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Long term trials with membrane bioreactor for enhanced wastewater treatment coupled with compact sludge treatment

-pilot Henriksdal 2040, results from 2018

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In cooperation with

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This report has been reviewed and approved in accordance with IVL's audited and approved management system.

Preface

This report presents work performed during 2018, within the long-term pilot study trials of municipal wastewater treatment with Membrane Bioreactors (MBR), including sludge treatment. The study is carried out in cooperation between IVL Swedish Environmental Research Institute and Stockholm Vatten och Avfall AB (Stockholm Water and Waste Company). The trials are performed at the R&D pilot facility Hammarby Sjöstadsverk in Stockholm, Sweden and they are jointly financed by the IVL foundation and Stockholm Vatten och Avfall AB.

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Summary

Henriksdal wastewater treatment plant (WWTP) in Stockholm is currently being extended and rebuilt for increased capacity and enhanced treatment efficiency. The new process configuration at the Henriksdal WWTP has been designed for a capacity of 1.6 million population equivalents (PE) which is about twice as much as today. The design maximum flow of the biological treatment is 10 m^3/s which is equivalent to 850 Mega Liters per Day (MLD). In addition, the treatment process has been designed to reach low nutrient concentrations in the effluent (5 mg BOD_7/L, 6 mg TN/L and 0.2 mg TP/L). The extension of the plant will include new primary treatment, new primary settlers and a new treatment step for thickening of primary and waste activated sludge. The reconstruction will include retrofitting of the existing conventional activated sludge (CAS) tanks with a new membrane bioreactor (MBR) process containing 1.6 million m^2 of membrane area. Digestion of thick sludge (~6% TS) will be done at thermophilic conditions instead of mesophilic digestion of thin sludge (~3-3.5%).

To increase the knowledge on membrane technology for wastewater treatment in Nordic conditions, Stockholm Vatten och Avfall (SVOA) decided, in 2013, to conduct long-term MBR pilot scale studies at the R&D facility Hammarby Sjöstadsverk, located on the premises of the Henriksdal WWTP. The pilot was completed by the end of 2013 and in full operation by early 2014. In 2017 it was decided to supplement the MBR pilot with a sludge treatment line in order to also have the possibility to study the future digestion process. The pilot scale studies are carried out in cooperation with IVL Swedish Environmental Research Institute. The studies will continue for as long as considered needed. This report presents the results from year 2018 (project year 5) of the pilot scale studies.

Results from previous years have verified that the process is able to treat a hydraulic load equivalent to the design load, and a nutrient load greater than the design load, to effluent concentrations below the future discharge limits. In addition, the function and resilience of the membrane design have been verified.

During 2018, a large focus was put on:

Mimicking the start-up operation of the first treatment line in full scale

The first full scale MBR treatment line at Henriksdal WWTP will be operated in a different way compared to the process design, during the first three years of operation. This include a high fixed flowrate, reject water connected to inlet, only ferrous sulphate (no ferric chloride) and no external carbon source. To gain knowledge of the treatment performance during these conditions the pilot was operated in this way for ten weeks (March to June). The results indicate that the process can manage TP below 0.2 mg/L and TN below 10 mg N/L in the effluent at the higher load expected for the first treatment line without use of ferric and external carbon. Also, no negative effects were observed regarding the membrane performance and permeability was similar before and after the trial.

Reduced resource consumption

Optimisation of resource consumption related to the membrane operation has also been in the spotlight during 2018. Trials to reduce the amount of scouring air used in the membrane tanks and the amount of chemicals used for membrane cleaning have been performed. Even though these trials were not finished by the end of 2018, and will continue in 2019, results indicate that there are

large potential savings in both chemical and energy use when operating the membrane tanks, without risking any decrease in membrane capacity.

Membrane cleaning

In order to study any possible differences in cleaning effect and membrane performance, the acid used for cleaning one of the membrane tanks (MT1), was, throughout 2018 oxalic acid, whereas the other membrane tank (MT2) was cleaned with citric acid. The results showed that the effect of cleaning with oxalic acid was at least as good as when cleaning with citric acid. Since oxalic acid is less expensive than citric acid, there is a large economic saving potential in switching to oxalic acid. Also, high phosphorus concentration peaks detected in the effluent during citric acid cleaning events, was not detected when using oxalic acid cleaning.

Before and after recovery cleaning the membranes were lifted for inspection (after two years of operation, new membranes were installed in 2016) and membrane fibers were sent for membrane autopsy to study the type of fouling on the membrane surface. The membranes were in good condition, the foulant before recovery cleaning contained mainly iron, some organic material and trace amounts of calcium phosphate. After recovery cleaning most fouling had been removed and the foulants left contained trace amounts of iron and organic material.

Phosphorus removal

The consumption of precipitation chemicals for phosphorus removal decreased significantly during 2017 leading to a hypothesis that enhanced biological phosphorus removal (EBPR) occurred in the pilot although there is no anaerobic zone presence. During 2018 this was confirmed by regular phosphate release tests showing high, but varying EBPR-activity over the year.

Sludge pilot operation

The sludge treatment line (including sludge thickening, anaerobic digestion and sludge dewatering) continued with mesophilic operation throughout the year. Due to several operational problems, much time was spent to upgrade, rebuild and adjust the different treatment steps. Trials on improved thickening and dewatering were performed.

Mapping of micro pollutans

A two-year long study on mapping of micro pollutants through the treatment process including pharmaceutical residues, micro plastics, bacteria, PFAS and chloro-organic halogens was started during autumn 2017 and the results of the second and third sampling campaign (out of a total of four planned campaigns, study is ending in 2019) is presented in this report.

Study visit to wastewater treatment plants with MBR in the USA

Visits to MBR WWTPs in the USA with similar regulations and recipient as Henriksdal WWTP was conducted. Some of the experiences from the plants is presented in this report.

Sammanfattning

Henriksdals avloppsreningsverk i Stockholm är under ombyggnad för att öka kapaciteten och avskiljningsgraden. Det nya reningsverket är designat för en kapacitet på 1,6 miljoner personekvivalenter (pe), vilket motsvarar ungefär dubbelt så mycket som 2018. Det nya reningsverket är också designat för att klara strikta utsläppskrav med avseende på fosfor, kväve och BOD_7 (5 mg BOD_7/L, 6 mg N-tot/L och 0,2 mg P-tot/L).

Uppgraderingen av Henriksdals reningsverk inkluderar ombyggnation av befintlig konventionell aktivslamprocess till en membranbioreaktorprocess (MBR) med 1,6 miljoner m^2 membranyta. Utöver detta byggs även en ny förbehandling, ny försedimentering och ett nytt behandlingssteg för primär- och överskottslam. Rötning av tjockt slam (ca 6 % TS) kommer ske vid termofila förhållanden istället för dagens mesofila rötning av tunt slam (ca 3-3,5 % TS).

MBR är en relativt väl beprövad teknik inom både industriell och kommunal avloppsrening men införandet i Henriksdal innebär en rad utmaningar för vilka tekniska och driftsmässiga lösningar utvecklas och testas i ett pilotprojekt på forskningsanläggningen Hammarby Sjöstadsverk. Projektet har pågått sedan 2013 och kommer att fortsätta så länge det bedöms att det finns ett behov av pilottester för Henriksdals framtida process. Under 2017 utökades projektet genom att MBR-piloten kompletterades med slambehandling för att även kunna studera framtida rötningsprocess för Henriksdal. Projektet är gemensamt finansierat av IVL Svenska Miljöinstitutet och Stockholm Vatten och Avfall. I den här delrapporten redovisas resultat från år 2018 (projektår 5) av pilotförsöksprojektet.

Resultat från tidigare års försök har visat att processen kan rena en hydraulisk belastning som motsvarar den dimensionerande belastningen och en näringsämnesbelastning som överstiger den dimensionerande belastningen till utgående koncentrationer som underskrider de framtida reningskraven. Även membranens funktion och uthållighet har verifierats tidigare.

Under 2018 hade pilotförsöken störst fokus på:

Imitation av uppstartsdrift av första MBR-linjen i fullskala på Henriksdals reningsverk

Första MBR-linjen (av totalt 7) på Henriksdals reningsverk kommer under de första åren att driftas under delvis annorlunda förhållanden än vad den dimensionerats för. Detta inkluderar att flödet kommer att styras till ett högt jämnt flöde, rejektvatten från avvattning av rötat slam kommer att ledas till inloppet, enbart järnsulfat (ingen järnklorid) kommer finnas tillgängligt som fällningskemikalie och ingen extern kolkälla kommer användas. För att få erfarenheter av hur processen fungerar vid dessa förhållanden gjordes ett 10 veckor långt försök i piloten under våren 2018. Försöket visade att processen klarar att nå under totalfosfor 0.20 mg P/L och under totalkväve 10 mg N/L i utgående vatten vid den höga belastningen, utan att använda järnklorid och extern kolkälla. Dessutom syntes inga negativa effekter på membranens drift eller funktion och permeabiliteten var likvärdig innan, under och efter försöket.

Minskad resursförbrukning

Under 2018 har stort fokus legat på att minska resursförbrukningen relaterad till driften av membranen. Försök att minska mängden luft som används i membrantankarna och mängden kemikalier som används för membranrengöring har genomförts. Försöken är inte slutförda utan kommer fortsätta under 2019, men redan nu finns indikationer på stora potentiella besparingar i både kemikalie- och energiförbrukning för membranen utan att risk för minskad kapacitet.

Membranrengöring

För att studera eventuella skillnader i rengöringseffekt och membranprestanda så användes oxalsyra för rengöring av ena membrantanken (MT1) under hela 2018 medan den andra membrantanken (MT2) rengjordes med citronsyra. Resultaten visade att effekten av rengöring med oxalsyra var minst lika bra som vid rengöring med citronsyra. Eftersom oxalsyra är billigare än citronsyra finns det en stor ekonomisk besparingspotential i att byta citronsyra mot oxalsyra. Höga fosforhalter i utgående behandlat vatten har noterats i samband med citronsyrarengöring. Motsvarande toppar har inte uppstått vid rengöring med oxalsyra.

Före och efter återhämtningsrengöring lyftes membranen ur sina tankar för visuell inspektion (efter två års drift, nya membran installerades 2016) och membrantrådar klipptes av och skickades för ”membranobduktion” för att studera typ av beläggning som fanns på membranytan. Membranen var i gott skick och den beläggning som fanns innan återhämtningsrengöring bestod av järn, vissa organiska material och små mängder kalciumfosfat. Efter rengöringen hade det mesta av beläggningen tvättats bort och kvar var små mängder järn och organiskt material.

Fosforrening

Förbrukning av fällningskemikalie för fosforrening minskade kraftigt under 2017 vilket resulterade i en hypotes om att utökad biologisk fosforrening (bio-P) utvecklats i processen trots avsaknaden av en anaerob zon. Under 2018 bekräftades detta med hjälp av regelbundna fosforsläppstester som visade på en hög varierad bio-P-aktivitet över året.

Slampbehandling

Slambehandlingslinjen (som inkluderar förtjockning, anaerob rötning och slamavvattning) var i drift med mesofil rötning under 2018. På grund av de många driftstörningar som uppstod spenderades stor del av året på att uppgradera, bygga om och justera de olika behandlingsstegen. Försök för att förbättra både förtjockning och avvattning genomfördes.

Kartläggning av mikroföroreningar

En tvååring studie för kartläggning av förekomsten av mikroföroreningar, såsom läkemedelsrester, mikroplast, bakterier, PFAS och klororganiska halogener i behandlingsprocessen startade under hösten 2017 och resultaten från den andra och tredje provtagningskampanjen (av totalt fyra, studien avslutas under 2019) presenteras i denna rapport.

Studiebesök på avloppsreningsverk med MBR i USA

Under 2018 genomfördes studiebesök i USA på avloppsreningsverk med MBR och liknande utsläppskrav och recipient som Henriksdals reningsverk. Erfarenheter från dessa anläggningar sammanfattas i denna rapport.

Terminology

ADM1	Anaerobic Digestion Model No. 1
Aerobic	Aerated
Anoxic	Non-aerated
AOX	Adsorbable organic halogensm (mg/L)
BB1	Bio-Block 1. First biological full scale treatment line to be reconstructed to MBR at Henriksdal WWTP.
BOD_7	Biochemical Oxygen Demand, 7 days (mg/L)
BR1 to BR6	Biological reactor 1 to 6, sampling points
BSM2	Benchmark Simulation Model No. 2
CAS	Conventional Activated Sludge
CFD	Computational Fluid Dynamics
COD	Chemical Oxygen Demand (mg/L)
cTOC	collodial Total Organic Carbon (mg/L)
DDMS	Dewatered digested mixed sludge, sampling point
DMS	Digested mixed sludge, sampling point
DO	Dissolved Oxygen (mg/L)
DS	Daily composite sample (flow proportional)
EFF	Effluent water, sampling point
EOX	Extractable organic halogens (mg/L)
Fe	Iron (mg/L)
F/M ratio	Food to Mass, incoming substrate in relation to the amount of microorganisms (kg BOD_7/kg SS, d)
Flux	Flow rate per unit area (L/(m^2·h)). Flux is a measurement of the load on the membranes
Fouling	Clogging of the pores in the membranes, causing reduced flow rate through the membranes
FS	Flat sheet (membrane type)
GS	Grab sample
Hepta	Iron(II)sulfate heptahydrate
HF	Hollow fibre (membrane type)
HFO	Hydrous ferric oxides
IN	Influent wastewater, sampling point
MBR	Membrane BioReactor, bio reactor with membrane separation
MLD	Million litres per day
MLSS	Mixed liquor suspended solids (mg SS/L)
MT1	Membrane tank 1 (of 2), sampling point
MT2	Membrane tank 2 (of 2), sampling point
MC	Maintenance cleaning
MS	Mixed sludge (PS+WAS), sampling point
NIT	Nitrification zone
NH_4-N	Ammonium nitrogen (mg/L)
NO_2-N	Nitrite nitrogen (mg/L)
NO_3-N	Nitrate nitrogen (mg/L)
Org-N	Organically bound nitrogen (mg/L)
PA	Pre-aeration tank
PE	Population equivalent (defined as 70 g BOD_7 per person and day)
Permeability	Flux per TMP (L/(m^2·h·bar)). Permeability is a measure of how well a specific flux permeates the membranes. The permeability gradually decreases with time due to fouling
Permeate	The treated wastewater that has passed through the membranes
PFAS	Perfluorinated Alkylated Substances
PIX	PIX 111, iron(III)chloride solution
PO_4-P	Phosphate phosphorus (mg/L)
Pre-DN	Pre-denitrification (Anoxic)
Post-DN	Post-denitrification (Anoxic)

PS	Primary sludge, sampling point
PTW	Primary treated water, water after primary settler, sampling point
RAS	Return activated sludge, sampling point
RAS-DeOx	Zone where return activated sludge (RAS) is led for reduction of DO concentration
RC	Recovery cleaning
RWD	Reject water from sludge dewatering, sampling point
RWT	Reject water from sludge thickening, sampling point
Scouring air	Constant air flow around the membranes to reduce fouling
SED	Pre-sedimentation (Primary settler)
SFA 2040	Stockholms Framtida Avloppsvattenrening år 2040 (name of reconstruction project)[1]
SS	Suspended Solids (mg/L)
SVOA	Stockholm Vatten och Avfall
TDS	Total Dissolved Solids (mg/L)
TOC	Total Organic Carbon (mg/L)
TMP	Transmembrane pressure (mbar). The pressure difference between two sides of a membrane, shows how much force is needed to push water through a membrane
TN	Total nitrogen (mg/L)
TP	Total phosphorus (mg/L)
TMS	Thickened mixed sludge, sampling point
TS	Total Solids (%)
TSS	Total Suspended Solids (%)
TTF	Time To Filter (s)
VS	Volatile Solids (%)
VSS	Volatile Suspended Solids (%)
WAS	Waste activated sludge, sampling point
WS	Weekly composite sample

[1] www.stockholmvattenochavfall.se/en/sfa-start/

1 Introduction

This report presents the results from year 2018 (project year 5), of the pilot scale trials with membrane biological treatment of municipal wastewater, carried out in cooperation between IVL Swedish Environmental Research Institute and Stockholm Vatten och Avfall AB (SVOA) at the R&D facility Hammarby Sjöstadsverk, in Stockholm, Sweden. In the trials, an activated sludge process with a new process configuration is combined with membrane filtration to reach a higher level of purification, operational stability and treatment capacity. In addition, the sludge is treated by thickening, digestion and dewatering with the goal to evaluate high loaded thermophilic digestion with short retention time. Project years 2014-2017 are presented in separate reports.

In the initial chapters (2-3), the project background and the configuration of the pilot plant are described. An overview of the experimental plan is presented in chapter 4, followed by a method description in chapter 5. Finally, all results are presented and discussed in chapter 6.

2 Background

Within the project Stockholm's Framtida Avloppsrening (SFA, *Stockholm's future wastewater treatment)*, the Henriksdal WWTP in Stockholm, Sweden, is being extended and rebuilt for increased capacity and enhanced treatment efficiency. The decision to extend and rebuild is based on several factors such as; (i) SVOA´s WWTP in Bromma (which is already over loaded with very limited space available for extension) will be decommissioned in 2025 to give space to new housing areas, and the wastewater will be led to the Henriksdal WWTP in a new 14 km long sewage tunnel, (ii) the population in the Stockholm region is increasing at a high rate, resulting in an increased influent load, and, (iii) the Swedish Environmental Court has decided to sharpen the effluent requirements on the WWTPs in the Stockholm region, which demands more efficient wastewater treatment processes.

The new process configuration at the Henriksdal WWTP has been designed for a capacity of 1.6 million population equivalents (PE) which is about twice as much as today. The design maximum flow of the biological treatment is 10 m^3/s which is equivalent to 850 MLD. In addition, the treatment process has been designed to reach low nutrient concentrations in the effluent (5 mg BOD_7/L, 6 mg TN/L and 0.20 mg TP/L). The extension of the plant will include new primary treatment, new primary settlers and a new treatment step for thickening of primary and waste activated sludge. The reconstruction will include retrofitting of the existing conventional activated sludge (CAS) tanks with a new MBR-process containing >1.6 million m^2 of membrane area. The first MBR-line, out of seven, will be taken into operation in 2020 and the retrofitting of all seven lines will take an additional 8-9 years. The sand filters, currently used as a final polishing step for phosphorus removal, will in the future be used for wet weather overflow treatment. Digestion of thick sludge will be done at thermophilic conditions instead of mesophilic digestion of thin sludge. Design data for the future Henriksdal WWTP can be found in Table 1, Table 2 and Table 3.

The MBR technology is well-known internationally with long term experiences from both industrial and municipal WWT. In Italy and Germany relatively large municipal WWTPs with MBR-technology have been in operation for around 15 years (Brepols, 2010; Judd, 2010). In USA, China, Japan, South Korea, France, Great Britain and Spain, there are several large MBR-plants (50,000-80,000 PE) which have been in operation for 5-10 years (Judd and Judd Limited, 2017). The largest MBR-plant in operation today is Huaifang Water Recycling Project in Beijing, China (commissioned in 2016), designed for an average inflow of 6.9 m^3/s, which is slightly larger than the capacity of the future Henriksdal WWTP (design average 6.1 m^3/s). Europe's largest MBR in operation, also the largest ZeeWeed (SUEZ) plant is Seine Aval in France (commissioned in 2016), with a design average inflow of 2.6 m^3/s (www.thembrsite.com, 2019-07-11).

Challenges for the future MBR-process at the Henriksdal WWTP include

- high seasonal variations in water temperature and inflow, affecting both the membrane performance and the nitrogen removal,
- to meet the strict effluent requirements for phosphorus (0.20 mg TP/L and 27 tons/year, which equals <0.15 mg/L before 2040) by means of pre- and simultaneous precipitation (no final polishing step), without affecting membrane performance and
- to minimize the resource consumption.

There are MBR-plants in the USA, eg. *Broad Run* and *King William County* in Virginia, *Ruidoso* in New Mexico and *Cauley Creek* and *Yellow River* in Georgia, that reach very low effluent nutrient concentration, 0.05-0.10 mg TP/L and 0-6 mg TN/L without final polishing steps (Pellegrin & Neethling, 2015). Phosphorus removal at these plants is achieved by a combination of biological phosphorus removal (EBPR) and precipitation using a trivalent metal ion (Al^{3+} or Fe^{3+}). However, none of these treatment plants use ferrous (Fe^{2+}), which is planned to be utilized at the Henriksdal WWTP, or have as low incoming water temperatures as the Henriksdal WWTP.

Membrane filtration requires aeration and chemicals for maintenance and cleaning of the membranes. However, each plant is unique, and the cleaning schedule can and should be optimized for the local conditions in order to save resources.

The SFA-project will also affect the sludge treatment. The load on the digesters is expected to double but the digester volume will not be expanded. Consequently, digestion must be performed with high organic load and short hydraulic retention time. To manage this, the raw sludge will be thickened, and digestion will be performed at thermophilic conditions. There are several uncertainties regarding the sludge handling, including: function of thickening of fine particulate MBR-sludge, stability of the digestion process, biogas production potential, smell, pumping of thick sludge, and function of dewatering of thermophilic digested sludge.

To increase the knowledge on membrane technology for wastewater treatment in Nordic conditions, SVOA decided in 2013 to conduct long-term pilot scale studies at the R&D facility Hammarby Sjöstadsverk, located on the premises of the Henriksdal WWTP. In 2017 SVOA decided to supplement the MBR-pilot with a sludge treatment line in order to study the future digestion process. The pilot scale studies are carried out in cooperation with IVL Swedish Environmental Research Institute.

3 Description of the pilot plant

The pilot plant is designed to be a small copy of the future Henriksdal WWTP plant, scale 1: 6,700. The incoming wastewater is pumped from the Henriksdal inlet with a mean flow of around 3.2 m^3/h. Primary treatment comprise a fine screen, pre-aeration, a primary settler and fine sieve. The biological treatment consists of a pre- and post-denitrification followed by two parallel membrane tanks and a return activated sludge DeOx zone. The sludge treatment consists of thickening, anaerobic digestion and dewatering. The pilot plant process set-up is shown in Figure 1. All equipment in the pilot has been linked to a control system and process control is highly automated.

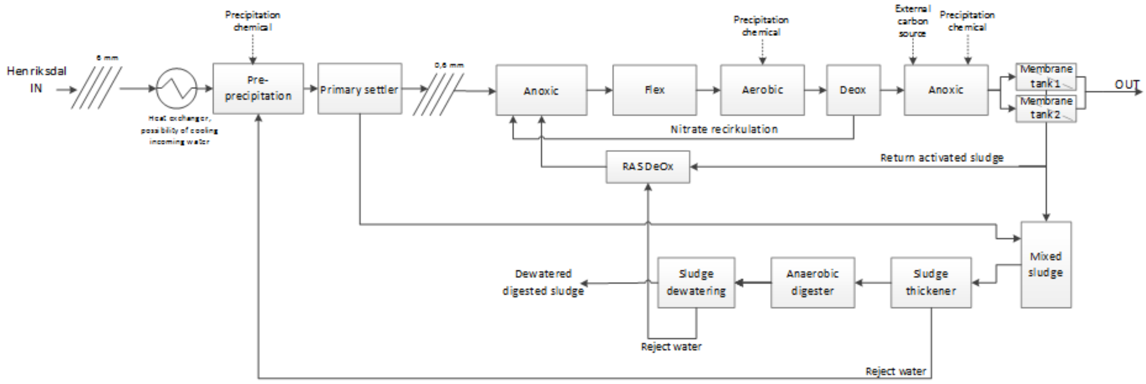


Figure 1. Flow scheme of the pilot WWTP. The biological treatment consists of 6 bioreactors (BR), BR1+BR2 are anoxic, BR3 is flexible, BR4+ a part of BR5 are aerobic, remaining part of BR5 is deox and BR6 is anoxic.

The reactor volumes of the pilot plant and the function of each reactor are specified in Table 1 together with a comparison to the future Henriksdal WWTP design.

Table 1. Reactor volumes in the wastewater treatment line in the pilot compared to the future Henriksdal WWTP (SFA).

Tank	Pilot (m^3)	Future H-dal (m^3)	Scale factor H-dal/Pilot	Specification
Pre-treatment				
PA (sand trap)	0.7	2 460	-	Pre-aeration. Dosing point 1 Fe^{2+}.
SED	3.3	30 000	9 200	Primary settler. Withdrawal of primary sludge.
Membrane bioreactor (MBR)				
BR1	4.8	33 500	7 000	Stirred. Pre-denitrification.
BR2	4.8	33 500	7 000	Stirred. Pre-denitrification.
BR3	4.8	40 000	8 300	Stirred/(aerated). Pre-denitrification/(nitrification). FLEX.
BR4	4.8	31 000	6 500	Aerated. Nitrification. Dosing point 2 Fe^{2+}
BR5ox	1.5	10 000	6 700	Aerated. Nitrification.
BR5DeOx	3.3	15 000	4 500	Stirred. DeOx.
BR6	4.8	24 000	5 000	Stirred. Post-denitrification. Dosing external carbon. Dosing point 3 Fe^{3+}.
MT1	1.45	9 750	6 700	Aerated. Membrane tank.
MT2	1.45	9 750	6 700	Aerated. Membrane tank.
RAS-DeOx	2.7	18 000	6 700	Stirred. DeOx. Addition of reject water (RWD). Withdrawal of WAS (before addition of RWD.
Summary MBR				
Total MBR	34.4	224 500	6 500	BR1-6, MT1-2, RAS-DeOx
Sludge treatment				
MS tank	0.4	1 060	2 650	Stirred. Tank for PS + WAS before thickening
Digester	5.9*	38 000	6 500	Stirred. Anaerobic digestion volume
DMS tank	0.2	9 000	45 000	Circulation mixing. Tank for digested mixed sludge before dewatering.

*The volume is set by choosing the liquid level in the digester and can be increased or decreased.

3.1 Process description water line

A schematic view of the wastewater treatment line is presented in Figure 2.

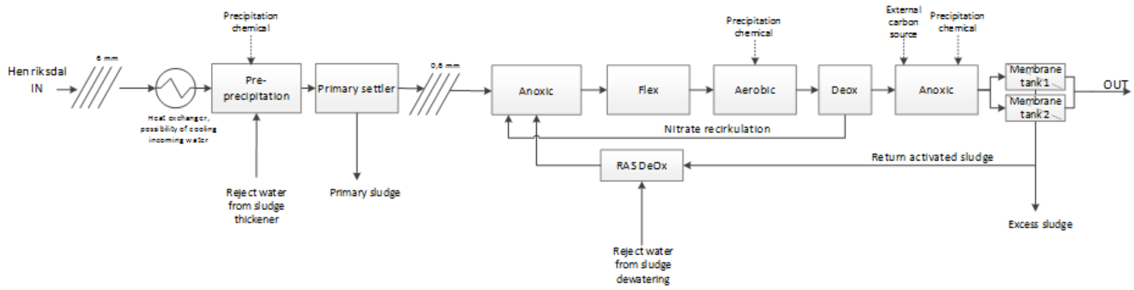


Figure 2. Process set-up for the wastewater treatment line.

3.1.1 Incoming wastewater

Incoming wastewater to the pilot plant is pumped from the Danviken tunnel, one of five inlet tunnels to Henriksdal WWTP plant. The pilot influent contains 10-20% higher concentration of organic matter (BOD_7) than the combined average inflow to the Henriksdal WWTP and about 60% higher BOD_7-concentration than in the inlet to the Bromma WWTP (the combined inlet from Henriksdal and Bromma will make up the future inlet to the Henriksdal WWTP, after reconstruction). The incoming flow rate to the pilot plant is proportional to the projected inflow to the Henriksdal WWTP year 2040. Flow variations in the inflow are proportional to the actual variations in inflow to the Henriksdal WWTP (controlled by signal from flow meters in the full-scale plant).

Since the influent to the pilot is set by the scaled down flow rate, and not a scaled down load, the incoming load on the pilot plant is proportionally higher than the corresponding design load for the Henriksdal WWTP, year 2040, see Table 2.

In addition, the incoming wastewater to the pilot has a higher temperature than incoming wastewater to Henriksdal. Previously, the incoming wastewater was during some periods cooled in heat exchangers. In April 2017 heat exchangers were taken out of operation due to repeated clogging after installation of a new coarser fine screen. At the end of March 2018 new heat exchangers were installed, however, due to continuous problems with clogging they were only in operation for shorter periods. The temperatures in the incoming wastewater to Henriksdal and to the pilot are presented in Figure 3. On average the temperature of the water to the pilot was about 2.5 °C higher than the influent wastewater to Henriksdal.

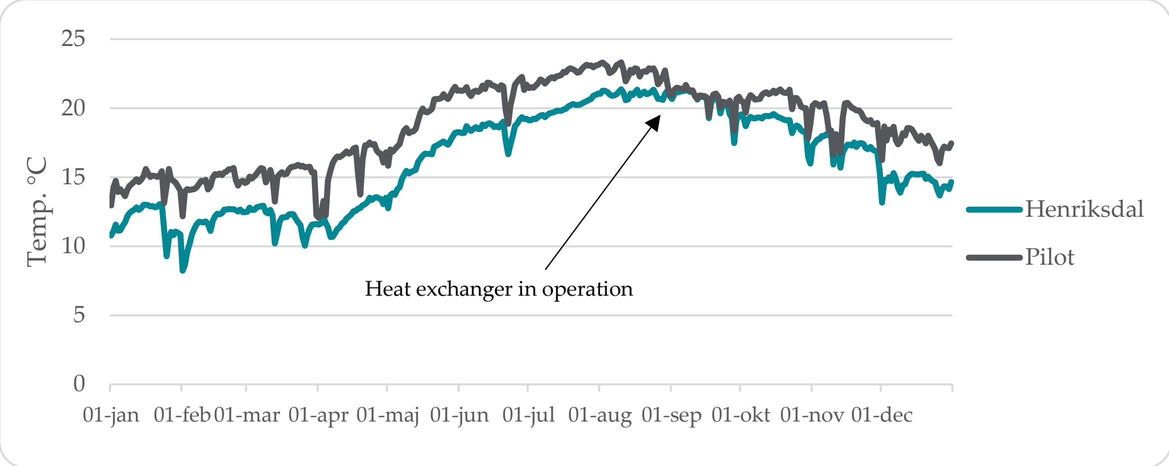


Figure 3. Influent temperatures to the MBR pilot (black line) and the Henriksdal WWTP (green line).

3.1.2 Pre-treatment

The pre-treatment steps in the pilot consisted of a 6 mm punch hole screen (efficiency similar to 2 mm step screen), a pre-aeration tank with ferrous sulfate dosing and a vertical flow primary settler, with a surface area of 1.13 m² and a water depth of 4.3 m. The primary settler in the pilot plant is proportionally smaller than the one in Henriksdal (scale 1:9,200 compared to the future Henriksdal design) since it was decided to use an existing installation, modified in 2017, instead of constructing a new. Finally, a 0.6 mm punch hole drum sieve was installed between the primary settling tank and the biology, see Figure 4. The small hole size of the drum sieve was chosen to enable the study of clogging tendencies (2 mm will be used in full scale).

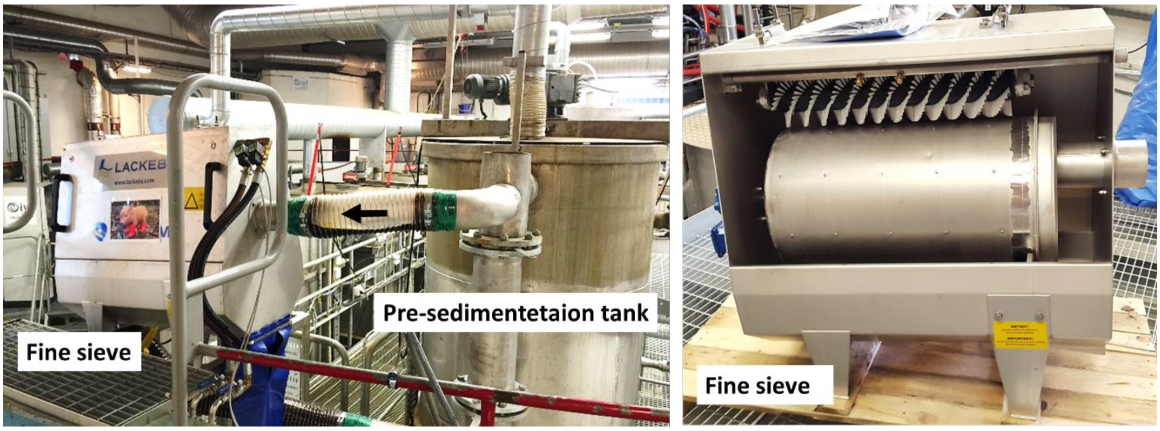


Figure 4. Photo of the fine sieve installation.

