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

## - pilot Henriksdal 2040, results from 2019

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ivl

**In cooperation with**

STOCKHOLM VATTEN OCH AVFALL

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# Preface

This report presents work performed during 2019, within the long-term pilot study trials of municipal wastewater treatment with Membrane Bioreactors (MBR) and sludge treatment with high loaded thermophilic digestion. 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.

Previous results from the project are presented in Swedish in Samuelsson et al. (2014), Westling et al. (2016) and Andersson et al. (2017) for project year 1, 2 and 3, respectively. For project year 4 and 5 the reports are in English, see Andersson et al. (2019) and Andersson et al. (2020).

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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 Million Litres 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 2019 (project year 6) 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 2019, a large focus was put on:

**Digester transition from mesophilic to thermophilic condition**

The sludge treatment line (including sludge thickening, anaerobic digestion and sludge dewatering) was operated with mesophilic operation until mid-March 2019. For 18 days the temperature was increased to reach thermophilic conditions. When exceeding 47 degrees, increases in ammonia and $H_2S$ (peaks at 1 300 mg $NH_4$-N/L and 16 ppm $H_2S$) were noted but both were far from the inhibitory levels (1 700 mg $NH_4$-N/L and 10 000 ppm $H_2S$). After the digester stabilized at thermophilic conditions, good results which were in line with the mesophilic reference period in terms of gas production (5 $m^3/d$ corresponding to 0.4 $m^3/kg$ VS) and methane content (60%) were obtained.

One of the main concerns of the future full-scale transition at Henriksdal WWTP, is the potentially strong smell when VFA or ammonia are accumulated, which might disturb the people living nearby. However, no strong odor was detected in the digester's surroundings during the transition.

**Increased efficiency in membrane operation**

Optimisation of resource consumption related to the membrane operation has been in the spotlight since 2017. 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 and 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.

This year it was tested to operate the membranes without relaxation for 6 weeks without any signs of negative impact on the membranes. This resulted in a new operation cycle with 15 min permeation and 1 minute relaxation instead of 10 min operation and 1 minute relaxation resulting in 31% less downtime.

**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 oxalic acid, throughout 2019, whereas the other membrane tank (MT2) was cleaned with citric acid. Results from 2019 have showed, that the effect of cleaning with oxalic acid using less than half of the standard consumption was at least as good as when cleaning with citric acid at standard consumption. Since oxalic acid is less expensive than citric acid, there is a large economic saving potential in switching to oxalic acid. During the autumn of 2019 both citric and oxalic acid cleanings were reduced with 50 to 60% still maintaining good permeability (>200 L/($m^2$·h·bar)) throughout the year.

Recovery cleaning of the membranes were performed in March 2019 by soaking the membranes in first sodium hypochlorite and then in citric or oxalic acid. During the soaking in sodium hypochlorite, a separate measuring campaign of chlorine gas emissions from the tank was performed. The results showed that most of the emissions occurred during the first hour. Measurement will be conducted during next recovery cleaning, using shorter intervals to better capture the dynamics in the emissions.

**Phosphorus removal**

The consumption of precipitation chemicals (iron(II)sulphate heptahydrate and iron(III)chloride) for phosphorus removal decreased significantly (from 20-30 mg Fe/L to as low as 6 mg Fe/L) during parts of 2017. The hypothesis that enhanced biological phosphorus removal (EBPR) occurred in the pilot although there is no deliberate anaerobic zone presence was confirmed during 2018. During 2019 the EBPR-activity was still contributing to the phosphorus removal resulting in decreased need for precipitation chemicals (yearly average total iron dosage was 13.7 mg Fe/L) which in turn resulted in reduced iron content in the activated sludge from above 10% to around 3% of TSS. Yearly average effluent total phosphorus concentration was 0.10 mg P/L.

**Testing of external carbon sources**

Methanol has been used as external carbon source added to the post denitrification since 2017. Methanol will not be available during the first years of operation of the first full-scale treatment line. Due to this, glycerol was tested in the pilot during 2019 as an alternative to methanol. Results show similar COD consumption with glycerol as with methanol which means that glycerol is a suitable candidate to replace methanol as external carbon source in terms of denitrification rate and consumption. However, potential long-term effects on colloidal total organic carbon (cTOC) and membrane performance are to be further evaluated. Acetic acid was tested for a couple of weeks but considered non-suitable for the full scale due to release of phosphorus in the post denitrification zone, most probably due to activity of phosphorous accumulating bacteria.

**Reducing HRT in digester**

As part of testing a compact sludge treatment, after transition from mesophilic to thermophilic condition, the hydraulic retention time (HRT) in the digester was lowered in steps, to monitor at what HRT the process would crash (VFA would accumulate decreasing the pH and gas production and methane content would deviate from normal). By the end of 2019 the digester HRT had been decreased to 6 days, still maintaining stable operation.

**Mapping of micro pollutants**

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 finalised during 2019. The results are presented in Närhi et al. (2020). From this study it was concluded that the concentrations of these micro pollutants, both in effluent and sludge, were comparable between the MBR pilot and the conventional activated sludge (CAS) process at full scale Henriksdal WWTP today. This indicates that sludge quality will not deteriorate by introducing a MBR, however, neither will effluent quality be improved significantly.

# 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 2019. Det nya reningsverket är också designat för att klara strikta utsläppskrav med avseende på fosfor, kväve och BOD7 (5 mg BOD7/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 2019 (projektår 6) 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 2019 hade pilotförsöken störst fokus på:

**Omställning från mesofila till termofila förhållanden vid rötning**

Slambehandlingslinjen (inkluderat förtjockning, rötning och avvattning) driftades under mesofila förhållanden (ca 37°C) fram till mitten av mars 2019. Temperaturen ökades sedan kontinuerligt under 18 dagar till termofila förhållanden (ca 55°C). Vid 47°C noterades en ökning i ammonium och svavelvätekoncentrationer (högsta uppmätta halter 1 300 $NH_4$-N/L respektive 16 ppm $H_2S$), men ingen av halterna var när inhiberande halter (1 700 mg $NH_4$-N/L och 10 000 ppm $H_2S$). När rötningen stabiliserats vid termofila förhållanden uppnåddes goda resultat i nivå med de som tidigare uppnåtts vid mesofila förhållanden både vad gällande gasproduktion (5 $m^3$/d motsvarande 0,4 $m^3$/kg VS) och metanhalt (60%).

