR was calculated to 23.6 hours, see equation (12) in Appendix H. The SRT for the full scale SBR was assumed to 32 days, as supposed for the lab scale SBR's. This resulted in a required sludge waste of 14.28 litres per day for each reactor, see equation (20) in Appendix H.

The  $\text{O}_2$  rate is decided from the requirement two to 3 mg  $\text{O}_2\text{/L}$ , where the upper boundary of 3 mg  $\text{O}_2\text{/L}$  is used to calculate the average oxygen transfer rate. It should be noted that the average oxygen transfer rate is optimised by multiplying it with a safety factor of 2.0 to make sure that there is enough oxygen transfer during the beginning of the cycle and to handle peak loads.

## 9 Conclusion

The lab scale Sequencing Batch Reactors *SBR*'s were evaluated to achieve nitrification and denitrification at the end of the study. The average removal of ammonium nitrogen  $NH_4-N$  reached 96 % for Reactor 1 and 94 % for Reactor 2 during the entire experimental trial. Concerning total nitrogen  $N_{tot}$ , the removal for the lab scale *SBR*'s were 51 % in Reactor 1 and 76 % in Reactor 2. The nitrification rates in Reactor 1 and Reactor 2 were assessed during the cycle analysis to 14 and 16 mg  $NH_4-N/L \cdot h$  respectively, whilst the denitrification rates were 42 and 39 mg nitrate nitrogen  $NO_3-N/L \cdot h$ . Generally, the outgoing concentration of  $NH_4-N$  was lower than the threshold value of 3 mg/L for discharge to the recipients evaluated by Ramböll regarding both reactors. The Moving Bed Biofilm Reactor *MBBR* had an average  $NH_4-N$  and  $N_{tot}$  removal rate of 99.8 % and 97 % respectively.

When comparing the average effluent concentrations from the lab scale *SBR*'s with the threshold values for potential discharge recipients, it could be concluded that the concentrations of  $NH_4-N$ , Biochemical Oxygen Demand  $BOD_7$  and Total Organic Carbon  $TOC$  were below the recommendations. The concentrations of  $N_{tot}$  from Reactor 1 were below the threshold value whereas Reactor 2 only achieved sufficient reduction of  $N_{tot}$  at the end of the experimental trial. This was due to that the denitrification process needed longer time to start functioning. The concentrations of total phosphorus  $P_{tot}$  for the reactors exceeded the threshold value significantly, which was due to overdosage of phosphoric acid.

The results from the experimental trial showed that the *SBR* technique is applicable to use for nitrogen removal in leachates from Brudaremossen. The design used for the lab scale *SBR*'s worked, however it could have been further optimised. Methanol and phosphorus dosages were added in too large quantities due to inaccurate assumptions of the incoming nitrogen concentrations. Also, the duration of the phases, exchange ratios and Sludge Retention Time *SRT*, could have been analysed and optimised. In order to improve the *SBR* performance, the aeration equipment should have been more robust to provide with a more even oxygen supply. Furthermore, the laboratory analyses applied in the study were sufficient for this thesis. However, to achieve more valid results it would have been preferable to delegate the analysing task to a commercial laboratory. This might have enabled more frequent sampling, such as further cycle analyses, which would have enabled continuous improvements of the lab scale *SBR*'s and provided more accurate results.

From the results for lab scale *SBR*'s, a full scale *SBR* was designed. It was decided that two reactors would provide with a more reliable treatment and the total cycle time was set to eight hours. The duration of the phases was adapted so that one reactor would be filled when all other phases occurred in the other reactor. The results showed no indication that toxic compounds in the leachate inhibited the biological treatment. Therefore, it was recommended that a full scale *SBR* could be installed directly after a chemical precipitation and lamellae sedimentation step in a permanent treatment facility at Brudaremossen.

Concerning the operational parameters, the pH varied within the recommended limits for achieving nitrification and denitrification. The addition of Dissolved Oxygen *DO* deteriorated throughout the experimental trial, but the biological processes still seemed to work. However, it is believed that higher removal rates could have been achieved if the *DO* concentrations would have been higher. The lab scale *SBR*'s was

operated at outdoor temperatures, which were considerably lower than the temperatures advised in the literature study. For a full scale SBR, it is recommended to operate at room temperature since this would result in a more efficient treatment. During operation, it is important to continuously regulate the chemical dosages and nitrogen loading according to incoming concentrations of nitrogen. Finally, for future similar trials, it is recommended to use an automatic measuring device for pH, temperature, flows and DO. This is believed to enable a more efficient operation with improved opportunities to regulate the technique if technical errors occur.

The MBBR was assessed to achieve more even removal rates compared to the lab scale SBR's, which was evaluated to be due to that it had 54 days of acclimatisation before sampling begun, whereas the SBR's had seven days. Also, the SBR technique was operated manually whilst the MBBR technique was run automatically with available operational personnel. Merely based on the results, the MBBR was considered most suitable for Brudaremsøen. However, an accurate comparison would require a longer trial period to optimise the SBR and more similar operational conditions. It is therefore recommended that the lab scale SBR's are operated longer at Brudaremsøen and continuously optimised or that the SBR technique is performed on a larger scale similar to the MBBR in the pilot plant.

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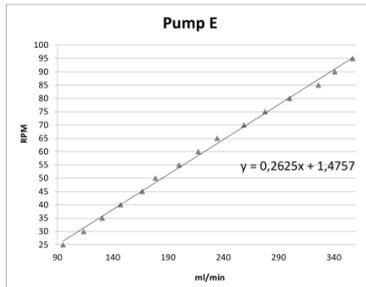
# 11 Appendices

Appendix A. RPM calculations	110
Appendix B. Average concentrations at exchange ratio 0.17 and 0.33	113
Appendix C. Measured parameters for the lab scale SBR's	115
Appendix D. Loading of NH <sub>4</sub> -N into Brudaremossen pilot plant 2015	122
Appendix E. Diagrams for removal of organic compounds and metals	123
Appendix F. Dissolved Oxygen & pH for the cycle analysis on Day 13	125
Appendix G. Nitrification & denitrification rates in Reactor 2 on Day 44	126
Appendix H. Calculations for a full scale SBR design	127

# Appendix A. RPM calculations

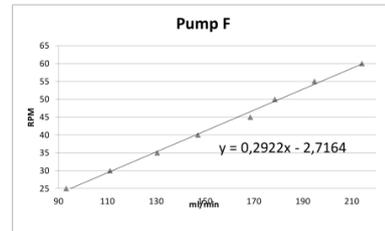
## Exchange ratio 0.17

Pump E	RPM	s	ml	ml/s	ml/min
	25	158	250	1,582	95
	30	132		1,894	114
	35	115		2,174	130
	40	102		2,451	147
	45	90		2,778	167
	50	84		2,976	179
	55	75		3,333	200
	60	69		3,623	217
	65	64		3,906	234
	70	58		4,31	259
	75	54		4,63	278
	80	50		5	300
	85	46		5,435	326
	90	44		5,682	341
	95	42		5,952	357



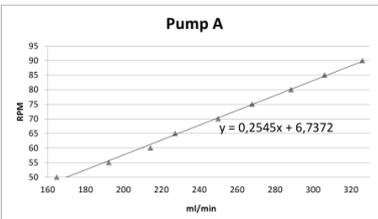
PUMP E	
Wanted flow	44,44444444 ml/min
k=	0,2625
m=	1,4757
Wanted RPM	13,1

Pump F	RPM	s	ml	ml/s	ml/min
	25	161		1,553	93
	30	135		1,852	111
	35	115		2,174	130
	40	102		2,451	147
	45	89		2,809	169
	50	84		2,976	179
	55	77		3,247	195
	60	70		3,571	214



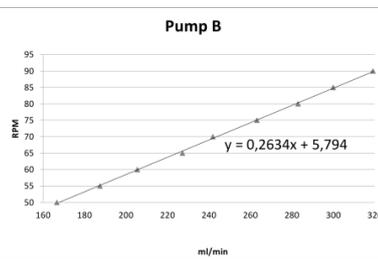
PUMP F	
Wanted flow	44,44444444 ml/min
k=	0,2922
m=	-2,7164
Wanted RPM	10,3

Pump A	RPM	s	ml	ml/s	ml/min
	50	91		2,747	165
	55	78		3,205	192
	60	70		3,571	214
	65	66		3,788	227
	70	60		4,167	250
	75	56		4,464	268
	80	52		4,808	288
	85	49		5,102	306
	90	46		5,435	326



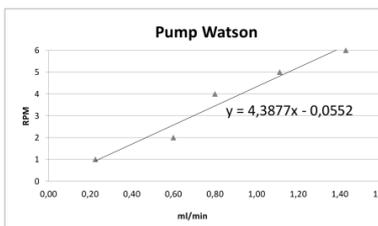
PUMP A	
Wanted flow	88,88888889 ml/min
k=	0,2545
m=	6,7372
Wanted RPM	29,4

Pump B	RPM	s	ml	ml/s	ml/min
	50	90		2,778	167
	55	80		3,125	188
	60	73		3,425	205
	65	66		3,788	227
	70	62		4,032	242
	75	57		4,386	263
	80	53		4,717	283
	85	50		5	300
	90	47		5,319	319



PUMP B	
Wanted flow	88,88888889 ml/min
k=	0,2634
m=	5,794
Wanted RPM	29,2

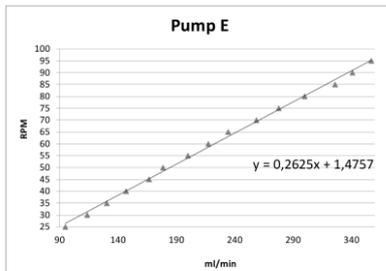
Pump Watson	RPM	s	ml	ml/s	ml/min
	1	1330	5	0,004	0,23
	2	200	2	0,01	0,60
	4	150	2	0,013	0,80
	5	108	2	0,019	1,11
	6	84	2	0,024	1,43



Calculation according to incoming nitrogen			
<b>PUMP Watson - Methanol diluted 3 times</b>		<b>PUMP Watson - Methanol without dilution</b>	
Wanted flow	0,842815002 ml M/min	Wanted flo	0,280938334 ml M/min
k=	4,3877	k=	4,3877
m=	-0,0552	m=	-0,0552
Wanted RPM	3,6	Wanted RP	1,2
<b>PUMP Watson - Phosphorus diluted 1000 times</b>		<b>PUMP Watson - Phosphorus diluted 100 times</b>	
Wanted flow	2,422040569 ml M/min	Wanted flo	0,242204057 ml M/min
k=	4,3877	k=	4,3877
m=	-0,0552	m=	-0,0552
Wanted RPM	10,6	Wanted RP	1,0

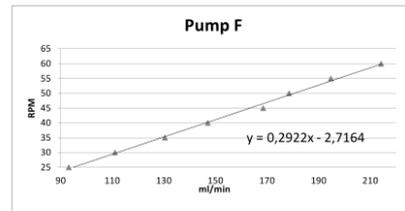
# Exchange ratio 0.33

Pump E	RPM	s	ml	ml/s	ml/min
	25	158	250	1,5823	95
	30	132		1,8939	114
	35	115		2,1739	130
	40	102		2,451	147
	45	90		2,7778	167
	50	84		2,9762	179
	55	75		3,3333	200
	60	69		3,6232	217
	65	64		3,9063	234
	70	58		4,3103	259
	75	54		4,6296	278
	80	50		5	300
	85	46		5,4348	326
	90	44		5,6818	341
	95	42		5,9524	357