3.1.3 Biological treatment

The biological treatment consisted of six identical biological reactor tanks, BR1-6, see Figure 5. All tanks were equipped with stirrers and BR3 to BR5 were equipped with membrane disc aerators. BR5 was divided into two zones where the first one was aerated and the second one was stirred. The biological process was operated with pre-denitrification, nitrification and post-denitrification with methanol as external carbon source. The oxygen-rich return activated sludge (RAS) flow (4×Q) passed a specific RAS-DeOx zone where RAS was mixed with ammonium-rich reject water from digested sludge dewatering before recirculation to the pre-denitrification zone. Waste activated sludge (WAS) was taken out from the return sludge stream, after the membrane tanks and prior to the RAS-DeOx. Precipitation chemicals for phosphorus removal were dosed in BR4 and BR6.

Figure 5. Photo of the top of biological treatment tanks BR2-4

The biological treatment set-up was almost identical to the design of the future Henriksdal WWTP in scale 1:6700, with few minor exceptions. The deox zone in BR5 and the post-denitrification zone in BR6 were slightly over dimensioned. The discrepancy is due to the size of the existing tanks in the pilot plant and the difficulties in creating zones within the tanks. When setting up the pilot, a correct volume of the aerated zones for nitrification was given priority (BR4 and BR5ox), as the size of these zones will be crucial for the nitrogen removal.

Another difference between the pilot and the future Henriksdal WWTP is that the pilot lacks a RAS-channel. Instead, the RAS flowed directly from the membrane tanks into the RAS-Deox from where it was pumped back into BR1. In the full-scale plant, the RAS will flow into a RAS-channel by gravity and then be pumped into the RAS-Deox zone from where it will flow to the pre-denitrification zone by gravity. The volume of the RAS-channel will be small (HRT ~ 2 minutes) which puts a lot of pressure on the RAS-pumps. This could not be tested in the pilot since the RAS-Deox volume is much larger (HRT ~ 10 min). Table 1 shows the size of the treatment volumes in the pilot plant compared to the design of the future full-scale system at Henriksdal.

3.1.4 Membrane tanks

In the pilot, hollow fiber membrane from Suez with a nominal pore size of 0.04 µm was used. The membrane pilot was made up of two cassettes (2.5 m x 1.0 m x 0.34 m) consisting of three membrane modules each, see Figure 6, immersed in two separate tanks. Each module had a membrane area of 34.4 m^2 and consisted of membrane fibers fastened at the top and bottom of the cassette frame. The filtered water (permeate) was transported on the inside of the fibers to connections in both the bottom and the top of the module. The membranes were kept clean during operation by aeration from below. As shown in Figure 6c, the membranes were not completely tensioned between the top and bottom, so that the air bubbles causes the fibers to move and thus more easily remove sludge stuck on the membrane fibers.

Figure 6. The membrane during installation in the pilot. a) Membrane cassette with three membrane modules, b) cassette lowered into the tank, view from above, c) mounting and aeration equipment at the bottom of the cassette, d) permeate connections (yellow) at the top of the cassette.

The total membrane surface area in the pilot (204 m^2) corresponds to the design membrane surface installed in six (out of seven) treatment lines in the full-scale plant. The reasons for this are both that it corresponds to two standard design pilot cassettes from the manufacturer and that the SFA design has accounted for that the design max flow rate to the biological treatment could be treated even if one of the seven treatment lines are out of operation.

In future Henriksdal WWTP, each treatment line (a total of 7) will have 12 membrane tanks each that can be taken into and out of operation depending on the influent flow rate. Each membrane tank is equipped with 12 cassettes, consisting of 48 modules. This provides good flexibility and an opportunity to always have a constant flux across the membrane surface. In the pilot there are only two membrane tanks, which give less flexibility than will be found at future Henriksdal WWTP. At design flow rate and normal operation, a membrane area of approximately 160 m^2 would have to be in operation in the pilot, which corresponds to 4.7 modules. However, as a pilot cassette contains three modules, the pilot could only be operated with three or six modules in operation. To enable operation at a constant flux, the pilot was equipped with permeate recirculation, meaning that the flow through the membranes was higher than the inflow but this was compensated by having a partial flow of the permeate recycled to the membrane tank.

The airflow requirement for membrane cleaning in the pilot plant is higher than the corresponding airflow in the Henriksdal design since both cassettes in the pilot plant must be aerated all the time. In future Henriksdal, only the number of membrane tanks required for the current flow will be in operation and only the membrane tanks in operation will be constantly aerated, which means a minimum air consumption.

The two membrane cassettes in the pilot were parallel to enable comparisons of different operational strategies.

3.2 Process description sludge treatment

During 2017 the MBR-pilot was supplemented with a sludge treatment line proportional to the sludge treatment of the future Henriksdal design, the configuration is shown in Figure 7. The aim of the sludge pilot is to test the future operation of digestion at Henriksdal with high organic load, short retention time and thermophilic conditions. During 2018, however, the sludge pilot was operated under mesophilic condition to obtain reference values at the same time as the process was modified and optimized in order to make it function and operate continuously.

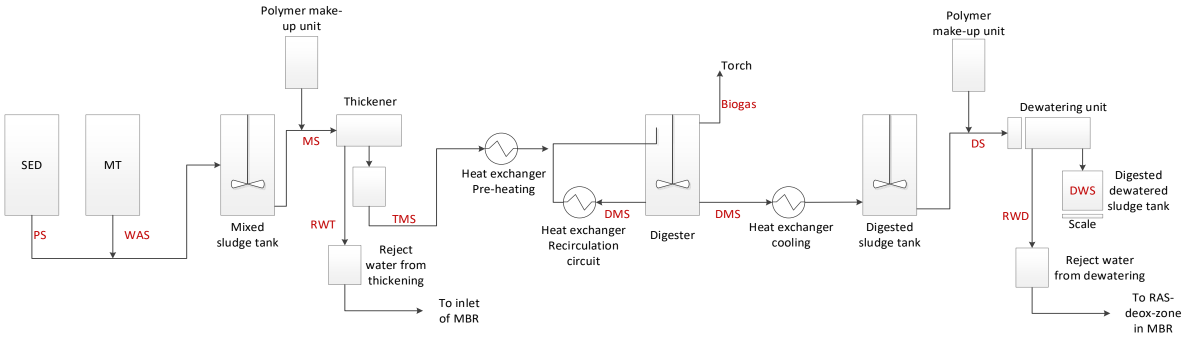


Figure 7. Process set-up for the sludge treatment line. PS=primary sludge, WAS=waste activated sludge, MS=mixed sludge, RWT=reject water from thickener, TMS=thickened mixed sludge, DMS=digested mixed sludge, DS=digested sludge, DWS=dewatered sludge, RWD=reject water from dewatering.

3.2.1 Thickening

Primary sludge and waste activated sludge was intermittently pumped to the stirred mixed sludge tank. Mixed sludge was then pumped to a rotating drum sieve thickener (Hjortkaer), se Figure 8. The goal was to reach 5-7% TS after thickening. Polymer was dosed inline in one of three possible dosing points. Reject water from the thickener flowed by gravity into a tank and was pumped back to the pre-aeration tank in the wastewater treatment line if the SS content was not too high. Thickened mixed sludge was injected into the digesters heat exchanger recirculation circuit by an eccentric screw pump and fed into the digester.

A major difference between the sludge treatment pilot and the future Henriksdal WWTP is that the primary and the waste activated sludge will be thickened separately at Henriksdal while the two sludge types are mixed before thickening in the pilot. This solution was chosen because of space and budget limitations and the fact that the main purpose with the pilot is to study high loaded digestion with short HRT. In addition, at Henriksdal, centrifuges and band thickeners will be used, not drum sieves. Choice of equipment for the pilot was done based on price and availability of small size machines.

During 2018 several attempts to optimize and control the function and result of the thickener were done, some of which were part of an MSc-project (Jirblom, 2019).

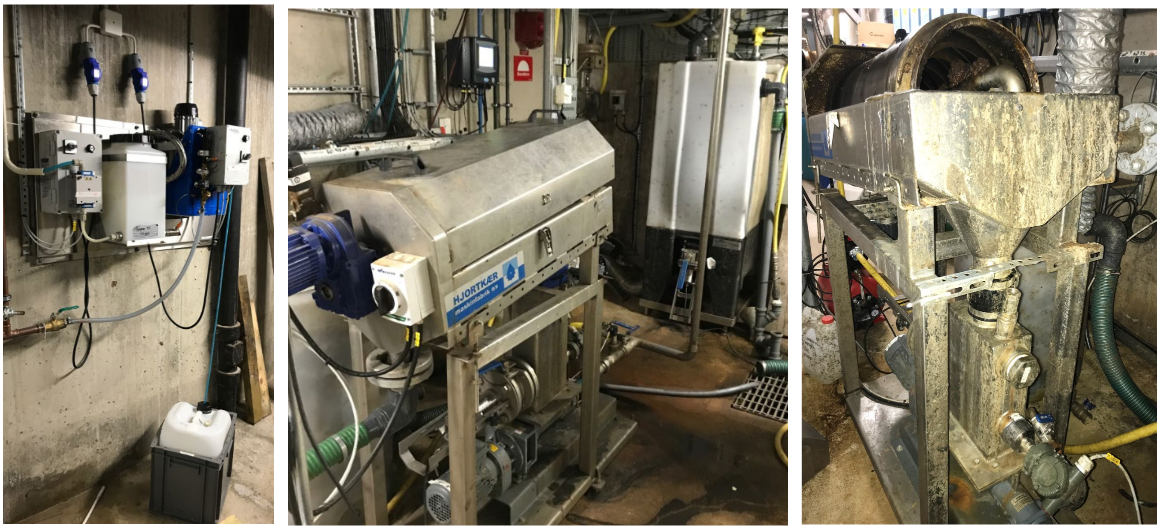


Figure 8. Photo of thickener and polymer make up unit.

3.2.2 Digestion

The digester is cylindrical with a base area of 2.54 m^2 and a variable water level. During 2018 the volume was kept at 5.9 m^3 which corresponds to full digester capacity in the future Henriksdal WWTP (scale 1:6,700). The sludge in the digester is homogenised and kept in suspension by a stirrer and by the recirculation flow. The recirculation circuit consist of a pump which is operated at its minimum capacity, approximately 3 m^3/h, and a heat exchanger controlled by a temperature meter in the digester. Digested sludge is pumped out of the digester, through a heat exchanger which can cool the sludge to a chosen temperature, and into an equalization tank (digested sludge tank).

Thickened mixed sludge was digested at mesophilic conditions. During 2018, the goal was to obtain a mesophilic (37 °C) reference period, and tests of increasing the temperature were done in order to verify that the heating system could reach thermophilic conditions (55 °C). During 2019, thermophilic digestion will be applied, which will be the mode of operation at the future Henriksdal WWTP.

No external organic material was fed to the digester during the pilot operation. In the future Henriksdal design, fat from restaurant drains and industrial byproducts like glycerol will be co-digested with wastewater treatment sludge.

3.2.3 Dewatering

Digested sludge was pumped out of the digested sludge tank into a pressurised, stirred mixing tank. Polymer was dosed inline just before the inlet to the mixing tank. From the mixing tank digested sludge was fed into a screw press. Dewatered sludge was collected in a vessel and weighted. The dewatering equipment is shown in Figure 9. The dewatering unit was not continuously operated during 2018.

Reject water from digested sludge dewatering was collected in a tank and was intended to be pumped through a filter into the RAS-Deox zone in the wastewater treatment line. Since the dewatering unit was not in constant operation in 2018, reject water was not returned to the treatment line during most of the year. Reject water from Henriksdal WWTP was used prior to primary sedimentation week 10-23 for imitation of first full scale treatment line operation and added to the RAS-Deox week 36-41 for separate evaluation of the RAS-Deox. Other times of the year the pilot was operated without reject water.

Figure 9. Photo of the dewatering equipment in the pilot.

3.3 Flow rate and load

Mean values for flow rates and loads in the pilot wastewater and sludge treatment lines during 2018 are shown in Table 2 and Table 3 respectively, together with the design values for the future Henriksdal WWTP. The design data for the pilot are also given in the table for comparison. The pilot was in operation during the entire year without any longer interruptions in operation.

The average incoming flowrate in 2018 was higher than the design flow rate; 3.5 m^3/h compared to the design average flow rate 3.16 m^3/h. This was done in accordance with the test plan for the pilot which included testing operational strategies with high load.

Table 2. Operation and design data for the wastewater treatment line in the pilot plant and design data (year 2040) for the future Henriksdal WWTP.

Parameter	Unit	Value Pilot	Design Pilot	Design future H-dal	Design future H-dal/ Value Pilot[i]
Flowrates					
Average influent flowrate, Q_{in}	m^3/h	3.5	3.16	20 880	6 000
Design flowrate, Q_{dim}	m^3/h		3.32	21 960	6 600
Max flowrate	m^3/h	5.5	5.44	36 000	6 500
Min flowrate	m^3/h	1.8	1.8	11 600	6 400
Nitrate recirculation flowrate	m^3/h	5.1-13.1	3.8-13.3	-	-
Nitrate recirculation flowrate	$\times Q_{in}$	2.6	1.2-4.2[ii]	0-4	-
RAS flowrate	m^3/h	4.1-19.6	3.6-19	-	-
RAS flowrate	$\times Q_{in}$	3.6	1.1-5.9[ii]	4 (3-5)	1.1
Incoming load					
BOD_7 influent	mg/L	265	206[iii]	216	0.8
SS influent	mg/L	322	201[iii]	280	0.9
TN influent	mg/L	46	44[iii]	37	0.8
TP influent	mg/L	6.1	5.7[iii]	4.9	0.8
Primary settler (SED)					
BOD_7 reduction over SED	%	25	46	50[iv]	2.0
SS-reduction over SED	%	35	60	60[iv]	1.7
TN reduction over SED	%	1	10	10[iv]	10
TP reduction over SED	%	12	40	40[iv]	3.3
BOD_7 PTW	mg/L	197	112	108	0.5
SS PTW	mg/L	200	80	112	0.6
TN PTW	mg/L	46	40	33	0.7
TP PTW	mg/L	5.4	3.4	3.0	0.6
SS removed over SED	kg SS/d	9.6	13.3[v]	89 300	9 300
Primary sludge production	kg SS/d	16.1	17.2[v]	115 000	7 100
VS-concentration PS	% of TS	88%	77%	77%	-
Biological treatment					
BOD_7-load PTW (at average flowrate)	kg BOD_7/d	16.3	8.6	57 500	3 500
Specific WAS-production [vi]	kg SS/kg BOD_7	0.87	1.02	1.02	1.2
WAS production, average	kg SS/d	14.1	8.8	58 600	4 200
VSS-concentration WAS	% of SS	73%	64%	64%	0.9
SS in biological tanks	mg/L	7 100	8 000	8 000	1.1
SS in membrane tanks	mg/L	9 800	10 000	10 000	1.0
Total sludge age	d	23.8	32.0	31.2	1.3
Membrane tanks					
Installed membrane area (gross)	m^2	206	206	1 600 000	7 800
Permeate recirculation	m^3/h	0.03-0.9	0.05-2	-	-
Net flux average (at average T)	$l/m^2,h$	19.7	17.9	20.9	1.1
Net flux max	$l/m^2.h$	28.2	30.8	30	1.1
Permeate pumping max	m^3/h	7.0	12.4	62 250	8 900
Permeate pumping min	m^3/h	0	0	0	-
Specific air demand at Leap-Lo [vii]	$Nm^3/h, m^2$	0.136	0.136	0.098	0.7
Specific air demand at Leap-Hi [vii]	$Nm^3/h, m^2$	0.252	0.252	0.196	0.8

[i] Design SFA divided by Value pilot. Value either 6 700 or 1 for complete compliance.
[ii] Based on average flowrate 3.2 m^3/h.
[iii] Design based on data from 2015.
[iv] Measured at Fe-dosage ca 10 g/m^3 in FL/sand trap.
[v] Calculated based on incoming load/scaled from SFA design with factor 6 700.
[vi] Excluding external carbon source.
[vii] Aeration of the membranes had two modes, one with lower (Leap-Lo) and one with higher air flowrate (Leap-Hi).

Table 3. Operation (average 2018, except week 25-33) and design data for the sludge treatment line in the pilot plant and design data (year 2040) for the future Henriksdal WWTP.

Parameter	Unit	Value Pilot	Design Pilot	Design future H-dal	Design future H-dal/ Value Pilot
Into thickener					
Flow mixed sludge (MS)	L/h	50	70	444 000[a]	8 900
TS-concentration MS	%	1.4%	1,6%	1.6%	1.1
VS-concentration MS	% of TS	83%	72%	72%	0.9
TS-load MS	kg TS/d	16.5	27.5	173 600	10 500
Polymer consumption	g/kg TS	6.4	5	6	0.9
After thickener (TMS into digester)					
Flow thickened mixed sludge (TMS)	L/h	12	16,6	118 000	9 800
TS-concentration TMS	%	5.1%	6.7%	6.0%	1.2
TS-load TMS	kg TS/d	12.6	27.0	172 000	13 700
VS-load TMS	kg VS/d	10.5	19.5	124 000	11 800
Flow reject RWT	L/h	46	53.4	326 000	7 100
SS-concentration reject RWT	mg/L	2 800	650	500	0.2
VSS-concentration reject RWT	% of SS	77	-	-	-
Digestion					
Digester temperature	°C	37	37/55	55	-
Retention time	d	20	5-20	13[c]	0.7
Specific VS-load	kg VS/m³,d	2.1	3.3	3.3[c]	1.6
Digestion efficiency	% of VS_{in}	46%	50%	42%[c]	0.9
VFA-concentration	mg/L	225	-	-	-
pH	-	7.2	-	7	1.0
Alkalinity	mg CaCO3/l	4 600	-	-	-
VFA/Alkalinity	mg/mg	0.05	-	-	-
NH_4-N	mg/L	949	-	-	-
Out of digester					
Flow DMS	L/h	13	16.6	123 000	9 500
TS-concentration DMS	%	2.9%	4.4%	3.9%	1.3
VS-concentration DMS	% of TS	72%	56%	60%	0.8
TS-load DMS	kg TS/d	9.0	18.0	124 000	13 800
VS-load DMS	kg VS/d	6.5	9.8	74 000	11 400
Specific biogas production	Nm³/kg $VS_{digested}$	1.0[d]	1.0	1.0	1
Flow biogas	Nm³/d	4.8[d]	9.7	52 000[c]	10 800
Methane content biogas	%	60%	65%	65%	1.1
Dewatering					
Flow DDMS	L/h	1.4	16.6	17 000	12 100
TS-concentration DDMS	%	27%	30%	30%	1.1
Flow reject RWD	L/h	11.6	14.2	114 000	10 300
SS-concentration reject RWD	mg/L	3 350	1 050	<900	0.3
Polymer consumption dewatering	g/kg TS	15	10	6-10	0.4-0.7

a) WAS and PS are thickened separately in the future Henriksdal process.
b) Not equal to the production of mixed sludge due to repeated operation failures.
c) Numbers without addition of external organic material (fat and glycerol)
d) m³/d not Nm³/d.

3.4 Chemicals

During 2018 methanol was used as external carbon source in the post denitrification zone. The phosphorus was precipitated using ferrous sulfate at two dosing points and ferric chloride in one point. For membrane cleaning sodium hypochlorite was used for both membrane tanks (MT) while one MT was cleaned using citric acid and the other one using oxalic acid.

3.4.1 External carbon source

Methanol was used as external carbon source since beginning of 2017. It was delivered in 25 L canisters and had a concentration of 1 200 g COD/L (concentration 100 % by weight). Previous years other external carbon sources were tested. The first and parts of the second year (until April 2015) sodium acetate was used, and the second and third year (2015-2016) Brenntaplus was used.

The dosing point of methanol was in BR5 deox-zone from mid-June 2017. Before that the dosing point was in the post-denitrification zone (BR6). The change was made in order to prolong the residence time for the added carbon and avoid leakage into the membrane tanks. However, it was noted that the carbon source consumption increased when dosed in the BR5 deox-zone. Week 45 2018 the dosing point was changed again, to a point in-between the BR5 deox-zone and BR6, to avoid recirculation back to BR1 from the BR5 deox-zone.

More about carbon source addition and treatment results can be found in section 6.2.2 Denitrification and section 6.10 Resource consumption.

3.4.2 Precipitation chemicals

Phosphorus was removed in the aqueous phase by precipitation with iron(II)sulfate heptahydrate (termed "hepta" in the report) and PIX 111 (iron(III)chloride; termed "PIX" in the report) in three dosing points; hepta in aerated pre-precipitation tank, hepta in the aerated part of the biological treatment (BR4) and PIX at the end of post-denitrification (BR6). Further details on the control of precipitation chemicals are given in section 6.3.1 Precipitation.

Hepta was collected in diluted form from Henriksdal treatment plant in batches of about 500 L. The iron content of the hepta solution varied during the experimental period between 14-71 g/L. For the batches used in the experiment, the iron content was determined by density measurement for each batch.

PIX was delivered as solution with a concentration of 35-45% by weight as specified by the supplier. The iron concentration used for control and dose calculation was 195.6 g Fe/L.

3.4.3 Chemicals for membrane cleaning

The membranes have been cleaned regularly with sodium hypochlorite and either citric acid or oxalic acid.

Sodium hypochlorite was delivered as a solution with a concentration of 10-20% by weight (150-185 g Cl_2/L), as specified by the supplier. The chlorine concentration in sodium hypochlorite decreases during storage. To prevent fast degradation the sodium hypochlorite has been stored in a closed, dark container. According to literature the rate of the degradation also decreases if the solution is diluted upon delivery (Svenskt Vatten, 2010a). During 2018, both diluted and concentrated sodium hypochlorite in the storage tank has been tested, and pumping have been adjusted to provide the right concentration in the solution entering the membranes during cleanings. Dilution was done with tap water to a concentration of about 60 g Cl_2/L. The concentration of sodium hypochlorite in the storage tank varied between 5 and 148 g Cl_2/ L during the year. A good correlation between conductivity and chlorine concentration was confirmed during 2017 and from the end of June 2018 an on-line sensor for conductivity was installed in the sodium hypochlorite container in order to monitor the chlorine degradation in detail.

Citric acid solution was delivered with 51% by weight as specified by the supplier.

Oxalic acid was delivered as powder which was dissolved in batches to a saturated solution (8% by weight).

For more information on how the cleanings were carried out, see section 6.5.3 Membrane cleaning.

3.4.4 Polymers

For thickening of mixed sludge different anionic polymers were tested, namely:

- Superfloc C-1592RS (Kemira)
- Superfloc C-1598RS (Kemira)
- Superfloc SD-6085 (Kemira)
- Flopam EM 640 HIB (SNF)

Towards the end of the year it was decided to continue with the Flopam EM 640 HIB polymer for sludge thickening. For dewatering of digested sludge, Superfloc C-1598 was used. Polymer was delivered in solution and prepared to the selected concentrations in % by weight solution in automated polymer make up units.

3.5 Control system

The pilot plant uses a control system consisting of a PLC (ABB AC800M) and a SCADA (UniView version 9.01). The control system is a standard system used at several treatment plants in Sweden. All equipment connected to the pilot, including the membranes, is controlled via the control system, with the exception of pumping of reject water that was locally controlled. Implementation of the control system has been carried out within the project, which provides great flexibility to adapt and optimize the control system.

4 Experimental plan year 2018

An overview of the experimental plan of year 2018 is presented in Table 4 and in more detail in later chapters of the report. During 2016-2017 the main goal of the project was to verify that the process design could meet the future effluent requirements for nitrogen (6 mg/l), BOD_7 (5 mg/l) and phosphorus (0.20 mg/l) and that the membranes functioned as expected. With this proven, the overall goals for 2018 was to continue with stable operation at different operational conditions, to minimize the resource consumption in the process, to test and evaluate specific processes/functions within the MBR-line and to achieve proper function of the sludge pilot.

Table 4. Experimental plan of year 2018.

Trial	Jan	Feb	Mar	Apr	May	Jun	Jul	Aug	Sep	Oct	Nov	Dec
Imitation of first phase operation SFA			X	X	X							
Trial with no Fe dosage							X	X	X			
Methanol used as carbon source	X	X								X	X	X
Nitrous oxide emission measurements				X	X					X		
Optimisation of scouring air use	X	X	X	X	X	X	X	X	X	X	X	X
Oxalic acid and citric acid comparison	X	X	X	X	X	X	X	X	X	X	X	X
Optimisation of membrane cleaning chemical use	X	X	X	X	X	X	X	X	X	X	X	X
Recovery cleaning and sampling					X							
Mapping of micro-pollutants				X								
RAS-DeOx evaluation										X		
Temperature increase in Digester	X				X		X					
Thickening tests				X						X	X	

During spring 2018, the pilot was operated with a high fixed flowrate, no addition of external carbon source or ferric chloride (PIX) and with reject water from sludge dewatering dosed prior to the primary sedimentation tank instead of to the RAS-deox. This was done to imitate the planned operation of the first full-scale MBR-line at Henriksdal during phase 2 of the SFA-project year 2020-2023. In phase 2, the first full-scale MBR line, out of seven, will be taken into operation and two of the remaining six activated sludge treatment lines will be shut down for rehabilitation to MBR. The total biological capacity at Henriksdal WWTP will be low and therefore the MBR-line will be operated above its design values. Tanks for new process chemicals will be constructed in a later phase of the project, thus external carbon source and PIX will not be available in phase 2. In addition, the new dewatering building at Henriksdal will not be taken into operation during phase 2 of the SFA-project, thus the reject water from dewatering of digested sludge will continue to be added to the inflow in Sickla and not added to the RAS-deox zone in the MBR-line as intended. The aim of the trial was to test if the treatment process could meet the current and future effluent requirements under these conditions and to study the effect on the membrane operation.

The RAS-deox zone has multiple functions, the most important ones are to remove the oxygen content in the sludge before recirculation to the pre-denitrification zone and to nitrify the ammonia from the reject water that is added to the zone using oxygen from membrane aeration. It has been proven previously that oxygen levels decrease in the RAS-deox zone but it is not clarified if it was removed by nitrification or other microbial activities. Therefore, the function of the RAS-deox zone was evaluated during autumn.

In order to determine the emission of greenhouse gasses from the membrane process, two measurement campaigns for nitrous oxide emissions carried out during 2018.

During 2017 it was noticed that the phosphorus removal and the precipitation chemical consumption did not add up and it was discovered that enhanced biological phosphorus removal (EBPR) mysteriously appeared in the process. Therefore, in 2018 the EBPR activity was monitored regularly and, in the summer, the iron dosage was completely stopped for three months in order to evaluate the dynamics of iron in sludge, effluent phosphorus and EBPR activity.

Optimisation of resource consumption has continuously been in the spotlight during 2018. Especially regarding the carbon source addition, chemicals used for membrane cleaning (focus this year was to minimize the consumption of oxalic acid), and energy for membrane aeration.

Throughout 2018 the sludge treatment line (including sludge thickening, anaerobic digestion and sludge dewatering) was in focus. The goal was to obtain steady operation at mesophilic conditions – a "reference period", and to evaluate the thickener and its function. However, the operation was characterized by technical problems and several minor adjustments and reconstructions were done in order to make it function over time. The optimization of the thickening step included testing different polymer doses and dosing points and a comparison of thickening MBR sludge and waste activated sludge from Henriksdal.

A two years long study on mapping of micro pollutants through the treatment process, such as pharmaceutical residues, micro plastics, bacteria, PFAS and chloro-organic halogens was started during autumn 2017. In 2018, the second sampling campaign (out of a total of four planned campaigns) was carried out and the results are presented in this report.

5 Method

5.1 Sampling and analyses

Eurofins Environment Sweden AB (Lidköping) conducted analyses of water samples from five different sampling points: IN (influent wastewater), PTW (primary treated water), EFF (effluent water), activated sludge from bioreactor BR4 (SLUDGE 1) and return sludge from RAS-DeOx (SLUDGE 2), and analyses of sludge samples from three different sampling points: PS (primary sludge), WAS (waste activated sludge) and DS (digested and dewatered sludge).

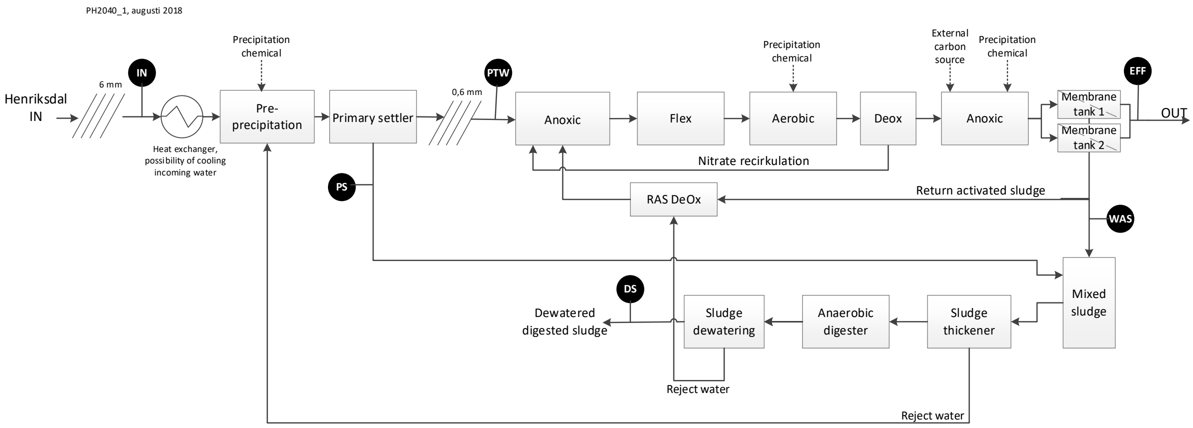


Figure 10. Sampling points in pilot process marked as black circles (SLUDGE1 and SLUDGE2 sampling points not included in figure).

Three different sampling types were used: daily composite samples, weekly composite samples and grab samples. Daily samples were taken with automatic samplers set for flow proportional sampling. Weekly samples were mixed from the daily samples proportionally to the mean flow during the respective days. Grab samples were an instantaneous sample taken from the respective tank. The weekly composite samples were conserved with 1 part 4M sulfuric acid to 100 parts sample volume, except for the samples analysed for TOC which were conserved with 2M hydrochloric acid in corresponding proportions.

Table 5 lists the parameters analysed at the accredited laboratory for the respective sampling points and sample types. One portion of the grab sample of sludge from the RAS-DeOx which was sent to accredited laboratory (Eurofins), was used to measure sludge volume (SVI) and time to filter (TTF) at IVL's internal laboratory at Hammarby Sjöstadsverk.

Table 5. Sampling points, parameters and number of samples sent per week for external analyses.