Det fans en oro att det skulle lukta illa under omställningsperioden från mesofila till termofila förhållanden, och att detta i samband med en omställning på en större anläggning skulle kunna störa närboende. Vid detta omställningsförsök noterades inga starka lukter.

**Minskad resursförbrukning**

Minskad resursförbrukning har varit ett fokusområde i projektet sedan 2017. Försök att minska mängden luft som används i membrantankarna och mängden kemikalier som används för membranrengöring har genomförts. Resultaten indikerar att det finns stor besparingspotential både vad gäller energi- och kemikalieförbrukning utan risk för minskad membrankapacitet.

Under 2019 genomfördes försök under sex veckor utan vilotid för membranen. Försöket visade inte på någon minskad membrankapacitet och resulterade i en ny driftcykel med 15 minuters drifttid följt av 1 minut vilotid jämfört med tidigare driftcykel med 10 minuters drifttid följt av 1 minuts drifttid, vilket resulterar i totalt 31 % mindre vilotid.

**Membranrengöring**

För att studera eventuella skillnader i rengöringseffekt och membranprestanda så användes oxalsyra för rengöring av ena membrantanken (MT1) under hela 2019 medan den andra membrantanken (MT2) rengjordes med citronsyra. Resultaten visade att rengöring genom att använda mindre än hälften av specificerad mängd oxalsyra har minst lika god effekt som rengöring med specificerad mängd citronsyra. Eftersom oxalsyra dessutom är billigare än citronsyra finns det en stor ekonomisk besparingspotential i att byta citronsyra mot oxalsyra. Under hösten 2019 och resten av året minskades både mängden oxalsyra och mängden citronsyra med 50 till 60 % jämfört med specificerade mängder utan försämrad permeabilitet (>200 $L/(m^2 \cdot h \cdot bar)$).

Återhämtningsrengöring av membranen genomfördes i mars 2019 genom att först dränka membranen i hypoklorit och sedan i citron- eller oxalsyra. I samband med hypokloritrengöring genomfördes mätning av kloremissioner till luftfas. Resultaten visade att högst halter emitteras under rengöringens första timme. Uppföljande mätningar med fler, kortare mätintervall, kommer att genomföras under kommande återhämtningsrengöring, för att bättre kunna följa emissionsdynamiken.

**Fosforrening**

Förbrukning av fällningskemikalie för fosforrening (järnsulfat och järnklorid) minskade kraftigt (från 20–30 mg Fe/L till så lågt som 6 mg Fe/L) 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. Under 2019 bidrog bio-P-aktiviteten fortsatt till fosforreningen vilket resulterade i fortsatt lågt behov av fällningskemikalie (medeldos över året 13,7 mg Fe/L). Den låga doseringen av järn resulterade även i minskade järnhalter i det aktiva slammet (från 10 % till ca 3 % av TSS). Utgående fosforhalter var under året ca 0,10 mg P/L.

**Försök med olika externa kolkällor**

Metanol har använts som extern kolkälla i efterdenitrifikationen sedan 2017. Under det första året av drift vid fullskaleanläggningen i Henriksdals reningsverk kommer metanol ännu inte finnas tillgängligt för dosering, och glycerol, som ett alternativ till metanol, har därför doserats på försök i

pilotanläggningen under 2019. Resultaten visar på en liknande COD-förbrukning vid dosering av glycerol som med metanol vilket indikerar på att glycerol är lämplig att ersätta metanol som extern kolkälla med avseende på denitrifikationshastighet och förbrukning. Fortsatta studier på eventuella långtidseffekter med avseende på kolloidalt organiskt kol (cTOC) och membrankapacitet kommer att genomföras under 2020.

Försök genomfördes även med ättiksyra under en kortare period men dessa försök avslutades efter att höga halter av fosfor noterats i efterdenitrifikationen, troligtvis orsakat av aktivitet av fosforackumulerande bakterier.

**Minskad uppehållstid i rötkammaren**

Som en del av försöket med ökad rötningskapacitet, genomfördes försök med minskad uppehållstid i rötkammaren vid termofila förhållanden. Uppehållstiden minskades stegvis för att studera vid vilken uppehållstid processen skulle krascha (ackumulering av flyktiga fettsyror (VFA) medför minskad pH och gasproduktion samt ändrad metanhalt i gasen). Vid slutet av 2019 hade uppehållstiden minskats till 6 dagar och rötningen var fortsatt stabil. Försöket fortsatte därför under 2020.

**Kartläggning av mikroföroreningar**

En tvååring studie för kartläggning av förekomsten av mikroföroreningar, såsom läkemedelsrester, mikroplast, bakterier, PFAS och klororganiska halogener i behandlingsprocessen startade under 2017 och slutfördes under 2019. Resultaten finns presenterade i Närhi et al. (2020). Studiens slutsatser var att halter av mikroföroreningar i MBR-pilotanläggningen var jämförbara med halter i aktiv slam-fullskaleanläggningen i Henriksdals reningsverk. Detta indikerar att slamkvaliteten inte försämras vid införande av en MBR-process, men att utgående vattenkvalitet inte heller avsevärt förbättras.

# Terminology

| | |
|---|---|
| AD | Anaerobic Digestion |
| Anoxic | Process condition without dissolved oxygen, but available $NO_3$ |
| Anoxic zone | Non-aerated zone |
| AOX | Adsorbable organic halogens (mg/L) |
| ATEX | Atmospheres Explosibles |
| $BOD_7$ | Biochemical Oxygen Demand, 7 days (mg/L) |
| BR1 to BR6 | Biological reactor 1 to 6, sampling points |
| 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) |
| EBPR | Enhanced Biological Phosphorus Removal |
| 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. In this report we use Fouling for both organic clogging and inorganic precipitation o membranes (sometimes referred to as scaling). |
| GS | Grab sample |
| Hepta | Iron(II)sulfate heptahydrate |
| IN | Influent wastewater, sampling point |
| Mesophilic | Temperature condition in anaerobic digester, in this project 37 °C |
| MBR | Membrane BioReactor, bio reactor with membrane separation |
| MLD | Million litres per day |
| 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) |
| 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, brand name of 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 |
| Thermophilic | Temperature condition in anaerobic digester, in this project 55 °C |
| 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 (mg/L) |
| TTF | Time To Filter (s) |
| VS | Volatile Solids (% of TS) |
| VSS | Volatile Suspended Solids (mg/L) |
| 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 2019 (project year 6), of the pilot scale trials with membrane biological treatment of municipal wastewater (operated since 2014) and sludge treatment (operated since 2018), carried out in cooperation between IVL Swedish Environmental Research Institute and Stockholm Vatten och Avfall AB 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 both purification and operational stability. Project years 2014-2018 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 wastewater treatment plant (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) SVOAs 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 6-8 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 (~6% TS) will be done at thermophilic conditions instead of mesophilic digestion of thin sludge (~3-3.5% TS). 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, 2020). 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 Beihu WWTP in Hubei, China (commissioned in 2019), designed for an average inflow of 9.3 $m^3/s$, which is significantly 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, 2020-05-07).