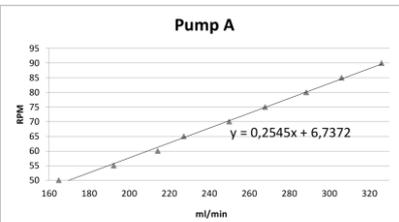
PUMP E	
Wanted flow	88,8888889 ml/min
k=	0,2625
m=	1,4757
Wanted RPM	24,8

Pump F	RPM	s	ml	ml/s	ml/min
	25	161		1,5528	93
	30	135		1,8519	111
	35	115		2,1739	130
	40	102		2,451	147
	45	89		2,809	169
	50	84		2,9762	179
	55	77		3,2468	195
	60	70		3,5714	214



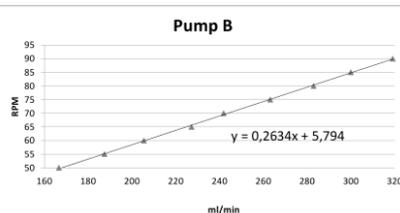
PUMP F	
Wanted flow	88,8888889 ml/min
k=	0,2922
m=	-2,7164
Wanted RPM	23,3

Pump A	RPM	s	ml	ml/s	ml/min
	50	91		2,7473	165
	55	78		3,2051	192
	60	70		3,5714	214
	65	66		3,7879	227
	70	60		4,1667	250
	75	56		4,4643	268
	80	52		4,8077	288
	85	49		5,102	306
	90	46		5,4348	326



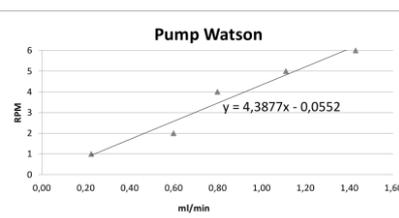
PUMP A	
Wanted flow	177,7777778 ml/min
k=	0,2545
m=	6,7372
Wanted RPM	52,0

Pump B	RPM	s	ml	ml/s	ml/min
	50	90		2,7778	167
	55	80		3,125	188
	60	73		3,4247	205
	65	66		3,7879	227
	70	62		4,0323	242
	75	57		4,386	263
	80	53		4,717	283
	85	50		5	300
	90	47		5,3191	319



PUMP B	
Wanted flow	177,7777778 ml/min
k=	0,2634
m=	5,794
Wanted RPM	52,6

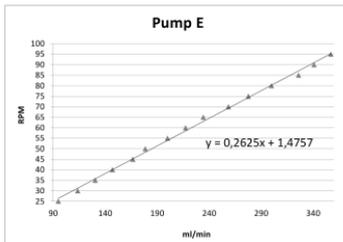
Pump Watsc	RPM	s	ml	ml/s	ml/min
	1	1330	5	0,0038	0,23
	2	200	2	0,01	0,60
	4	150	2	0,0133	0,80
	5	108	2	0,0185	1,11
	6	84	2	0,0238	1,43



Calculation according to incoming nitrogen			
<b>PUMP Watson - Methanol diluted 3 times</b>		<b>PUMP Watson - Methanol without dilution</b>	
Wanted flow	1,685630004 ml M/min	Wanted flow	0,561876668 ml M/min
k=	4,3877	k=	4,3877
m=	-0,0552	m=	-0,0552
Wanted RPM	7,3	Wanted RPM	2,4
<b>PUMP Watson - Phosphorus diluted 1000</b>		<b>PUMP Watson - Phosphorus diluted 100 times</b>	
Wanted flow	4,844081138 ml M/min	Wanted flow	0,484408114 ml M/min
k=	4,3877	k=	4,3877
m=	-0,0552	m=	-0,0552
Wanted RPM	21,2	Wanted RPM	2,1

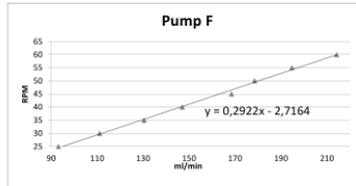
# Exchange ratio 0.50

Pump E	RPM	s	ml	ml/s	ml/min
	25	158	250	1,5823	95
	30	132		1,8939	114
	35	115		2,1739	130
	40	102		2,451	147
	45	90		2,7778	167
	50	84		2,9762	179
	55	75		3,3333	200
	60	69		3,6232	217
	65	64		3,9063	234
	70	58		4,3103	259
	75	54		4,6296	278
	80	50		5	300
	85	46		5,4348	326
	90	44		5,6818	341
	95	42		5,9524	357



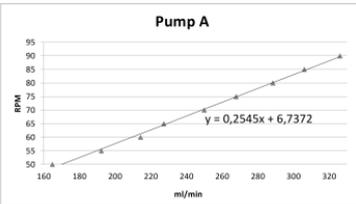
PUMP E	
Wanted flow	133,33 ml/min
k=	0,2625
m=	1,4757
Wanted RPM	36,2

Pump F	RPM	s	ml	ml/s	ml/min
	25	161		1,5528	93
	30	135		1,8519	111
	35	115		2,1739	130
	40	102		2,451	147
	45	89		2,809	169
	50	84		2,9762	179
	55	77		3,2468	195
	60	70		3,5714	214



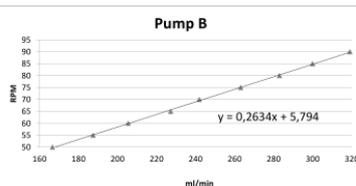
PUMP F	
Wanted flow	133,33 ml/min
k=	0,2922
m=	-2,7164
Wanted RPM	36,2

Pump A	RPM	s	ml	ml/s	ml/min
	50	91		2,7473	165
	55	78		3,2051	192
	60	70		3,5714	214
	65	66		3,7879	227
	70	60		4,1667	250
	75	56		4,4643	268
	80	52		4,8077	288
	85	49		5,102	306
	90	46		5,4348	326



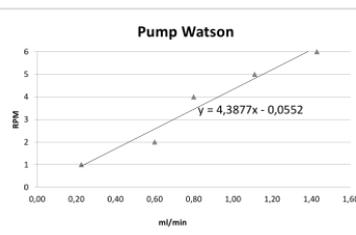
PUMP A	
Wanted flow	266,67 ml/min
k=	0,2545
m=	6,7372
Wanted RPM	74,6

Pump B	RPM	s	ml	ml/s	ml/min
	50	90		2,7778	167
	55	80		3,125	188
	60	73		3,4247	205
	65	66		3,7879	227
	70	62		4,0323	242
	75	57		4,386	263
	80	53		4,717	283
	85	50		5	300
	90	47		5,3191	319



PUMP B	
Wanted flow	266,67 ml/min
k=	0,2634
m=	5,794
Wanted RPM	76,0

Pump Watson	RPM	s	ml	ml/s	ml/min
	1	1330	5	0,0038	0,23
	2	200	2	0,01	0,60
	4	150	2	0,0133	0,80
	5	108	2	0,0185	1,11
	6	84	2	0,0238	1,43



Calculation according to incoming nitrogen			
<b>PUMP Watson - Methanol diluted 3 times</b>	<b>PUMP Watson - Methanol without dilution</b>		
Wanted flow	2,5284 ml M/min	Wanted flow	0,842815002 ml M/min
k=	4,3877	k=	4,3877
m=	-0,0552	m=	-0,0552
Wanted RPM	11,0	Wanted RPM	3,6
<b>PUMP Watson - Phosphorus diluted 1000 times</b>	<b>PUMP Watson - Phosphorus diluted 100 times</b>		
Wanted flow	7,2661 ml M/min	Wanted flow	0,726612171 ml M/min
k=	4,3877	k=	4,3877
m=	-0,0552	m=	-0,0552
Wanted RPM	31,8	Wanted RPM	3,1

## Appendix B. Average concentrations at exchange ratio 0.17 and 0.33

Table 1 Average concentrations in the influent and effluent for the lab scale SBR's, with an exchange ratio of 0.17, and of the MBBR process in the Brudaremassen pilot plant 2015.

Exchange ratio 0.17	Unit	Lab scale SBR's - REACTOR 1		Lab scale SBR's - REACTOR 2	
		Influent	Effluent	Influent	Effluent
Parameter	Unit	Influent	Effluent	Influent	Effluent
pH	-	7.38	8.05	7.49	8.14
Temperature	°C	8.95	10.53	9.12	10.55
NH <sub>4</sub> -N	mg/L	56.19	1.59	45.69	2.78
NO <sub>2</sub> -N	mg/L	2.51	1.54	1.17	1.85
NO <sub>3</sub> -N	mg/L	2.25	30.37	1.75	6.05
N <sub>tot</sub>	mg/L	58.13	32.64	48.16	8.96
P <sub>tot</sub>	mg/L	0.03	4.35	0.33	10.31
BOD <sub>7</sub>	mg/L	3.13	0.31	0.53	-0.08
COD	mg/L	100.45	71.67	81.42	52.01
TOC	mg/L	17.08	21.54	10.74	14.97
DOC	mg/L	15.83	22.17	9.89	14.41
Fe	mg/L	1.80	0.26	0.12	0.21
Fe filtered	mg/L	0.06	0.04	0.05	0.03
Mn	mg/L	0.52	0.45	0.38	0.35
Al	mg/L	0.28	0.24	0.04	0.25
Ti	mg/L	0.37	0.32	0.32	0.33
Pb	µg/L	1.22	1.40	1.18	1.51
Cr	µg/L	1.45	0.77	0.42	0.28
Co*	µg/L	3.09	3.37	2.13	2.29
Ni	µg/L	5.18	5.19	4.14	4.21
Li	µg/L	0.02	0.02	0.02	0.02
Cu*	µg/L	11.23	18.50	9.40	15.00
Zn	µg/L	15.24	17.79	5.58	16.44
Ag*	µg/L	<0.10	<0.10	<0.10	<0.10
Cd	µg/L	1.05	1.18	1.06	1.19
Hg	µg/L	0.02	0.02	0.02	0.02

\*Based on detected measured concentrations at the exchange ratio 0.17

Table 2 Average concentrations in the influent and effluent for the lab scale SBR's, with an exchange ratio of 0.33, and of the MBBR process in the Brudaremassen pilot plant 2015.