	Parameters												
Sampling point	TOC	COD	BOD_7	TP	PO_4-P	SS	VSS	cTOC	NH_4-N	NO_3-N + NO_2-N	TN	Fe (digested)	P (digested)
Daily composite samples													
IN	1		1	1	1	1	1					1	
PTW	1		1	1	1	1	1					1	
EFF	1			1	1	1	1					1	
Grab samples													
RAS-DeOx						1	1	1				1	1
Reject water mixed sludge thickening						1	1						
Reject water digested sludge dewatering			1	1	1	1	1		1		1	1	
Weekly composite samples													
IN	1	1		1					1	1	1	1	
PTW	1	1		1					1	1	1	1	
EFF	1			1					1	1	1	1	
Total number	6	2	3	7	4	6	7	1	4	3	4	8	1

In addition to the samples and analyses presented in Table 5, a monthly composite sample of dewatered digested sludge (DDMS) was sent to external accredited laboratory for analysis of TS, VS, pH, nitrogen, phosphorus, chlorine, and 15 different metals. Multiple organic parameters and three more metals were analysed each quarter, including Polybrominated diphenyl ethers (PBDE, 24), Triclosan, Polychlorinated biphenyls (PCB, 7), Polycyclic aromatic hydrocarbons (PAH, 6), organotin compounds (10), Phenols (19), Perfluorooctanoic acid (PFOA), Perfluorooctanesulfonic acid (PFOS) and Per- and polyfluoroalkyl substances (PFAS).

In addition to the external analyses, analyses were also performed internally at IVL's laboratory at Hammarby Sjöstadsverk. Water phase samples were analysed by means of colorimetric methods using a spectrophotometer (WTW photolab 6600) and standard cuvette tests. The daily composite samples were analysed according to Table 6. Additional analyses of daily composite samples or grab samples were also done in order to calibrate process instruments.

Table 6. Internal analyses on daily composite samples from effluent water samples.

	Weekday		
Analysis	Monday	Wednesday	Friday
EFF COD		X	
EFF NH_4-N		X	
EFF NO_3-N	X	X	X
EFF TN		X	
EFF PO_4-P	X	X	X
EFF TP		X	

Sludge phase samples were analysed regarding total solids (TS (%)) and volatile solids (VS (%)) between 2-3 times per week. This applies to all different sludges; primary sludge, waste activated sludge, mixed sludge, thickened mixed sludge, digested sludge and dewatered and digested sludge. The reject water from sludge thickening and sludge dewatering was internally analysed at the Hammarby Sjöstadsverk´s laboratory with the same approximate frequency regarding total suspended solids (mg/L). To monitor the digestion process, a sample from the digester was taken once per week and pH, VFA, alkalinity and ammonium were analysed. Measurements of methane, carbon dioxide and hydrogen sulphide in the produced biogas was conducted weekly with a portable gas meter.

5.2 Online measurements

The process was controlled and/or monitored with a number of online sensors installed in the treatment line. Dynamic values from online measurements supplemented information from the analysis results and were used for continuous follow-up and control of the process. A summary of the most important online measurements is shown in Table 7 and Table 8. In addition to online sensors, there was also an on-line analyser for PO_4-P sampling from the effluent.

Table 7. Placement of on-line sensors in the water treatment line.

Placement	Parameter	Function
General	Flowrate (water)	Measure inflow, permeate flow and all recirculation streams
IN	Temperature	Measure the incoming wastewater temperature. Sometimes used for control
IN	Flowrate (water)	Measure the influent water
IN	SS	Monitor influent suspended solids
PTW	NH_4-N	Measure incoming ammonium
BR1	DO	Monitor Dissolved Oxygen
BR2	DO	Monitor Dissolved Oxygen
BR2	NH_4-N	Measure ammonium in to aerated part of biological treatment. Sometimes used for control
BR3	DO	Controlling Dissolved Oxygen
BR3	Flowrate (air)	Measure air consumption
BR4	DO	Controlling Dissolved Oxygen
BR4	Flowrate (air)	Measure air consumption
BR4	SS	Measure suspended solids
BR5	DO	Controlling Dissolved Oxygen
BR5	Flowrate (air)	Measure air consumption
BR5	NO_3-N	Measure nitrate, monitor function of post-denitrification
BR6	NO_3-N	Measure nitrate, control dosage of external carbon
BR6	pH	Measure pH in the biological treatment
MT1/MT2	Temperature	Measure temperature in membrane tank (x2)
MT1/MT2	DO	Measure Dissolved Oxygen in membrane tank (x2)
MT1/MT2	Pressure	Level and pressure measurements for calculation of TMP (4 sensors)
MT1/MT2	Flowrate (water)	Effluent of permeate from membrane 1 and 2 (x2)
MT1/MT2	Flowrate (air)	Measure air consumption (x2)
RAS-DeOx	SS	Measure suspended solids
RAS-DeOx	DO	Monitoring Dissolved Oxygen
RAS-DeOx	NH_4-N	Measure ammonium concentration (after addition of reject water)
EFF	PO_4-P	Measure effluent phosphate concentration and control dosage of precipitation chemicals
EFF	NO_3-N	Measure effluent nitrate

Table 8. Placement of on-line sensors in sludge treatment line.

Placement	Parameter	Function
PS	Flowrate	Measure flowrate of primary sludge
PS	TS	Measure total solids in primary sludge
WAS	Flowrate	Measure flowrate of waste activated sludge
MS	TS	Measure total solids in mixed sludge, used to control dosage of polymer to sludge entering the thickener
TMS	TS	Measure total solids in thickened mixed sludge
AD	Temperature	Monitor temperature in anaerobic digester, used to control heating of sludge
AD	Level	Measure the level in the anaerobic digester, used to test variable volumes
AD	Pressure	Measure the pressure of the gas
AD	pH	Monitor pH in the anaerobic digester
AD	Gas flow	Measure biogas production
DMS	TS	Measure total solids in digested mixed sludge, used to control dosage of polymer to sludge entering the dewatering

5.3 Evaluation parameters

5.3.1 Membrane performance

The membranes were evaluated using several parameters described in this section.

As the membranes are operated in cycles with 10 minutes of permeate withdrawal and 1 minute relaxation, the membrane performance parameters can be calculated as *gross* values (using only data from the 10 minutes of actual permeate withdrawal) or as *net* values (using average data from the full operation cycle, permeation and relaxation = 11 minutes). The gross values are higher than the net values, however the net values corresponds better to the average operation. All values for the parameters described below are given as net values in this report.

1) Flux: Flowrate per membrane area, unit L/(m^2·h). The flux is describing the hydraulic load on the membranes. Flux is calculated as permeate flow divided by membrane area.

2) TMP: Transmembrane pressure, unit mbar. The difference in pressure before and after the membranes. TMP is the driving force for transportation through the membrane. TMP is measured using online pressure transmitters in the membrane tank and on the permeate pipe.

3) Permeability: Flux per TMP, unit L/(m^2·h·bar). Permeability is a measurement of how well a certain flux is withdrawn through the membranes. The permeability is gradually decreasing with time due to fouling.

The permeability is affected by the temperature. Because of this, temperature compensated permeability (normalised to a standard temperature of 20 °C) was used for evaluation. The normalisation equation is shown below and was provided by the membrane supplier.

$$Normalised\ permeability\left[\frac{L}{m^2 \cdot h \cdot bar}\ at\ temperature\ 20\ °C\right] = Permeability\ \cdot \theta^{(20-T)}$$

where

$T = Temperature, \theta = 1.025\ if\ T \geq 20\ °C$ and $\theta = 1.033\ if\ T < 20\ °C$.

5.3.2 Sludge quality

In addition to the parameters analysed at the external and internal laboratory, listed in Table 5 and Table 6 above, a number of additional analyses were performed on the sludge from the RAS-DeOx. These included sludge volume index (SVI), Time To Filter (TTF), and trash content.

Sludge volume index (SVI)

Sludge volume index were analysed according to APHA's standard method (2005) with dilution of the sludge as described by Svenskt Vatten (2010b).

Time To Filter (TTF)

TTF was analysed according to instructions from the membrane supplier. 25, 50 and 100 mL of the sludge (TTF-25, TTF-50 and TTF-100 respectively) was filtered through 1.5 micron filter (particle retention 1.5 µm) and the filtration time was noted. The filtrate and the permeate were sent to the external laboratory for analysis with respect to TOC (mg/L) to evaluate the amount of colloidal TOC in permeate, that is, the difference between TOC in the filtrate after TTF and the TOC in the permeate. According to the membrane supplier, the concentration of colloidal TOC (cTOC) should be less than 10 mg/L.

Trash content

The method for Trash content is described in detail in last year's report (Andersson *et al.* 2017). In short, the sludge is filtered through sievs with different slot width and the amount of trash captured in the sieves is measured. This analysis was carried out once every month in order to assure that particles larger than 2 mm, which could harm the membranes, would not accumulate in the treatment line. For a well-functioning process, the amount of trash content in the sludge, at a sieve size of 2 mm, should not exceed 2 mg/L (information from the membrane supplier).

6 Results and discussion

6.1 Primary treatment

A summary of the most relevant trials related to the primary treatment is presented in Table 9 and includes the trial of imitating the first phase operation of the full-scale plant and a trial operating without any precipitation chemicals.

Table 9. Trials related to primary treatment.

Trial	Jan	Feb	Mar	Apr	May	Jun	Jul	Aug	Sep	Oct	Nov	Dec
Imitation of first phase operation SFA												
Trial with no Fe dosage												

6.1.1 Inlet screen

During 2018 a 6 mm punch hole screen was used as the first treatment step except in December when it was exchanged to a 3 mm punch hole screen. This was made to avoid clogging of the inlet heat exchanger. Previously even finer screens have been used. The change to a coarser screen was done in order to increase the SS concentration and the particle size into the primary settler and thereby obtain increased primary sludge production, which is required for a representative operation of the sludge pilot. Figure 11 shows that the SS concentration increased slightly after installing the 6 mm screen and the SS concentration in the inlet to the pilot was similar as the concentration in the inlet to Henriksdal WWTP.

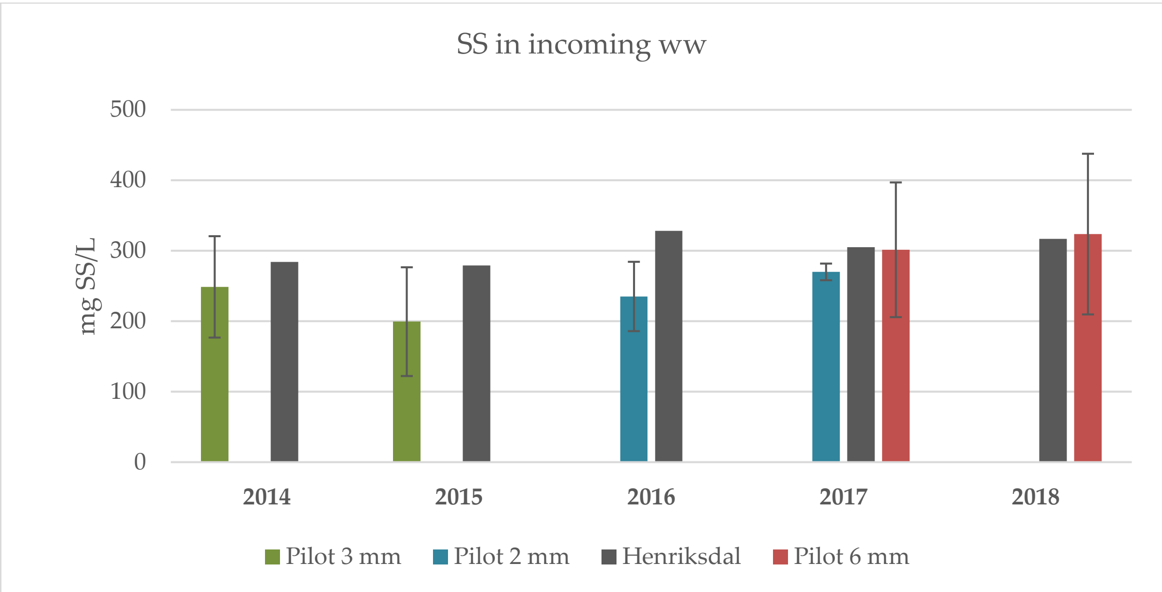


Figure 11. Incoming SS concentration to the pilot line after passing through 2, 3 or 6 mm punch hole screen and to Henriksdal (about half of the flow not screened, half screened through 3 mm step-screen).

6.1.2 Efficiency of primary settler

The primary settling volume (3.3 m³) and capacity is smaller than it should be (design value 4.5 m³, see section 3.1.2) which resulted in poor reduction and insufficient primary sludge production compared to present and future Henriksdal WWTP. The reduction rate over the primary settler is showed in Table 10 and Figure 12 below. The PS production increased between 2017 and 2018 which is a result of the reconstruction of the primary settler that was done in 2017 but it did not reach 17.2 kg/d which corresponds to the design value for future Henriksdal WWTP.

Table 10. Reduction over the primary settler and primary sludge production

Parameter	2018	2017	2018: Henriksdal WWTP	Design future Henriksdal WWTP
SS (%)	35%	37%	53%	60%
BOD_7 (%)	25%	30%	48%	50%
TP (%)	10%	14%	30%	40%
TN (%)	1%	4%	9%	10%
TOC (%)	18%	17%	29%	-
PS production (kg/d)	16.1	13.1	82 000	115 000

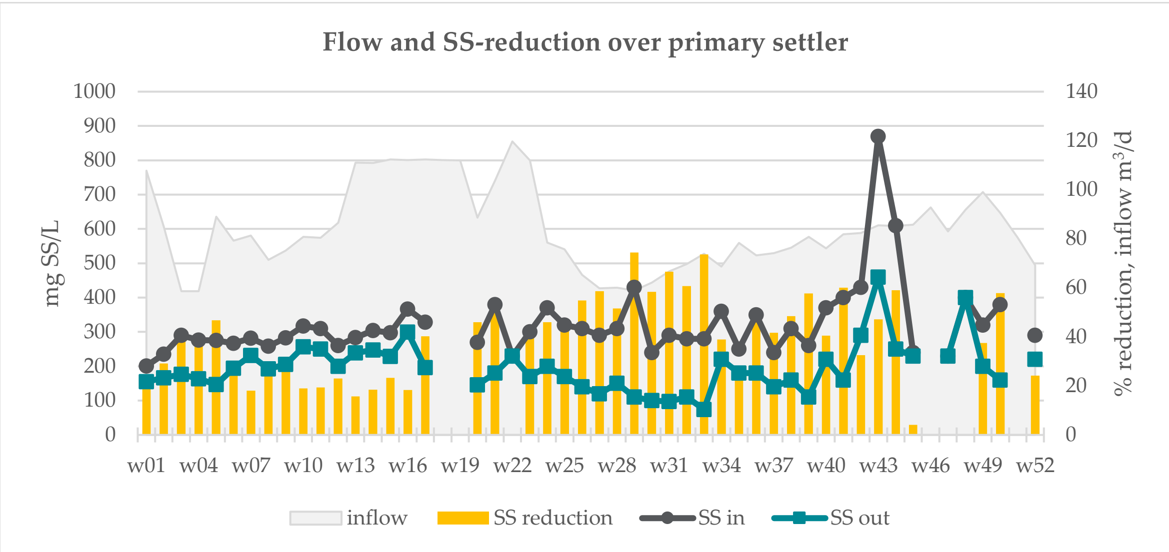


Figure 12. Reduction of SS over primary settler.

6.1.3 Screen and sieve – effect on trash content

During 2018 the pilot line was operated with a 6 mm punch hole inlet screen (except for December, see 6.1.1 Inlet screen) and a 0.6 mm mesh fine sieve before the biological treatment. Over the years different screen/sieve configurations have been used. In order to monitor the amount of particles, fibres and hair that accumulates in the activated sludge with potential to cause problems in the membrane tanks, analysis of trash content (see 5.3.2 Sludge quality) was made on activated sludge once a month. Results are presented in Table 11.

The results have been consistent since the current screen/sieve configuration was implemented. In addition, visual inspection of the membrane cassettes (see 6.5.4 Membrane autopsy) show very little build-up of trash indicating that the measured values are good. The 2017 data after changing to 6 mm fine sceen and installing the 0.6 mm fine sieve is presented as one average from March to

December and one average from May to December. The exclusion of the first two months of data gives about half as much trash content caught by the 1 mm sieve. This can be explained by the fact that it takes at least 3 sludge retention times to exchange the sludge filled with larger particles from the previous period. The trash content measurements from 2018 show similar values as to the later part of 2017. The finest trash content (1 mm) were slightly reduced compared to 2017.

Table 11. Trash content in waste activated sludge (WAS) with various screen/sieve-configurations.

Screens/sieves and hole size	Date	Trash content - 1 mm sieve mg/L	Trash content - 2 mm sieve mg/L
3 mm fine screen at inlet pump	Dec 2013	11.6 ± 5.4	1.0 ± 0.7
2 mm fine screen at inlet pump	Nov 2016 – Feb 2017	6.4 ± 2.4	1.1 ± 0.7
6 mm fine screen at inlet pump and 0.6 mm fine sieve before biology	Mar 2017 → Dec 2017	4.1 ± 3.8	0.6 ± 0.3
6 mm fine screen at inlet pump and 0.6 mm fine sieve before biology	May 2017 → Dec 2017	2.2 ± 1.6	0.6 ± 0.3
6 mm fine screen at inlet pump and 0.6 mm fine sieve before biology	Jan 2018 → Nov 2018	2.0 ± 1.1	0.6 ± 0.5

6.1.4 Pre-treated wastewater

The quality of the pre-treated wastewater (PTW) is presented in Table 11. The concentrations measured in the pilot were higher than the corresponding concentrations measured in the Henriksdal WWTP for SS, BOD_7 and TN. This is mainly due to the poor performance of the primary settler in the pilot. The concentrations were also higher than the design values for the future plant.

The concentration difference will affect the biological treatment including WAS production (and thereby SRT and the amount of phosphorus assimilated in sludge), pre-denitrification capacity and simultaneous precipitation. It can be noted that the iron dosage in the primary settler was low in the pilot compared to the full-scale and the future Henriksdal. This is due to enhanced biological phosphorus removal (EBPR) which is described in chapter Table 12.

Table 12. Data on pre-treated wastewater (PTW) from the pilot compared to data from Henriksdal 2018 and the design data for the future Henriksdal WWTP.

Parameter	Value Pilot 2018	Value Henriksdal 2018	Design future Henriksdal	Design H-dal/ Value Pilot
Pre-treated wastewater (PTW) – into biological treatment				
SS (mg/L)	200 ± 74	186	113	0.6
BOD_7 (mg/L)	197 ± 45	148	108	0.5
TN (mg/L)*	46 ± 8	41	33	0.7
TP (mg/L)	5.3 ± 0.9	6.7	3.0	0.6
Fe (mg/L)	7.3 ± 4.9	16	12	1.6
BOD_7/TN (mg/mg)	4.3	4.2	3.3	0.8

*At Henriksdal the reject water from dewatering of digested sludge is added to the inlet while in the pilot and future Henriksdal it will be added to the RAS-deox zone in the biological treatment.

6.2 Nitrogen removal

A summary of the trials related to the nitrogen removal is presented in Table 13. The trial of imitating the first phase operation of the full-scale plant affected the nitrogen removal as the load was high and no external carbon source was used. In addition, a measurement campaign for nitrous oxide emissions from the process was conducted and a separate evaluation of the function of the RAS-DeOx zone was carried out.

Table 13. Trials related to the nitrogen removal.

Trial	Jan	Feb	Mar	Apr	May	Jun	Jul	Aug	Sep	Oct	Nov	Dec
Imitation of first phase operation SFA												
Methanol used as carbon source												
Nitrous oxide emission measurements												
RAS-DeOx evaluation												

Nitrogen concentrations in the incoming water to the biological treatment (PTW, primary treated water) and in the effluent are presented in Table 14. On average the effluent total nitrogen concentration was 4.6 mg/L and 3 out of 52 weekly composite samples was above the limit concentration of 6 mg N/L. The reduction of total nitrogen (measured in primary treated water) including reject water was 90.1%.

Table 14. Nitrogen concentrations in primary treated water (PTW) and effluent (permeate) during 2018.

Parameter	Limit	Average	Min	Max	No. of weekly samples
TN PTW (mg/L)	-	46	30	67	52
TN EFF (mg/L)	6	4.6	2.8	6.6	52

Effluent nitrogen concentrations as weekly composite samples are presented in Figure 13. From week 10 to week 22 the load was high and no external carbon source was used (see description of trial imitating the first full scale treatment line operation in section6.6. During this period effluent TN was close to 6 mg/L while ammonium concentrations were low. The increase in TN until week 41 was mainly related to reduced nitrification due to limited aeration capacity in the biology.

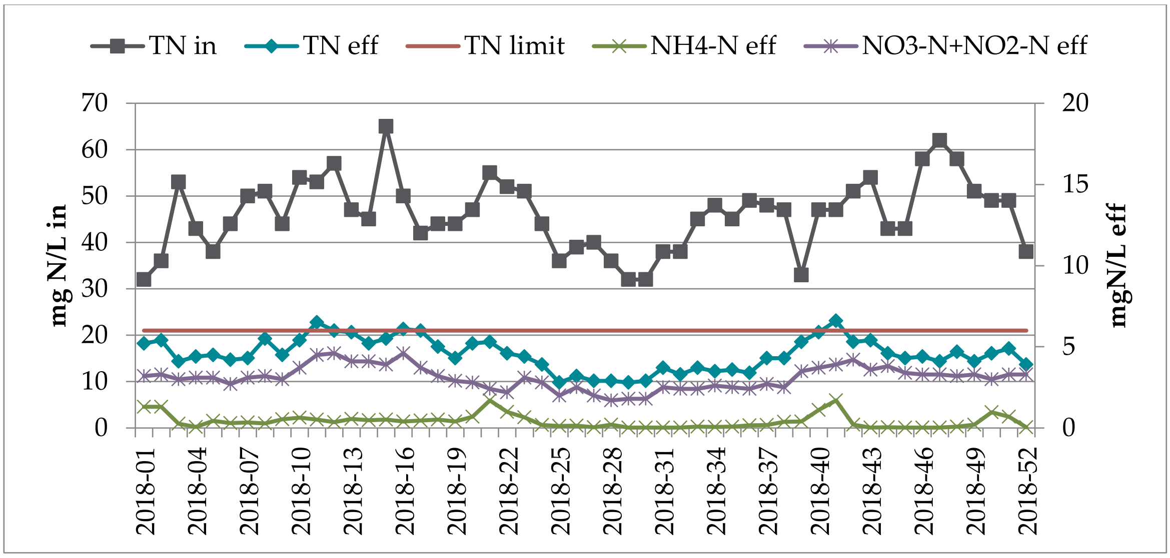


Figure 13. Incoming and effluent nitrogen concentrations from analysis of weekly composite samples. Limit for effluent total nitrogen was set to 6 mg N/L.

In Table 15, key values for the nitrogen removal in the pilot are presented and compared to the design value for the future Henriksdal WWTP. The amount of removed total nitrogen was 15% higher compared to the design. The external carbon source used throughout this year was methanol, however, it should be noted that the external carbon dosage was not in use for 26 weeks (from week 10 to week 36) due to separate trials where the biological treatment was operated without use of external carbon source. On yearly average the methanol consumption was 4.8 g COD/m^3 (148 kg COD/year). If excluding week 10 to week 36 when the methanol dosage was deliberately off, the dosage was on average 10 g COD/m^3 corresponding to about 297 kg COD/year, or 45% of SFA design. Although yearly effluent nitrogen concentrations were low with low methanol dosage, the dosage is not comparable with the dosage required if effluent nitrate should have been controlled throughout the year. The yearly average of nitrogen removal in the pilot was slightly lower this year (3.6 kg N/d) compared to last year (3.7 kg N/d) but the water temperature this year was about 1.7 degrees higher because the heat exchangers on the inlet water was out of operation.

The nitrogen removal rate presented in Table 15 was lower in the pilot which comes from the fact that the VSS content in the sludge was 76% while it was estimated to 63% in the design. The parameter that stood out the most was the aeration of the biology which was almost 6 times higher than the design. One reason for a higher value is that the basins in the pilot were about 4 times less deep than the design. The aeration of the biology is about the same as previous year (2017) when it was 54 m^3/h. From comparing aeration in biology with the first year when membrane aeration was higher, reduced aeration of the membranes result in a higher demand for oxygen in the biology which indicates that some nitrification occurred in the membrane tanks.

Table 15. Comparison of parameters related to the nitrogen removal between operational data from the pilot and the SFA design.

Parameter	Unit	Value Pilot	SFA design	Value pilot / Scaled[vi] SFA design
Removed nitrogen (including reject water)	kg N/d	3.6	21 000	115%
Nitrogen removal rate	g N/kg VSS, d	19.6	22	89%
Aerated sludge age (including membrane tanks)	d	7.5	9,4[vii]	79%
Air consumption biology (activated sludge)	Nm³/h	52.4[viii]	62 000	566%
Specific oxygen demand (SOTR)	kg O_2/d	60[ix]	260 000	155%
Consumption of external carbon	kg COD/d	0.8[x]	12 000	45%

[vi] The value of SFA-design divided by the scale factor 6 700
[vii] Assumed that ¾ of all membrane tanks are in operation as a yearly average.
[viii] m³/h not Nm³/h
[ix] SOTR was calculated from the measured airflow and a water depth (aerator surface to water surface) of 3,19 m and a specific oxygenation capacity of 0,015 kg O_2/Nm³, m.
[x] Excluding week 10 to week 36.

6.2.1 Nitrification

The total load of ammonium nitrogen (including reject water) compared to the reduction and related to total aeration (including the membrane tanks) is presented in Figure 14. The aeration trend follows the load although more aeration was used per NH_4-N reduced in the beginning of the year and during the summer period.

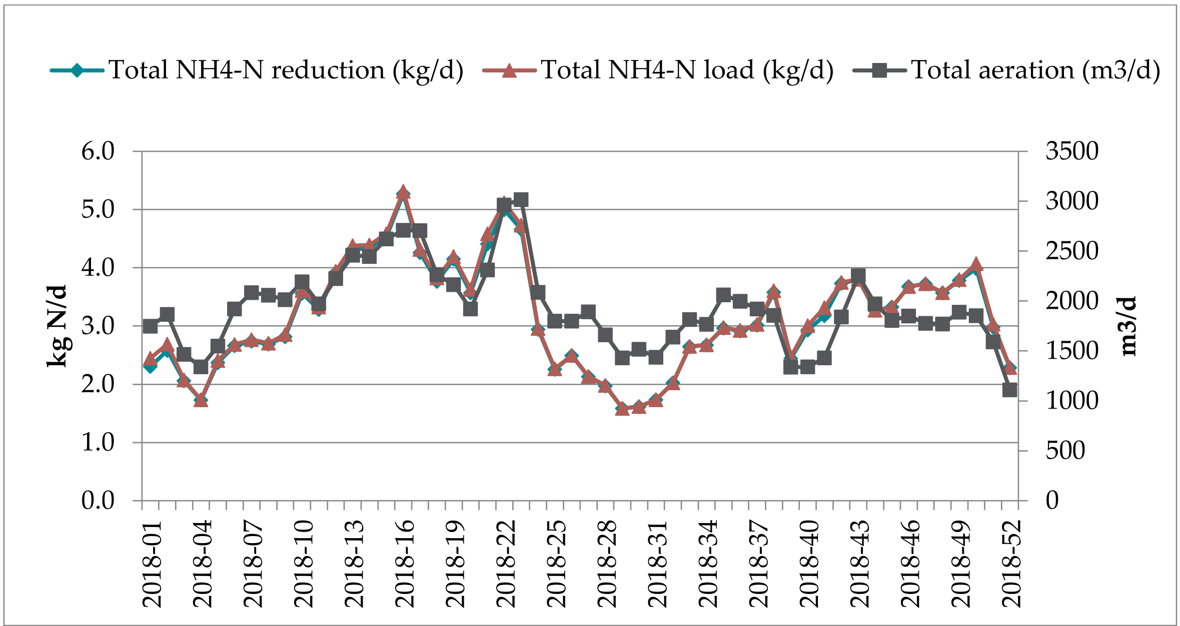


Figure 14. The total load and reduction of NH_4-N together with total aeration of biology and membrane tanks. Note that the load and reduction of NH_4-N is almost the same.

When comparing the aeration of the biology with aeration of the membrane tanks (Figure 15) it can be observed that the membranes most of the time was operated at the lower aeration level (Leap-Lo corresponding to 14 m³/h each, 28 m³/h in total) with only a few peaks above 28 m³/h. The aeration of the biology varied as weekly average between 27 and 87 m³/h and was on average 52.4 m³/h.

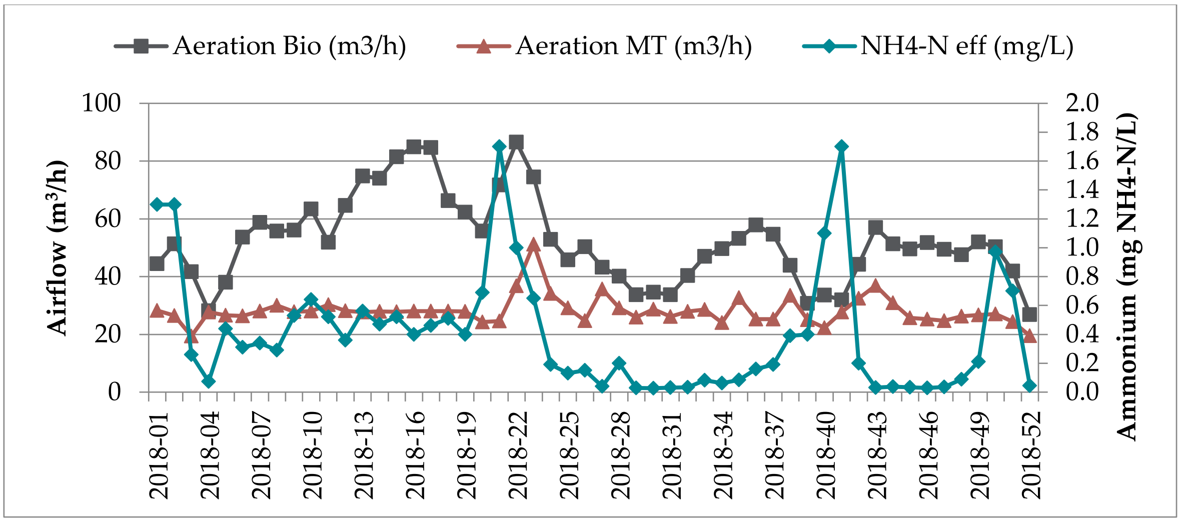


Figure 15. Aeration need in biology and membrane tanks (MT) together with effluent NH_4-N.

On average, the aeration of the membranes accounted for 36% of the total aeration, same as last year (2017). This is a significant reduction compared to the first year (2016) when 54% of total aeration was used for the membranes. In week 47 in 2016 fouling control was implemented for the membrane aeration. Fouling control is part of the membrane suppliers' strategy to optimize the air consumption for cleaning of the membranes, and it controls how much air to be used in the membrane tanks. With fouling control, the air flow set point to the membrane tank automatically switches between two air flow rates based on how fouled the membranes are (higher aeration when indication of fouling).

The initial aim for the pilot was to operate at a total sludge age of 25 days. However, as the membrane supplier terms states that the membranes should not be operated in sludge concentration above 10 000 mg SS/L for longer periods, the sludge age has not been controlled. Waste activated sludge (WAS) flowrate has been manually adjusted to keep the suspended solids concentrations in the RAS-DeOx normally between 8 000 and 10 000 mg SS/L, and during special trials at higher concentrations to study fouling of the membranes.

Total and aerated (including membrane tanks) sludge age is presented in Figure 16 together with suspended solids concentration in RAS as well as the WAS flowrate. The calculations of sludge age are uncertain because of foaming in the aerated bioreactors leading to overflow and loss of sludge not accounted for in the calculations. The sludge concentration decreased during June and July and by the end of July the WAS pump was stopped for about one week to increase the sludge concentration in the system. The low WAS flow resulted in a high peak in the sludge age during summer (July to September). The total sludge age was on average 24 days and the aerated sludge age (including membrane tanks) was 7.5 days.