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 low effluent requirements for phosphorus (0.20 mg TP/L and 27 tons TP/year) by means of pre- and simultaneous precipitation without affecting membrane performance.
- To minimize the resource consumption.

There are MBR-plants in the USA, e.g. *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 et al., 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 higher organic load and shorter 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 (IVL).

# 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 m³/h. Primary treatment comprise a fine screen (6 mm), pre-aeration, a primary settler and fine sieve (0.6 mm). The biological treatment consists of a pre- and post-denitrification followed by two parallel membrane tanks. The return activated sludge (RAS) passes a deoxygenation zone (RAS-Deox). The purpose of this zone is to lower the oxygen concentrations in the RAS stream, so this do not disturb the pre-denitrification. 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.

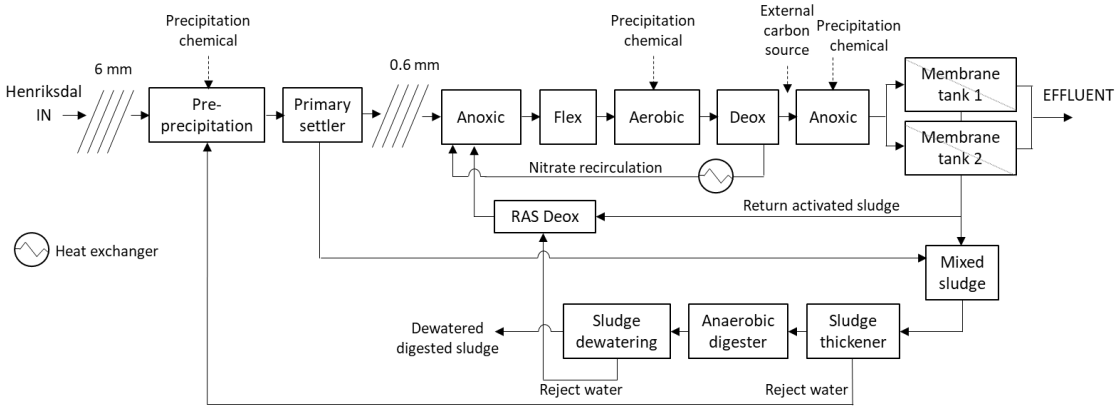



**Figure 1. Flow scheme of the pilot WWTP.**

The 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 (m³) | Future H-dal (m³) | 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 | Anoxic conditions. Stirred. Pre-denitrification. |
| BR2 | 4.8 | 33 500 | 7 000 | Anoxic conditions. Stirred. Pre-denitrification. |
| BR3 | 4.8 | 40 000 | 8 300 | Flex. Stirred/(aerated). Pre-denitrification/(nitrification). |
| BR4 | 4.8 | 31 000 | 6 500 | Aerated. Nitrification. Dosing point 2 $Fe^{2+}$ |
| BR5ox | 1.5 | 10 000 | 6 700 | Aerated. Nitrification. |

| Tank | Pilot (m³) | Future H-dal (m³) | Scale factor H-dal/Pilot | Specification |
|---|---|---|---|---|
| BR5DeOx | 3.3 | 15 000 | 4 500 | DeOx. Stirred. |
| BR6 | 4.8 | 24 000 | 5 000 | Anoxic conditions. Stirred. Post-denitrification. Dosing external carbon. Dosing point 3 $Fe^{3+}$. |
| MT1 | 1.45 | 9 750 | 6 700 | Membrane tank. Aerated. |
| MT2 | 1.45 | 9 750 | 6 700 | Membrane tank. Aerated. |
| RAS-DeOx | 2.7 | 18 000 | 6 700 | Deoxygenation of the RAS. Stirred. 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 | Tank for PS + WAS before thickening. Stirred. |
| Digester | 5.9* | 38 000 | 6 500 | Anaerobic digestion volume. Stirred. |
| 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.

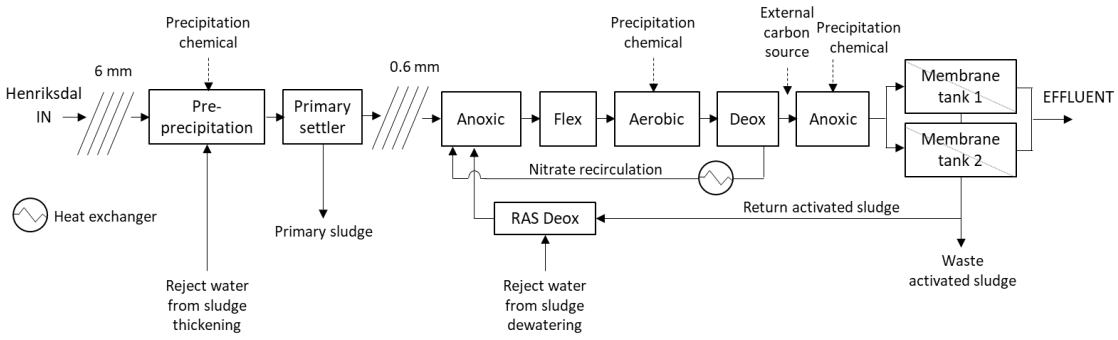



**Figure 2. Process set-up for the wastewater treatment line.**

### 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 (measured as $BOD_7$) than the combined average inflow to the Henriksdal WWTP. It has about 60% higher $BOD_7$-concentration than 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 pilot inflow are proportional to the actual

inflow variations to the Henriksdal WWTP, as the pilot inflow is controlled by a signal from flow meters in the full-scale plant.