Exchange ratio 0.33	Unit	Lab scale SBR's - REACTOR 1		Lab scale SBR's - REACTOR 2	
		Influent	Effluent	Influent	Effluent
Parameter	Unit	Influent	Effluent	Influent	Effluent
pH	-	7.29	7.95	7.32	7.97
Temperature	°C	10.60	11.70	10.50	11.95
NH <sub>4</sub> -N	mg/L	43.78	0.67	53.16	1.15
NO <sub>2</sub> -N	mg/L	1.47	1.33	1.47	1.33
NO <sub>3</sub> -N	mg/L	2.29	31.08	3.91	10.69
N <sub>tot</sub>	mg/L	54.46	28.12	49.97	12.29
P <sub>tot</sub>	mg/L	0.34	8.16	0.11	11.01
BOD <sub>7</sub>	mg/L	1.55	0.17	1.37	0.26
COD	mg/L	82.47	93.99	61.63	84.95
TOC	mg/L	11.09	29.01	4.14	25.99
DOC	mg/L	13.07	27.38	10.49	28.04
Fe	mg/L	0.47	0.24	0.10	0.20
Fe filtered	mg/L	0.09	0.05	0.08	0.05
Mn	mg/L	0.47	0.45	0.47	0.45
Al	mg/L	0.39	0.11	0.06	0.25
Ti	mg/L	0.45	0.43	0.46	0.44
Pb	µg/L	2.55	2.26	2.63	2.40
Cr	µg/L	0.00	0.00	0.00	0.00
Co*	µg/L	3.09	3.37	2.13	2.29
Ni	µg/L	4.62	3.74	3.64	3.32
Li	µg/L	0.02	0.02	0.02	0.02
Cu*	µg/L	11.23	18.50	9.40	15.00
Zn	µg/L	177.43	10.55	11.84	17.66
Ag*	µg/L	<0.10	<0.10	<0.10	<0.10
Cd	µg/L	1.78	1.77	1.81	1.75
Hg	µg/L	0.02	0.02	0.02	0.04

\*Based on detected measured concentrations at the exchange ratio 0.17

# Appendix C. Measured parameters for the lab scale SBR's

## Reactor 1

REACTOR 1		= < DETECTION LIMIT		*Ntot & NH4-N taken from Ramböll results 26/3									
		2015-03-18		2015-03-20		2015-03-25		2015-03-27		2015-04-01			
		Day 8		Day 10		Day 15		Day 17		Day 22			
Parameter	Unit	Before*	After	Before	After	Before	After	Before	After	Before	After		
pH	-	7,73	7,96	7,39	7,96	7,39	8,06	7,41	8,14	7,07	8,11		
Cond	µS/cm	n.a.	1530	2040	1646	2120	1681	2110	1705	1522	1736		
Temp	°C	8,7	13	10	11,3	9,3	10,3	8,3	8,8	7,8	8,8		
BOD7	mg O2/l	7	2,36	2,48	0,4	n.a.	0,09	2,31	-0,35	3,14	-0,31		
Alk	meq/l	18,33	9,28	15,76	10,16	18,33	9,44	16,32	9,28	11,12	9,28		
Ptot	mg/l	0,023	n.a.	n.a.	n.a.	0,022	4,05	0,03	4,58	0,05	4,87		
Ptot-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.		
NH4-N	mg/l	63,0	1,5	51,8	2,3	63,0	1,1	63,2	1,5	44,3	1,5		
Removal NH4-N	%	97,58		95,51		98,31		97,56		96,53			
NO3-N	mg/l	0,1	21,6	4,9	25,9	n.a.	26,0	1,5	38,0	1,5	43,6		
NO2-N	mg/l	0,0	1,5	7,9	1,5	n.a.	1,5	1,5	1,5	1,5	1,5		
Total nit fractions	mg/L	63,1	24,6	64,7	29,8	#VALUE!	28,6	66,3	41,1	47,4	46,7		
TOC	mg/l	33	22,71	16,04	20,26	n.a.	20,63	13,42	22,58	12,69	22,34		
COD	mg/L	148,2	75,2	97,3	67,8	#VALUE!	68,9	89,5	74,8	87,3	74,0		
DOC	mg/l	33	24,27	n.a.	23	n.a.	17,23	12,52	23,23	9,355	23,46		
TC	mg/l	n.a.	124,30	196,2	140,1	n.a.	128	211,7	137,1	150,6	127,9		
IC	mg/l	n.a.	101,60	180,2	168,5	n.a.	107,3	198,3	114,5	137,9	105,6		
Ntot	mg/l	68,0	23,8	57,5	30,3	74,0	31,0	62,4	37,2	39,1	43,1		
Removal Ntot	%	65,04		47,26		58,07		40,48		-10,18			
Ntot-tot nit frac	mg/l	4,94	-0,87	-7,15	0,58	#VALUE!	2,39	-3,90	-3,90	-8,34	-3,66		
Li	mg/l	n.a.	0,05	0,04	0,04	n.a.	0,04	0,76	0,76	0,76	0,05		
Na	mg/l	n.a.	205,72	207,57	210,98	n.a.	199,78	199,43	243,30	162,71	222,14		
K	mg/l	n.a.	61,29	65,20	64,34	n.a.	61,80	56,17	66,40	42,28	65,29		
Mg	mg/l	n.a.	35,29	35,83	36,36	n.a.	33,95	33,15	39,03	25,40	37,16		
Mn	mg/l	n.a.	0,85	0,91	0,91	n.a.	0,85	6,04	6,04	0,18	0,42		
Ca	mg/l	n.a.	93,37	94,96	112,29	n.a.	110,17	48,80	85,84	79,18	133,54		
Cl	mg/l	n.a.	128,73	128,23	131,93	n.a.	124,21	137,70	158,72	108,18	151,59		
Br	mg/l	n.a.	19,25	26,28	19,25	n.a.	19,25	8,79	8,79	8,79	8,79		
PO <sub>4</sub> <sup>3-</sup>	mg/l	n.a.	29,15	10,45	10,45	n.a.	33,33	10,45	10,45	10,45	35,52		

REACTOR 1		= < DETECTION LIMIT											
		2015-03-18		2015-03-20		2015-03-25		2015-03-27		2015-04-01			
		Day 8		Day 10		Day 15		Day 17		Day 22			
Parameter	Unit	Before*	After	Before	After	Before	After	Before	After	Before	After		
Ni	µg/l	7,4	10,54	9,21	10,02	n.a.	2,27	5,36	1,81	2,08	3,69		
Ni-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.		
Co	µg/l	n.a.	3,38	3,09	3,36	n.a.	-1,24	-0,27	-1,51	-1,67	-0,88		
Co-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.		
Cu	µg/l	3,6	22,43	18,86	14,50	n.a.	-11,94	n.a.	-11,75	n.a.	-10,27		
Cu-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.		
Zn	µg/l	19	37,36	13,11	21,60	n.a.	11,19	19,75	10,07	11,73	11,32		
Zn-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.		
As	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	0,03	0,04	0,03	0,04	0,03		
As-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.		
Se	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	0,26	0,31	0,24	0,22	0,29		
Se-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.		
Sr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	0,27	0,37	0,23	0,30	0,30		
Sr-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.		
Ag	µg/l	<0,1	n.a.	n.a.	n.a.	n.a.	-2,38	-2,38	-2,38	-2,38	-2,38		
Ag-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.		
Cd	µg/l	0,02	0,07	0,10	0,14	n.a.	1,72	1,71	1,72	1,72	1,73		
Cd-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.		
Sn	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	5,71	4,13	5,96	1,87	6,41		
Sn-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.		
Sb	µg/l	n.a.	0,49	0,36	0,45	n.a.	5,20	5,25	5,17	5,51	5,20		
Sb-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.		
Ba	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	0,12	0,20	0,11	0,28	0,14		
Ba-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.		
Hg	µg/l	<0,005	n.a.	n.a.	n.a.	n.a.	0,02	0,02	0,02	0,02	0,02		
Hg-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.		
Tl	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	0,003	0,003	0,003	0,004	0,003		
Pt	µg/l	n.a.	0,0000	0,0000	0,0004	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.		
Pb	µg/l	0,2	0,50	0,57	0,73	n.a.	1,80	1,63	1,69	1,95	1,83		

REACTOR 1		= < DETECTION LIMIT									
		2015-03-18		2015-03-20		2015-03-25		2015-03-27		2015-04-01	
		Day 8		Day 10		Day 15		Day 17		Day 22	
Parameter	Unit	Before*	After	Before	After	Before	After	Before	After	Before	After
SO42-	mg/l	n.a.	36,24	35,29	35,93	n.a.	37,72	33,18	31,28	36,67	32,86
Li	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	0,02	0,03	0,02	0,02	0,02
Li-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Be	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	0,002	0,001	0,002	0,002	0,002
Be-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
B	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	0,41	0,54	0,35	0,35	0,48
B-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Na	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	58,84	59,41	55,26	57,36	57,00
Na-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Mg	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	25,75	35,25	22,55	25,65	30,26
Mg-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Al	mg/l	n.a.	0,66	0,30	0,32	n.a.	0,11	0,11	0,11	0,66	0,11
Al-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Si	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	4,57	7,22	4,21	8,47	6,13
Si-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
K	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	51,45	65,48	44,39	44,70	57,48
K-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Ca	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	16,11	22,02	15,50	18,45	18,80
Ca-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Ti	mg/l	n.a.	0,27	0,26	0,26	n.a.	0,35	0,49	0,30	0,41	0,41
Ti-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
V	mg/l	n.a.	0,00	0,00	0,00	n.a.	0,00	0,00	0,00	0,00	0,00
V-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Cr	µg/l	3,8	4,12	4,21	3,99	n.a.	-0,91	0,37	-1,06	-0,26	-0,54
Cr-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Mn	mg/l	n.a.	0,74	0,85	0,82	n.a.	0,30	0,48	0,29	0,39	0,34
Mn-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Fe	mg/l	1	0,53	0,53	0,33	n.a.	0,16	0,25	0,17	7,06	0,15
Fe-filtr	mg/l	0,03	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.