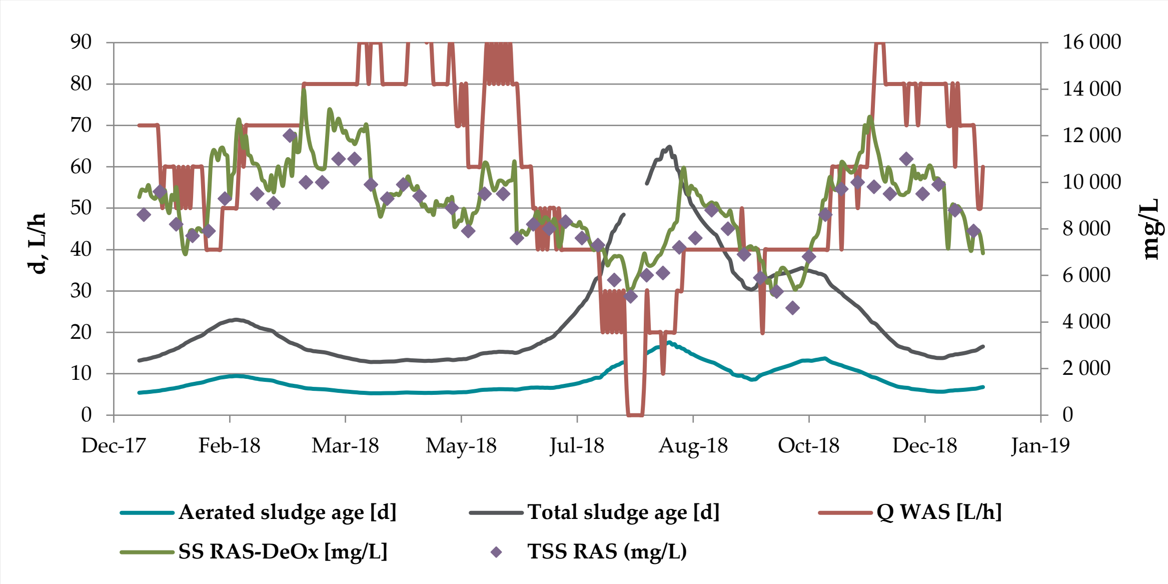


Figure 16. Total and aerated (incl. MTs) sludge age (moving average one month back in time) together with WAS flowrate and online SS in the RAS DeOx as well as lab analysis of TSS on grab samples from the RAS-DeOx.

The aeration of the biology has been problematic this year due to excessive foaming in the bioreactors and some equipment failures. Previously, mainly dissolved oxygen (DO) control using fixed setpoints have been used. In 2017 the first aerated bioreactor (BR3) was intermittently aerated to a fixed DO set point based on a threshold value on ammonium entering the aerated part of the biological treatment (measured in BR2). In the beginning of 2018 ammonium feedback control was implemented to control the DO setpoint for BR3 and BR4. DO setpoints could vary between 1 mg/L and 3.5 mg/L to reach setpoint of 2 mg NH_4-N/L in BR5. The aim of this control strategy was to save aeration in the biology, especially during night when the load was low, by allowing a lower DO concentration. As can be seen in Figure 17, the DO concentrations in BR3 and BR4 show a large variation from week 3, compared to previous control strategy.

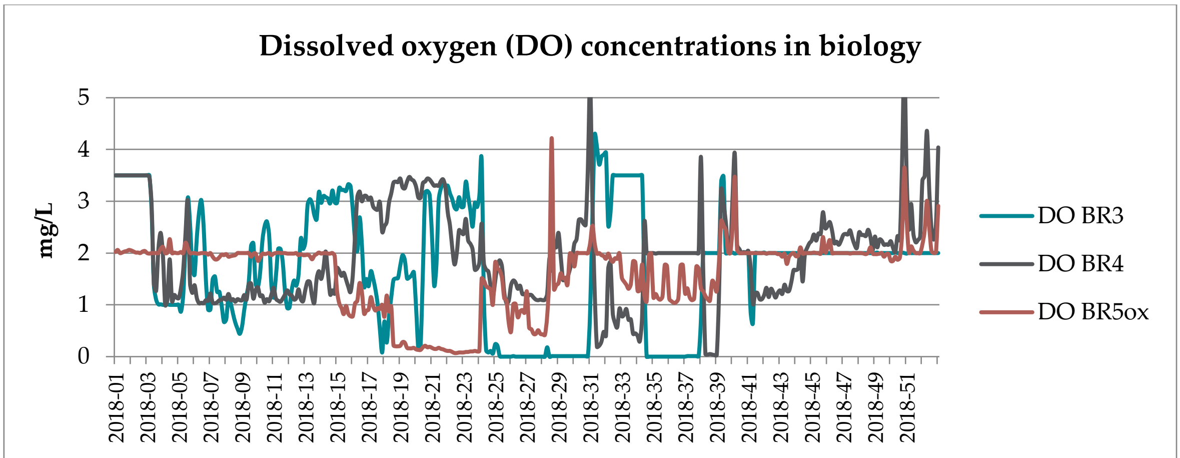


Figure 17. Dissolved oxygen (DO) concentrations in the three aerated bioreactors (BR3, BR4 and BR5ox).

BR5 was initially aerated to keep a DO concentration of 2 mg/L. In week 10 the load was increased which also increased the need for aeration and problems with foaming in the aerated reactors started in week 11. In week 15 the aeration in BR5 could not maintain DO concentration 2 mg/L and during that period there were severe problems with foaming in BR4. In week 18 the aeration in

BR5 was set to a low fixed flowrate as the aeration control was not functioning. BR4 and BR5 were drained for inspection in week 24 and it was found that the wall separating the aerated compartment and the deox-compartment in BR5 had collapsed, blocking the diffusers in BR5. Once the wall was fixed, aeration could be started again in BR4 and BR5. At the same time, due to warmer temperatures and less need for aeration, BR3 was used as anoxic zone.

As part of an EU funded research project (R3Water) a different aeration equipment was tested since 2016 in BR4. This aeration equipment is based on so called oscillating aeration with the target of getting more efficient oxygen transfer from air to water. Week 34, after some tubing problems in aeration equipment in BR4, the test equipment was removed, and the standard aeration equipment was put in operation. After this, aeration of BR4 was started and another attempt to use ammonium feedback control was made, this time only in BR4. However, the foaming and overflow of sludge occurred as soon as aeration was started. Aeration in BR3 was started in week 31 with at fixed airflow to evaluate if this could help the problems with foaming and overflow in BR4. Aeration of BR4 had to be either manually controlled with fixed flowrate or completely stopped in the period from week 29 to week 41 to avoid too much overflow of foaming sludge.

Because of the problems with foaming, unreliable DO sensor readings in the foamy sludge and several aeration equipment failures this year, no deeper evaluation of the aeration control strategies have been made.

6.2.2 Denitrification

For the post denitrification (BR6) external carbon source has been added using different carbon sources and control strategies in previous trials. This year only methanol has been used as carbon source. For a long period this year, the post denitrification was operated without dosage of methanol as to mimic the first years of the full scale treatment line operation (when it will not be possible to dose methanol).

On Tuesday 6th of March (week 10) the methanol dosage was stopped. Until then the dosage had been controlled based on effluent nitrate concentrations with a setpoint of 4 mg NO_3-N/L in the effluent. The setpoint was increased from 3 mg/L to 4 mg/L at the beginning of the year as previous operation showed that the effluent concentration during night was very low and therefore the daily average was always lower than the setpoint. As can be seen in Figure 18, the setpoint of 4 mg/L resulted in an average nitrate concentration of 3 mg/L or less. When methanol dosage was stopped in week 10, the nitrate concentration increased at first and then decreased as temperature increased towards summer. The methanol dosage was started again in week 36, again with setpoint 4 mg NO_3-N/L. Dosage was only needed during short periods each day at the daily peak load which resulted in low average consumption of methanol. In week 45 the dosage point was changed from dosing directly into BR5 to a point in the piping between BR5 and BR6. This was done in order to avoid carbon source recycling back to the pre-denitrification tanks with the nitrate recirculation flow. After this change the required dosage decreased even more.

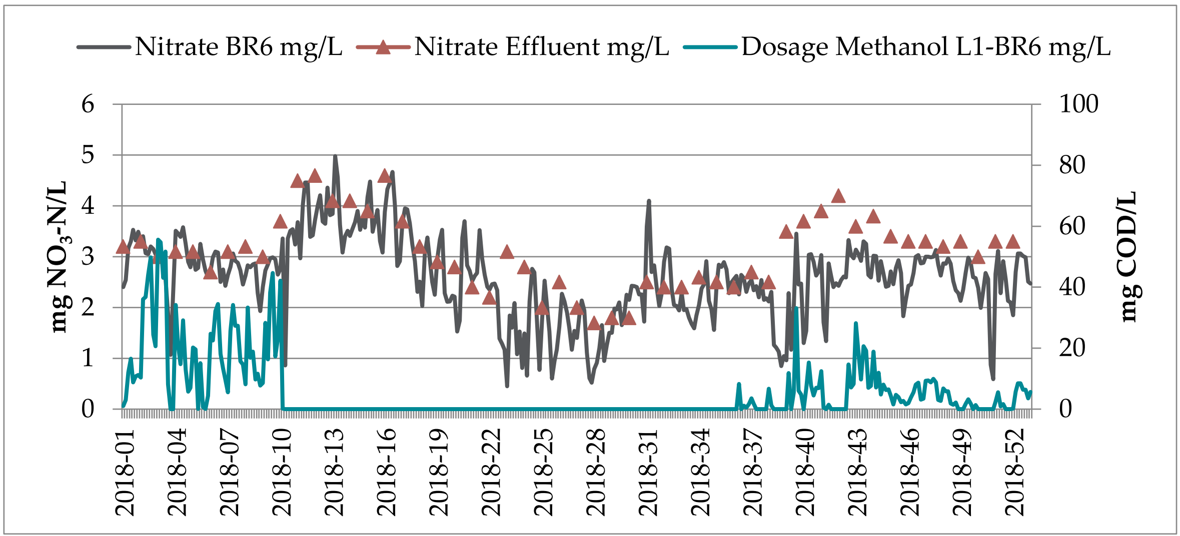


Figure 18. Nitrate in BR6 and dosage of methanol as daily average values and effluent nitrate analysed in weekly composite samples.

The nitrate sensor in BR5deox was taken out of operation in October 2017 and was replaced with a new sensor by the end of February 2018, about at the same time as the methanol dosage was stopped. From the online measurements of nitrate as weekly average data (Figure 19) the effect of starting methanol dosage in week 36 is not clear as the dosage only was active a few hours per day.

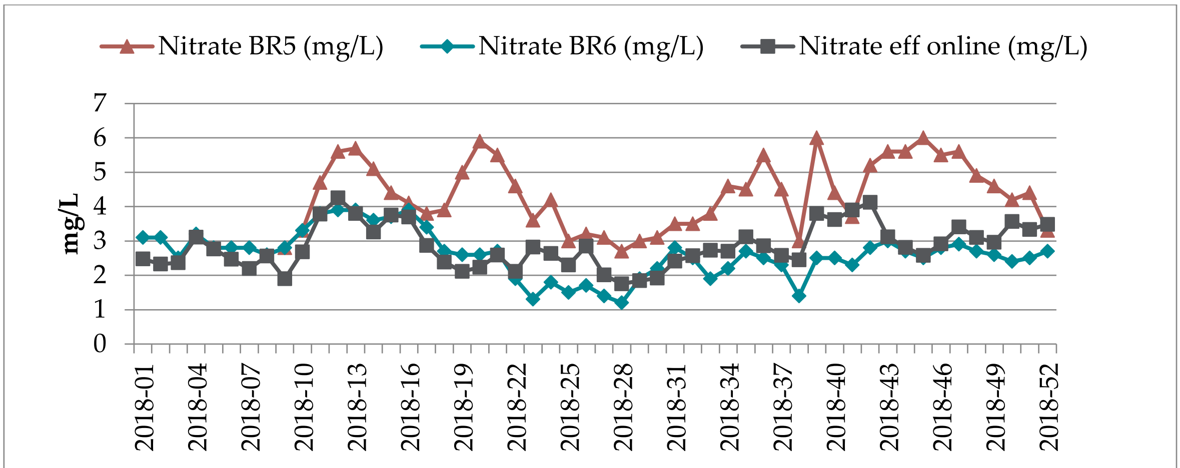


Figure 19. Online nitrate concentrations as weekly average. Methanol dosage was in operation week 1-10, then stopped and started again in week 36.

For the period of week 36 to week 52 2018 the average COD dosage was 4.8 g COD/m^3 influent to the treatment line. This corresponds to 0.82 kg COD per day. In 2017 the methanol consumption was about 30 g COD/m^3 based on online measurements. Some reasons for the large difference in dosage are the lower setpoint for nitrate (3 mg/L) used in 2017, the lower average temperature in 2017 (yearly average was 1.7 degrees lower compared to 2018) and the change in dosage point, first changed from BR6 to BR5deox in 2017 which resulted in increased consumption and then changed to the pipe in-between BR5deox and BR6 which reduced consumption.

6.2.3 Zone for deoxygenation of return sludge - RAS-Deox

This year, a deeper evaluation of what is happening with nitrogen and oxygen in the RAS-Deox zone and the treatment capacity of this zone, was carried out. A full report of the evaluation including a sampling campaign and tests in the pilot as well as process simulations are presented in a MSc project (Taylor, 2019).

The idea of the RAS-Deox zone is to collect the return sludge with a high oxygen concentration from the membrane tanks and provide enough retention time for the oxygen concentrations to decrease before the return sludge is pumped back to the pre-denitrification zone (BR1). As an additional feature, reject water will be added in this zone in order to make use of the excess oxygen to nitrify some of the ammonium in the reject water from dewatering of digested sludge. The RAS-Deox tank is equipped with an oxygen sensor and an ammonium sensor in order to monitor to what extent the oxygenated return sludge could be used to nitrify ammonium in the reject water.

The evaluation looked at different aeration levels of the membrane tanks (Leap-Hi and Leap-Lo) and three different loads of ammonium from reject water to the RAS-Deox zone corresponding to 5, 10 and 15% of influent ammonium load.

The least favorable situation, with high aeration and low reject water addition (5%) resulted in DO peaks reaching a maximum concentration above 2 mg/L. With higher reject water addition, the DO decreased well below 0.2 mg/L (Figure 20).

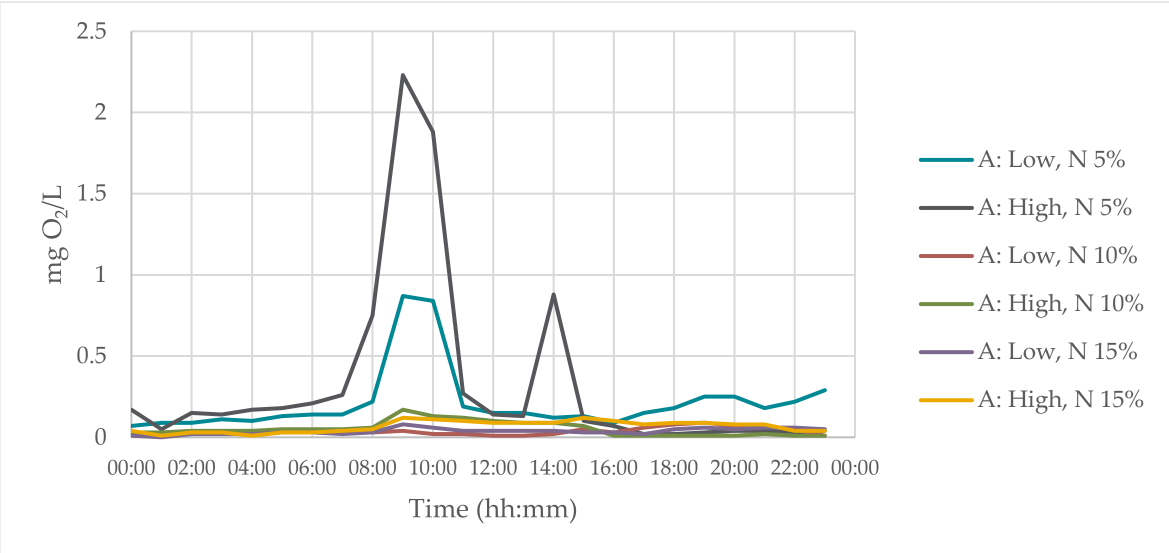


Figure 20. Dissolved oxygen from online sensor in the RAS-DeOx for six days with varying aeration (A) between low and high and ammonium (N) load.

The lower aeration level for the membranes was used most of the time. This means less oxygen entering the RAS-Deox and low DO concentrations in the RAS-Deox could be achieved most of the time even without addition of reject water (Figure 21).

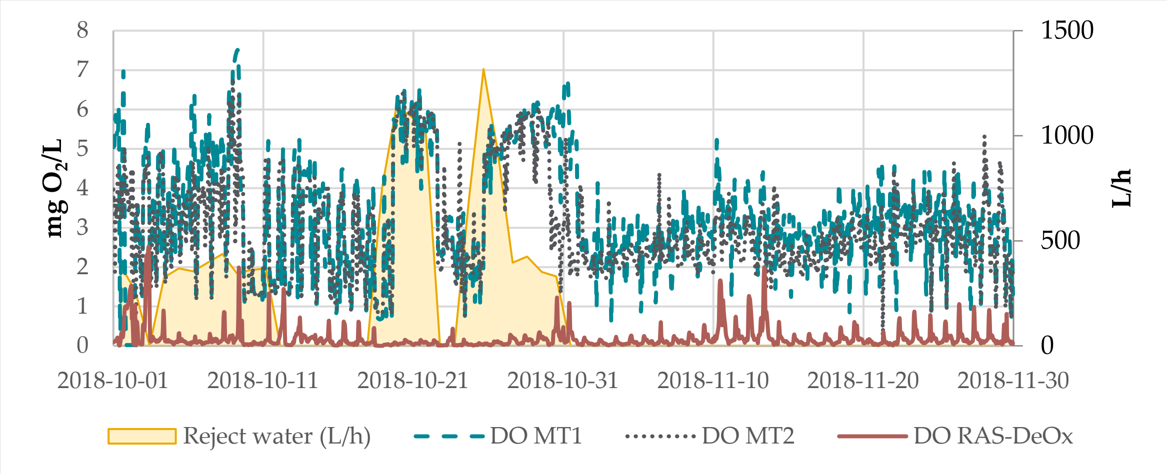


Figure 21. DO concentrations in membrane tanks (MT) and in the RAS-DeOx zone with and without addition of reject water.

Both sampling and simulations showed that there was simultaneous nitrification and denitrification in the zone when reject water was added. From the sampling campaign it was observed that the ammonium reduction varied, however most of the time it was above 60% (Figure 22). This indicates that nitrification occured, however no conclusion or performance indicator could be calculated regarding how much oxygen was used for ammonium conversion. For nitrate (assuming a decrease in NH_4-N result in a corresponding NO_3-N increase) a reduction of about 20 to 50% was observed.

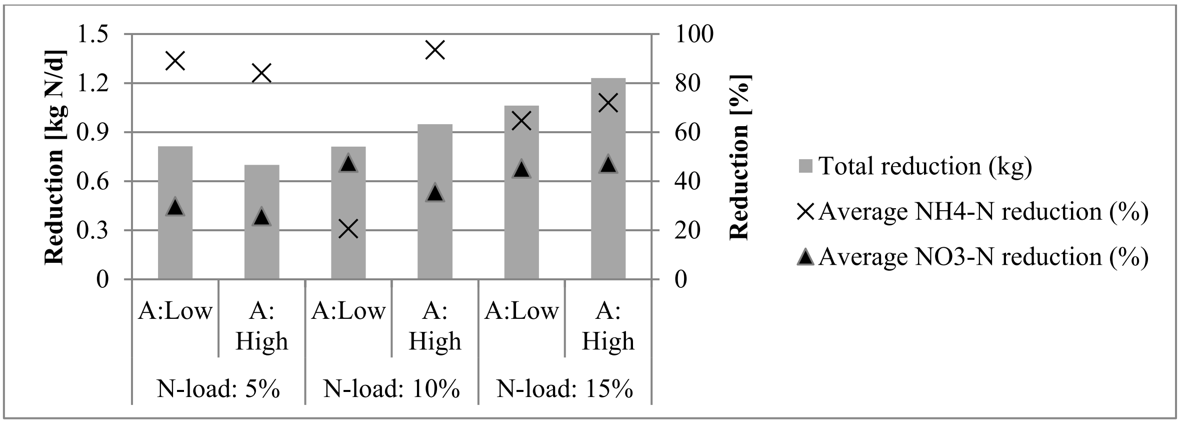


Figure 22. Results from sampling campaign in the RAS-Deox.

When comparing to process simulations (10% N load from reject water) a reduction of ammonium nitrogen of about 50% was observed in the RAS-Deox (concentration from 1 mg NH_4-N/L to 0.5 mg NH_4-N/L) and 10% reduction of nitrate (from 3.57 NO_3-N + 0.5 N from ammonium conversion, to 3.62 mg NO_3-N exiting the RAS-Deox zone).

Based on the DO readings at sampling, a larger decrease in DO was observed when the ammonium load increased (Figure 23). For the first three days the DO was higher in the first sample (at time 09:00) because of lower load. With reject load corresponding to 5% of the inlet load the morning samples indicated that peaks with remaining DO concentrations in the return sludge of 1-2 mg O_2/L will occur. This indicates peaks in return sludge if reject water load is low or if no reject water is added to the RAS-Deox zone. However, when comparing to a longer period (Figure 21) with and without reject water addition, peaks above 1 mg/L in the RAS-Deox zone seldom occur.

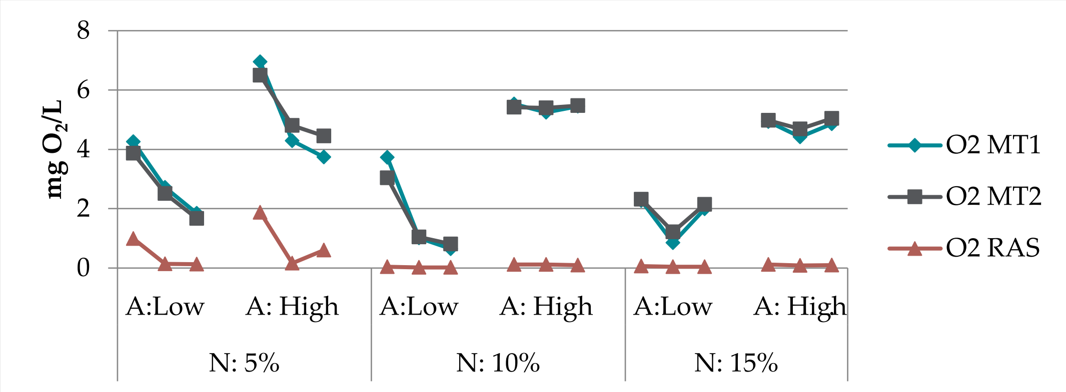


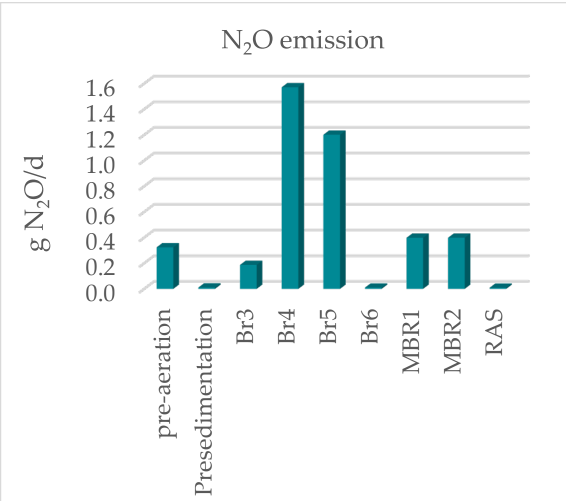
Figure 23. DO concentrations in membrane tanks and RAS-DeOx at the times for sampling.

6.2.4 Greenhouse gas emissions

A measurement campaign of nitrous oxide and methane emissions from the process was carried out from 26th of April to 6th of May. Process air from each reactor was collected and analysed with an on-line instrument (Fresenius, GA2020) which could measure in six points at the same time. The campaign was divided into two parts, first measurements were carried out in Pre-aeration, Primary clarifier, BR3, BR4, BR5 and BR6 for one week (26th of April to 3rd of May), followed by measurements in MT1, MT2 and RAS-Deox for one day (5th to 6th of May).

As a previous measurement campaign in 2014 indicated that the main part of the emissions occurred in the aerated zones, these zones were in focus for this campaign.

Preliminary results showed that BR4 and BR5 were the zones where most of the nitrous oxide was emitted (Figure 24). Approximately 20% of the total nitrous oxide emissions came from the membrane tanks. Most of the methane emissions originated from the Pre-aeration and BR4.



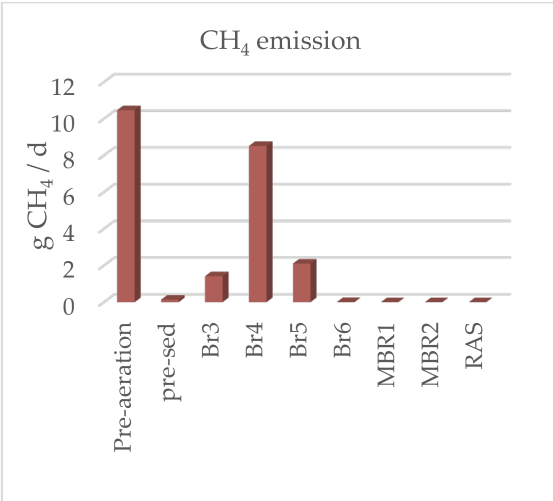


Figure 24. Preliminary results of nitrous oxide emissions (left) and methane emissions (right) from the pilot process.

Based on the preliminary data from 2018 approximately 0.07% of the total NH_4-N load was converted to nitrous oxide nitrogen in the treatment line.

6.3 Phosphorus removal

A summary of the trials related to the nitrogen removal is presented in Table 16. The trial of imitating the first phase operation of the full-scale plant affected the phosphorus removal as the load was high and no iron chloride was used. Throughout 2018 test of enhances biological phosphorus removal (EBPR) have been carried out. A separate test where no precipitation chemicals were added was conducted during summer. Towards the end of the year a phosphate analyser was in use and the dosage strategy based on effluent phosphate was put into operation.

Table 16. Trials related to the phosphorus removal.

Trials	Jan	Feb	Mar	Apr	May	Jun	Jul	Aug	Sep	Oct	Nov	Dec
Normal load, fixed dose												
Imitation of first phase operation SFA												
Precipitation chemical addition stopped												
Effluent phosphate analyser in use												
EBPR tests												

The goal of reaching stable effluent phosphorus concentrations below 0.15 mg P/L had been achieved previous year by using a control strategy with dosage of ferrous sulfate and ferric chloride in three points in the process, where the first dosage was flow proportional and the dosage in the other two points were controlled using feedback control from online effluent phosphate measurements.

For the first part of 2018 (w.1- w.25) only ferrous sulfate was used as precipitation chemical and no phosphate control of the dosage was used. After this period all dosage was stopped in order to evaluate the potential of biological phosphorus removal during week 26 to week 42. The last part of the year, dosage in three points was used, with ferrous and ferric dosages based on feedback control from the phosphate analyser.

The phosphorus concentrations in and out from the biological treatment is presented in Table 17 and Figure 25 below. As no precipitation was used for a long time, the yearly average effluent concentration was high, 0.75 mg P/L, however if excluding the weeks without precipitation (w.26-w.41) the average effluent concentration was 0.14 mg P/L.

Table 17. Phosphorus concentrations in primary treated water (PTW) and effluent during 2018.

Parameter	Limit	Average	Min	Max	Nr of weekly samples
TP PTW (mg/L)	-	5.4	3.7	9.0	52
TP EFF (mg/L)	0.20	0.75	0.058	3.5	52

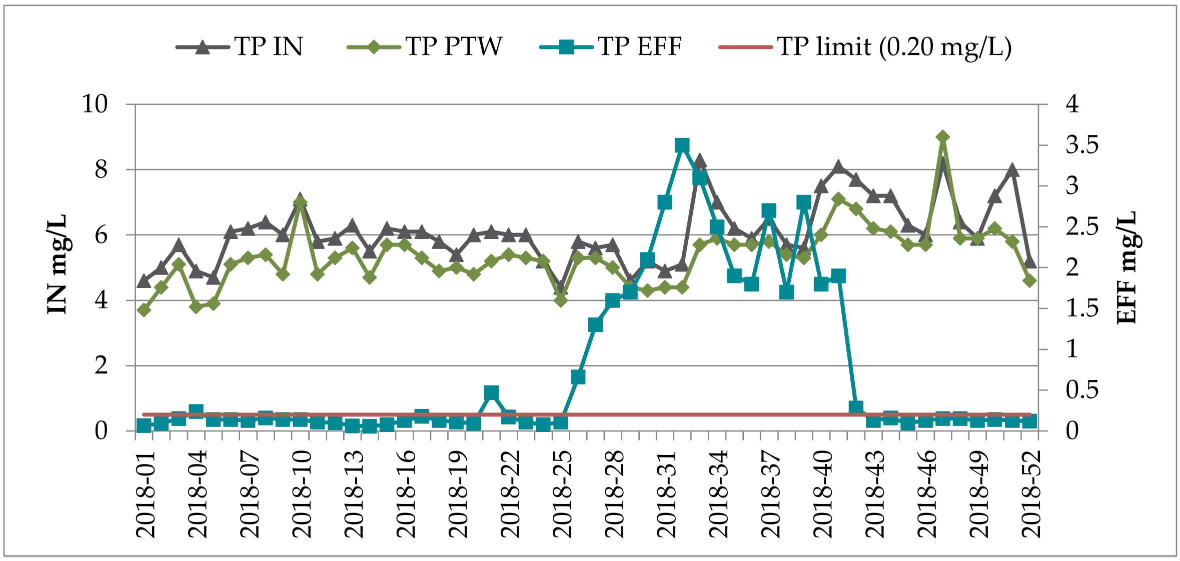


Figure 25. Influent and effluent total phosphorus analysed on weekly composite samples.

To evaluate and compare phosphorus removal with different precipitation and operational strategies, the year 2018 has been divided into four periods.

The first period (T1, week 1 to 9) was when the load and operation was normal except for not using ferric, not using online phosphate to control precipitation chemicals and the treatment line was operated without addition of reject water. The background for having this period was that the phosphate analyser broke in October 2017 and a new analyser was not installed until the end of January 2018. Due to some problems the first months after installation of the analyser, precipitation chemical dosage was kept as flow proportional dose in pre-aeration, prior to pre-sedimentation tank, and a fixed flow to BR4.

The second period (T2, week 10 to 24) was during the imitation of the first full scale treatment line operation. This operational strategy included phosphorus removal using only ferrous sulfate and the dosage was flow proportional to pre-aeration and a fixed flow to BR4. The reason this period is separated from the first one is because it was operated at high load (high inflow and reject water addition).

In between the second and third period, there was one week of continuation of the precipitation dosage, however the inflow was reduced to normal and no reject water was added. The precipitation chemical dosage was then stopped in the beginning of week 26. This short period (week 25-26) was excluded in the evaluation of the four periods.

The third period (T3, week 27 to 41) started during summer when no precipitation chemicals were used in order to evaluate the potential for enhanced biological phosphorus removal, EBPR, without interference from chemical dosage. During this trial it was also of interest to study how the concentration of iron in the sludge behaved.

The fourth period (T4, week 42 to 52) was during the last couple of months of the year when all three dosing points were used, both ferrous and ferric was added and the online phosphorus control was in operation.

Key parameters for the phosphorus removal, both for the pilot and for the SFA design, are presented in Table 18. The phosphorus load on the biological treatment in the pilot was 1.5-2.3

times bigger than in the SFA design. In all experimental periods except T3 (no Fe dose) the uptake in the biology was also bigger than in the SFA design. At the same time the iron to phosphorus ratio was low, 1.5-1.9 mole/mole in the pilot which indicates enhanced biological phosphorus uptake (see 6.3.2).