Since the influent to the pilot is set by a 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. However, due to continuous problems with clogging, the heat exchangers were taken out of operation on the 12th of February 2019. On the 8th of April 2019 heat exchangers were again taken into operation, this time placed on the nitrate recirculation (flow from BR5 to BR1). It was unfortunate for the evaluation that the heat exchanges were out of operation during the period with lowest temperatures. However, the processes have been tested during cold inlet temperatures previous years. Initially the temperature of inflow to Henriksdal was used as setpoint. In the mid of June this strategy was changed, and the heat exchanger was controlled to maintain a temperature in MT1 corresponding to 1°C higher than inlet to Henriksdal, since normal temperature increase from inlet to biology in Henriksdal is about 1°C. The temperatures in the incoming wastewater to Henriksdal and to the pilot are presented in Figure 3. On average the temperature of the inlet water to the pilot was 18.6°C, which is 2.5°C higher than the influent wastewater to Henriksdal (16.1°C). The daily average temperature in the pilot inlet varied between 10.1 °C and 23.1 °C. With cooling of the influent, there was still a temperature increase in the pilot processes with about 2.2°C to the membrane tanks. With cooling on the nitrate recirculation using the temperature in MT1, process temperature could be controlled to match the temperatures in Henriksdal better.

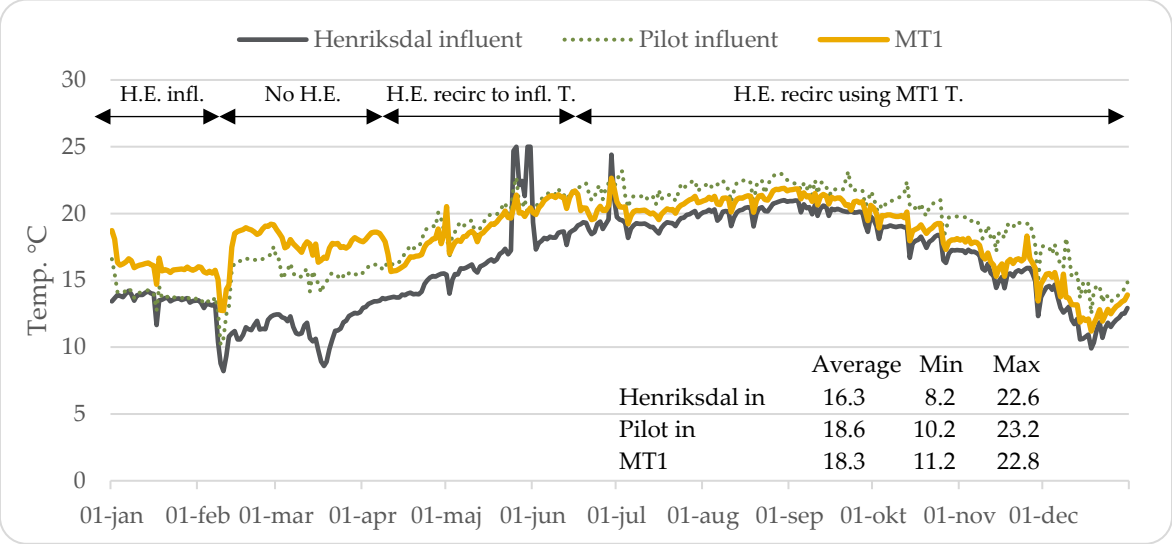



**Figure 3. Influent temperatures to the MBR pilot (dotted line) and the Henriksdal WWTP (black line) together with temperature in Membrane tank 1 (MT1). H.E. = heat exchanging, infl. = influent, T. = temperature, recirc. = nitrate recirculation.**

### 3.1.2 Pre-treatment

The pre-treatment steps in the pilot consisted of a 6 mm punch hole sieve (with screen capture rates similar to 3 mm step screen, UKWIR (2015)), a pre-aeration tank with Ferrous dosing and a vertical flow primary settler, with a surface area of 1.13 m² and a water depth of 4.3 m (scale 1:9,200 compared to the future Henriksdal design), followed by a 0.6 mm punch hole drum sieve before the biology, see Figure 4. The small hole size of the drum sieve was chosen to enable the study of clogging tendencies.

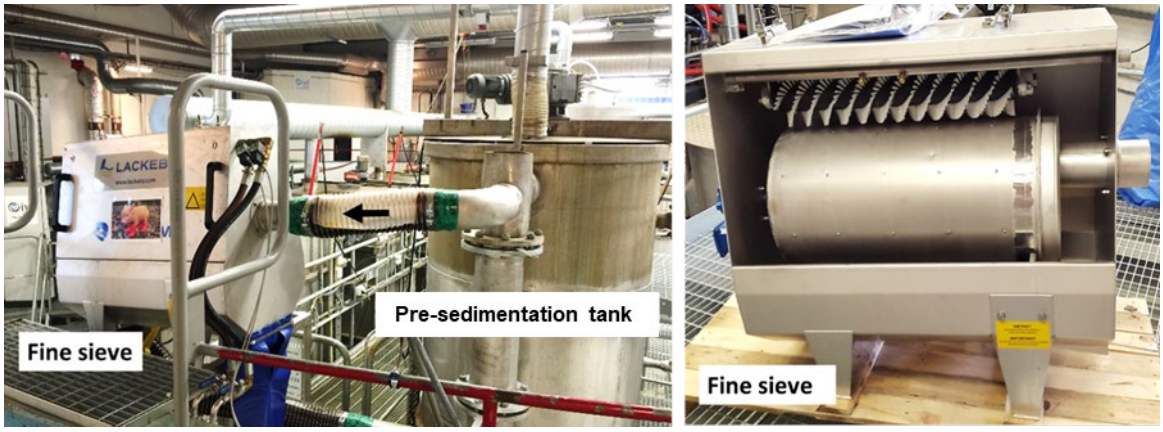



**Figure 4. Photo of the fine sieve installation.**

### 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, BR4 and 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 primarily 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 was 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 1:6,700, with few minor exceptions. The Deox zone in BR5 and the post-denitrification zone in BR6 were slightly over dimensioned. The discrepancy depends on 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 this zone will be crucial for the nitrogen removal during winter.