REACTOR 1		= < DETECTION LIMIT										*Aeration fault	
		2015-04-08		2015-04-10		2015-04-15		2015-04-17		2015-04-23		2015-04-29	
		Day 29		Day 31		Day 36		Day 38		Day 44		Day 50	
Parameter	Unit	Before	After	Before	After	Before	After	Before	After	Before	After	Before	After
pH	-	7,3	8,05	7,24	7,91	7,33	7,98	7,33	7,98	7,27	8,08	7,34	7,57
Cond	µS/cm	1702	1406	1919	1557	1961	1537	1883	1533	1753	1361	2040	1985
Temp	°C	9,6	11	11,2	12,2	10	11,2	9,8	10,3	12,7	14,7	12,7	13
BOD7	mg O2/l	0,74	-0,31	1,88	0,29	1,22	0,04	0,08	0,08	0,73	0,09	1,39	0,06
Alk	meq/l	12,56	7,52	15,2	8,16	16	9,52	14,88	10,56	14,56	10,56	16,56	16,56
Ptot	mg/l	0,02	3,91	0,65	6,15	0,03	10,16	0,03	14,16	0,03	10,92	0,06	13,23
Ptot-filtr	mg/l	0,02	3,78	0,03	6,62	0,02	9,69	0,07	14,09	0,02	11,15	0,04	13,00
NH4-N	mg/l	51,7	1,5	41,1	0,2	46,4	1,1	32,4	5,9	21,8	1,1	59,8	40,5
Removal NH4-N	%	97,02		99,48		97,59		81,82		94,86		32,27	
NO3-N	mg/l	3,2	27,1	3,2	32,8	1,4	29,4	2,4	9,3	10,3	2,1	7,2	1,4
NO2-N	mg/l	1,5	1,5	1,5	1,5	1,4	1,1	1,4	1,1	1,1	1,1	1,4	1,4
Total nit frac	mg/L	56,5	30,2	45,8	34,5	49,2	31,6	36,3	16,3	33,3	4,3	68,4	43,3
TOC	mg/l	10,26	20,74	14,18		8,001	29,01	13,96	37,52	13,59	30,31	13,1	68,31
COD	mg/L	80,0	69,3	91,7		73,2	94,0	91,1	119,4	90,0	97,9	88,5	211,5
DOC	mg/l	8,426	21,8	16,32	34,67	9,827	20,08	12,5	32,42	11,51	35,13	16,85	55,16
TC	mg/l	171,3	117,2	197,2	n.a.	215,4	142,6	174,5	150,5	184,9	127,5	197,1	254,2
IC	mg/l	161,1	96,44	183	n.a.	207,4	113,6	160,5	113	171,3	97,16	184	185,9
Ntot	mg/l	47,8	30,5	56,0	n.a.	52,9	28,1	54,6	21,4	50,7	1,3	61,1	55,6
Removal Ntot	%	36,12		n.a.		46,86		60,70		97,52		9,11	
Ntot-tot nit frac	mg/l	-8,69	0,30	10,17	#VALUE!	3,67	-3,50	18,29	5,15	17,45	-3,08	-7,30	12,24
Li	mg/l	0,76	0,76	0,76	0,76	0,69	0,56	0,69	0,56	0,56	0,56	0,69	0,69
Na	mg/l	174,57	169,94	136,21	205,72	138,08	185,00	103,71	112,13	67,92	170,09	203,78	141,87
K	mg/l	42,28	46,75	38,54	59,74	41,25	52,17	32,84	31,87	24,87	47,92	57,79	43,95
Mg	mg/l	28,34	27,27	24,06	33,42	21,63	28,97	15,07	17,50	10,31	25,08	30,62	21,87
Mn	mg/l	6,04	0,18	6,04	0,36	5,49	4,40	5,49	4,40	4,40	4,40	5,49	5,49
Ca	mg/l	55,06	97,69	50,13	104,22	34,31	68,00	24,29	41,84	35,53	49,15	30,82	66,93
Cl	mg/l	113,52	114,03	88,57	132,48	101,96	134,78	72,36	80,58	48,19	120,68	133,87	96,54
Br	mg/l	8,79	8,79	8,79	8,79	7,99	11,38	7,99	6,39	6,39	6,39	7,99	7,99
PO4 <sup>3-</sup>	mg/l	10,45	10,45	14,05	145,07	9,50	20,51	9,50	21,27	7,60	7,60	9,50	9,50

REACTOR 1		= < DETECTION LIMIT												*Aeration fault	
		2015-04-08		2015-04-10		2015-04-15		2015-04-17		2015-04-23		2015-04-29			
		Day 29		Day 31		Day 36		Day 38		Day 44		Day 50			
Parameter	Unit	Before	After	Before	After	Before	After	Before	After	Before	After	Before	After	Before	After
SO42-	mg/l	34,98	37,62	29,38	31,38	30,26	29,66	32,18	22,98	23,21	32,35	30,45	31,41		
Li	mg/l	0,02	0,02	0,03	0,02	0,02	0,02	0,02	0,02	0,02	0,02	0,03	0,02		
Li-filtr	mg/l	0,02	0,02	0,03	0,02	0,02	0,02	0,02	0,02	0,02	0,02	0,02	0,02		
Be	µg/l	0,001	0,002	0,008	0,002	0,001	0,001	0,001	0,002	0,001	0,002	0,001	0,002		
Be-filtr	µg/l	0,001	0,001	0,002	0,001	0,001	0,001	0,001	0,001	0,001	0,001	0,001	0,001		
B	mg/l	0,33	0,34	0,50	0,45	0,46	0,50	0,47	0,45	0,37	0,41	0,49	0,48		
B-filtr	mg/l	0,43	0,37	0,50	0,49	0,48	0,47	0,47	0,46	0,40	0,41	0,48	0,49		
Na	mg/l	55,06	55,50	57,78	57,08	57,10	56,98	55,78	53,94	54,72	55,12	56,22	56,23		
Na-filtr	mg/l	56,10	55,31	58,74	58,08	56,55	55,73	56,52	55,05	54,70	55,62	56,90	56,02		
Mg	mg/l	21,38	22,34	33,13	29,08	30,79	32,20	30,64	29,49	24,67	27,01	31,77	30,45		
Mg-filtr	mg/l	29,75	24,14	32,96	32,46	32,74	31,11	30,88	29,80	27,37	27,36	31,22	30,80		
Al	mg/l	0,04	0,15	0,66	0,09	0,12	0,12	0,34	0,17	0,12	0,06	0,29	0,15		
Al-filtr	mg/l	0,04	0,03	0,06	0,03	0,06	0,03	0,07	0,02	0,06	0,03	0,07	0,02		
Si	mg/l	6,85	5,87	10,36	7,68	9,13	8,94	10,58	8,70	9,38	9,70	12,35	10,72		
Si-filtr	mg/l	9,68	7,15	9,99	9,40	10,20	9,56	10,67	10,07	10,38	8,86	12,59	12,01		
K	mg/l	40,96	41,86	60,24	53,31	56,42	58,24	57,15	55,21	45,80	49,61	60,03	57,66		
K-filtr	mg/l	52,70	45,24	60,01	59,91	59,43	57,20	57,65	55,23	49,83	50,92	59,64	58,42		
Ca	mg/l	15,95	15,64	21,16	20,06	21,04	27,74	21,39	21,17	17,47	19,81	21,99	21,53		
Ca-filtr	mg/l	21,14	17,59	21,73	21,47	22,36	22,73	21,45	20,73	19,77	19,91	20,44	21,05		
Ti	mg/l	0,32	0,32	0,48	0,42	0,43	0,45	0,42	0,41	0,36	0,40	0,43	0,42		
Ti-filtr	mg/l	0,45	0,35	0,46	0,46	0,45	0,44	0,44	0,40	0,40	0,39	0,43	0,42		
V	mg/l	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00		
V-filtr	mg/l	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00		
Cr	µg/l	-0,89	-0,99	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00		
Cr-filtr	µg/l	-0,19	-1,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00		
Mn	mg/l	0,34	0,24	0,47	0,37	0,46	0,54	0,44	0,43	0,37	0,37	0,44	0,44		
Mn-filtr	mg/l	0,44	0,25	0,48	0,38	0,48	0,43	0,45	0,41	0,42	0,37	0,42	0,42		
Fe	mg/l	0,16	0,21	0,66	0,24	0,28	0,24	0,82	0,37	0,35	0,15	0,35	0,55		
Fe-filtr	mg/l	0,08	0,04	0,10	0,05	0,08	0,05	0,09	0,05	0,08	0,05	0,08	0,12		

REACTOR 1		= < DETECTION LIMIT												*Aeration fault	
		2015-04-08		2015-04-10		2015-04-15		2015-04-17		2015-04-23		2015-04-29			
		Day 29		Day 31		Day 36		Day 38		Day 44		Day 50			
Parameter	Unit	Before	After	Before	After	Before	After	Before	After	Before	After	Before	After	Before	After
Ni	µg/l	1,86	2,80	5,15	3,52	4,10	3,97	3,82	3,47	2,50	3,14	4,21	4,50		
Ni-filtr	µg/l	3,90	2,93	5,34	4,23	4,89	4,28	4,11	3,68	3,24	3,23	19,63	4,59		
Co	µg/l	-1,33	-1,48	-0,64	-0,99	-0,66	-0,70	-0,67	-0,79	-1,19	-0,98	-0,51	-0,59		
Co-filtr	µg/l	-0,54	-1,39	-0,58	-0,70	-0,45	-0,71	-0,65	-0,74	-0,90	-0,97	-0,46	-0,48		
Cu	µg/l	n.a.	-8,64	40,33	-9,78	-10,34	-11,51	-13,72	-12,37	-14,37	-14,97	-14,01	-15,94		
Cu-filtr	µg/l	3,61	-12,69	-10,63	-12,37	-9,74	-13,33	-13,65	-15,11	-14,01	-15,93	-11,68	-17,16		
Zn	µg/l	12,62	15,21	342,45	11,13	12,41	9,97	9,39	10,44	8,82	5,80	9,52	8,69		
Zn-filtr	µg/l	16,14	7,17	13,09	47,93	12,87	5,93	9,21	5,44	9,56	3,98	25,09	12,16		
As	µg/l	0,03	0,03	0,04	0,04	0,03	0,05	0,04	0,04	0,03	0,03	0,05	0,04		
As-filtr	µg/l	0,03	0,03	0,04	0,04	0,04	0,04	0,04	0,04	0,04	0,03	0,04	0,04		
Se	µg/l	0,21	0,21	0,27	0,26	0,27	0,28	0,31	0,30	0,28	0,28	0,34	0,34		
Se-filtr	µg/l	0,25	0,24	0,29	0,28	0,31	0,30	0,31	0,30	0,28	0,30	0,33	0,33		
Sr	mg/l	0,23	0,23	0,33	0,29	0,31	0,32	0,31	0,29	0,25	0,27	0,30	0,28		
Sr-filtr	mg/l	0,31	0,25	0,33	0,32	0,32	0,30	0,30	0,28	0,27	0,27	0,29	0,29		
Ag	µg/l	-2,38	-2,38	-2,16	-2,25	-2,38	-2,38	-2,38	-2,38	-2,39	-2,39	-2,28	-2,38		
Ag-filtr	µg/l	-2,38	-2,38	-2,23	-2,37	-2,38	-2,38	-2,39	-2,38	-2,39	-2,39	-2,37	-2,34		
Cd	µg/l	1,72	1,73	1,84	1,81	1,72	1,72	1,71	1,74	1,72	1,83	1,75	1,72		
Cd-filtr	µg/l	1,73	1,71	1,80	1,81	1,72	1,71	1,71	1,71	1,73	1,83	1,75	1,72		
Sn	µg/l	3,68	5,73	3,08	5,50	4,14	7,83	2,79	3,97	2,70	6,52	1,95	2,81		
Sn-filtr	µg/l	4,72	6,50	2,89	5,61	4,14	6,29	2,64	3,82	3,00	6,55	1,76	2,81		
Ba	mg/l	0,17	0,15	0,22	0,17	0,22	0,22	0,23	0,18	0,20	0,16	0,22	0,19		
Ba-filtr	mg/l	0,21	0,15	0,22	0,17	0,22	0,18	0,22	0,16	0,21	0,16	0,21	0,17		
Hg	µg/l	0,02	0,02	0,02	0,02	0,02	0,02	0,01	0,01	0,01	0,01	0,01	0,01		
Hg-filtr	µg/l	0,02	0,02	0,02	0,02	0,02	0,02	0,01	0,01	0,01	0,01	0,01	0,01		
Tl	µg/l	0,003	0,004	0,002	0,003	0,002	0,003	0,002	0,003	0,002	0,002	0,002	0,001		
Tl-filtr	µg/l	0,003	0,003	0,003	0,003	0,002	0,002	0,002	0,002	0,002	0,001	0,002	0,001		
Pt	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.		
Pb	µg/l	1,72	1,83	3,34	2,71	1,76	1,81	1,77	1,85	1,73	2,12	1,91	1,90		
Pb-filtr	µg/l	1,73	1,70	2,56	2,71	1,71	1,72	1,69	1,72	1,69	1,88	1,97	1,76		