Table 18. Comparison of operational data from the pilot with data for the SFA design, yearly average values.

Parameter	Unit	T1 2018	T2 2018	T3 2018	T4 2018	SFA design	Scaled SFA design*
Phosphorus load influent	kg P/d	0.43	0.61	0.43	0.59	2 594	0.39
Phosphorus load biology	kg P/d	0.36	0.54	0.38	0.53	1 556	0.23
Phosphorus load effluent	kg P/d	0.010	0.014	0.155	0.013	79	0.012
Phosphorus removed in biology	kg P/d	0.35	0.53	0.22	0.52	1 477	0.22
Iron consumption (PS+BR4+BR6)	kg Fe/d	1.21	1.43	0	1.70	10 000	1.49
Iron consumption per removed phosphorus	mole Fe/mole P	1.9	1.5	0	1.8	2.8	-
Phosphorus in sludge	% of SS	3.0	3.0	3.5	3.7	5.4	-
Iron in sludge	% of SS	9.1	7.8	2.8	7.4	-	-
VSS in sludge	% of SS	73	75	80	76	63	-

*SFA divided by 6700. Values comparable to pilot data.

6.3.1 Precipitation

The total amount of iron dosed is presented in Figure 26. The base dose of Fe^{2+} was added to the pre-aeration which, when in operation, was controlled flow proportionally to a dose of 12 mg Fe/L (except two shorter periods when 10 mg Fe/L was used due to low concentrations of effluent phosphate). A supplementary dosage of Fe^{2+} was added to the aerated part of the biological treatment (BR4). This dose was during most of the year controlled using a manually adjusted fixed flow however a slow (time constant 1 h) online feedback control from effluent phosphate with a maximum dosage of 17 mg Fe/L was used from week 42. The weekly average of the dose to BR4 varied in the range 0-11.5 mg Fe/L. A third and final polishing dose using Fe^{3+} was added in BR6 (just prior to the membrane tanks) during the last months of the year. This dose was added during shorter peaks in effluent phosphate with a maximum dosage of 7 mg Fe/L (weekly average dosage was 1.6 mg Fe/L during the period when it was used).

As this year included separate trials for enhanced biological phosphorus removal, the yearly average cannot be used to compare to the SFA design. However, the total iron dosage for week 42 to 52 was 20 mg Fe/L which is the same as assumed in the SFA design and then an effluent concentration of 0.15 mg P/L was reached. Iron dosage and effluent phosphate concentration is presented as daily average values in Figure 26.

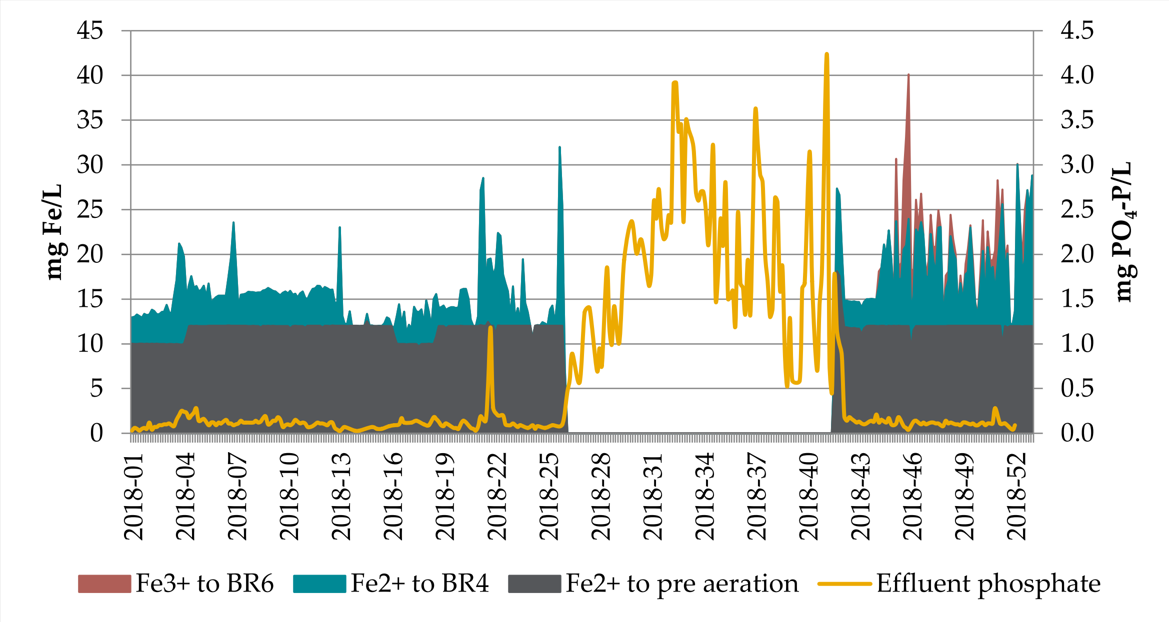


Figure 26. Iron dosage (the sum of three dosing points) as daily average and effluent phosphate analysed in daily composite samples.

Normal fluctuations of iron content in sludge was between 6.5 and 10% of TSS during normal iron dosage to reach effluent phosphorus concentrations below 0.2 mg P/L (Figure 27). When iron dosage was stopped, the iron content slowly decreased, reaching a minimum of 1.1% after 12 weeks without iron dosage. The phosphorus content in sludge (biologically and chemically bound) varied between 2 and 3.7% in the period before iron dosage was stopped with an average of 3.0%. During the period without iron dosage, phosphorus content in sludge was slightly higher, on average 3.5%, and peaking at 4.3% of TSS. After starting the iron dosage again, phosphorus content in sludge increased further, on average 3.7% of TSS.

Although the phosphorus concentration in the activated sludge increased when the iron dose was stopped, the effluent phosphorus concentration increased directly when the dose was cut. The iron buffer in the sludge when the Fe-dose was cut was 7-8% of TSS. The higher phosphorus concentration in sludge without iron dosing can be explain by a higher TP concentration into the biological treatment due to decreased pre-precipitation. These results can be compared with the results from 2016 when the iron dose was cut for two weeks during the autumn and it took one week before the effluent concentration of phosphorus increased. The initial concentration of iron in sludge at that time was also around 8% of TSS and the phosphorus concentration in sludge was 3% of TSS which was a slight decrease from the values before the dose was cut (3.2-3.4%).

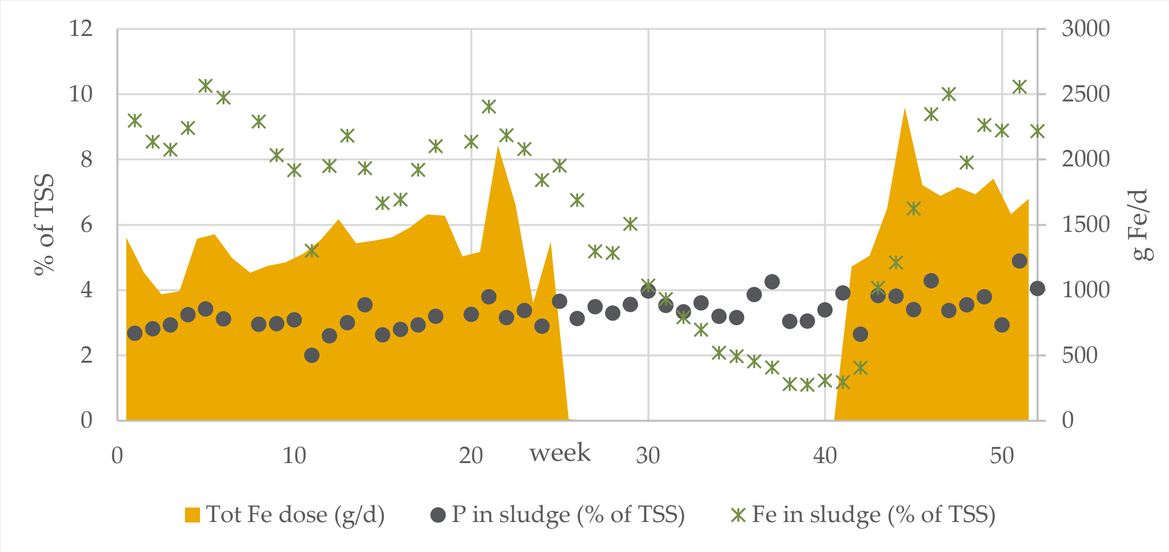


Figure 27. Iron and phosphorus in sludge.

6.3.2 Enhanced Biological Phosphorus Removal (EBPR)

The biological process in the future Henriksdal WWTP, and the MBR-pilot, was designed for chemical phosphorus removal and therefore, the anaerobic zone required for enhanced biological phosphorus removal (EBPR) was not included in the design. Still, operational data regarding Fe-dosing and phosphorus removal in combination with the phosphate release during acid cleaning of the membranes indicated EBPR-activity within the biological process. Therefore, P-release tests were conducted according to Tykesson & la Cour Jansen (2005) approximately once per month starting in February 2018. Sludge from the conventional activated sludge (CAS) process at Henriksdal WWTP was used as a negative reference.

Results from the P-release tests together with data on TP and Fe are shown in Figure 28 and Figure 29. In April the sludge from the pilot showed a higher P-release rate than the Henriksdal sludge even though the Fe-dose was higher in the pilot than in Henriksdal (10-12 g Fe/m^3 in Henriksdal). In July to October when the Fe-dose was shut off completely in the pilot, P-release rates increased to above 7 g P/kg VSS, h (with exception for august) which indicates a high EBPR-activity (Janssen *et al.* 2002). Also, after the startup of the Fe-dosing the P-release rate remained significantly higher than in the Henriksdal sludge. The results show a moderate correlation between the iron content in WAS and the EBPR activity (Figure 30). However, more data is required to determine if there is a true correlation or not. These results show that EBPR arose spontaneously in the process despite the absence of a designated anaerobic zone and the relatively high dosage of Fe.

In order to determine where in the process the P-release and –uptake takes place, TP and PO_4-P was measured along the treatment line on three occasions; in February, June and August. Results from February and August did not display results that indicated clear P-release or –uptake zones. The results from June (see Figure 31), however, showed a small release in the second pre-denitrification zone, BR2, where the nitrate levels generally are low (normally <1 mg N/L) and VFA may be produced by endogenous hydrolysis, and an uptake in the two fist aerated zones, BR3 and BR4. A second large release was seen in the post-denitrification zone, BR6, where the nitrate level at the time was 2 mg N/L and methanol was dosed. In the following aerated membrane tanks, MT1 and MT2, phosphorus was taken up again, resulting in an effluent concentration of 0.06 mg PO_4-P/L.

Since there is only one profiling showing these results, more studies are needed in order to be able to draw any conclusions. Therefore, more detailed studies of the EBPR and the concurrence with chemical phosphorus removal are planned for the coming years.

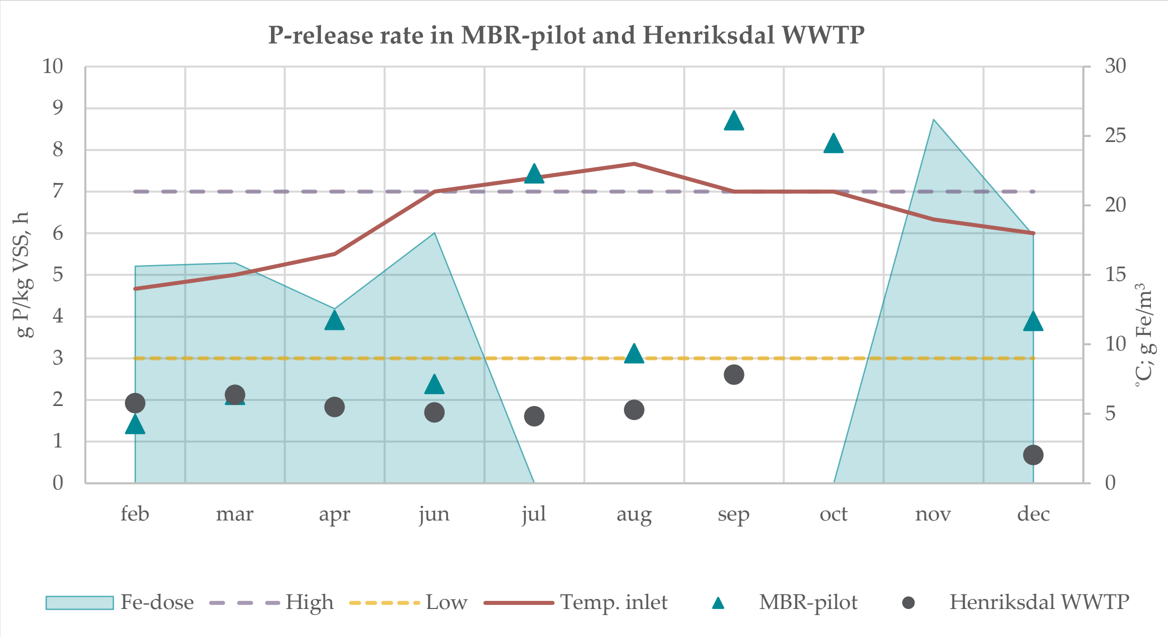


Figure 28. EBPR activity in the activated sludge from the MBR-pilot and Henriksdal WWTP (CAS). Dotted lines show limits for EBPR-activity according to Janssen *et al.* 2002, values below the yellow line indicate poor EBPR and values above the purple line indicate high EBPR.

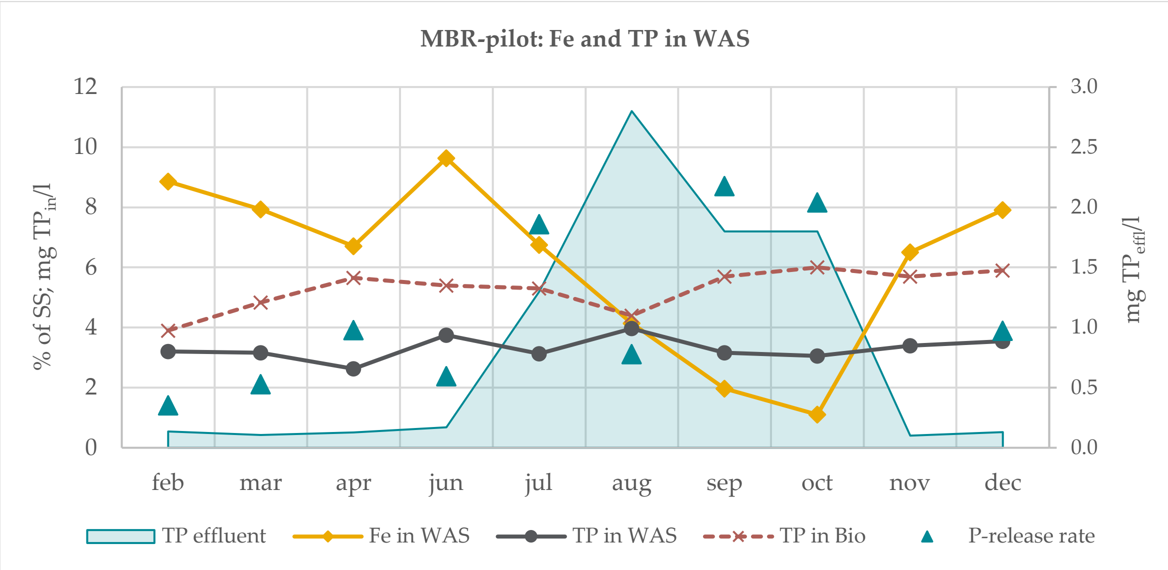


Figure 29. The content of phosphorus and iron in waste activated sludge (WAS) compared to P-release rate and TP in and out of the biology.

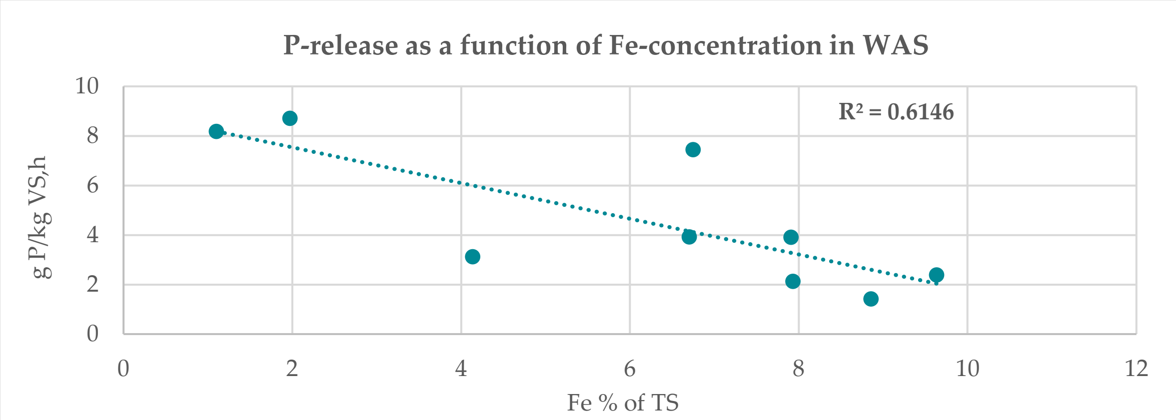


Figure 30. The relationship between iron in waste activated sludge (WAS) and P-release rate in the MBR-pilot.

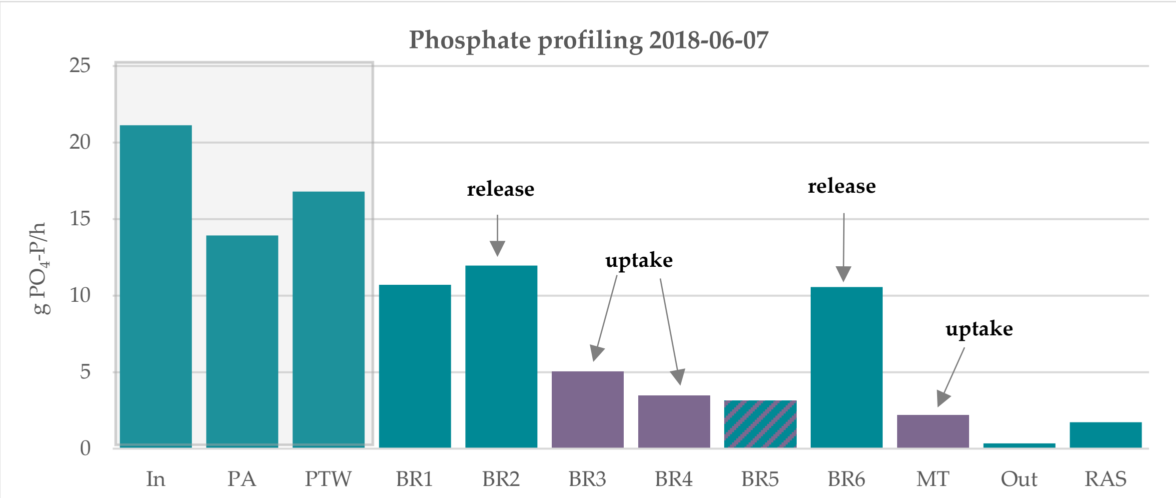


Figure 31. Phosphate profiling along the MBR-line. The three first steps are the pre-treatment (no EBPR). Purple bars are aerated volumes. The arrows indicate possible EBPR induced P-release and P-uptake.

6.3.3 Phosphate analysers

Since Henriksdal WWTP was offered phosphate analysers from EndressHauser (Liquiline System CA80PH) for the future plant through the turnkey contract with the machine entrepreneur, it was decided to test the specified analyser in the MBR-pilot and compare its performance to the existing Metrohm-analyser. Most phosphate analysers on the market have a lower detection limit of 0.05-0.10 mg PO_4-P/l (molybdenum blue method) with an accuracy in that range of 0.05 mg PO_4-P/l. The MBR-pilot is operated at PO_4-P levels in the effluent below 0.1 mg/L most of the time. It is therefore important to verify that the analysers used can provide reliable results in that low range. Results from a trial period in November and December is shown in Figure 32. Both analysers showed acceptable values and the EndressHauser was actually closer to the laboratory values than the Metrohm during this period. Trials will continue in 2019.

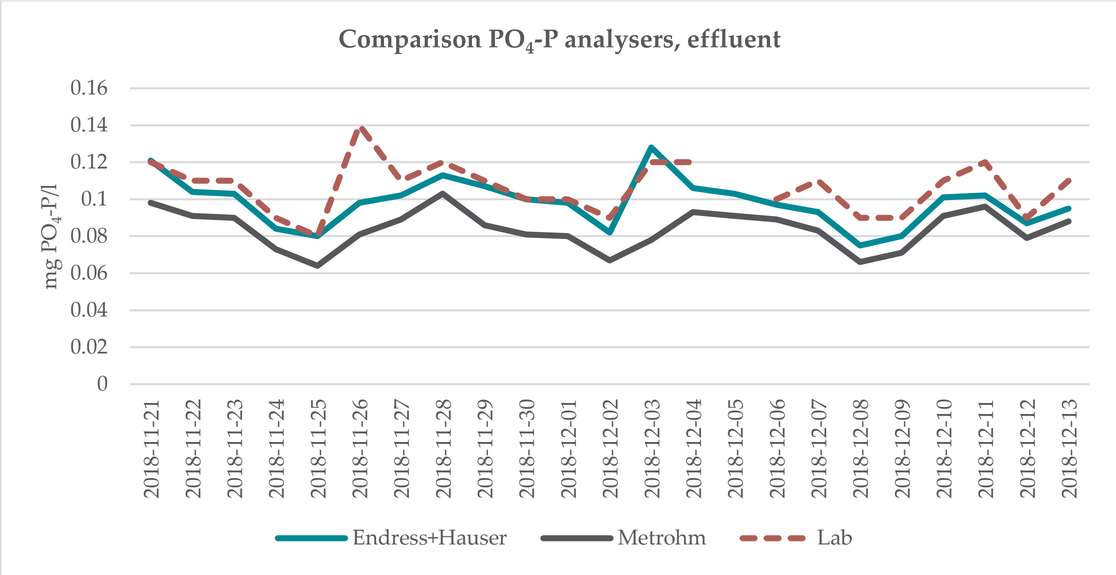


Figure 32. Daily average values from the two phosphate analysers compared to lab values of daily samples.

6.4 BOD reduction

Analysis on BOD_7 from daily composite samples have, since start-up of the MBR pilot in 2013, shown values of <2 mg O_2/L, except for one sample where the analysed concentration was 3 mg O_2/L. Since the expected effluent requirement of BOD_7 in year 2040 is 6 mg O_2/L as an annual average, there is no reason to assume that the effluent requirement will not be met. Analysis of BOD was not carried out in 2018 and no specific measures have been taken to achieve a higher BOD reduction.

6.5 Membrane performance

A summary of the trials related to the membrane performance is presented in Table 19. Throughout 2018 the algorithm for fouling control has been in use to optimise the scouring air used for membrane cleaning. The cleaning procedure, minimising the amount of chemicals used for cleaning and a comparison of citric and oxalic acids for cleaning has been in focus. During recovery cleaning some membrane fibers were sent for membrane autopsy.

Table 19. Trials related to the membrane performance.

Trial	Jan	Feb	Mar	Apr	May	Jun	Jul	Aug	Sep	Oct	Nov	Dec
Optimisation of scouring air use												
Oxalic acid and citric acid comparison												
Optimisation of membrane cleaning chemical use												
Recovery cleaning and sampling												

The membranes have been operated in cycles with 10 minutes of permeation followed by 1 minute of relaxation. Both feed (pumping from BR6) and aeration was on during the normal operation cycle. In order to manage the varying flowrate with only two membrane tanks, the pumping of permeate was proportional to the feed, which in turn was proportional to the level in BR6.

As it is inefficient to operate the membranes at too low fluxes, the membrane tank longest in operation went into standby mode at low influent flowrates (normally during night). In standby mode the membranes were aerated intermittently 5 minutes every half hour.

The membranes had slightly too large surface area compared to the scale of the rest of the treatment line. In order to maintain representative flux over the membranes a fraction of the permeate was recycled back to the membrane tank. At normal flowrates one third of the permeate was recycled. In order to also manage peaks in the flowrate, the permeate recirculation was reduced with increased influent flow rate. The varying permeate recirculation affected the sludge concentration in the membrane tanks and thus also the sludge concentration in return and waste sludge. On average the permeate recirculation was about 17 % of withdrawn permeate during 2018.

6.5.1 Permeability

Permeability above 200 L/(m²·h·bar) is considered good according to the supplier. As can be seen in Figure 33, the permeability was above 300 L/(m²·h·bar) throughout most of the year for both membranes. During the beginning of the year (week 1 to 19) permeability was slightly higher for MT1 (using oxalic acid for cleaning) and relatively stable permeability for both membranes. After recovery cleanings performed in week 20 and 21 (see details regarding recovery cleaning in section 6.5.3.) the permeability increased. A decrease from week 22 can be seen and was considered related to a peak flow test where influent was fixed at 5.5 m³/h for one week. For MT1 permeability continued to increase during summer with higher temperature, while permeability for MT2 continued to decrease. Week 26 it was decided to switch acids for the two membranes, so MT2 was cleaned with oxalic acid and MT1 was cleaned with citric acid. After four weeks with cleaning using oxalic acid, the permeability for MT2 was increased to values close to MT1 and the acids was switched back.

The increases in permeability that can be seen for MT1 in week 43 and week 51 are not related to recovery cleaning, but the effect of maintenance cleaning with oxalic acid.

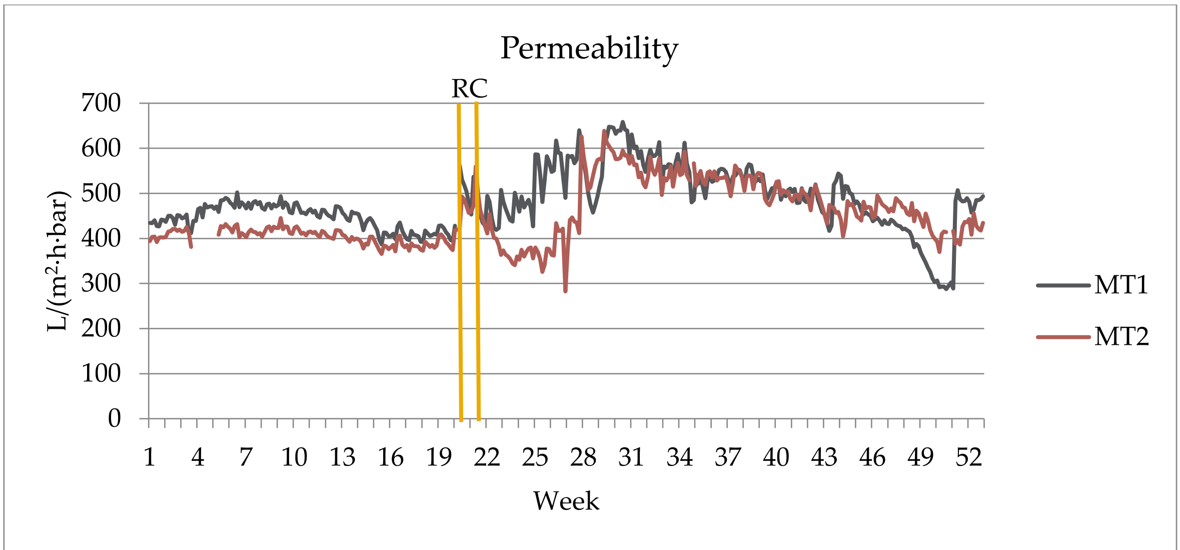


Figure 33. Permeability (temperature compensated) for membrane 1 (MT1) and 2 (MT2) during project year 5 (2018). Recovery cleaning (RC) was carried out in week 20 and 21.

6.5.2 Flux and TMP

Fluxes for the two membranes are presented in Figure 34. Normally the membranes were operated with net flux around 20 to 25 L/(m^2·h). After finishing the test with high fixed influent flow rate in week 24, the permeate recirculation was not adjusted until after the summer (week 34) resulting in lower flux than intended.

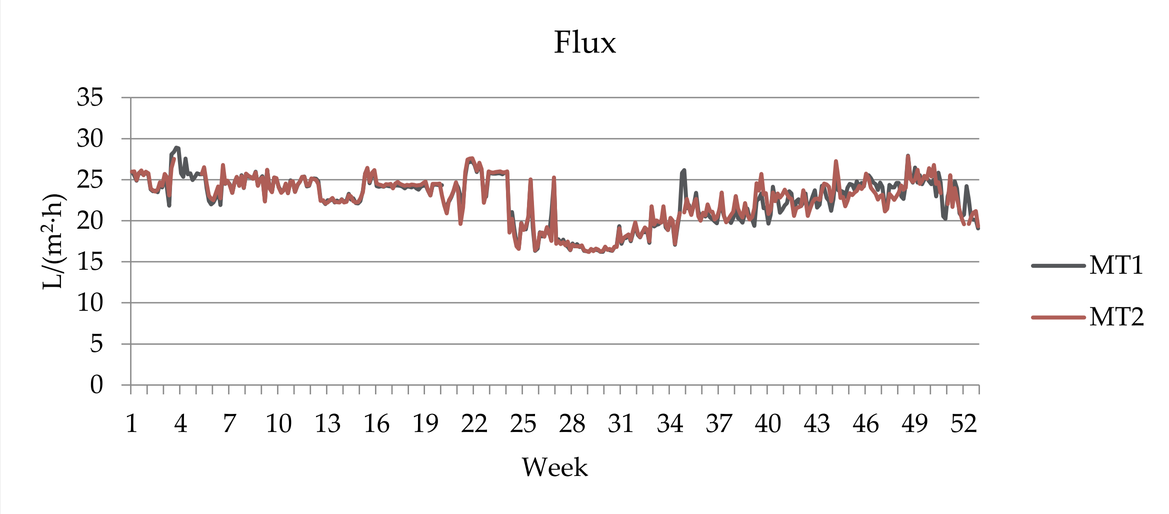


Figure 34. Net flux during 2018.

Throughout the period with low flux (below 20 L/(m^2·h)) the transmembrane pressure (TMP) was at its lowest, see Figure 35. Based on daily average data, TMP varied between 20 and 88 mbar during 2018. The TMP decreased only slightly (from about 60-65 mbar to 45 mbar) after recovery cleaning. This is likely due to the low initial TMP. Before the last recovery cleaning (RC) in 2017 TMP was 186 and 140 mbar for MT1 and MT2 respectively. TMP increased from mid-summer towards the end of the year. The last weeks of 2018, TMP decreased again, for MT1 this was the result of one oxalic acid maintenance cleaning (MC).

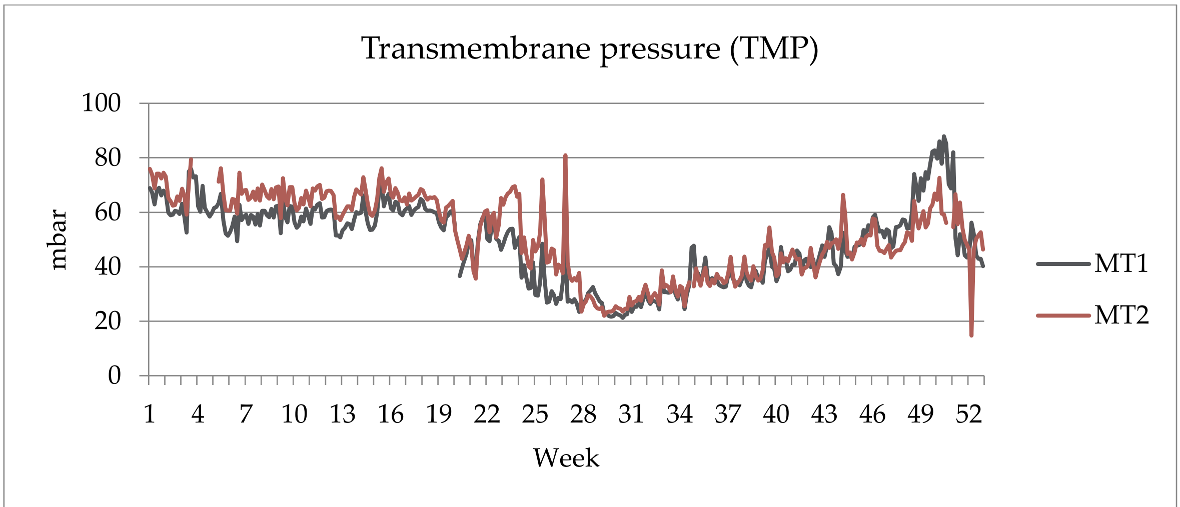


Figure 35. Net TMP during 2018.