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 (ZeeWeed 500D-Leap). 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 (air scouring). 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.

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


**Figure 6. The membrane during installation of 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 treatment lines (out of seven in total) in the full-scale plant. There two reasons for this. First, it corresponds to two standard design pilot cassettes from the manufacturer. Secondly, the design max flow rate to the biological treatment, according to the SFA design, could be treated even if one of the seven treatment lines are out of operation.

In future Henriksdal, each treatment line (a total of seven) 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, with 48 modules in each cassette. 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 and six modules, which gives less flexibility. 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, the pilot could only be operated with three or six modules in operation, as a pilot cassette contains three modules. To enable operation at a constant flux, the pilot was equipped with permeate recirculation. This means that the flow through the

membranes was higher than the inflow by having a partial flow of the permeate recycled back to the membrane tank.

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

## 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 sludge treatment pilot is visualized in Figure 7.

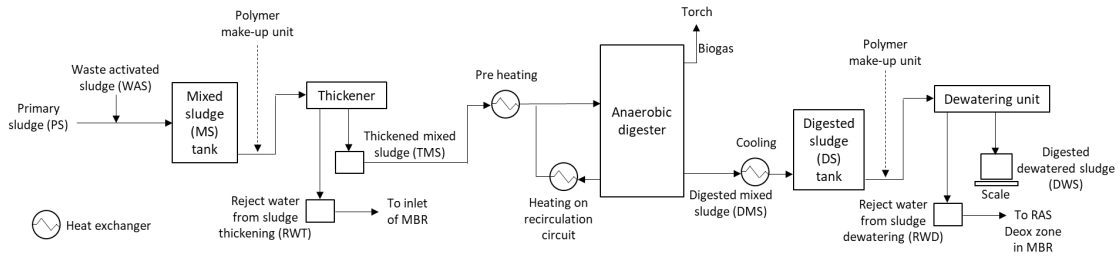



**Figure 7. Process set-up for the sludge treatment line.**

### 3.2.1 Thickening

Primary sludge and waste activated sludge was intermittently pumped to the mixed sludge tank. Mixed sludge was then pumped to a rotating drum sieve thickener. Polymer was dosed inline in one of three possible dosing points, see Figure 8. 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. Thickened mixed sludge was pumped directly into the digesters heat exchanger recirculation circuit and fed into the digester. The heat exchanger for pre-heating was bypassed this year due to problems with clogging.

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.

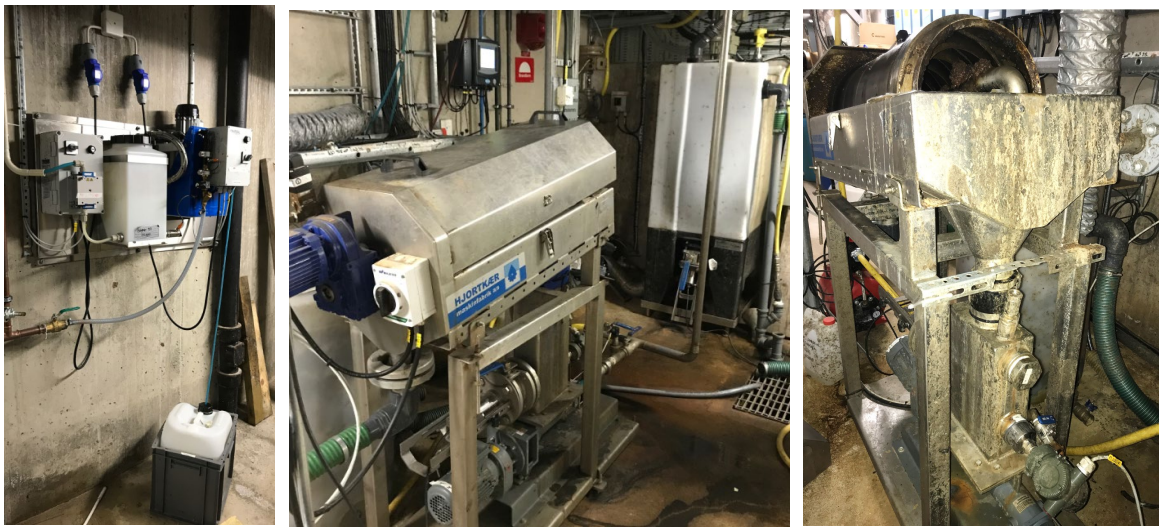



**Figure 8. Photo of thickener and polymer dosing points.**

### 3.2.2 Digestion

The digester is cylindrical with a base area of 2.54 $m^2$ and a variable water level. A volume of 5.7 $m^3$ corresponds to full digester capacity in the future Henriksdal WWTP (scale 1:6,700). During 2019 the volume was kept between 4 and 5 $m^3$ (instead of 5.7 $m^3$) in order to achieve lower HRT (down to 6 days) than the design (13 days). The sludge is 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 thereafter into an equalization tank (digested sludge tank).

The thickened mixed sludge was digested at mesophilic and thermophilic conditions. During 2018, a mesophilic (37°C) reference period was sought, and tests of increasing the temperature was done in order to verify that thermophilic conditions (55°C) could be achieved. During the main part of 2019, thermophilic digestion was applied, which will be the mode of operation at the future Henriksdal WWTP. Special attention was given to the transition from mesophilic to thermophilic conditions, with a master thesis student performing additional monitoring and evaluating the transition period.

In the future Henriksdal design, fat from grease traps at restaurants and industrial by-products like glycerol will be co-digested with WWT-sludge. However, no external organic material was fed to the pilot digester.

### 3.2.3 Dewatering

Digested sludge was stored in the digested sludge tank and pumped into a pressurized, 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.

Reject water from digestate dewatering was collected in a tank for continuously pumping through a filter into the RAS-Deox zone, via a filter, in the wastewater treatment line. Due to operational problems the reject water was not continuously returned to the process.