# Reactor 2

TANK 2		= < DETECTION LIMIT				*Ntot taken from Ramböll results 18/3				2015-04-01	
Parameter	Day	2015-03-18		2015-03-20		2015-03-25		2015-03-27		2015-04-01	
		Day 8		Day 10		Day 15		Day 17		Day 22	
		Before	After	Before	After	Before	After	Before	After	Before	After
pH	-	7,73	7,96	7,78	8,25	7,52	8,27	7,49	8,21	7,09	8,15
Cond	µS/cm	n.a.	1557	1717	1530	1961	1570	1965	1579	1516	1614
Temp	°C	9,3	12,2	9,9	11,2	9,7	10,5	7,9	9,2	8,4	9,2
BOD7	mg O2/l	n.a.	#VALUE!	0,14	0,55	n.a.	0,01	0,09	-0,65	0,76	-0,21
Alk	meq/l	16,50	11,44	13,12	11,28	16,50	11,28	14,4	11,2	10,88	10,48
Ptot	mg/l	0,05	n.a.	n.a.	n.a.	n.a.	9,00	1,19	9,77	0,02	11,71
Ptot-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
NH4-N	mg/l	61	6,9	18,1	0,8	63,0	5,8	48,3	1,5	44,9	0,2
Removal NH4-N	%	88,71		95,83		90,80		96,81		99,59	
NO3-N	mg/l	0,0	8,6	3,7	4,9	0,3	5,4	1,5	3,3	1,5	4,6
NO2-N	mg/l	0,9	3,4	1,5	1,5	0,0	1,5	1,5	1,5	1,5	1,5
TOC	mg/l	16	18,67	13,85	16,13	n.a.	15,33	6,90	13,66	6,21	11,76
COD	mg/L	97,20	63,07	90,75	55,48	#VALUE!	53,09	69,89	48,09	67,82	42,41
DOC	mg/l	16	17,64	15,77	12,04	n.a.	16,72	9,243	15,61	4,331	11,59
TC	mg/l	n.a.	140,50	162,9	130	n.a.	108,2	191	131,6	138,6	133,7
IC	mg/l	n.a.	121,80	149	113,8	n.a.	90,93	184,1	117,9	132,4	121,9
Ntot	mg/l	64	12,1	20,6	1,7	64	26,0	44,1	2,0	38,3	3,9
Removal Ntot	%	81,06		91,81		59,44		95,37		89,76	
Ntot-Nit frac	mg/l	2,12	-6,77	-2,78	-5,54	0,71	13,23	-7,26	-4,37	-9,67	-2,39
Li	mg/l	n.a.	0,04	0,04	0,04	n.a.	0,04	0,76	0,14	0,76	0,76
Na	mg/l	n.a.	202,34	205,77	202,41	n.a.	202,87	207,47	216,04	161,85	220,32
K	mg/l	n.a.	59,65	63,74	60,43	n.a.	61,80	59,70	61,67	42,19	64,04
Mg	mg/l	n.a.	35,02	33,69	33,42	n.a.	33,42	33,95	34,49	25,40	35,56
Mn	mg/l	0,56	0,66	0,60	0,60	n.a.	0,73	6,04	6,04	6,04	6,04
Ca	mg/l	n.a.	89,41	110,21	107,30	n.a.	105,81	71,29	119,60	70,54	108,41
Cl	mg/l	n.a.	123,70	130,45	126,32	n.a.	128,58	146,05	136,07	106,08	146,09
Br	mg/l	n.a.	19,34	19,51	24,96	n.a.	19,78	8,79	8,79	8,79	25,58
PO <sub>4</sub> <sup>3-</sup>	mg/l	n.a.	28,00	10,45	32,80	n.a.	42,52	10,45	49,10	10,45	29,25
SO4 <sup>2-</sup>	mg/l	n.a.	36,45	36,03	34,87	n.a.	34,87	31,17	29,90	37,62	28,95
Li	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	0,02	0,02	0,02	0,01	0,02
Li-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Be	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	0,003	0,001	0,003	0,002	0,002
Be-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.

TANK 2		= < DETECTION LIMIT				*Ntot taken from Ramböll results 18/3				2015-04-01	
Parameter	Day	2015-03-18		2015-03-20		2015-03-25		2015-03-27		2015-04-01	
		Day 8		Day 10		Day 15		Day 17		Day 22	
		Before	After	Before	After	Before	After	Before	After	Before	After
B-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Na	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	59,10	58,42	56,58	54,59	56,86
Na-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Mg	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	27,07	30,28	22,19	18,37	31,21
Mg-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Al	mg/l	n.a.	0,52	0,08	0,37	n.a.	0,21	0,04	0,19	0,03	0,08
Al-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Si	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	3,65	6,98	4,32	7,10	7,13
Si-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
K	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	56,29	62,25	43,86	34,46	59,66
K-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Ca	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	19,02	19,96	14,80	13,80	21,12
Ca-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Ti	mg/l	n.a.	0,26	0,23	0,30	n.a.	0,39	0,43	0,31	0,29	0,43
Ti-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
V	mg/l	n.a.	0,00	0,00	0,00	n.a.	0,00	0,00	0,00	0,00	0,00
V-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Cr	µg/l	2,7	3,24	2,62	2,76	n.a.	-1,20	-0,82	-1,21	-1,38	-0,79
Cr-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Mn	mg/l	0,56	0,71	0,40	0,44	n.a.	0,23	0,33	0,16	0,30	0,28
Mn-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Fe	mg/l	0,08	0,35	0,21	0,31	n.a.	0,18	0,14	0,17	0,08	0,11
Fe-filtr	mg/l	0,03	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Ni	µg/l	7,3	9,15	8,47	8,14	n.a.	2,22	2,83	1,31	0,45	3,07
Ni-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Co	µg/l	n.a.	2,50	2,13	2,08	n.a.	-1,52	-1,19	-1,76	-2,32	-1,01
Co-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Cu	µg/l	0,5	19,24	9,39	11,20	n.a.	-10,68	-11,11	-11,11	-14,77	-14,77
Cu-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Zn	µg/l	3	30,32	5,44	19,73	n.a.	14,98	5,19	12,89	3,31	9,41
Zn-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
As	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	0,03	0,03	0,03	0,02	0,03
As-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Se	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	0,25	0,28	0,22	0,17	0,28

TANK 2		= < DETECTION LIMIT											
Parameter	Day	2015-03-18		2015-03-20		2015-03-25		2015-03-27		2015-04-01			
		Day 8		Day 10		Day 15		Day 17		Day 22			
		Before	After	Before	After	Before	After	Before	After	Before	After		
Se-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	
Sr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	0,31	0,32	0,23	0,21	0,31		
Sr-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	
Mo	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	0,00	0,00	0,00	0,00	0,00		
Mo-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	
Ag	µg/l	<0,1	n.a.	n.a.	n.a.	n.a.	-2,37	-2,39	-2,37	-2,38	-2,38		
Ag-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	
Cd	µg/l	0,02	0,06	0,10	0,16	n.a.	1,73	1,70	1,72	1,75	1,75		
Cd-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	
Sn	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	3,96	0,42	2,61	0,49	3,00		
Sn-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	
Sb	µg/l	n.a.	0,44	0,53	0,49	n.a.	5,25	5,24	5,18	5,37	5,23		
Sb-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	
Ba	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	0,12	0,16	0,09	0,17	0,11		
Ba-filtr	mg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	
Hg	µg/l	<0,005	n.a.	n.a.	n.a.	n.a.	0,02	0,02	0,02	0,01	0,02		
Hg-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	
Tl	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	0,007	0,002	0,004	0,003	0,003		
Tl-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	
Pt	µg/l	n.a.	0,0007	0,0004	0,0015	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	
Pb	µg/l	0,2	0,45	0,67	1,18	n.a.	1,94	1,60	1,79	1,74	1,83		
Pb-filtr	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	

TANK 2		* Ntot taken from Ramböll results 15/4												*Aeration not working	
Parameter	Day	2015-04-08		2015-04-10		2015-04-15		2015-04-17		2015-04-23		2015-04-29			
		Day 29		Day 31		Day 36		Day 38		Day 44		Day 50			
		Before	After	Before	After	Before	After	Before	After	Before	After	Before	After		
pH	-	7,3	8,02	7,27	7,92	7,37	8,02	7,37	8,02	7,35	8,27	7,35	7,32		
Cond	µS/cm	1512	1209	1861	1456	1923	1510	1859	1505	1769	1361	1970	1930		
Temp	°C	9,5	11	11,2	12,2	9,8	11,7	9,8	10,4	12,1	14,8	12,1	13		
BOD7	mg O2/l	1,11	-0,08	1,46	0,12	1,27	0,4	1,22	0,04	0,64	0,01	0,02	0,04		
Alk	meq/l	10,88	7,52	15,36	10	15,44	10,4	15,52	11,52	15,2	10,16	16,56	16		
Ptot	mg/l	0,05	10,77	0,18	15,20	0,05	6,83	0,04	15,06	0,04	14,06	0,21	12,69		
Ptot-filtr	mg/l	0,04	10,11	0,05	12,64	0,04	7,30	0,02	15,23	0,04	13,24	0,12	15,74		
NH4-N	mg/l	38,8	1,5	55,0	1,2	51,3	1,1	24,5	4,7	31,1	1,1	35,1	34,4		
Removal NH4-N	%	96,03		97,84		97,82		80,86		96,40		2,11			
NO3-N	mg/l	3,4	9,4	2,8	14,1	5,0	7,3	16,2	3,9	4,4	2,6	2,4	2,6		
NO2-N	mg/l	1,5	1,5	1,5	1,5	1,4	1,1	1,4	1,1	1,1	1,1	1,4	1,4		
TOC	mg/l	n.a.	14,26	7,066	29,96	1,223	22,01	14,02	39,01	9,185	34,68	10,7	77,27		
COD	mg/L	#VALUE!	49,89	70,40	96,83	52,87	73,06	91,26	123,89	76,76	110,94	81,30	238,29		
DOC	mg/l	4,118	12,84	7,944	27,74	13,03	28,34	6,741	37,68	9,886	39,38	8,285	78,75		
TC	mg/l	n.a.	112,5	182,4	146,7	202,7	152,7	163,5	151,3	165,5	136,7	180,4	251,5		
IC	mg/l	n.a.	98,29	175,4	116,7	201,4	130,7	149,5	112,2	156,3	102	169,7	174,3		
Ntot	mg/l	58	8,1	53,3	12,9	46,7	11,6	52,9	11,5	49,7	1,4	54,3	53,5		
Removal Ntot	%	86,12		75,72		75,07		78,24		97,14		1,33			
Ntot-Nit frac	mg/l	14,18	-4,44	-6,04	-3,87	-11,10	2,09	10,80	1,81	13,06	-3,38	15,35	15,20		
Li	mg/l	0,11	0,76	0,76	0,07	0,69	0,56	0,69	0,56	0,56	0,56	0,69	0,69		
Na	mg/l	147,71	166,40	178,61	203,52	156,81	96,54	81,06	109,89	94,18	102,17	112,01	118,08		
K	mg/l	39,48	42,49	48,90	59,57	44,26	26,81	30,54	30,78	27,06	30,06	32,30	32,22		
Mg	mg/l	22,99	25,40	29,14	33,15	23,58	15,56	11,18	17,50	15,17	15,94	17,26	17,26		
Mn	mg/l	6,04	0,18	6,04	0,79	5,49	4,40	5,49	4,40	4,40	4,40	5,49	5,49		
Ca	mg/l	110,43	90,68	65,78	111,89	59,44	53,45	20,24	40,17	21,74	53,35	24,05	50,70		
Cl	mg/l	98,70	106,78	119,33	134,86	112,32	71,05	56,16	78,25	65,66	68,01	85,69	86,36		
Br	mg/l	8,79	8,79	8,79	8,79	7,99	6,39	7,99	6,39	6,39	6,39	7,99	7,99		
PO4 <sup>3-</sup>	mg/l	10,45	43,88	12,28	62,34	9,50	7,60	9,50	7,60	7,60	7,60	9,50	9,50		
SO4 <sup>2-</sup>	mg/l	38,57	37,30	34,34	34,02	29,97	22,90	29,30	22,06	22,52	26,05	26,99	26,80		
Li	mg/l	0,02	0,02	0,02	0,02	0,02	0,02	0,02	0,02	0,02	0,02	0,02	0,02		
Li-filtr	mg/l	0,02	0,02	0,02	0,02	0,02	0,02	0,02	0,02	0,02	0,02	0,02	0,02		
Be	µg/l	0,002	0,003	0,004	0,007	0,002	0,001	0,002	0,002	0,002	0,002	0,003	0,003		
Be-filtr	µg/l	0,001	0,001	0,002	0,001	0,002	0,001	0,002	0,001	0,002	0,001	0,002	0,001		
B	mg/l	0,31	0,29	0,48	0,46	0,46	0,44	0,44	0,45	0,41	0,39	0,51	0,36		