6.5.3 Membrane cleaning

The membranes were cleaned with sodium hypochlorite and citric or oxalic acid. Two types of cleaning procedures were carried out; maintenance cleaning (MC) and recovery cleaning (RC).

Maintenance cleaning

The maintenance cleanings (MC) were automatically carried out every week. In order to keep the treatment line in operation, each membrane was cleaned separately, and the cleanings were scheduled at night when the influent flow rate was low. In order to assure that the influent flow rate was not too high for the one membrane tank in operation, the influent flow set-point was set to half of the current value, although never lower than 1.8 m^3/h.

The MC takes about one hour and according to the cleaning schedule provided by the supplier these cleanings should be carried out with acid about once per week (after 345 m^3 of permeate were produced by that membrane) and with sodium hypochlorite about twice per week (after 173 m^3 of permeate was produced). The cleaning chemical was mixed with permeate and back pumped in pulses through the membranes. Normally there were nine back pulses (BPs), the first one a bit longer (2-5 minutes) followed by eight shorter with relaxation in-between (30 seconds followed by 4.5 minutes of relaxation). The chemical solution was pumped with a back flux of 20 L/(m^2·h) and the target concentrations of the solution entering the membranes (after dilution with permeate) were 200 mg Cl_2/L for sodium hypochlorite, 2000 mg/L for citric acid and 1300 mg/L for oxalic acid.

Last year attempts of reducing the chemicals used for maintenance cleaning started. The time of the initial backpulse was reduced from 5 minutes to 2 minutes and later the number of backpulses were reduced from 9 (incl. the first longer one) to 7 in total.

During this year further work on primarily reducing the oxalic acid usage by altering the cleaning intervals in between cleaning events. The operational settings have been divided into five trial periods. An overview of the trials is presented in Table 20.

Table 20. Overview of trials with reduced acid for membrane cleaning. BP=backpulses.

Trial	Start	Trial
T1	Sept 2017	Citric vs Oxalic - Reduced BP (both MT)
T2	June 2018	Recovery Period (short switch between chemicals)
T3	July 2018	Trial reduced nr of BP oxalic acid, standard citric acid
T4	Aug 2018	Trial reduced nr of BP and 20% longer time in-between oxalic acid cleanings.
T5	Oct 2018	No oxalic acid cleanings
T5	Dec 2018	One oxalic acid cleaning
T5	Dec 2018	No oxalic acid cleanings

The amount of chemicals used normalized to the initial settings[2] (back pulse duration 2 minutes + 8 x 30 seconds carried out after 345 m^3 permeate produced) are presented in Figure 36 together with the permeability.

Trial 1 (T1) started in September 2017 when the number of backpulses were reduced from 9 (in total) to 7. As M1 (cleaned with oxalic acid) had higher permeability than M2, the acids were

[2] Inital settings previous year (2017) was calculated as 5 min initial backpulse, followed by 8 x 30 s backpulses.

switched (Trial 2, T2) so M1 was cleaned with citric acid and M2 was cleaned with oxalic acid for a short period (19 days).

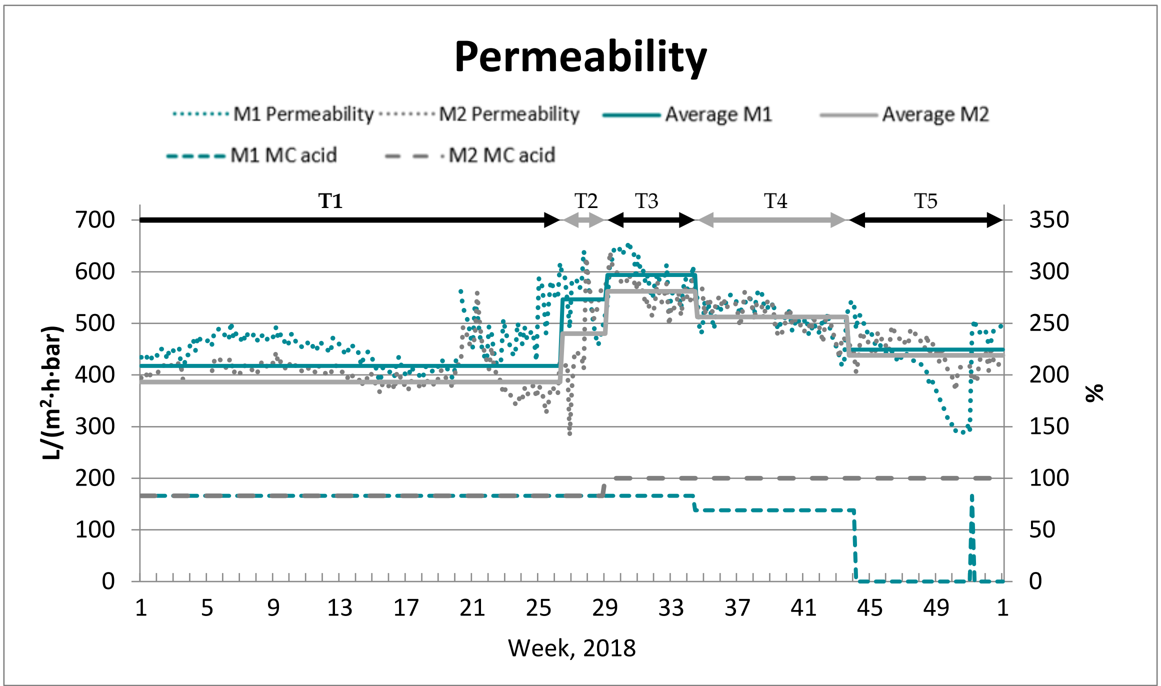


Figure 36. Amount of acid used for maintenance cleaning (MC), normalized to back pulse duration of 5 minutes + 8 x 30 seconds carried out with interval of 345 m³ of permeate produced. M1 was cleaned with oxalic acid, M2 was cleaned with citric acid. T1 – T5 are trial periods. T1 started in September 2017.

When permeability was similar for the two membrane tanks, trial 3 (T3) started in July (week 29). M1 was again cleaned with oxalic acid with 7 backpulses (oxalic acid consumption corresponding to 83 % of suppliers' specification) and M2 where cleaned with citric acid and 9 backpulses (100%). Trial 4 started in August, where the interval in-between acid cleanings were 20% longer for M1 (corresponding to 69% of specified consumption). As permeability for both membrane tanks were similar during T4, it was decided to completely stop doing the acid cleanings for M1 (Trial 5) and await a permeability decrease. M1 was operated for 53 days without acid cleaning, permeability decreased from 544 to 289 L/(m²·h·bar). One maintenance cleaning with oxalic acid (7 backpulses) restored permeability to 507 L/(m²·h·bar). M1 continued to operate without any oxalic acid into 2019.

Throughout 2018 maintenance cleaning with sodium hypochlorite has been carried out with interval according to supplier, but with reduced backpulses; 2 min initial pumping followed by 6 x 30 seconds.

Recovery cleaning

During recovery cleaning (RC) the membrane tank was emptied, then filled with chemical solution and the membranes where left to soak overnight.

According to the supplier the RC should be carried out twice every year with both sodium hypochlorite and acid. Previously permeability has been good, and RC has only been needed once per year. The previous RCs were carried out in Oct/Nov 2017 but this year it was decided not to wait a full year but do the RCs in May although permeability was good (around 400 L/(m²·h·bar)).

The reason was to get more information from sampling during RC and to send some membrane treads for membrane autopsy to characterize the fouling.

Since the previous RCs about 10 700 m^3 permeate was produced by MT1 and 10 050 m^3 permeate by MT2. This corresponds to about 104 and 97 m^3 per m^2 membrane area for MT1 and MT2, respectively.

Before and after the RC this time the membrane cassettes were lifted, the membranes were inspected, and some threads were cut and sent for membrane autopsy. A technician from the supplier was attending when the first cassette was lifted to give instructions for the inspection and on how to cut the membrane treads and reseal the cut with special glue.

The cleanings were carried out first with sodium hypochlorite and then with acids (oxalic acid for MT1 and citric acid for MT2) one week later.

The schedule for cleanings can be seen in Table 21 together with the amount of chemicals used and conditions at start and end of the soaking. With oxalic acid the pH was lower (pH 2.16) compared to using citric acid (pH 2.52). About five times larger volume of concentrated acid was needed for the oxalic compared to citric, which for storage reasons is something the full scale plant will need to consider.

Table 21. Results from recovery cleaning (RC).

Date	Membrane tank	Chemical	Amount	Measurements in tank at the start of soak	Soaking time	Measurements in tank at end of soaking
2018-05-15 to 2018-05-16	MT1	Sodium hypochlorite (126 g/L)	10 L	pH 9.5 Cl_2 555 mg/L	22.5 h	pH 8.2 Cl_2 346 mg/L
2018-05-16 to 2018-05-17	MT2	Sodium hypochlorite (126 g/L)	10 L	pH 9.5 Cl_2 660 mg/L	21 h	pH 7.9 Cl_2 282 mg/L
2018-05-21 to 2018-05-22	MT1	Oxalic acid (8%)	23 L	pH 2.16 COD 190 mg/L	19 h	pH 2.24 COD 260 mg/L
2018-05-22 to 2018-05-23	MT2	Citric acid (51%)	4.3 L	pH 2.52 COD 1800 mg/L	19 h	pH 2.57 COD 1500 mg/L

When comparing the effect on permeability there was no great difference between oxalic and citric acid (Figure 37). One week after the citric acid cleanings the permeability was almost the same for the two membrane tanks, and about the same as prior to starting the sodium hypochlorite RC, around 400 L/(m^2·h·bar). One reason for no visible long term improvement of the permeability was likely the high permeability prior to the RCs and that there were not much fouling or scaling to be removed.

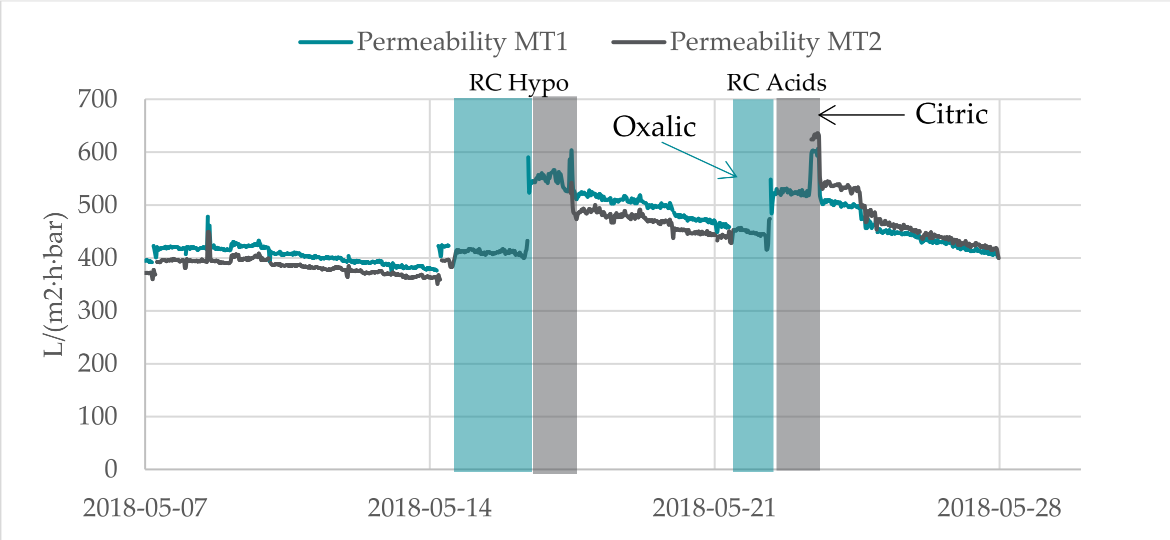


Figure 37. Permeability before and after recovery cleaning (RC).

Analysis of the first permeate after RC is presented in Table 22. When comparing the two acids the permeate after cleaning with citric acid contained higher concentrations of all analysed parameters. Compared to previous RCs in 2017 the phosphate concentration after citric acid cleaning was much lower; 2.2 mg/L 2018 compared to 12 mg/L 2017. One reason for this could be a longer sludge recirculation time through the membrane tank prior to starting the permeation.

Table 22. Analysis of first permeate after recovery cleaning (RC) 2018 (2017 values in parenthesis). n.a., not analysed.

Permeate analysis after RC	Chloride (mg/L)	COD-Cr (mg/L)	PO_4-P (mg/L)	Fe (mg/L)	Mn (mg/L)
RC NaOCl MT1	160 (450)	67 (130)	0.21 (0.39)	0.21 (0.34)	0.014 (0.032)
RC NaOCl MT2	140 (460)	62 (120)	0 21 (0.38)	0.14 (0.29)	0.021 (0.025)
RC Oxalic acid MT1	n.a.	36 (69)	1.3 (0.46)	14 (12)	0.12 (0.31)
RC Citric acid MT2	n.a.	320 (470)	2.2 (12)	32 (55)	0.14 (0.74)

As observed before, the phosphate peak after oxalic acid cleaning was lower than after cleaning with citric acid (Figure 38).

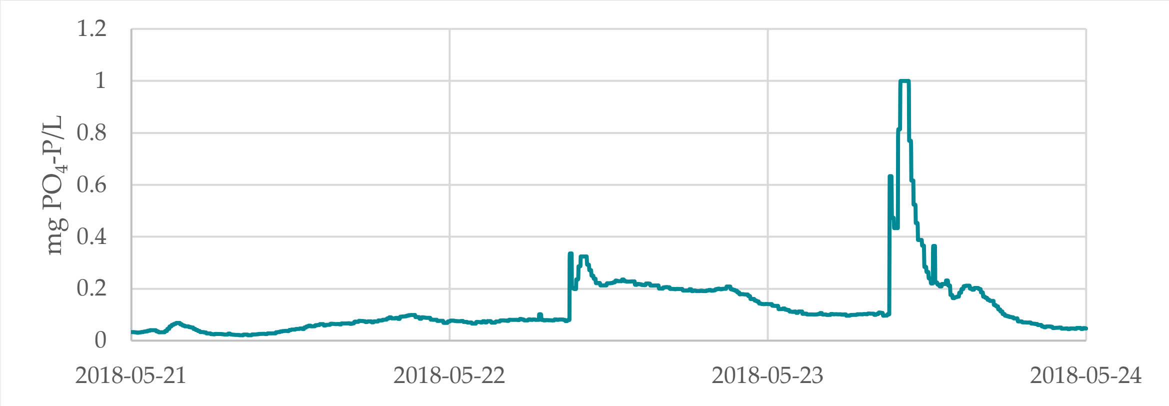


Figure 38. Phosphate concentration in effluent measured by automatic online analyser after RC with oxalic acid (22nd of May) and after RC with citric acid (23rd of May).

6.5.4 Membrane autopsy

In connection to the 2018 recovery cleaning, in May 2018, the two membrane cassettes were lifted and inspected, and three membrane fibers were cut from each cassette, one from each side of the cassette and one from the center of the cassette. Three membrane fibers were cut at three different occasions; before start of recovery cleaning, after cleaning with sodium hypochlorite but before cleaning with acid, and after cleaning with acid. In total, 18 membrane fibers were cut, nine from each membrane cassette, and the top of each fiber was marked with a knot. The 18 membrane fibers were sent to external laboratory for autopsy. The cut membrane parts remaining on the cassette were sealed with a dedicated glue.

The inspection of the cassettes showed that the stainless-steel frame was in good condition. There was no corrosion, no cracks, no bending or any other damage to the frame could be identified. Also, the plastic parts were tight and in good condition with no cracks, bending or any other damage. The membranes were not sludged up and not stuck together, and there was no black colouring (which could indicate anaerobic conditions). No broken fibers were seen. The colour of the fibers were light grey-brown with a slight bio-fouling on the surface. A small amount of sludge and fibric material was stuck in the top and bottom header (Figure 39). The membranes were flexible and with a good slack. Generally, the membrane cassettes were in really good condition after two years of operation.

Figure 39. Photos of the membranes during inspection.

Microscopic images of the membrane fibers showed decreased fouling on the membranes as the recovery cleaning process progressed. The foulant before recovery cleaning contained mainly iron, some organic material and trace amounts of calcium phosphate. The foulants on the samples after cleaning contained trace amounts of iron and organic material, see Figure 40 and Figure 41.

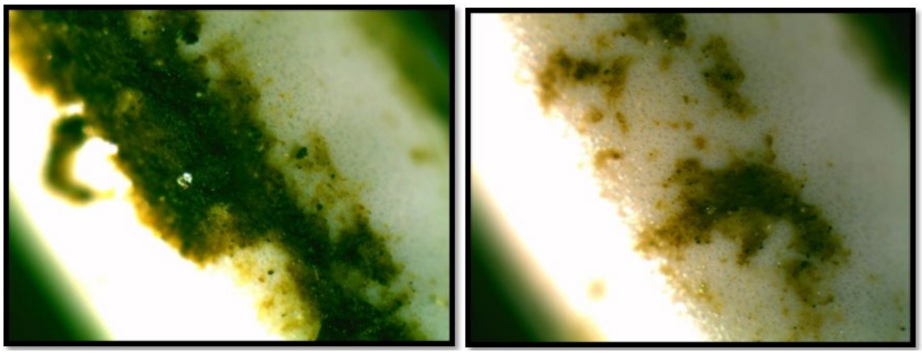


Figure 40. Microscopic image of membrane fouling before recovery cleaning.

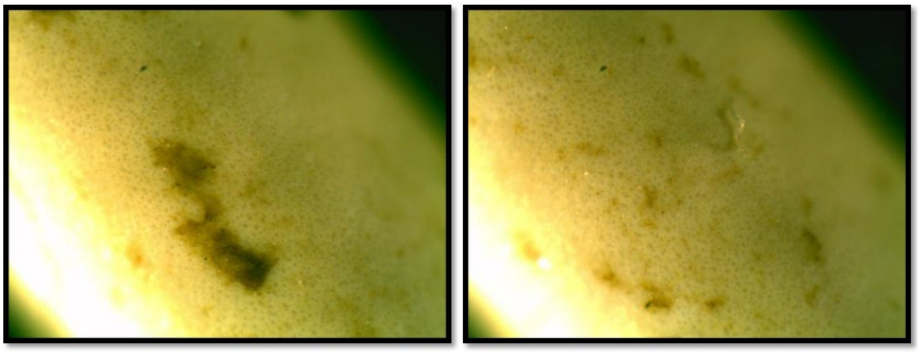


Figure 41. Microscopic image of membrane fouling after recovery cleaning (with both sodium hypochlorite and acid).

Fourier Transform Infrared Spectroscopy (FT-IR) of the membrane fibers before cleaning resulted in several peaks related to the membrane material (PVDF), proteins and phosphate containing compounds, whereas membrane fibers after cleaning only resulted in peaks related to PVDF, i.e. most fouling had been removed.

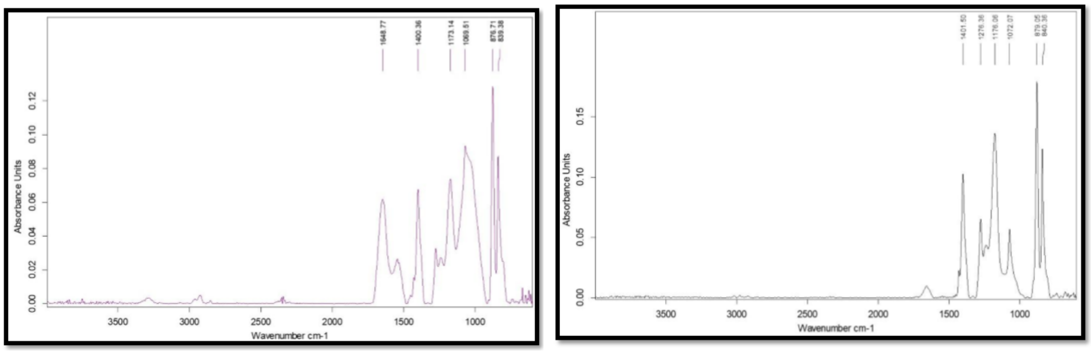


Figure 42. FT-IR of membrane fibre before cleaning (left) and after cleaning (right). The peaks displayed around 1400, 1200-1100 and 850 cm^{-1} are all related to the fibre material (PVDF) and can be seen in both figures. In the left figure there can also be seen peaks around 1650 cm^{-1} which are associated with presence of proteins and a broad peak around 1000 cm^{-1} which is associated with presence of phosphate containing compounds. These peaks cannot be seen in the right figure.

6.6 Operation according to phase two of SFA

For 10 weeks the pilot was operated in a way to mimic how the first treatment line with MBR (BB1) will be operated at Henriksdal during the first 2-3 years (phase 2 of the SFA-project year 2020-2023). This trial was called the BB1-trial.

As two activated sludge lines will be taken out of operation once the first MBR line is commissioned, a high flow to the first MBR line will be required. The pilot operation involved a high fixed inflow, reject water connected to inlet and ferrous sulfate to the inlet pipe prior to the pre-aeration tank and BR4 as chemical addition. No ferric chloride and no external carbon source were used.

The chemical dosages in BR6 was stopped on March 6th and reject from Sickla was added starting on the 7th of March.

The test period was started on March 23th (Friday week 12) when the inflow was increased to a constant flow of approximately 4.8 m^3/h (Figure 43) which corresponds to 150% of design flow. The test period ended on June 11th (Monday week 24). A separate high flow test (Qin=5.5 m^3/h) was also conducted during one week towards the end of the test period after RC was performed on both membranes.

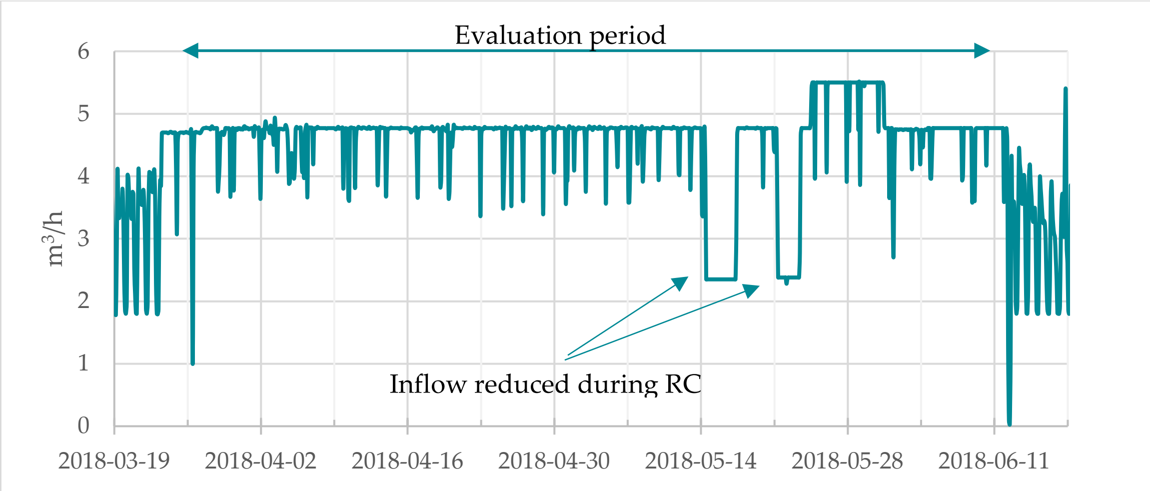


Figure 43. Influent flowrate during BB1 operation test. After seven weeks RC was carried out, first with sodium hypochlorite, a week later with acids. After RC the maximum flow was tested for one week.

Membrane performance during high flow

During the first two and a half weeks, no changes were made in the operation of the membranes. However, at high inflow, the high permeate withdrawal resulted in a lower return sludge flowrate (Q_{RAS}) in relation to the inflow (Q_{IN}) than the recommended 3-4 times Q_{IN} (Figure 44). This caused high MLSS concentrations in the membrane tank and RAS (above 12 000 mg SS/L) although the MLSS in the bio rectors were kept at normal concentration of about 8 000 mg/L (Figure 45). It was decided to adjust the permeate withdrawal in relation to the feed, and to adjust the permeate recirculation to keep the sludge from getting too concentrated in the membrane tank and keeping a high flux. After the adjustments, permeate recirculation was fixed at 15% and flux increased slightly (Figure 46).

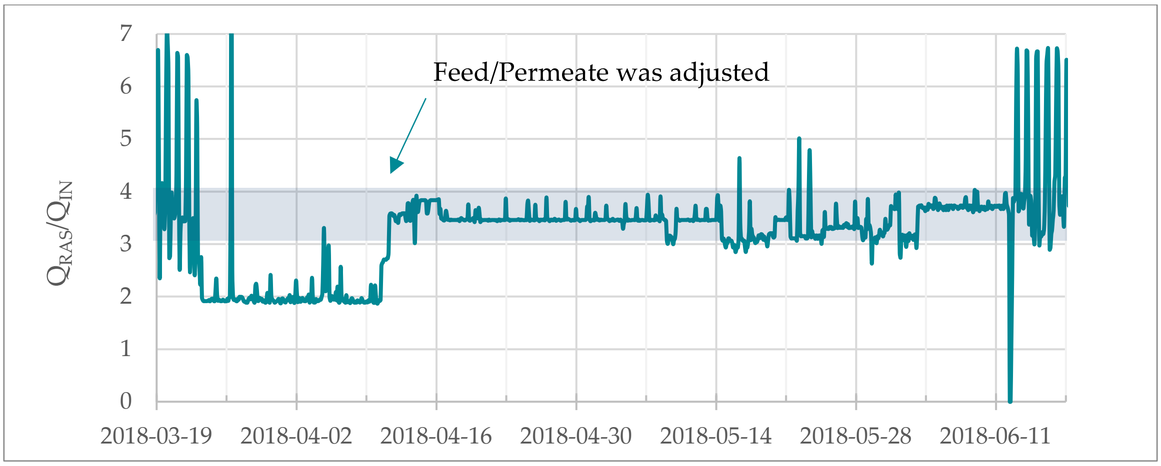


Figure 44. Return sludge flowrate related to influent. Recommended operating range is between 3-4 xQ_{IN} (shadowed).

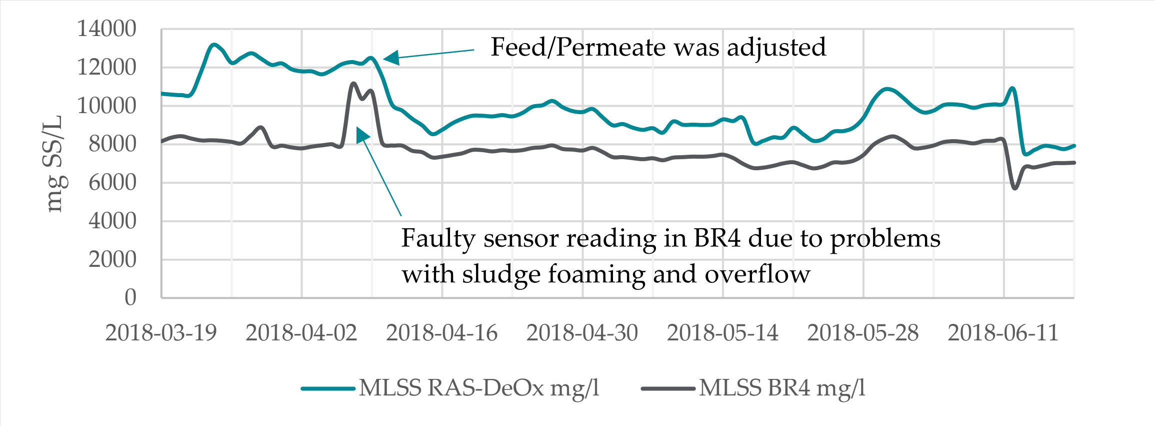


Figure 45. Online measurements of sludge concentration in the bioreactors (BR4) and in the return sludge (RAS-DeOx).

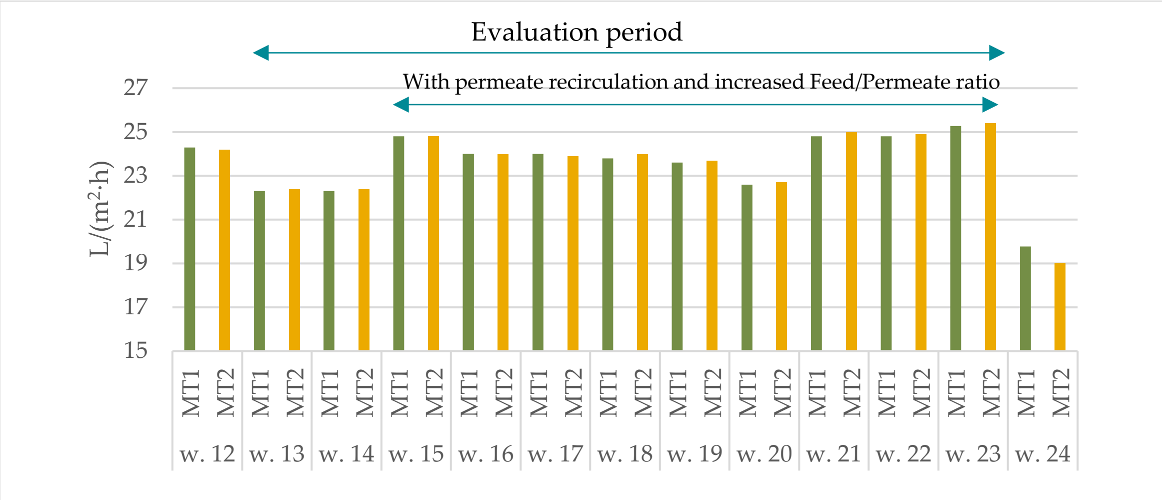


Figure 46. Net flux. The first two weeks of BB1 operation there was no permeate recirculation. With permeate recirculation flux increased.

The permeability was even throughout the trial (Figure 47). No immediate effect of the higher MLSS concentrations in the membrane tank during week 13-14 was observed. The permeability decreased from 448 and 401 L/(m^2·h·bar) the week before the trial started to 405 and 384 L/(m^2·h·bar) prior to RC was conducted (RC with sodium hypochlorite week 20 and RC with acid week 21) for MT1 and MT2 respectively. Only a small increase in permeability was observed after RC, possibly due to a) the high permeability prior to RC and b) the week after RC (Thursday w.21 to Thursday w.22) the line was operated at peak flow (5.5 m^3/h).

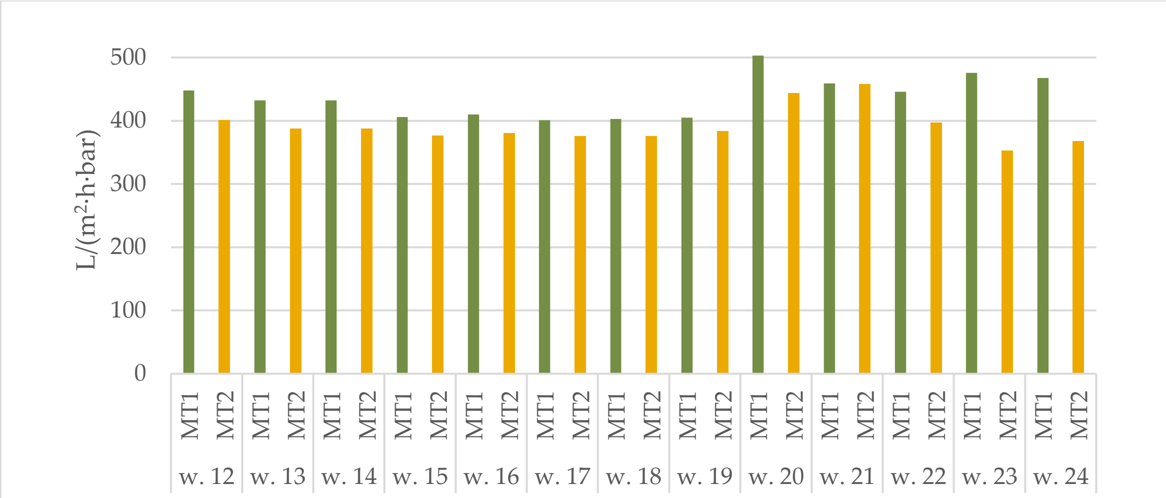


Figure 47. Permeability. RC with sodium hypochlorite was carried out in week 20, RC with acid week 21.