**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 2019 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 2019 was higher than the design flow rate; 3.58 $m^3$/ 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 2019 | Design Pilot | Design future H-dal | Design H-dal/ Value Pilot[i] |
|---|---|---|---|---|---|
| Flowrates | | | | | |
| Average influent flowrate, $Q_{in}$ | $m^3/h$ | 3.58 | 3.16 | 20 880 | 5 800 |
| 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 | × $Q_{in}$ | 2.8 | 1.2-4.2[ii] | 0-4 | - |
| RAS flowrate | $m^3/h$ | 4.1-23.3 | 3.6-19 | - | - |
| RAS flowrate | × $Q_{in}$ | 3.6 | 1.1-5.9[ii] | 4 (3-5) | 1.1 |
| Temperatures | | | | | |
| Temperature influent | °C | 18.6 | | | |
| Temperature biology | °C | 18.3 | | | |
| Incoming load | | | | | |
| $BOD_7$ influent | mg/L | 288 | 206[iii] | 216 | 0.8 |
| SS influent | mg/L | 325 | 201[iii] | 280 | 0.9 |
| TN influent | mg/L | 46 | 44[iii] | 37 | 0.8 |
| TP influent | mg/L | 6.3 | 5.7[iii] | 4.9 | 0.8 |
| Primary settler (SED) | | | | | |
| $BOD_7$ reduction over SED | % | 27 | 46 | 50[iv] | 1.9 |
| SS-reduction over SED | % | 36 | 60 | 60[iv] | 1.7 |
| TN reduction over SED | % | 4 | 10 | 10[iv] | 2.5 |
| TP reduction over SED | % | 13 | 40 | 40[iv] | 3.1 |
| $BOD_7$ PTW | mg/L | 207 | 112 | 108 | 0.5 |
| SS PTW | mg/L | 191 | 80 | 112 | 0.6 |
| TN PTW | mg/L | 45 | 40 | 33 | 0.7 |
| TP PTW | mg/L | 5.5 | 3.4 | 3.0 | 0.5 |
| SS removed over SED | kg SS/d | 11.0 | 13.3[v] | 89 300 | 8 100 |
| Primary sludge production | kg SS/d | 18.5 | 17.2[v] | 115 000 | 6 200 |
| VS-concentration PS | % of TS | 88% | 77% | 77% | 0.9 |
| Biological treatment | | | | | |
| $BOD_7$-load PTW (at average flowrate) | kg $BOD_7$/d | 17.4 | 8.6 | 57 500 | 3 300 |
| Specific WAS-production [vi] | kg SS/kg $BOD_7$ | 0.76 | 1.02 | 1.02 | 1.3 |
| WAS production, average | kg SS/d | 13.3 | 8.8 | 58 600 | 4 400 |
| VSS-concentration WAS | % of TSS | 74% | 64% | 64% | 0.9 |
| SS in biological tanks | mg/L | 7 900 | 8 000 | 8 000 | 1.0 |
| SS in membrane tanks | mg/L | 10 100 | 10 000 | 10 000 | 1.0 |
| Total sludge age | d | 19.1 | 32.0 | 31.2 | 1.6 |
| 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) | $l/m^2,h$ | 22.8 | 17.9 | 20.9 | 0.9 |
| Net flux max | $l/m^2.h$ | 27.8 | 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/m³in FL/sand trap.
[v] Calculated based on incoming load/scaled from SFA design with factor 6 700.
[vi] Excluding external carbon source. Calculated from process data for Values Pilot 2019. Design values from German standard ATV DVWK-A 131E (2000) based in incoming SS and BOD, and $SRT_{tot}$
[vii] Aeration of the membranes had two modes, one with lower (Leap-Lo) and one with higher air flowrate (Leap-Hi).

**Table 3. Operation and design data for the sludge treatment line in the pilot plant and design data (year 2040) for the future Henriksdal WWTP. The data is presented as average before decrease in HRT (1 Jan-14 July) and during trials with decreased HRT in the digester (15 July-31 Dec).**

| Parameter | Unit | Value pilot 1 Jan-14 July | Value pilot 15 July-31 Dec | Design future H-dal | Design future H-dal / Value Pilot 1 Jan-14 Jul |
|---|---|---|---|---|---|
| Into thickener (from 15 July thickener was bypassed) | | | | | |
| Flow mixed sludge (MS) | L/h | 83 | | 444 000[a] | 5 350 |
| TS-concentration MS | % | 1.7% | | 1.6% | 0.9 |
| VS-concentration MS | % of TS | 75 | | 72 | 1.0 |
| TS-load MS | kg TS/d | 34.5 | | 173 600 | 5 000 |
| Polymer consumption | g/kg TS | varied | | 6 | |
| Sludge into digester (before 15July TMS, after 15 July MS) | | | | | |
| Flow into digester (TMS/MS) | L/h | 13.2[b] | 26.4[b] | 118 000 | 8 900 |
| TS-concentration TMS/MS | % | 5.0% | 2.0% | 6.0% | 1.2 |
| TS-load TMS/MS | kg TS/d | 15.4[b] | 12.8 | 172 000 | 11 200 |
| VS-load TMS/MS | kg VS/d | 11.3 | 9.8 | 124 000 | 8 600 |
| Flow reject RWT | L/h | 24.8[b] | 0 | 326 000 | 11 000 |
| SS-concentration reject RWT | mg/L | 2 035 | - | 500 | 0.3 |
| Flow external organic material (EOM) | L/h | 0 | 0 | 11 400 | - |
| VS-load EOM | kg VS/d | 0 | 0 | 44 000 | - |
| Digestion | | | | | |
| Digester temperature | °C | 37-55 | 55 | 55 | - |
| Hydraulic Retention time, HRT | d | 18 | 8 | 13[c] | 0.7 |
| Specific VS-load | kg VS/m³,d | 2.6 | 2.0 | 3.3[c] | 1.3 |
| Digestion efficiency | % of $VS_{in}$ | 44% | 40% | 42%[c] | 1.0 |
| VFA/Alkalinity | mg $CH_3COO$-eq/mg $CaCO_3$ | 0.11 | 0.08 | - | - |
| Out of digester | | | | | |
| Flow DMS | L/h | 13.5 | 28.8 | 123 000 | 9 100 |
| TS-concentration DMS | % | 2.9% | 1.2% | 3.9 | 1.3 |
| VS-concentration DMS | % of TS | 68% | 73% | 60% | 0.9 |
| TS-load DMS | kg TS/d | 9.4 | 8.3 | 124 000 | 13 200 |
| VS-load DMS | kg VS/d | 6.4 | 6.1 | 74 000 | 11 600 |
| Specific biogas production | $Nm^3/kg\ VS_{digest.}$ | 0.95 | 0.93 | 1.0 | 1.1 |
| Flow biogas | $Nm^3/d$ | 3.9[d] | 3.0[d] | 52 000[c] | 13 300 |
| Methane content biogas | % | 58% | 60% | 65% | 1.1 |
| After dewatering | | | | | |
| Flow DDMS | L/h | 1.3 | 1.1 | 17 000 | 13 000 |
| TS-concentration DDMS | % | 27% | 25% | 30% | 1.1 |
| Flow reject RWD | L/h | 12.2 | 27.7 | 114 000 | 9 300 |
| SS-concentration reject RWD | mg/L | 4 000 | 2 800 | 900 | 0.2 |
| $NH_4$-N in reject water | mg/L | 560 | 240 | 1 500 | 2.7 |
| Polymer consumption dewatering | g/kg TS | 15 | 16 | 6-10 | -[e] |