TANK 2		= < DETECTION LIMIT										*Aeration not working	
Parameter	Day	2015-04-08		2015-04-10		2015-04-15		2015-04-17		2015-04-23		2015-04-29	
		Day 29		Day 31		Day 36		Day 38		Day 44		Day 50	
		Before	After	Before	After	Before	After	Before	After	Before	After	Before	After
B-filtr	mg/l	0,29	0,31	0,42	0,44	0,42	0,48	0,41	0,47	0,38	0,37	0,38	0,50
Na	mg/l	55,58	57,55	59,52	57,03	56,69	56,98	53,91	53,84	57,50	54,66	56,34	53,63
Na-filtr	mg/l	55,12	54,75	56,70	57,19	54,34	56,55	52,74	55,10	54,14	52,95	55,64	56,63
Mg	mg/l	21,65	20,00	32,24	29,47	31,24	29,52	28,67	29,72	29,53	27,36	29,90	22,32
Mg-filtr	mg/l	20,38	19,87	27,50	28,43	28,46	31,65	27,75	29,88	26,83	26,32	23,28	28,75
Al	mg/l	0,03	0,14	0,09	0,46	0,04	0,03	0,13	0,09	0,07	0,10	0,17	0,13
Al-filtr	mg/l	0,03	0,02	0,03	0,02	0,04	0,03	0,07	0,02	0,06	0,03	0,06	0,02
Si	mg/l	8,19	5,71	11,44	6,13	10,06	9,15	10,55	9,35	11,48	9,91	12,40	9,00
Si-filtr	mg/l	7,84	6,46	9,27	8,58	10,00	10,62	10,16	10,42	11,28	10,34	9,99	11,95
K	mg/l	39,72	36,99	57,63	55,03	56,33	54,72	52,79	53,66	52,85	49,49	58,47	43,88
K-filtr	mg/l	37,97	39,34	51,61	52,81	52,55	58,08	51,54	55,10	49,80	47,09	46,02	56,78
Ca	mg/l	15,53	13,79	19,92	21,52	22,61	21,59	19,86	20,78	20,38	19,90	23,58	15,74
Ca-filtr	mg/l	21,75	15,56	19,88	21,26	20,51	21,34	20,94	20,08	20,69	20,37	16,19	20,89
Ti	mg/l	0,34	0,31	0,47	0,45	0,44	0,43	0,39	0,39	0,42	0,39	0,41	0,30
Ti-filtr	mg/l	0,32	0,31	0,40	0,41	0,41	0,45	0,37	0,40	0,40	0,38	0,32	0,39
V	mg/l	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00
V-filtr	mg/l	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00
Cr	µg/l	-1,04	-1,11	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00
Cr-filtr	µg/l	-1,18	-1,38	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00
Mn	mg/l	0,33	0,27	0,44	0,47	0,49	0,42	0,40	0,40	0,43	0,35	0,50	0,37
Mn-filtr	mg/l	0,44	0,27	0,43	0,40	0,44	0,42	0,42	0,39	0,43	0,34	0,34	0,46
Fe	mg/l	0,08	0,14	0,11	0,34	0,10	0,05	0,17	0,10	0,16	0,10	0,22	0,27
Fe-filtr	mg/l	0,07	0,03	0,08	0,05	0,08	0,05	0,09	0,05	0,08	0,04	0,09	0,09
Ni	µg/l	1,66	1,34	3,60	3,49	3,68	3,15	3,59	4,54	2,85	2,83	4,12	2,69
Ni-filtr	µg/l	1,52	0,98	2,85	2,74	3,25	4,21	2,88	4,37	2,72	2,47	2,77	5,49
Co	µg/l	-1,87	-2,02	-1,12	-1,19	-1,02	-1,21	-1,10	-1,07	-1,09	-1,25	-0,39	-1,21
Co-filtr	µg/l	-1,91	-2,10	-1,40	-1,33	-1,19	-1,03	-1,09	-1,00	-1,19	-1,33	-1,03	-0,54
Cu	µg/l	-11,71	2,30	0,23	-6,59	-14,65	-13,74	-14,04	-10,62	-12,00	-14,41	-12,10	
Cu-filtr	µg/l	-11,60	-16,21	-11,88	-13,17	-7,48	-14,67	-13,93	-16,12	-10,84	-15,73	2,36	-17,38
Zn	µg/l	10,94	11,34	15,67	29,07	8,00	6,25	5,96	7,62	8,68	13,79	8,05	16,67
Zn-filtr	µg/l	9,84	2,92	14,56	6,15	7,24	4,75	5,12	3,84	8,05	4,54	19,18	2,36
As	µg/l	0,02	0,03	0,03	0,04	0,03	0,03	0,04	0,04	0,03	0,03	0,05	0,03
As-filtr	µg/l	0,03	0,02	0,03	0,03	0,03	0,03	0,04	0,04	0,04	0,04	0,03	0,04
Se	µg/l	0,18	0,19	0,27	0,26	0,28	0,29	0,29	0,28	0,29	0,28	0,34	0,27

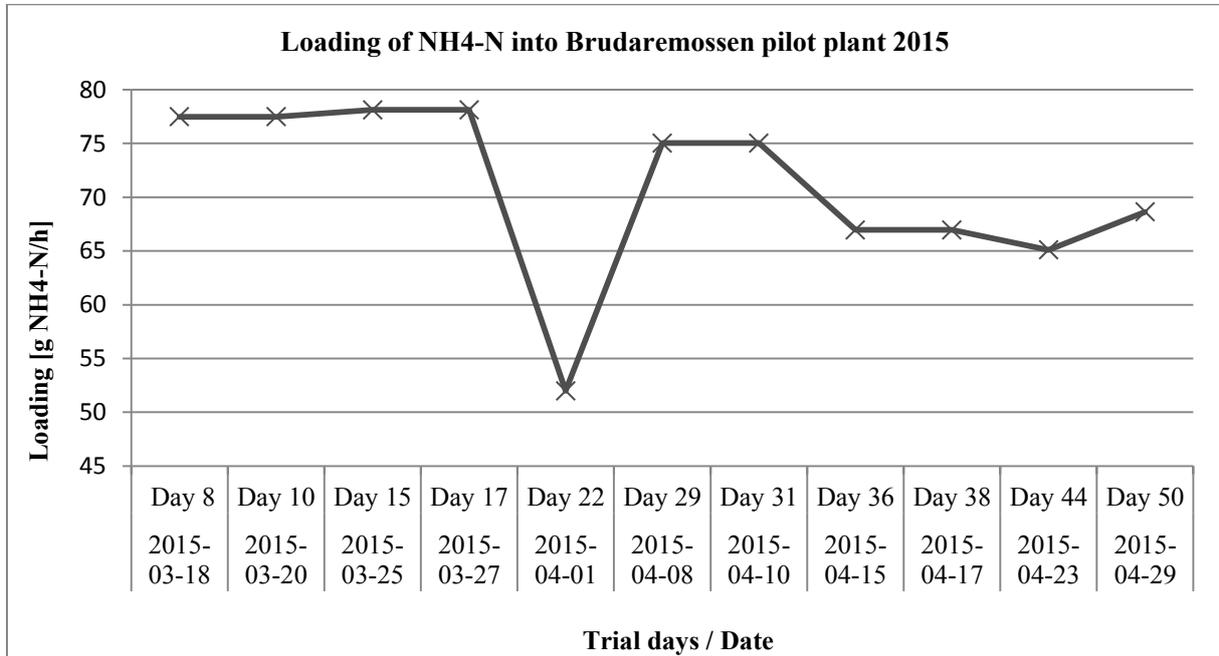
TANK 2		= < DETECTION LIMIT										*Aeration not working	
Parameter	Day	2015-04-08		2015-04-10		2015-04-15		2015-04-17		2015-04-23		2015-04-29	
		Day 29		Day 31		Day 36		Day 38		Day 44		Day 50	
		Before	After	Before	After	Before	After	Before	After	Before	After	Before	After
Se-filtr	µg/l	0,18	0,20	0,24	0,26	0,28	0,31	0,28	0,29	0,26	0,28	0,27	0,33
Sr	mg/l	0,24	0,22	0,33	0,30	0,31	0,30	0,28	0,28	0,29	0,27	0,28	0,21
Sr-filtr	mg/l	0,23	0,22	0,28	0,29	0,28	0,31	0,28	0,29	0,27	0,25	0,22	0,27
Mo	mg/l	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00
Mo-filtr	mg/l	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00	0,00
Ag	µg/l	-2,38	-2,38	-2,21	-2,13	-2,39	-2,38	-2,38	-2,38	-2,39	-2,38	-2,37	-2,38
Ag-filtr	µg/l	-2,38	-2,38	-2,01	-2,09	-2,38	-2,38	-2,39	-2,38	-2,39	-2,39	-2,37	-2,34
Cd	µg/l	1,71	1,72	1,90	1,79	1,71	1,71	1,75	1,75	1,75	1,76	1,73	1,72
Cd-filtr	µg/l	1,71	1,71	1,78	1,77	1,71	1,71	1,72	1,74	1,76	1,74	1,84	1,71
Sn	µg/l	0,43	2,45	0,91	3,25	0,70	2,42	0,85	2,21	0,83	2,68	0,87	0,80
Sn-filtr	µg/l	0,60	2,48	0,82	2,80	0,62	2,33	0,82	1,95	0,76	2,65	0,61	0,94
Sb	µg/l	5,44	5,37	5,38	5,34	5,29	5,28	5,40	5,38	5,36	5,34	5,29	5,17
Sb-filtr	µg/l	5,42	5,40	5,35	5,37	5,27	5,30	5,39	5,42	5,33	5,33	5,24	5,23
Ba	mg/l	0,18	0,14	0,20	0,24	0,22	0,19	0,20	0,18	0,21	0,18	0,23	0,18
Ba-filtr	mg/l	0,26	0,13	0,19	0,16	0,19	0,19	0,21	0,17	0,22	0,17	0,17	0,19
Hg	µg/l	0,02	0,01	0,02	0,07	0,02	0,01	0,01	0,01	0,01	0,01	0,01	0,01
Hg-filtr	µg/l	0,02	0,02	0,02	0,09	0,02	0,01	0,01	0,01	0,01	0,01	0,01	0,01
Tl	µg/l	0,004	0,005	0,003	0,008	0,003	0,003	0,002	0,003	0,003	0,002	0,003	0,001
Tl-filtr	µg/l	0,004	0,004	0,003	0,003	0,003	0,003	0,002	0,002	0,002	0,002	0,002	0,001
Pt	µg/l	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.	n.a.
Pb	µg/l	1,71	1,87	3,51	3,06	1,74	1,75	1,75	1,98	2,02	2,45	1,92	2,19
Pb-filtr	µg/l	1,70	1,70	2,47	2,48	1,72	1,73	1,71	1,85	1,88	1,99	1,83	1,77