The conclusions regarding membrane performance, during the BB1 operation test, was that a high permeability was maintained at flux up to 25 L/(m^2·h) with normal MLSS concentrations (8 000 mg/L in bioreactors and 10 000 in MT). No decreased permeability was seen when operating at higher MLSS concentration (12 000 mg/L in MT, Q_{RAS}/Q_{IN}=2) when flux was between 22 and 24 L/(m^2·h).

6.7 Biological treatment during high load

The food to mass (F/M ratio) varied during the trial, however an increase from 0.06 kg BOD_7/kg SS to about 1.0 kg BOD_7/kg SS was kept during at least 7 weeks of the trial period (Figure 48).

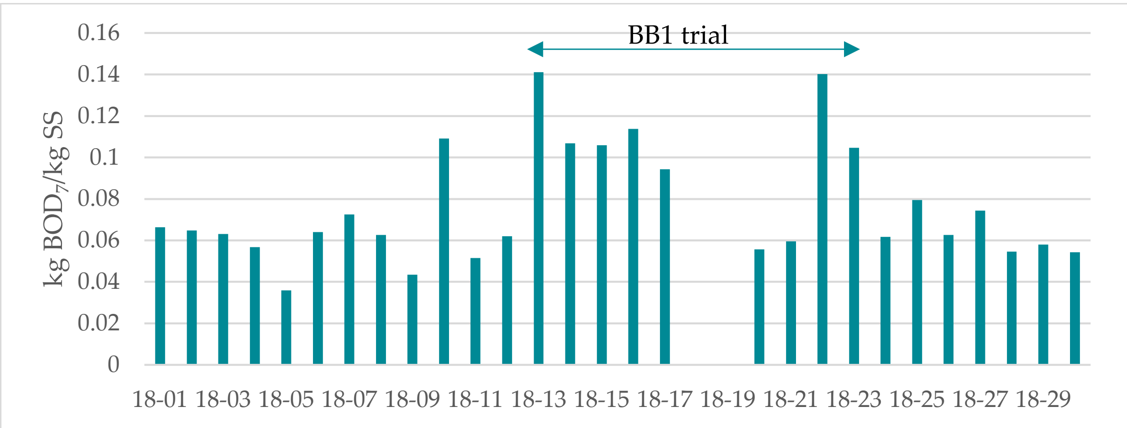


Figure 48. Food to mass. Only bioreactors 1-6 was used as biological volume when calculating the F/M ratio. Week 18-19 data is missing.

Effluent nitrogen concentrations are presented in Figure 49. The nitrate and total nitrogen concentrations decreased during the trial (as temperature increased). The total nitrogen concentration was maintained below 7 mg N/L reaching as low as 3 mg N/L during the warmer summer period although no external carbon source was used. Effluent ammonium concentration peaked in week 21 at 1.7 mg NH_4-N/L but was on average 0.7 mg NH_4/L during the high flow period (w.10 to w.23). The return sludge flow rate increased from 2xQ_{IN} to 4xQ_{IN} in week 15, resulting in returning more nitrate to the pre denitrification zone, however it is not clear what effect this had in the nitrate removal.

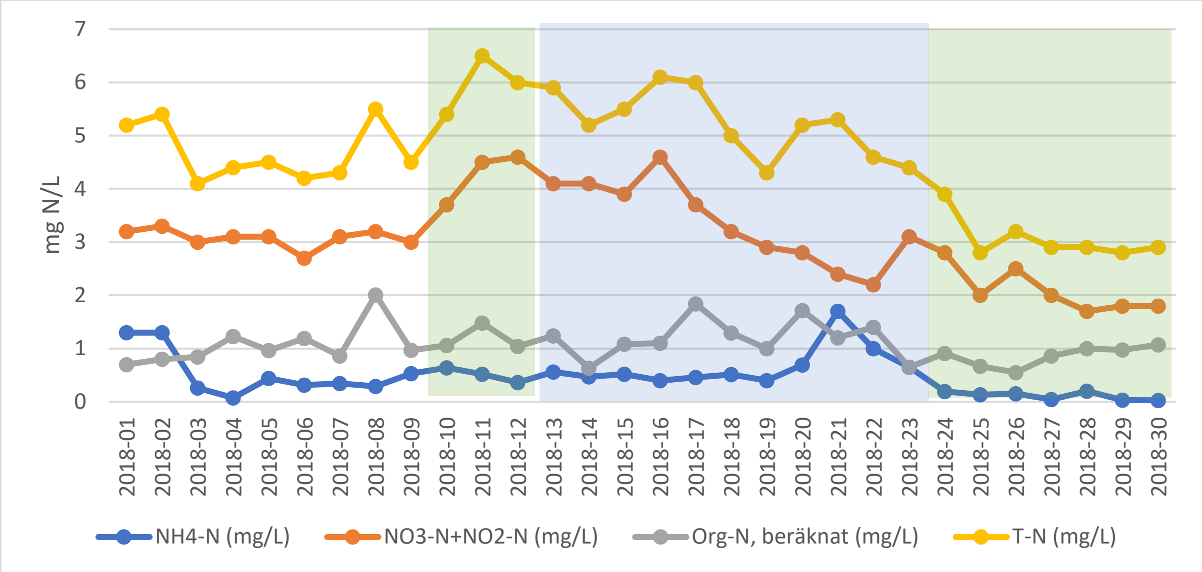


Figure 49. Effluent nitrogen. Carbon source was not added from the mid of week 10. Not restarted until week 36. Flow was high between week 13 and 24. Green box – no carbon source. Blue box – no carbon source and high load.

The effluent phosphorus concentration was kept low (below 0.2 mg P/L) during most of the BB1-trial (marked with blue box in Figure 50) although only ferrous sulfate was used as precipitation chemical.

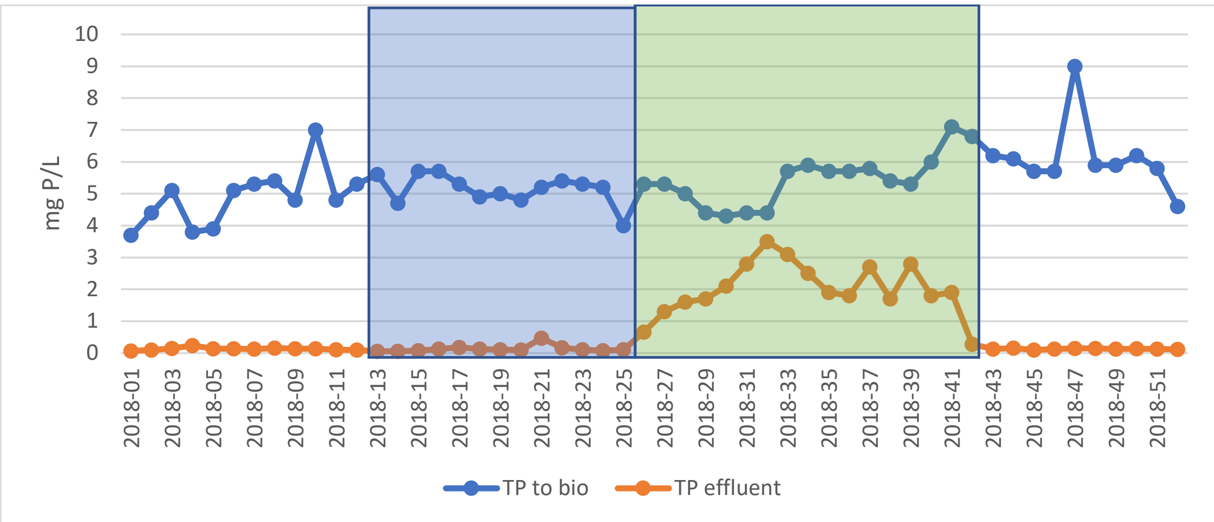


Figure 50. Total phosphours. Blue box – only ferrous sulfate. Green box – no precipitation chemical.

One weekly composite sample (w.21) showed a higher effluent concentration (0.47 mg P/L) (Figure 51).

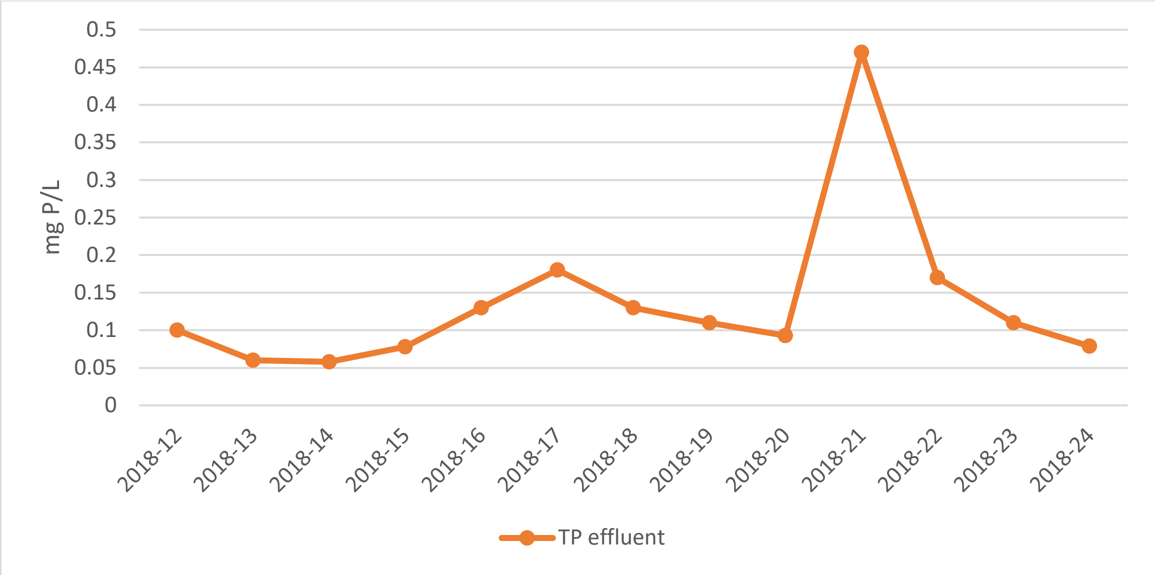


Figure 51. Total phosphorus in effluent during the imitation of BB1 operation.

The trial indicates that the process can manage TP below 0.2 mg/L and TN below 10 mg N/L in the effluent at the higher load expected for the first treatment line without use of ferric and external carbon.

6.8 Sludge production and sludge properties

The sludge production in the treatment process (primary sludge and WAS) is an important parameter for the MBR treatment process as well as input data for the sludge pilot. Table 23 shows some of the sludge data relevant for the MBR-process; sludge production, sludge age (SRT), and Sludge Volume Index (SVI), where data from the pilot is compared to design data for the pilot, data from the Henriksdal WWTP (annular average 2018) and design data for the future Henriksdal WWTP according to SFA.

The primary settlers in the Henriksdal WWTP are much more efficient than the one in the pilot. Therefore, the production of primary sludge in the pilot corresponds to about half of the total sludge production whereas it makes up 66-75% of the total sludge production in the full-scale plant. The total sludge production in the pilot is significantly higher than the design data. This is due to higher incoming load compared to design (see section 3.3). The WAS-production in the pilot is almost twice as high as designed, which corresponds to the high BOD-load on the biology (due to high incoming load and poor reduction over primary settlers). The high WAS-production causes a short SRT (see section 3.3).

The difference in sludge production and composition between the pilot and full-scale plant will affect results from thickening, digestion and dewatering of mixed sludge.

Similar to the results from 2016 and 2017, the SVI-value in the pilot was higher than in the full-scale plant with conventional activated sludge (CAS) process. This is expected since the membranes retain all sludge, also the floating and foaming sludge, in the process. In a CAS-process, sludge that does not settle will be washed out from the secondary clarifiers and a natural selection of sludge with good settling properties is achieved (at least in theory).

Table 23. Sludge data from the pilot year 2018 compared to design data for the pilot, data from the Henriksdal WWTP 2018 and design data for the future Henriksdal WWTP (SFA-design).

Parameter	Pilot data 2018	Pilot design data*	Henriksdal WWTP data 2018	SFA-design 2040
WAS production (kg SS/d)	**14.1**	**8.8**	**23 800**	**59 000**
Part of total sludge production (%)	**47**	**34**	**23**	**34**
VSS in WAS (% of SS)	**73**	**63**	**68**	**63**
Fe in WAS (% of SS)	**6.3**	**-**	**11**	**-**
PS-production (kg TS/d)	**16.1**	**17.2**	**80 300**	**117 000**
Part of total sludge production (%)	**53**	**66**	**77**	**66**
VS in PS (% of TS)	**88**	**80**	**78**	**80**
Total sludge production (kg TS/d)	**30.2**	**26.3**	**104 100**	**176 000**
Total sludge age, SRTtot (d)	**24**	**28**	**11**	**28**
Aerated sludge age, SRTox (d)**	**7.5*** **	**7**	**5**	**7**
SVI jan-jun (mL/g)	**211**	**-**	**170**	**-**
SVI jul-dec (mL/g)	**138**	**-**	**141**	**-**

*pilot design data is the scaled down SFA-design data (1:6700)

**yearly average, the aerated volume is adjusted based on water temperature using the flex-zones.

***including membrane tanks, without membrane tanks SRTox = 6.0 d

The function and efficiency of the membrane filtration depends on several parameters, amongst them the sludge properties. Therefore, weekly analyses of time to filter (TTF) and colloidal TOC in filtered sample (cTOC) was done, as suggested by the membrane supplier. Data on sludge properties, membrane performance (transmembrane pressure, TMP) and temperature is shown in Figure 52.

Like previous years, no strong correlations between any of the sludge properties and the membrane performance (presented as TMP) was found. However, a moderate correlation (R^2=0.61) between TTF (normalised based on TSS) and TMP was found. This indicates that TTF could be a good parameter to monitor in relation to membrane operation. In addition, weak correlations between TMP and TSS (mg/L), VSS (% of TSS) and Fe (% of TSS) respectively was found. TTF and the Fe content in WAS gave a moderate correlation (R^2=0.53) showing that addition of ferrous increases the filterability of the sludge, thus indicating that addition of ferrous is beneficial for membrane operation.

No correlation at all was seen between TMP and SVI or TMP and cTOC. Similarly, no correlation between SVI and TTF was seen while a weak correlation (R^2=0.38) was found between the SVI and cTOC values. Thus, analysis of SVI cannot replace analysis of TTF.

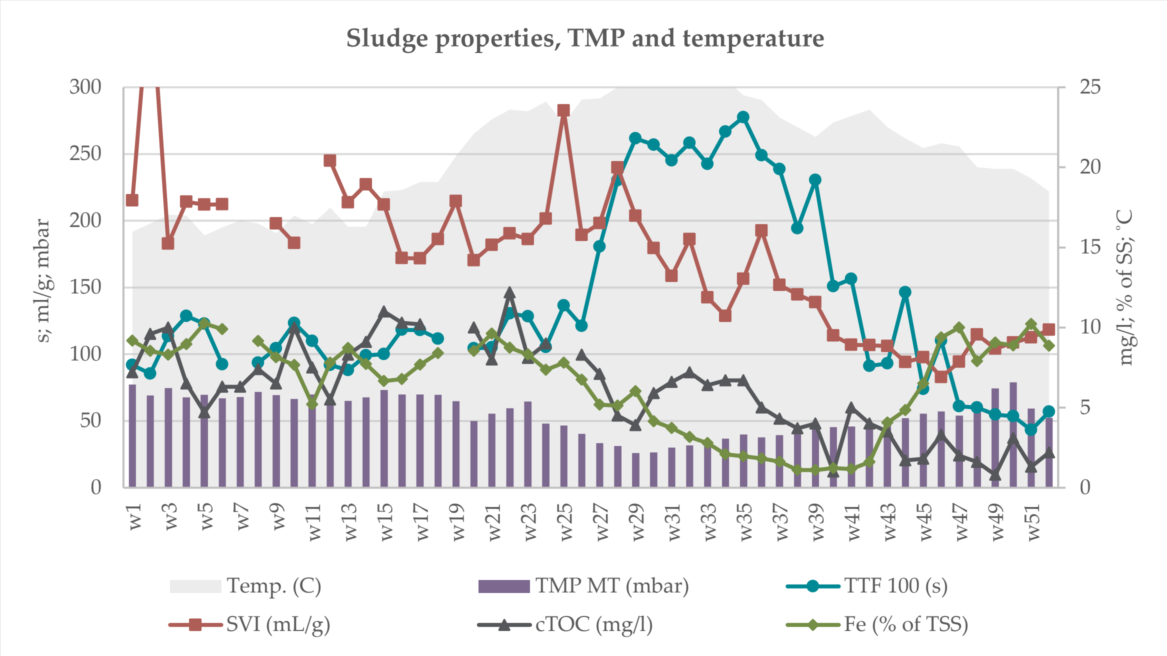


Figure 52. Sludge properties, TMP and temperature in the pilot.

In Table 24 the TSS concentration in waste activated sludge (WAS) as well as the content of iron, phosphorus and VSS is listed for the four years the pilot plant has been in operation. The design value for TSS in WAS is 10 000 mg/L. The values are quite similar from year to year, a part from the iron concentration in the sludge which decreased over the first three years due to lower Fe dose achieved by process optimization and development of precipitation strategies. The increase in 2017 is due to a lower production goal for phosphorus in the effluent, 0.15 mg TP/l compared to 0.20 mg TP/l previous years, and the decrease during 2018 is due to EBPR activity (see 6.3.2).

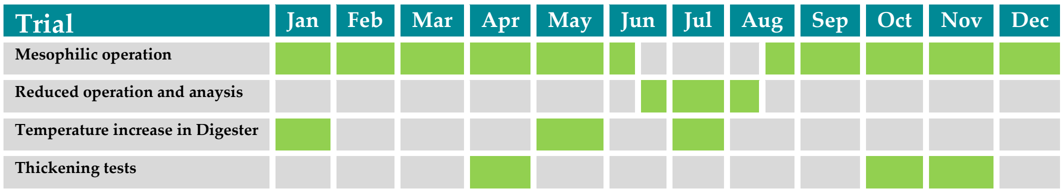
Table 24. WAS composition (annual average) in the pilot over the 5 years of operation.

Year	TSS (mg/L)	Fe in sludge (% of TSS)	P in sludge (% of TSS)	VSS (% of SS)	Fe/P in sludge (mole/mole)
2018	8480	6.4	3.3	77	1.1
n	*50*	*50*	*50*	*50*	*50*
2017	9632	10.3	3.0	71	1.9
n	*50*	*47*	*47*	*47*	*47*
2016	8126	8.3	3.4	74	1.3
n	*31*	*31*	*31*	*31*	*31*
2015	9910	10.1	3.3	71	1.7
n	*44*	*44*	*42*	*44*	*42*
2014	9263	11.9	3.1	69	2.3
n	*38*	*38*	*27*	*38*	*27*

6.9 Sludge treatment

A summary of the trials related to the sludge treatment is presented in Table 25. During 2018 the operational mode for the digester was mesophilic conditions with some shorter tests at higher temperatures to test the heating equipment. Separate tests of thickener operation have also been conducted.

Table 25. Trials related to the sludge treatment.



The sludge pilot, comprising thickening of mixed sludge (PS and WAS), digestion and dewatering of digested sludge, was taken into operation in September 2017. The year 2018 was characterized by problem solving and optimization. During week 25-33 the sludge pilot operation and analysis was reduced due to summer holiday period.

In January the heating system was tested to ensure that thermophilic conditions could be reached, but the temperature stopped at 50˚C. After several improvements, including insulation and installation of a new heat element, 55˚C could be reached in July. The thickener suffered from various recurrent problems resulting in discontinuous operation of the digester. The problems included clogging of the mesh in the thickener drum, problems with the output of thickened sludge into the collecting trough, malfunctioning level transmission in the trough, inaccurate online TS-measurements causing random polymer dosing, air in the flush water (effecting the polymer concentration and the cleaning of the drum) etc. The trough was reconstructed in July which improved the function slightly. During the spring the function of the thickener was optimized and in the autumn tests of thickening of WAS from the pilot (MBR) and Henriksdal (CAS) were done as part of an MSc-project.

Thickener

The thickener was operated discontinuously, as can be seen in Figure 53, over the year due to several technical problems described in the section above. Figure 54 shows the weekly average values for the flow and TS in and out of the thickener. The thickener was designed for a mixed sludge average inflow of 70 L/h and a TS-concentration of 6.7% in the thickened sludge. Due to repeated stops of the thickener, the weekly average inflow was <60 L/h 28 weeks out of the 41 when it was in operation. The TS of the thickened sludge varied a lot over the year. Out of 65 grab samples analysed, 11 had a TS-concentration >6.0%.

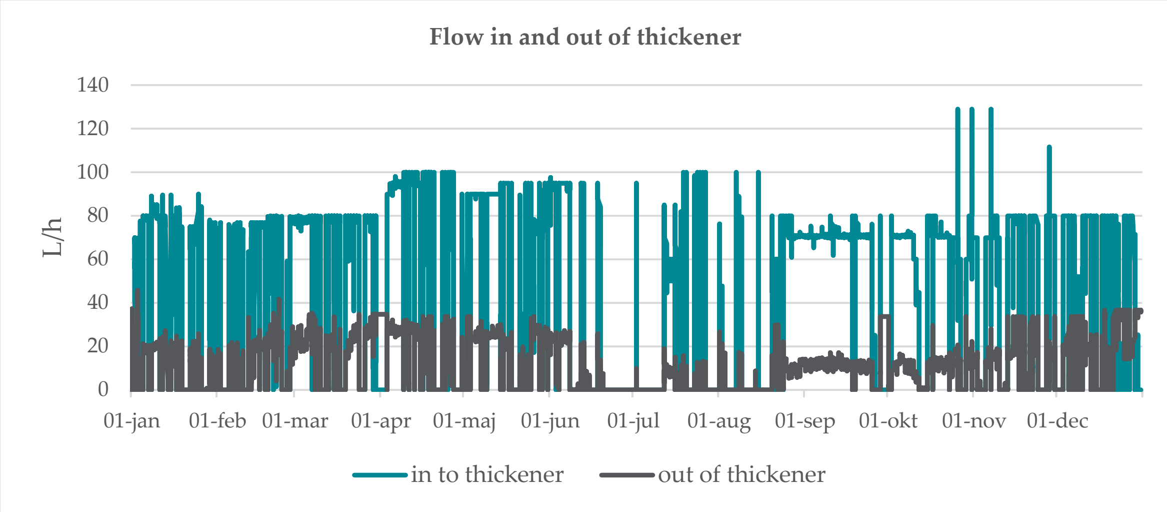


Figure 53. Flow in and out of the thickener. Hourly averages.

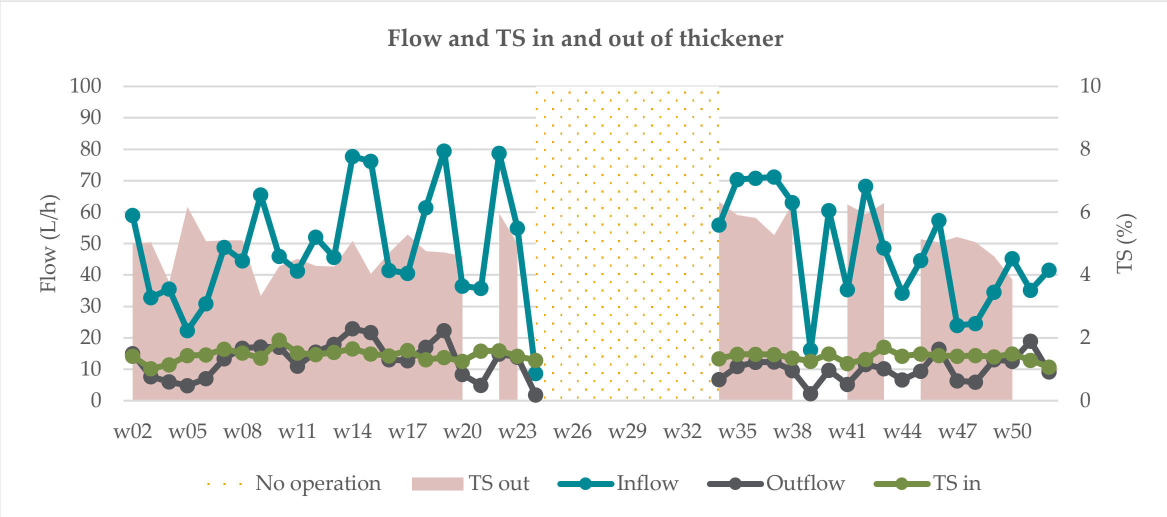


Figure 54. Flow and TS-concentration into and out of the thickener. Weekly averages.

The quality of the reject water from the thickener was monitored over time. Results are shown in Table 21. Data was used for optimisation of polymer dose and operation as well as to control the internal load on the biological process derived from recirculation flows.

Table 26. The quality of the reject water from the thickening unit.

Parameter	Average value	Min value	Max value
SS (mg/L)	3 300	300	16 000
VSS (%)	77	61	86
Tot-P (mg/L)	62	7	190
PO_4-P (mg/L)	22	2	51

An MSc-project study was conducted in the autumn with focus on the quality difference between MBR and CAS sludge and the corresponding effect on the thickening process. The settling properties (SVI) and filterability (TTF) of the MBR and CAS sludge respectively was analysed and thickening trials were performed where the TS in thickened sludge and the SS in the reject water were the main indicators. The study showed that the MBR sludge displayed slightly lower SVI-values but poorer filterability compared to the CAS sludge. The poor filterability could be caused by the higher TS-concentration (1% compared to 0.7%) or the presence of more fine particles in the sludge. Thickening of MBR sludge resulted in higher TS in the thickened sludge compared to CAS sludge. In addition, it also resulted in higher SS in the reject water, which was expected since the MBR sludge theoretically should contain more fine particles than CAS sludge. The methods, results and conclusions are described in more detail in the thesis report (Jirblom, 2019).

Thickener optimization trial

A thickener optimization trial was conducted in mid-April. A test plan was created using the Umetrics software MODDE. Three manipulated variables were used; polymer dose (range 6 to 16 g/kg TS), frequency of the drum rotation (range 25 to 40 Hz) and the angle of the drum related to the horizontal plane (two settings tested; high slope, 7.8 degrees and low slope, 6.3 degrees). Although three different dosing points are available only the dosing point closest to the thickener was used in order to get a reasonable amount of test runs. Two non-manipulated factors were also considered; TS of the mixed sludge entering the thickener (varying between 1.33% and 1.47%) and the flowrate of sludge to the thickener (100 L/h). The test plan included 11 experiments carried out in random order during two days. For each setting (test) the TS of thickened sludge and the suspended solids of the reject water was analysed. The test settings and results are presented in Table 22.

Experiment N2 and N6 both had the highest TS in thickened sludge and low SS in the reject water. These two experiments both combined the highest polymer dosage (16 g/kg TS) and the lowest rotational frequency of the drum (25 Hz) but different angle of the drum. The varying angle of the drum had minor effect (compare the upper and lower graphs in Figure 55). Looking at the SS in the reject water, the dominating factor was the polymer dosage, where a high dosage resulted in lower SS in the reject and vice versa. For the thickened sludge a combination of high polymer and low frequency gave increased TS. If only comparing the different slopes of the drum a lower slope more often resulted in higher TS than a higher slope when the other settings were the same.

Table 27. Settings and results from the 11 experiments on thickener optimization. Red=bad, blue=good results.

Exp Name	Run Order	TS in (%)	Qin (L/h)	Polymer dose (g/kg TS)	Frequency drum (Hz)	Angle drum	TS thickened sludge (%)	SS in reject (mg/L)
N4	1	1.46	100	16	40	Low	4.71	343
N10	2	1.47	100	11	32.5	Low	4.84	1195
N7	3	1.33	100	6	40	High	3.99	4022
N8	4	1.45	100	16	40	High	3.79	337
N9	5	1.47	100	11	32.5	Low	5.10	1429
N5	6	1.45	100	6	25	High	4.61	3383
N2	7	1.39	100	16	25	Low	5.32	394
N3	8	1.42	100	6	40	Low	4.62	4127
N6	9	1.42	100	16	25	High	5.32	481
N1	10	1.42	100	6	25	Low	4.34	3480
N11	11	1.42	100	11	32.5	Low	4.99	1567

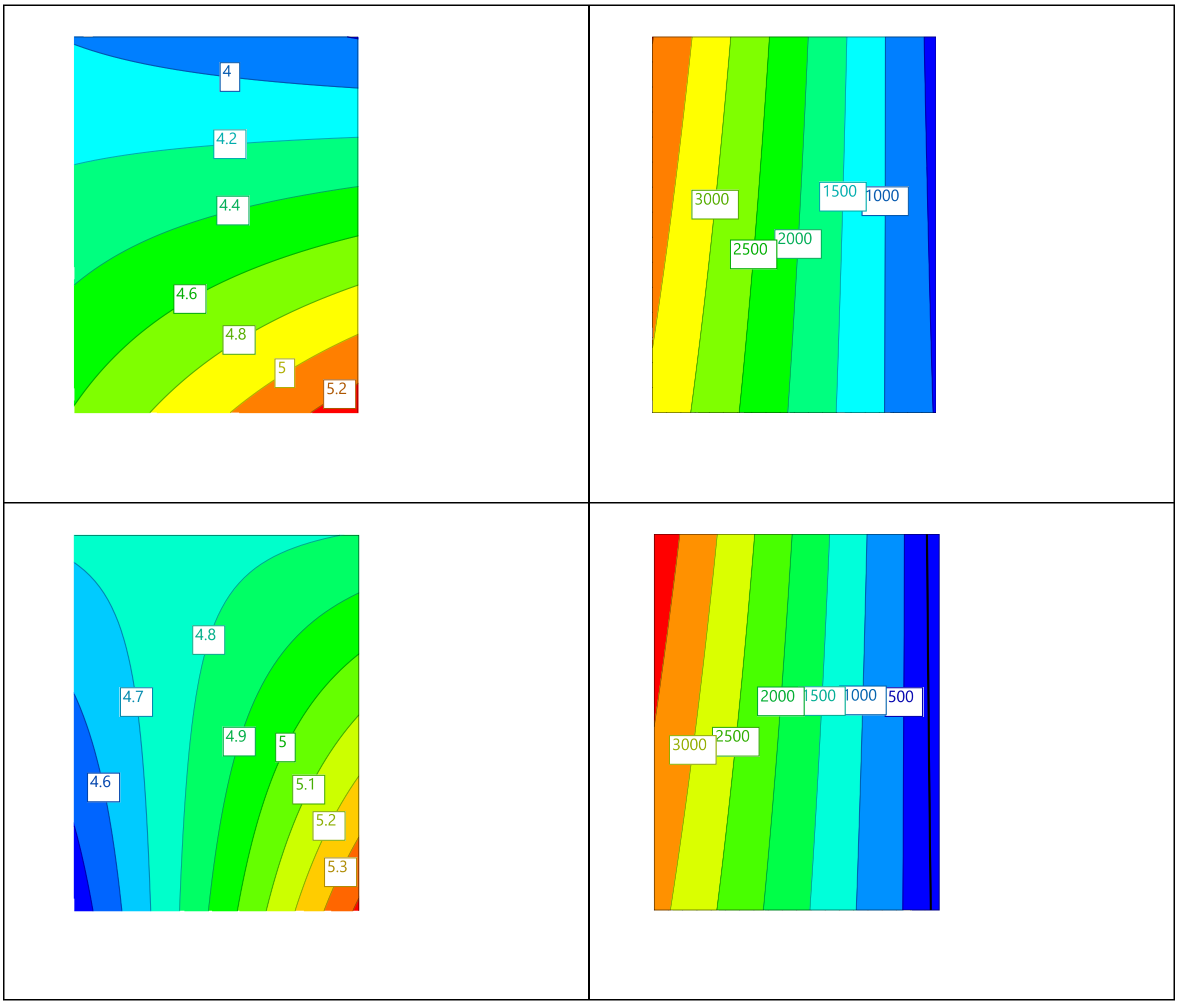


Figure 55. Response contour plots from the MODDE software showing TS in thickened sludge as % (left) and SS in reject water (right) in relation to Frequency (Hz) on the y-axis and polymer dosage on the x-axis. The two upper plots are for the tests with a higher slope of the drum and the two lower plots are for the low slope of the drum.