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 and deliberate wasting of WAS during autumn.

c) Numbers excluding addition of external organic material in order to be comparable to data from the pilot.
d) $m^3/d$ not $Nm^3/d$.
e) Different methods of dewatering. Not comparable.

# 3.4 Chemicals

During 2019, methanol, acetic acid and glycerol were used as external carbon source (only one at a time) in the post denitrification zone. The phosphorus was precipitated using ferrous(II)sulphate at two dosing points and ferric(III)chloride in one point. For membrane cleaning sodium hypochlorite was used for both MTs 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 until the end of August 2019. It was delivered in 25 L canisters and had a concentration of 1 200 g COD/L (concentration 100 % by weight). Before 2017, other external carbon sources were tested. During 2014 until April 2015 (Project year 1 and 2) sodium acetate was used, after this until the end of 2016 (Project year 2 and 3) Brenntaplus was used. During the last part of 2019, acetic acid was tested for 3 weeks in September and after this trial Glycerol was used throughout the year.

The acetic acid (75%) was a waste product from food industry. It was delivered in 25 L containers from the manufacturer Helm. The COD concentration was about 850 g COD/L.

Glycerol is used by Scandinavian Biogas operating on the Henriksdal site area next to Hammarby Sjöstadsverk. The proximity and easy access to this carbon source made it attractive to test in the pilot as alternative to methanol. It was collected in 25 L containers from Scandinavian Biogas' storage. The measured COD concentration was about 850-900 g COD/L.

The dosing point of external carbon source was in-between the BR5 Deox-zone and BR6. This point was tested out previously and provided longer residence time compared to dosing directly in BR6 (which also led to a higher risk of carbon source leakage to the membrane tanks) while avoiding risk of recirculating carbon source to BR1 via the nitrate recirculation from the BR5 Deox-zone.

More about carbon source addition and treatment results can be found in section 6.2.2 Denitrification and section 6.9 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 Phosphorus removal.

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 25 and 57 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. An iron concentration of 195.6 g Fe/L has been used for control and dose calculation.

### 3.4.3 Chemicals for membrane cleaning

The membranes have been cleaned regularly with sodium hypochlorite and either citric acid or oxalic acid. For more information on how the cleanings were carried out, see section 6.5.3.

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 (p.68. Svenskt Vatten, 2010a). During 2019, both diluted and non-diluted 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 40 and 90 g $Cl_2$/ L during the year.

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).

### 3.4.4 Polymers

For thickening of mixed sludge Flopam EM 640 HIB (SNF) polymer was used.

During 2019 polymer for dewatering of digested sludge, was changed from Superfloc C-1598 (Kemira) to the same as for thickening, i.e. Flopam EM 640 HIB (SNF). Polymer was delivered in solution and prepared to decided 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, except for pumping of reject water that was locally controlled. Implementation of the

control has been carried out within the project, which provides great flexibility to adapt and optimize control.

# 4 Experimental plan year 2019

An overview of the experimental plan year 2019 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. In 2017 the performance was tested with inlet temperatures <10°C for four weeks. With the first goals 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.

During 2019, the main theme was "**how low can we go**" – regarding membrane cleaning chemicals, membrane air scouring, membrane relaxation, nitrogen and phosphorus in the effluent and the retention time in the digester. In addition, a transition from mesophilic to thermophilic digestion was done and a trial mimicking the first years of operation of the first full scale MBR-line at Henriksdal, adapted to adjustments in the SFA-project compared to the 2018 trial, started by the end of 2019.

**Table 4. Experimental plan of year 2019.**

| Trial | J | F | M | A | M | J | J | A | S | O | N | D |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Mesophilic operation of the digester | ■ | ■ | | | | | | | | | | |
| Transition to thermophilic digestion | | | ■ | | | | | | | | | |
| Thermophilic operation of the digester | | | | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ |
| Sludge thickener in operation | ■ | ■ | ■ | ■ | ■ | ■ | | | | | | |
| Trial with decreasing digester HRT | | | | | | | ■ | ■ | ■ | ■ | ■ | ■ |
| Minimising membrane scouring air use | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ |
| Oxalic acid and citric acid comparison | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ |
| Minimising membrane cleaning chemical use | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ |
| Recovery cleaning and sampling | | | ■ | | | | | | | | | |
| Trail without membrane relaxation | | | | | | | ■ | ■ | | | | |
| Extended membrane operational cycle | | | | | | | | | ■ | ■ | ■ | ■ |
| Inlet cooling | ■ | ■ | | | | | | | | | | |
| Cooling in the biology | | | | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ |
| Effluent phosphorous target 0.10 mg P/L | | | | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ |
| Increased inflow | | | | | | | ■ | ■ | | | | ■ |
| Methanol as carbon source | ■ | ■ | ■ | ■ | ■ | ■ | ■ | ■ | | | | |
| Acetic acid as carbon source | | | | | | | | | ■ | | | |
| Glycerol as carbon source | | | | | | | | | | ■ | ■ | ■ |
| Imitation of first phase operation SFA | | | | | | | | | | | | ■ |
| Mapping of micro-pollutants | ■ | | | | | | | | | ■ | | |

Previous years, the sludge treatment line have had operational problems and therefore this line has been in focus during 2019. The first months focused on getting a good reference period with mesophilic conditions before the transition to thermophilic conditions which was conducted in

March. The thermophilic conditions were kept for the remainder of the year. By July a new trial started where the thickener was bypassed, and the focus was on pushing the process by slowly lowering the HRT until the process crashed.