## Sludge samples for Reactor 1 & Reactor 2

REACTOR 1		*Aeration not working						
Parameter	Unit	20-mar	27-mar	01-apr	08-apr	15-apr	23-apr	29-apr
		Day 10	Day 17	Day 22	Day 29	Day 36	Day 44	Day 50
SV	ml/L	460	340	290	220	218	200	150
MLSS	mg/L	3760	3460	1950	3500	2540	3670	2000
SVI	ml/g	122	98	216	63	88	54	78
MLVSS	mg/L	3000	3200	3160	2610	2910	3620	2740
F/M ratio	kg BOD7/kg MLSS*day	0,98	0,99	3,47	0,31	1,46	0,88	3,19

REACTOR 2		*Aeration not working						
Parameter	Unit	20-mar	27-mar	01-apr	08-apr	15-apr	23-apr	29-apr
		Day 10	Day 17	Day 22	Day 29	Day 36	Day 44	Day 50
SV	ml/L	300	220	230	160	191	180	160
MLSS	mg/L	3550	3260	2080	3510	3400	3610	550
SVI	ml/g	85	68	121	46	56	50	308
MLVSS	mg/L	2500	3100	2020	2850	3520	3500	2940
F/M ratio	kg BOD7/kg MLSS*day	0,06	0,04	0,59	0,47	1,11	0,79	0,17

## Appendix D. Loading of NH<sub>4</sub>-N into Brudaremossen pilot plant 2015



## Appendix E. Diagrams for removal of organic compounds and metals

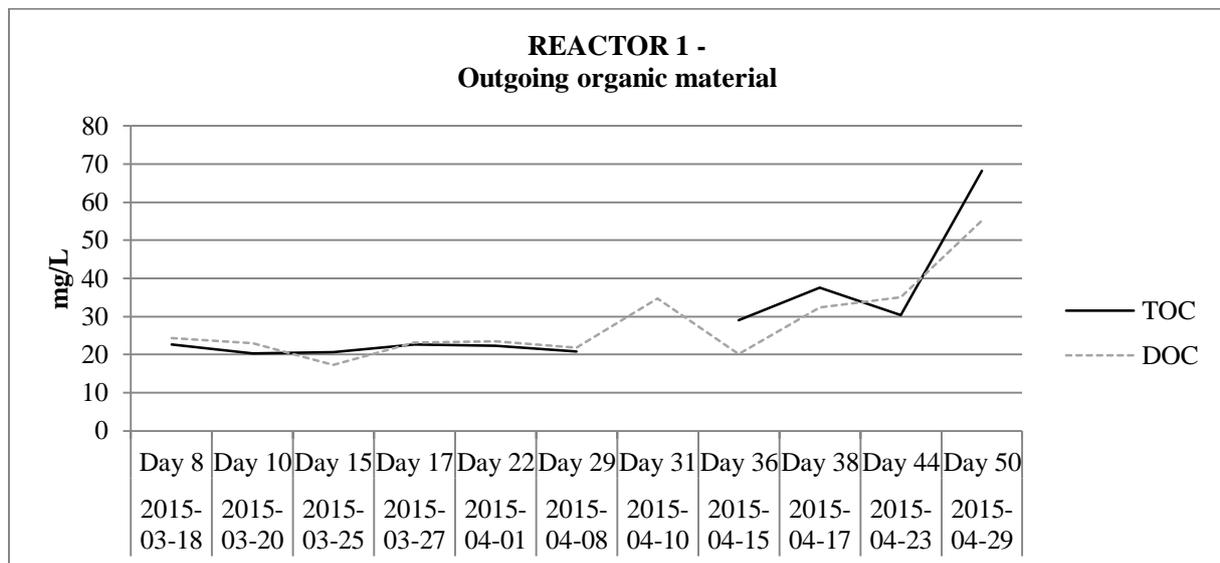


Figure 1 Concentrations of TOC, DOC and COD in the effluent from Reactor 1 throughout the experimental trial.

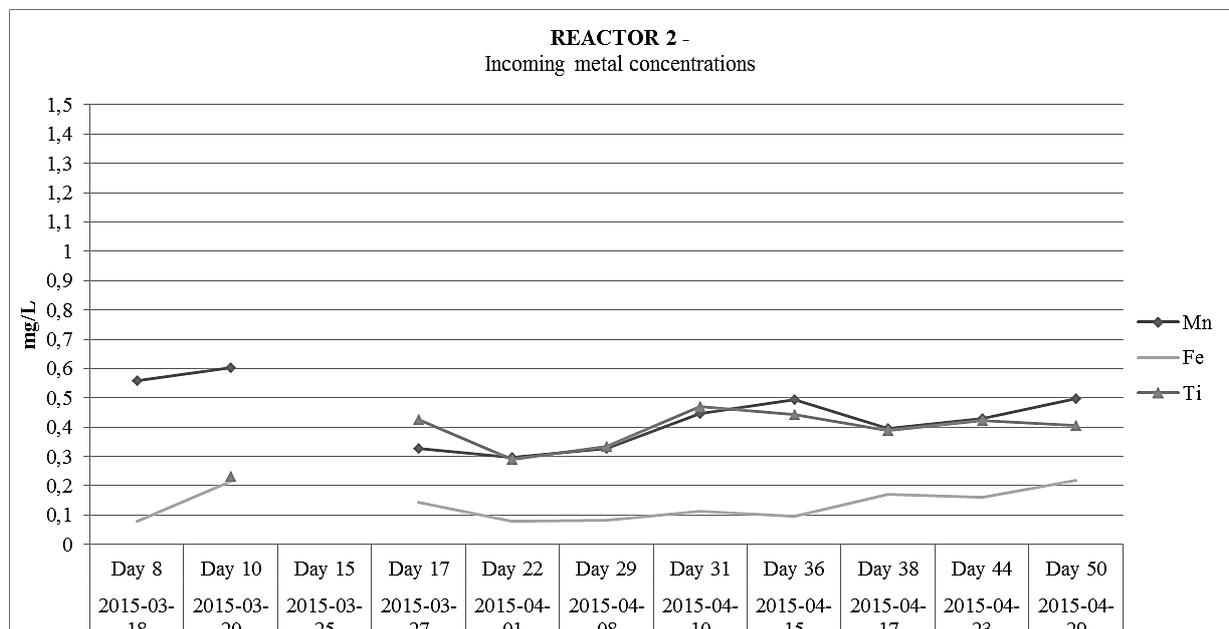
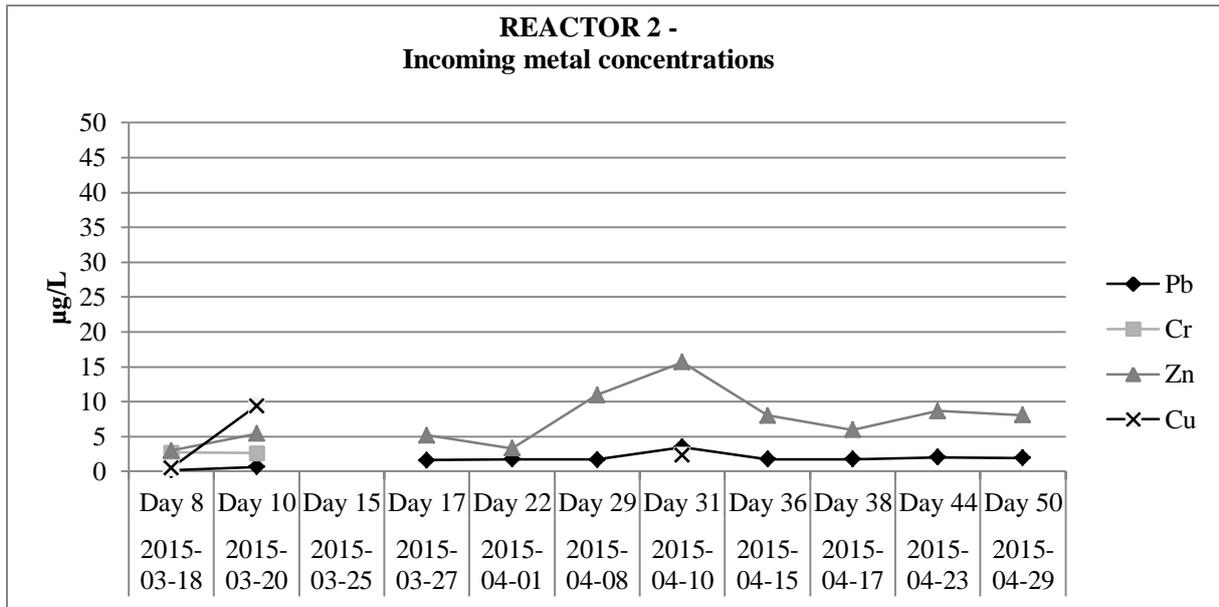


Figure 2 Concentrations of Mn, Fe and Ti in the influent to Reactor 2 throughout the experimental trial.