Digestion

The 5.9 m^3 digester was operated under mesophilic conditions the whole year, with exception of a few shorter time periods when the thermophilic heating system was tested. The unintentional intermittent operation of the thickener led to uneven loading of the digester, 4-23 kg VS/d, and varying retention times, 9-40 d (weekly average values). The target values were 20 kg VS/d and a retention time of 13 days. Results from the mesophilic digestion are shown in Figure 56 and previously presented in section 3.3, Table 3.

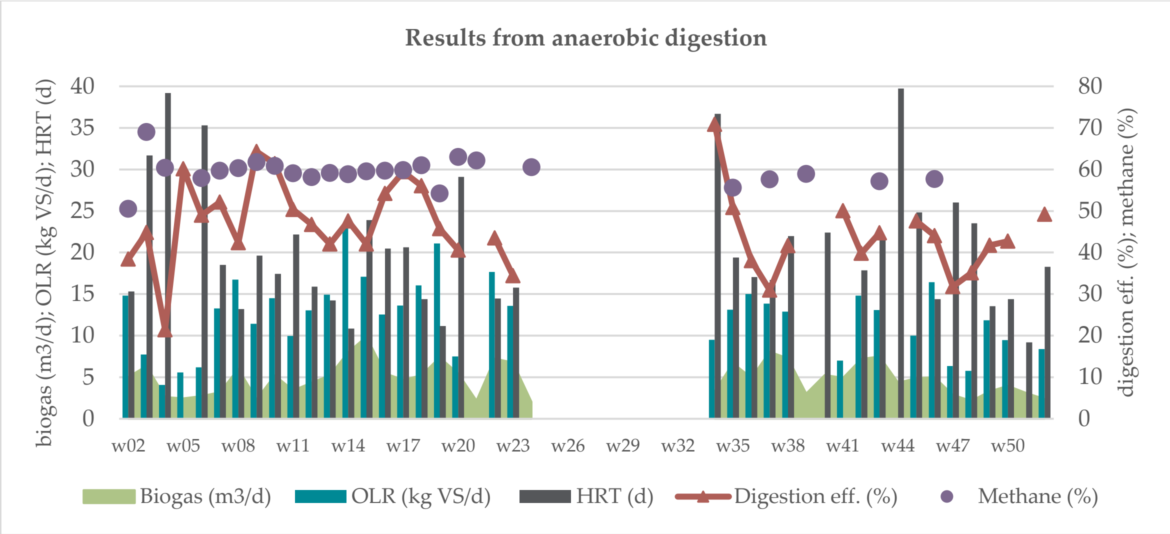


Figure 56. Data from mesophilic digestion of thickened mixed sludge. Weekly averages.

Key performance indicators for the anaerobic digestion process is shown in Table 23. They are all in line with common design figures (Metcalf & Eddy, 2014).

Table 28. Average values on digestion efficiency.

Parameter	Average value
Specific OLR (kg VS/m³, d)	2.1
Digestion efficiency (% $VS_{destroyed}$ of VS_{in})	46
Specific biogas production (m³/kg $VS_{destroyed}$)	1.0
Specific biogas production (m³/kg VS_{in})	0.43
Specific methane production (m³/kg VS_{in})	0.25

Using the specific biogas production key figure 1 Nm³/kg $VS_{destroyed}$ (Metcalf & Eddy, 2014) the theoretical biogas production was calculated based on weekly average data on VS (kg/d) in and out of the digester. This was done primarily to control the gas meter function. As can be seen in Figure 57, the theoretical values correspond well to the measured values.

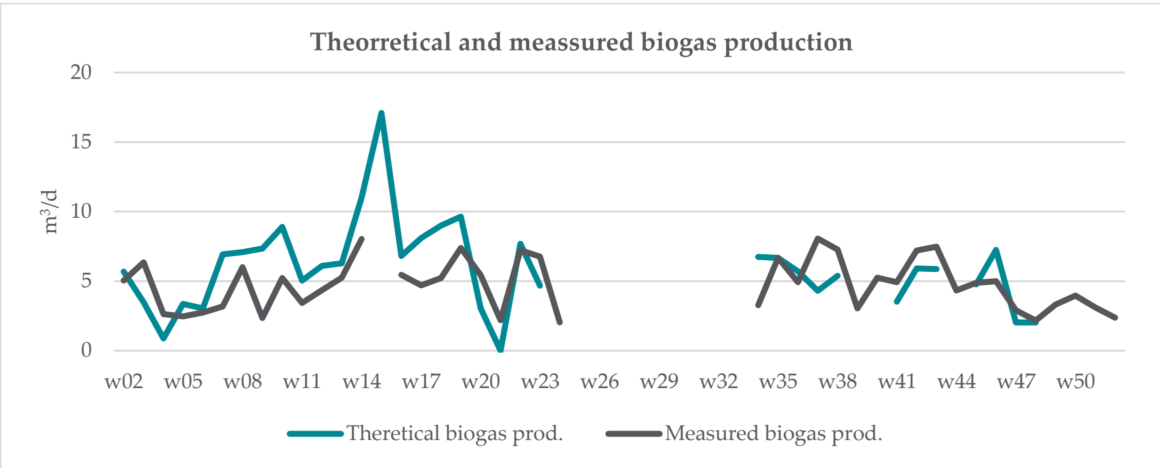


Figure 57. The calculated theoretical and the actual measured biogas production.

Based on the available data from digester operation, there was no clear correlation between the biogas or methane production and the OLR or HRT (Figure 58). This is most likely due to the discontinuous operation of the digester.

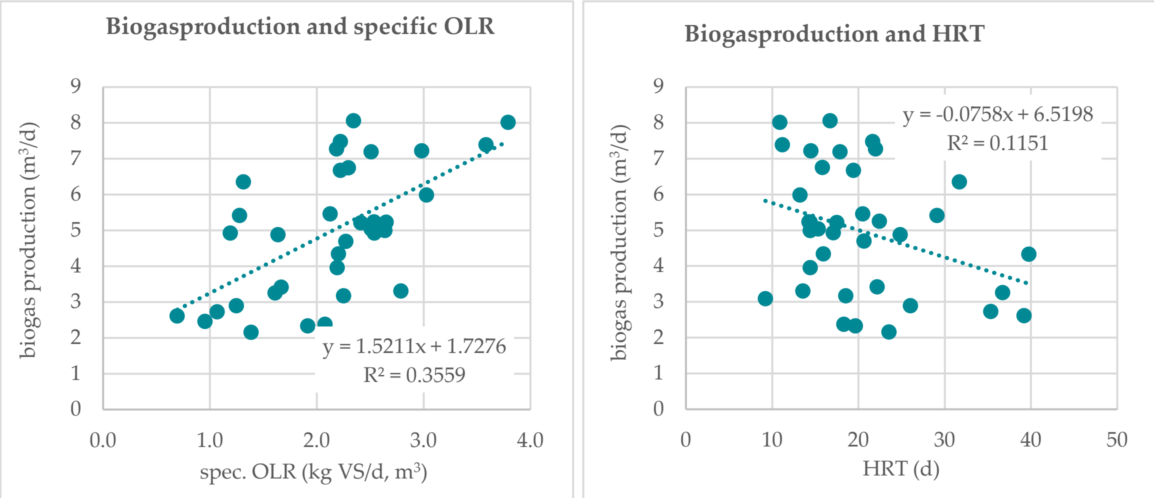


Figure 58. Biogas production as a function of the specific organic loading rate (OLR) and HRT. Weekly averages.

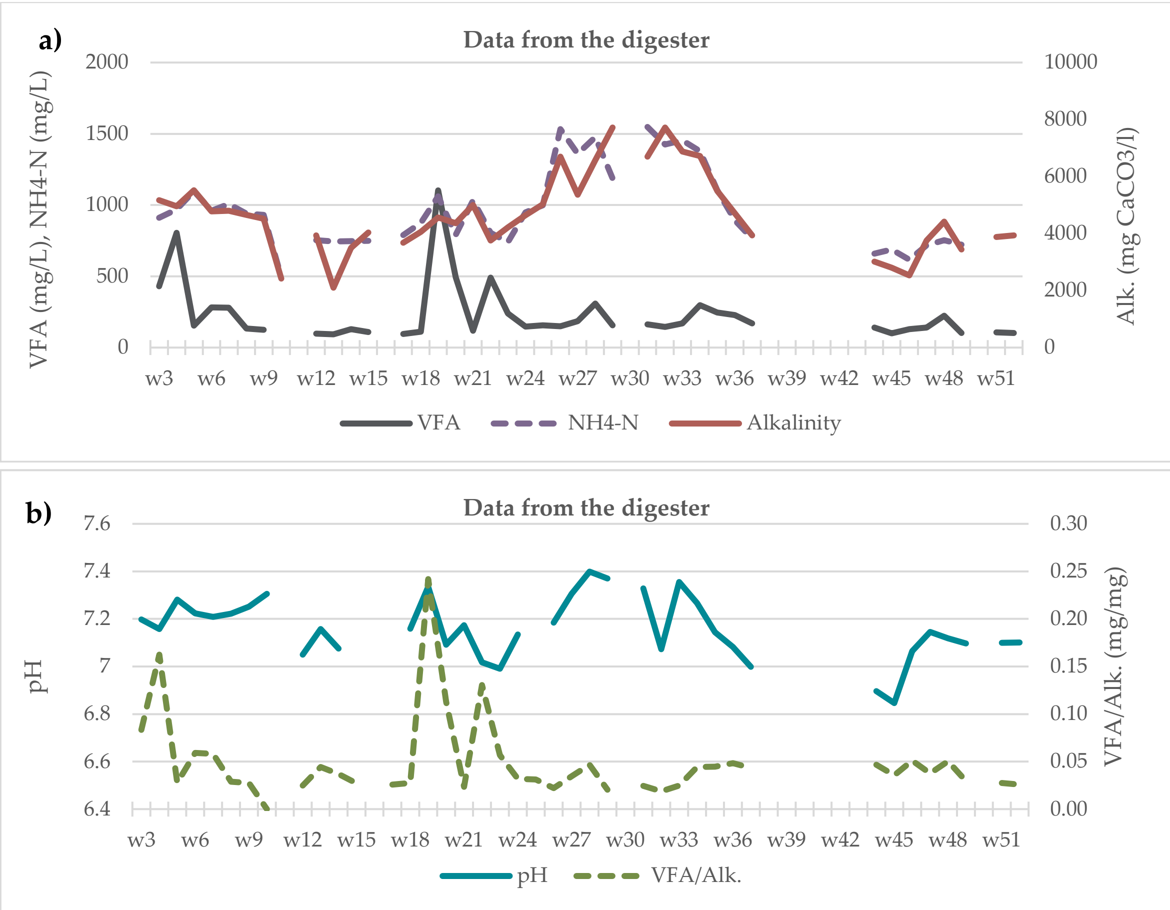


Figure 59. Data from the digester; a) shows VFA, NH_4-N and alkalinity and b) shows pH and the VFA/alkalinity ratio. Grab sample once per week.

The analyses of pH, alkalinity and NH4-N (Figure 59) gave stable results during the weeks with relatively continuous feed (all except week 24-33). During the period with limited feeding (w 24-33), alkalinity and ammonia increased while pH and VFA did not change significantly. VFA varied more between the samples but in general the occasional high values coincided with high OLR-values.

Dewatering

The dewatering unit was in operation with exception of some shorter periods during the beginning of 2018 and summer holidays (week 24 to 35). The digested sludge was dewatered to TS concentrations varying between 20% and 33% (Figure 60).

As weekly average the flow to the dewatering ranged from 6 to 17 L/h digested sludge. Different polymer dosages were tested, from 8 to 20 g/kg TS.

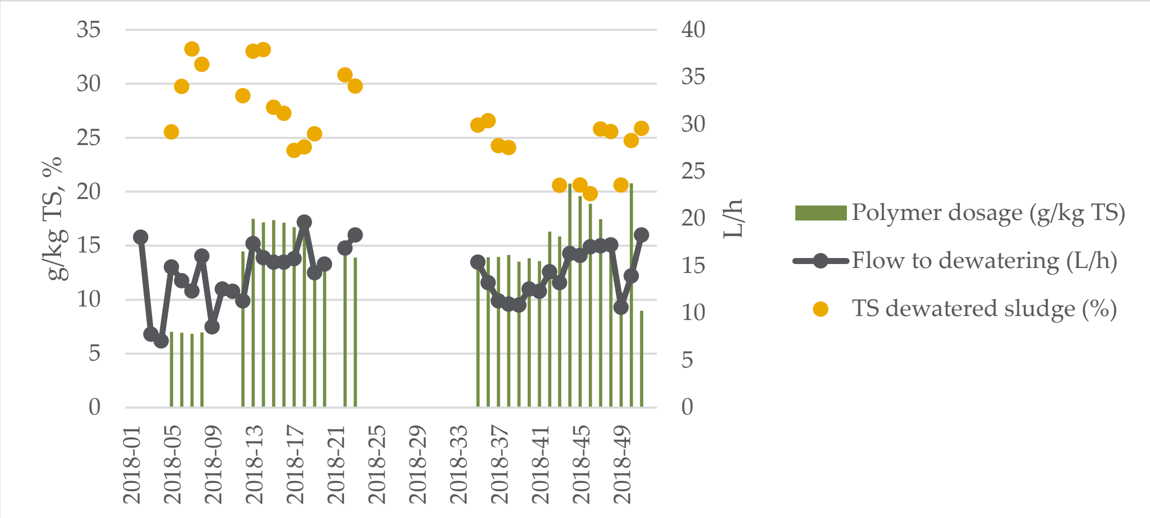


Figure 60. Flow of digested sludge to dewatering, polymer dosage and resulting TS concentration after dewatering.

Varying quality of the reject water from the dewatering unit has been one issue throughout the year. If suspended solids in the reject water from dewatering of sludge was below 2000 mg SS/L the quality was considered acceptable and returned to the water treatment line. Obtained quality concentrations in reject water when suspended solids were less than 2000 mg SS/L are presented in Table 24.

Table 29. Reject water quality when dewatering unit operated with less than 2000 mg SS/L in reject water.

	Min	Max	Median
SS (mg/L)	110	2000	580
VSS (mg/L)	65	1400	380
BOD_7 (mg/L)	110	980	170
NH_4-N (mg N/L)	88	690	400
PO_4-P (mg P/L)	0.36	63	1.9
Total phosphorus (mg P/L)	5.4	160	20.7
Total nitrogen (mg N/L)	83	770	510
Iron (mg Fe/L)	17	150	51

Dewatered digested sludge

Monthly composite samples of the dewatered digested sludge were collected and sent for analysis of TS, VS, pH, nitrogen, phosphorus, chlorine, and 15 different metals. In addition, multiple organic parameters and three more metals were analysed each quarter, including Polybrominated diphenyl ethers (PBDE, 24), Triclosan, Polychlorinated biphenyls (PCB, 7), Polycyclic aromatic hydrocarbons (PAH, 6), organotin compounds (10), Phenols (19), Perfluorooctanoic acid (PFOA), Perfluorooctanesulfonic acid (PFOS) and Per- and polyfluoroalkyl substances (PFAS).

Due to some interruptions in the operation of the dewatering only 8 monthly samples were analysed during 2018 (Feb to May and Sept to Dec). The extended analysis was made on samples from February, May and November.

A summary and comparison to Henriksdal values will be presented in the next yearly report.

6.10 Resource consumption

Resource consumption in the pilot as yearly average is presented in Table 25. A comparison with the SFA design was made, however pilot values contain great uncertainties. Problems with pumps and air in the pipes, difficulties in manually reading levels and degradation of some chemicals are some reasons why consumption is difficult to measure.

External carbon source consumption 2018 (0.81 kg COD/d and 0.28 g COD/g N) was similar to 2016 when Brenntaplus was used (0.7 kg COD/d and 0.33 g COD/g N) and much lower than 2017 (2.14 kg COD/d and 0.84 g COD/g N) which was the year methanol dosage started. Effluent nitrate was approximately the same as yearly average 2018 as previous years (3 mg NO_3-N/L). The reduced consumption of methanol from 2017 to 2018 is likely because external carbon dosage was not in use for 26 weeks (from week 10 to week 36) as part of separate trials. On yearly average the methanol consumption was 4.8 g COD/m^3 (148 kg COD/year). If excluding week 10 to week 36 when the methanol dosage was deliberately off, the dosage was on average 10 g COD/m^3 corresponding to about 297 kg COD/year, or 45% of SFA design. Other aspects contributing to reduced consumption was less problems with the methanol pump, possibly improved performance once the bacteria had time to adjust to methanol and a change of dosing point.

The iron consumption was calculated excluding the trial without iron dosage (week 27-41). The values for 2018 (1.09 kg Fe/d and 1.67 mole Fe/mole P) was lower than 2017 (1.6 kg Fe/d and 2.1 mole Fe/mole P) and lower compared to the SFA-design. Effluent phosphate concentration was in line with target concentration of 0.15 mg P/L.

The chemical consumption for membrane cleaning showed that sodium hypochlorite was 10% higher than SFA design although the number of backpulses were reduced from 8 to 6 (corresponds to 83 % of design consumption) and only one RC was performed instead of two. One explanation to not having lower consumption is the high inflow (50% higher than design for almost 3 months and then 10 % higher than design – in total 9,6% higher than 2017 when the sodium hypochlorite consumption was 97% of the scaled SFA design) which means increased frequency of MC.

Compared to cleaning according to the membrane supplier recommendation[3], the amount of sodium hypochlorite used was 68% of recommended.

For the citric acid, as only one out of two membranes were cleaned with citric acid, the consumption was 87% of the SFA design value (43% of design for two membrane tanks). Compared to 2017 the citric acid consumption was reduced with 7%. Compared to cleaning according to the membrane supplier recommendation[3], the amount of citric used was 83% of recommended.

The main efforts on reducing the amount of chemicals for MC have been focused on the oxalic acid consumption. Compared to 2017 the consumption was reduced with 25%. As the oxalic was not included in the SFA design consumption was only compared to the membrane supplier recommendation[3]. The consumption 2018 was then 68% of recommended.

Table 30. Resourceconsumption during 2018.

Resource	Unit	Value pilot	SFA design	Value pilot/scaled SFA design
External carbon source (methanol)	kg COD/d	0.81	12 000	45%
	g COD/g N	0.28*	-	-
Iron (FV+BR4+BR6)	kg Fe/d	1.09**	10 000	73%
	mole Fe/mole P	1.67**	2.8	60%
Citric Acid (51%)	L/d	0.091***	1 400	43%
Sodium hypochlorite (12%)	L/d	0.197***	1 200	110%
Oxalic acid (8%)	L/d	0.377***	-	-
Aeration biology	m^3/d	1 260	1 488 000	567%
Aeration MT	m^3/d	676	2 660 000	170%

* NO_3-N removed in total, from inlet to effluent.
**Excluding week 27 to 41 when no iron was dosed.
*** Number of MCs with each chemical multiplied with time settings and number of backpulses using design flowrate of chemical. Measured consumption for the RCs preformed with each chemical.

When comparing the impact of RC, two RC cleanings per year would mean that 25% of total acid consumption and 45% of total sodium hypochlorite consumption would be used for RC. RC has a larger impact on yearly sodium hypochlorite consumption than on the acid consumptions.

The air flow to the biological treatment line was 567% higher in the MBR-pilot than in the SFA design. This is partly due to the water depth of the tanks which is 4.7 m compared to 12 m in full scale. However, when calculating the theoretical air flow demand in the pilot based on the yearly average BOD- and TN-load to the biology and the tank geometry the result was 1 027 Nm^3/d (462% of scaled SFA-design) which is lower than the measured 1 260 m^3/h.

Also, the air flow to the membrane tanks were higher in the MBR-pilot than the scaled SFA-design air flow. This is explained by the difference in number of membrane tanks in operation. In the full scale plant, the number of tanks/trains in operation at average flow is 8 of 12 (67%) while in the pilot it is 2 of 2 (100%).

[3] Membrane supplier recommendation for cleaning: MC backpulsing 120 s + 8 x 30 s (every 173 m^3 for hypo and every 345 m^3 for acid) and RC two times per year (2 with hypo and two with acid for each tank).

6.11 Mapping of micro pollutants

Micropollutants and microplastics are commonly detected in WWTPs, both in water and sludge samples from various locations in the process. When upgrading to MBR, the distribution of these compounds is likely to change, mainly due to the stricter separation of water and solids over the membranes in the MBR but also due the higher solids concentration, and possibly higher biological activity, in MBR activated sludge compared to the CAS activated sludge.

Due to the possible distributional change of micropollutants and microplastics when upgrading to MBR, which might affect the possible future compilation of micropollutant removal regulations, there is a need to further study the fate of micropollutants and microplastics in the MBR process, compared to the CAS process. Funding for this specific study has been received by the Swedish Water and Wastewater Association.

6.11.1 Method

A total of four separate sampling campaigns were planned in the study. The first two campaigns were conducted in the autumn of 2017 and in the spring of 2018, including samples only from the MBR pilot process. The remaining two sampling campaigns are planned for 2019 and are planned to also include samples from the Henriksdal WWTP CAS process, for comparison. Both processes treating the same influent wastewater.

Both water phase and sludge phase samples were included, 3 water phase sampling points (IN, PTW and EFF) and 3 sludge phase sampling points (PS, WAS and DDMS), see Figure 61. Water samples were taken as composite weekly samples and sludge samples were taken as daily grab samples and then mixed.

The samples were analysed regarding pharmaceuticals, antibiotics, hormones, microplastics, PFAS and chloro-organic halogens, which all are commonly found in WWTPs. Concentrations of chloro-organic halogens are usually neglectable but here of interest since the membranes are being cleaned with sodium hypochlorite which has been shown to be a source for different chlorinated compounds (Ma *et al.* 2013).

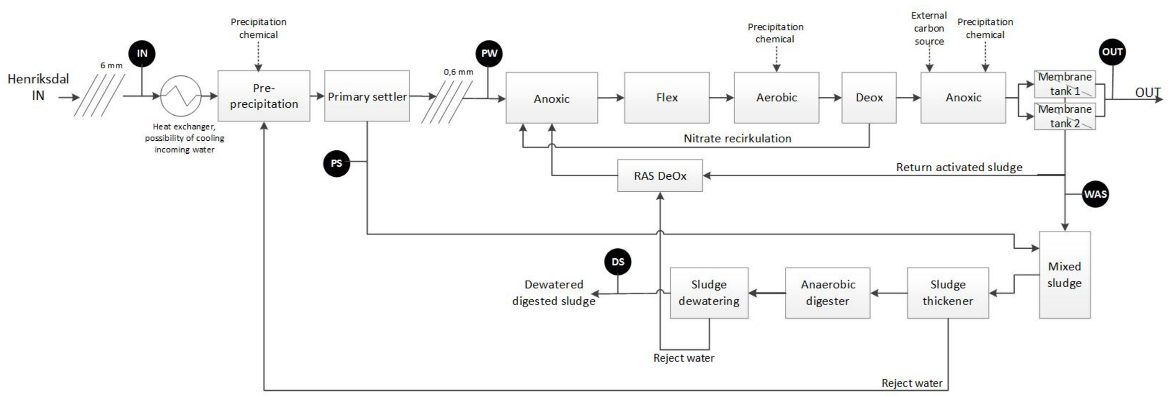


Figure 61. Process scheme with sampling points (IN, PTW, EFF, PS, WAS, DDMS).

Table 26 shows from which sampling point samples were taken for each analysed parameter. The decision to not include all sampling points for all parameters was taken in internal discussions between the project group and the laboratories conducting the analysis.

Table 31. Sampling points for each analysed parameter.

	Water samples			Sludge samples		
Parameter/Sample	**IN**	**PTW**	**EFF**	**PS**	**WAS**	**DDMS**
Pharmaceuticals	X	X	X	X	X	X
Antibiotics	X	X	X	X	X	X
Hormones	X		X			X
Micro plastics		X	X			
PFAS	X		X			X
AOX/EOX	X	X	X	X	X	X

6.11.2 Results

The results from the 2017 sampling campaign was presented in the yearly report of 2017 (Andersson *et al.* 2019) and also in the Master Thesis report connected to this specific study (Murad, 2018). The samples from the 2018 sampling campaign were frozen after sampling and stored for later analysis together with the 2019 samples. Therefore, no new results are presented in this report. A full report from all four sampling campaigns will be presented in the 2019 yearly report.

7 Field trip to five MBR plants with strict effluent requirements

Since there are no municipal WWTPs with membrane technology in operation in Sweden, it is important to learn from other plants in other countries. With this as background, five MBR WWTPs in the USA, were visited in March 2018. All visited plants were selected to be as similar as possible to Swedish WWTPs concerning climate, effluent regulations and recipient. The outcome from the visits are presented in detail in a Svenskt Vatten research report (Westling & Andersson, 2019) but the main experiences are presented below.

The MBR process was generally working well for all visited plants. The operation was stable, and the treatment plants are complying with their effluent regulations. However, there are some common general challenges that several of the visited plants were experiencing. It was difficult to find material for decking of the membrane tanks that is light enough to be lifted but at the same time stable enough to walk on. Corrosion has been noted on equipment handling hypochlorite, but not if the equipment is made entirely of plastic material. Since the hypochlorite is degrading with time storage should not exceed one month of usage. All visited plants have experienced problems with foaming in the biological treatment step, but this has been reduced by installing sprinklers containing water or a chlorine solution. For some plants, the power of the crane used to lift membranes had to be increased, since wet, fouled membranes weighed more than initially expected. The membranes in all plants were continuously cleaned using hypochlorite and citric acid, based on a specific cleaning schedule. After some time, the plants have adjusted the cleaning schedule to their specific needs, and they are all satisfied with the cleaning effect on the membrane capacity. The two plants continuously monitoring effluent phosphate have both noted an increase in effluent phosphate concentrations in connection to membrane cleaning with citric acid. This effect has also been noted within this pilot project. The main focus of operation at the visited plants was to comply with the effluent requirements and not much focus had been put on increasing the resource and energy efficiency, such as chemical used for phosphorus and nitrogen removal and energy used for aeration in the biological treatment step and in the membrane tanks.

8 Conclusions

The fifth year of the pilot plant operation of the future Henriksdal WWTP process has been completed. The water treatment line was in operation throughout the year, without any longer disturbances. Several shorter trials have been conducted and some longer trials and measurement campaigns will continue into the next year. The sludge treatment line, although with discontinuous operation, has provided some initial baseline data for mesophilic conditions and the equipment has been modified to ensure stable operation also at thermophilic conditions in future tests.

The most important conclusions and results from project year five are listed below:

- In general, and as concluded previous years, operation of the water treatment line shows stable effluent concentrations below target values of 5 mg N/L and 0.15 mg P/L for long periods.
- The membrane permeability has continuously maintained good performance, above 300 L/(m^2·h·bar) most of the year compared to reference value of 200 L/(m^2·h·bar) which is considered good according to the supplier.
- Effluent concentrations below limits of 10 mg N/L and 0.2 mg P/L could be met during a trial with 150% inflow compared to design and without the use of external carbon source and without ferric chloride addition.
- Attempts to reduce resource consumption related to membrane operation, both aeration in the membrane tanks and chemicals used for membrane cleaning, has been in the spotlight this year. Even though the trials are not finished, and will continue in 2019, results indicate that there are large potential savings in both chemical and energy use when operating the membrane tanks, without risking membrane capacity.
- Although measurements of chemical consumption are difficult and include great uncertainties estimations indicate lower consumption of both external carbon source and precipitation chemicals compared to the SFA design.
- Visual inspection of the membranes was conducted before and after recovery cleaning and membrane fibers were sent for membrane autopsy. The membranes were in good condition, the foulant before recovery cleaning contained mainly iron, some organic material and trace amounts of calcium phosphate. After recovery cleaning most fouling had been removed.
- When comparing thickening of MBR sludge with thickening of CAS sludge, no clear difference was observed.
- To obtain a high TS of the thickened sludge, the polymer dosage was more important than adjusting the thicker unit operational settings.
- Measurements of nitrous oxide emissions from the water treatment showed that 0.07 % of the NH_4-N load was converted to nitrous oxide.
- Measurements have shown high EBPR activity in periods. EBPR is present although iron is dosed to the process but increased as iron dosage was stopped. EBPR is not enough to manage the total phosphorus removal required to reach target effluent concentrations.
- Special attention has been given to reduction of oxalic acid and by the end of 2018 the maintenance cleaning had been reduced to one cleaning per 8 weeks compared to design 1 cleaning per week while maintaining good permeability.

9 Further studies

Some of the planned activities for 2019 are to:

- Make a transition from mesophilic to thermophilic conditions in the anaerobic digester with extra sampling and monitoring of the transition period.
- After stabilized thermophilic operation, conduct a trial with reduced HRT in the digester.
- Continue the optimization of membrane cleaning chemical use.
- Change membrane operation to increase treatment capacity and reduce energy consumption (aeration).
- Trials to reach even lower effluent concentrations of nitrogen and phosphorous.
- Try out different external carbon sources (non EX-classified) for possible use during the first phase operation of the full scale plant.
- Finalization of the mapping of micro pollutants measurement campaign.

10 Related publications

Jirblom, M. (2019) *Egenskaper och förtjockningspotential hos slam från MBR- respektive CAS-process.* Master Thesis. Uppsala university, Sweden.

Taylor, S. (2019) *Utvärdering av return activated sludge deoxygenation (RAS-DeOx) i membranbioreaktor pilotlinje vid Hammarby Sjöstadsverk*. Master Thesis. Uppsala university, Sweden

Westling K., Andersson, S. L., Andersson, S. (2018) *The world's largest membrane bioreactor: Henriksdal WWTP, Stockholm, Sweden – results from pilot scale trials*, Membrane Technology Conference and Exhibition, West Palm Beach, FL, USA, March 12-16, 2018.

Westling, K. & Andersson, S. (2019) *Fem avloppsreningsverk med MBR-process och strikta reningskrav – Rapport från en studieresa i USA*, SVU-rapport 2019-5.

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Svenskt Vatten (2010a) *Avloppsteknik 2 - Reningsprocessen.* Stockholm: Svenskt Vatten AB.

Svenskt Vatten (2010b) *Avloppsteknik 3 - Slamhantering.* Stockholm: Svenskt Vatten AB.

Taylor, S. (2019) *Utvärdering av return activated sludge deoxygenation (RAS-DeOx) i membranbioreaktor pilotlinje vid Hammarby Sjöstadsverk*. Master Thesis. Uppsala university, Sweden

Tykesson E., la Cour Jansen J. (2005) Evaluation of laboratory batch tests for enhanced biological phosphorus removal. *Vatten 61*, pp. 43-50.

Westling, K. & Andersson, S. (2019). *Fem avloppsreningsverk med MBR-process och strikta utsläppskrav - Rapport från en studieresa till USA.* Svenskt Vatten Utveckling, Rapport Nr 2019-5.

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