*Figure 3 Concentrations of Pb, Cr, Zn and Cu in the influent to Reactor 2 throughout the experimental trial.*

## Appendix F. Dissolved Oxygen & pH for the cycle analysis on Day 13

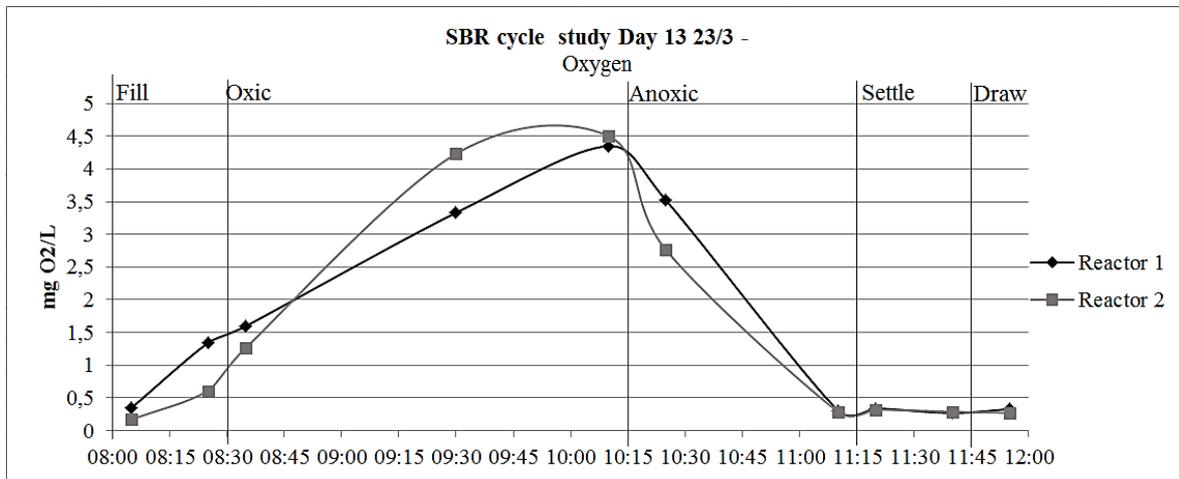


Figure 1 The DO concentrations on Day 13, 2015-03-23, for Reactor 1 and Reactor 2.

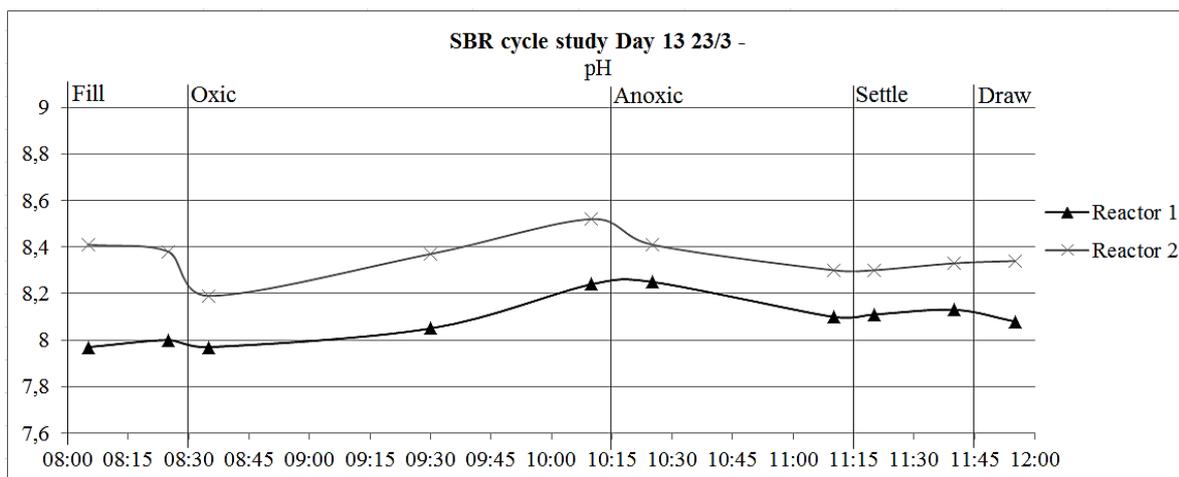


Figure 2 The pH measured on Day 13, 2015-03-23, for Reactor 1 and Reactor 2.

## Appendix G. Nitrification & denitrification rates in Reactor 2 on Day 44

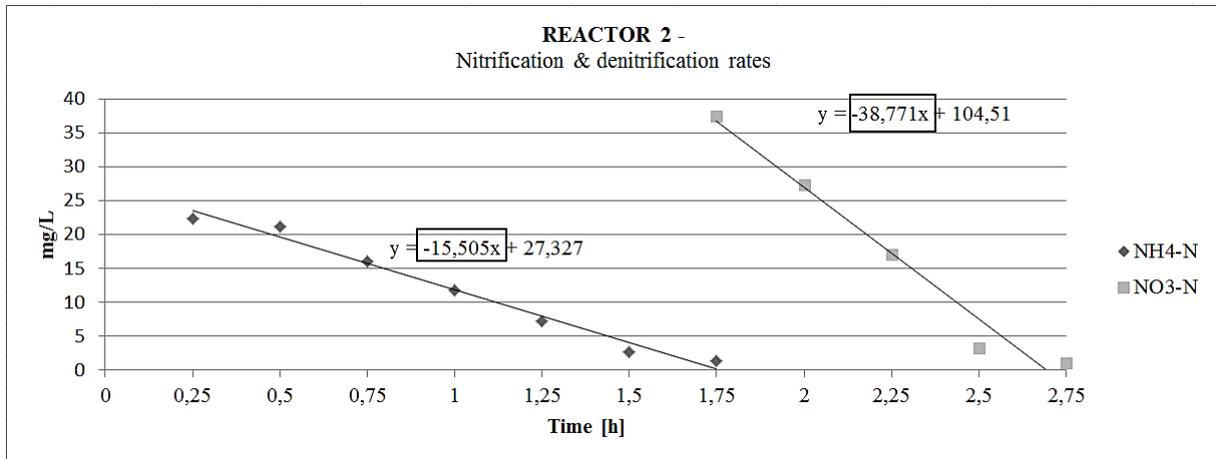


Figure 1 Nitrification and denitrification rates during the cycle analysis on Day 44, 2015-04-23, for Reactor2.

## Appendix H. Calculations for a full scale SBR design

### Input data

$$Q_{average} = 40 [m^3/day/reactor] = 960,000 [L/reactor]$$

$$NH_4 - N_{load} = 55 [mg/L] = 52.8 [kg/day]$$

$$R_{nitrification} = 13.75 [mg NH_4 - N/L \cdot h]$$

$$R_{denitrification} = 42.22 [mg NH_4 - N/L \cdot h]$$

$$No. cycles per day = 3 [Cycles]$$

$$SVI = 60 [mL/g]$$

$$MLSS = X = 3,670 [mg/L]$$

$$t_{settle} = 0.5 [h] (assumed)$$

$$t_{draw} = 0.5 [h] (assumed)$$

### Equations

$$Q_{average} = 960,000 [L/reactor] = 320,000 [L/cycle] \quad (1)$$

$$NH_4 - N_{load} = 52.8/3 = 17.6 [kg/cycle] = 17,600,000 [mg/cycle] \quad (2)$$

$$t_{aeration} = 17,600,000/(13.75 \cdot 320,000) = 4 [hours] \quad (3)$$

$$t_{anoxic} = 17,600,000/(42.22 \cdot 320,000) \approx 1.3 [hours] \quad (4)$$

It was decided to aerate during two hours of the  $t_{fill}$ , resulting in following phase times.

$$t_{fill} = t_{oxic} + t_{anoxic} + t_{settle} + t_{draw} = 2 + 1 + 0.5 + 0.5 = 4 [h] \quad (5)$$

$$t_{tot} = t_{fill} + t_{oxic} + t_{anoxic} + t_{settle} + t_{draw} = 4 + 4 = 8 [h] \quad (6)$$

Equation (7) describes the relationship between volume and biomass concentration whilst equation (8) and equation (9) determines the MLSS concentration in settled volume and the settled fraction respectively.

$$V_t \cdot X = V_s \cdot X_s \quad (7)$$

$V_t$  = Total volume [m<sup>3</sup>]

$X$  = MLSS concentration at full volume [g/m<sup>3</sup>]

$V_s$  = Settled volume after draw [m<sup>3</sup>]

$X_s$  = MLSS concentration in settled volume [g/m<sup>3</sup>]

$$X_s = \frac{V_t \cdot X}{V_s} = \frac{1}{SVI} = \frac{10^3 [mg/g] \cdot 10^3 [mL/L]}{60 [mL/g]} \approx 16,666 [g/m^3] \quad (8)$$

$$\text{Settled fraction} = \frac{V_s}{V_t} = \frac{X}{X_s} = \frac{3,670}{16,666} \approx 0.220 [-] \quad (9)$$

Provide with 20 percent of liquid above the sludge blanket so that solids are not removed during draw, which results in equation (10) and equation (11).

$$\text{Settled fraction} = \frac{V_s}{V_t} \cdot 1.2 \approx 0.264 [-] \quad (10)$$

$$\text{Fill fraction} = 1.0 - 0.264 = 0.736 \approx 0.7 [-] \quad (11)$$

The height of the draw and the required volume of the reactors are calculated in equation (12) and equation (13) respectively. A full liquid depth,  $H_{reactor} = 6.0 m$ , was assumed.

$$H_{draw} = 0.7 \cdot 6.0 = 4.2 [m] \quad (12)$$

$$V_{reactor} = \frac{Q_{average}}{0.7} = \frac{320 [m^3/tank]}{0.7} \approx 457 [m^3/tank] \quad (13)$$

Determination of overall hydraulic retention time  $\tau$  is explained in equation (14).

$$\tau = \frac{V_{reactor} \cdot 2 [reactors] \cdot 24 [h/day]}{Q_{average} [m^3/day]} \approx 23.6 [h] \quad (14)$$

The required area for each reactor is calculated in equation (15).

$$A_{reactor} = \frac{V_{reactor}}{H_{reactor}} = \frac{457}{6.0} \approx 76.2 [m^2/tank] \quad (15)$$

The pumping rates for the fill and draw pumps are calculated below - see equation (16) and equation (17).

$$P_{fill} = \frac{Q_{average}}{t_{fill}} = \frac{320}{240} \approx 1.333 [m^3/min] = 1,333 [L/min] \quad (16)$$

$$P_{draw} = \frac{Q_{average}}{t_{draw}} = \frac{320}{15} \approx 21.333 [m^3/min] = 21,333 [L/min] \quad (17)$$

The average oxygen transfer rate required for nitrification is based on the requirement of 3 mgO<sub>2</sub>/L and a safety factor of 2.0 to ensure enough oxygen, see equation (18).

$$O_2 \text{ rate} = 3 [mgO_2/L] \cdot Q_{average} \cdot 2.0 = 0.48 [kg O^2/h] \quad (18)$$

The sludge produced during the processes is calculated in equation (19). The required mixed sludge waste can thereafter be calculated according to equation (20).

$$\text{Sludge prod} = \frac{V_{reactor} \cdot X}{SRT} = \frac{457 \cdot 3,670}{32} \approx 52.4 [kg/day] \quad (19)$$

$$\text{Sludge waste} = \frac{\text{Sludge prod}}{X} = \frac{52.4}{3,670} \approx 14.28 [L/day] \quad (20)$$