Reductions of the resource consumption, especially for the membranes was continuously in the spotlight this year. Further reductions in acid cleaning chemical consumption, both oxalic and citric have been evaluated. In addition, forced low aeration in combination with less chemicals for cleaning just started by the end of the year.

The capacity of the membranes has also been tried by operating without the normal relaxation period of 1 minute after each cycle of 10 minutes of producing permeate. This was tested for 6 weeks.

Recovery cleaning of the membranes was carried out in March. This time focus was on chlorine gas emissions from the process which was measured during soaking of the membranes in sodium hypochlorite.

The new permit for future Henriksdal will include both a maximum amount of phosphorous to be released with the effluent as well as the maximum effluent concentration. The amount specified in the permit will in the future, with expected high flows, mean that the effluent phosphorous must be even lower than the concentration limit. To test the capacity of the process the target concentration in the pilot operation has been lowered to 0.10 mg P/L. The strategy with three precipitation dosing points remains, where a flow proportional dose of ferrous sulphate is added at the inlet combined with simultaneous precipitation using ferrous sulphate and ferric chloride. The later doses are controlled by a phosphate feedback controller from online effluent phosphate concentrations.

For the first full-scale treatment line it will not be possible to use methanol during the first phase of operation. However, it could be possible to use a non-explosive external carbon source. As a result of this, two other external carbon sources have been tested in the pilot this year; acetic acid and glycerol. By the end of this year another trial to imitate the first phase operation of the full scale (BB1 trial) was initiated, this time with the use of external carbon source (glycerol).

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 and finalized in October 2019.

# 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). The sampling points (except SLUDGE 1 and 2) are illustrated in Figure 10.

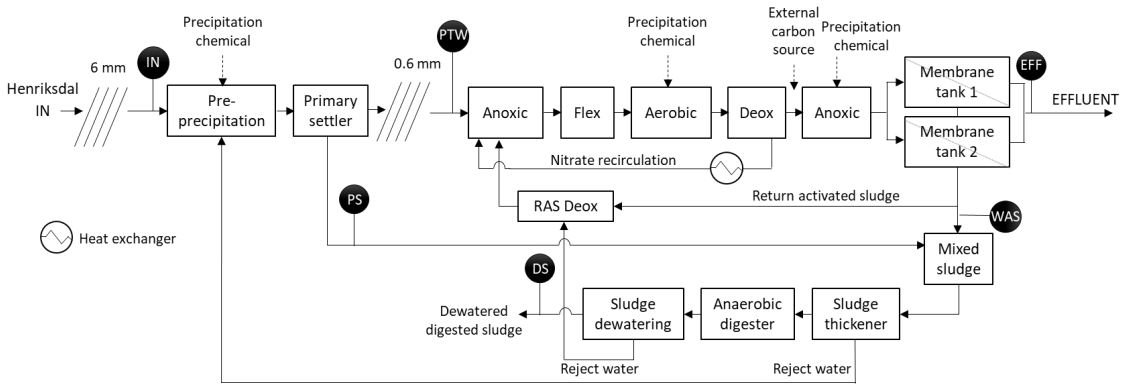



**Figure 10. Sampling points in pilot process marked as black circles (SLUDGE1 and SLUDGE2 sampling points not included in figure).**

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 Eurofins, was used to measure sludge volume (SVI) and time to filter (TTF) at IVL's laboratory at Hammarby Sjöstadsverk. The filtrate from the TTF analysis was also sent for analysis of TOC. This was done in order to calculate the colloidal TOC (cTOC, see section 5.3.2 Sludge quality) which, according to the membrane supplier, could relate to membrane performance.

In addition to the samples and analyses presented in Table 5, a monthly composite sample of dewatered digested sludge (DDMS) which was stored at -30°C during the sampling period, was sent to external laboratory 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).

**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 | Alkalinity | Fe (digested) | P (digested) |
| Daily composite samples | | | | | | | | | | | | | | |
| IN | 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 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 further observe the process (for example measurements of $NO_2$-N during disturbances) and to calibrate process instruments.

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

| | Weekday | | |
|---|---|---|---|
| Analysis | Monday | Wednesday | Friday |
| 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 regards 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 with the same approximate frequency regarding total suspended solids (mg/L). To monitor the digestion process, a sample from the digester was taken at least once per week and pH, VFA (mg $CH_3COO$-eq/l), alkalinity (mg $CaCO_3$/l) and ammonium (mg $NH_4$-N/l) were analysed. Measurements of methane, carbon dioxide and hydrogen sulphide in the produced biogas was conducted several times per week with a hand-held gas meter (Sewerin Multitec 54).

## 5.2 Online measurements

The process was controlled and/or monitored with several 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 online analyser for $PO_4$-P sampling from the effluent.

**Table 7. Placement of online sensors in MBR.**

| Placement | Parameter | Function |
|---|---|---|
| General | Flowrate (water) | Measure all recirculation flows |
| IN | Temperature | Measure the incoming wastewater temperature. Sometimes used for control |
| IN | Flowrate (water) | Measure the influent water flow |
| IN | SS | Monitor influent suspended solids concentration |
| PTW | $NH_4$-N | Measure incoming ammonium concentration |
| BR1 | DO | Monitor Dissolved Oxygen |
| BR2 | DO | Monitor Dissolved Oxygen |
| BR2 | $NH_4$-N | Measure ammonium concentration in to aerated part of biological treatment. Sometimes used for control. |
| BR3 | DO | Controlling Dissolved Oxygen |
| BR3 | Flowrate (air) | Measure air flow |
| BR4 | DO | Controlling Dissolved Oxygen |
| BR4 | Flowrate (air) | Measure air flow |
| BR4 | SS | Measure suspended solids concentration in activated sludge |
| BR5 | DO | Controlling Dissolved Oxygen |
| BR5 | Flowrate (air) | Measure air flow |
| BR5 | $NO_3$-N | Measure nitrate concentration, monitor function of post-denitrification |
| BR6 | $NO_3$-N | Measure nitrate concentration, 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 flow of permeate from membrane 1 and 2 (x2) |
| MT1/MT2 | Flowrate (air) | Measure air flow (x2) |
| MT1/MT2 | pH | Monitor pH, especially during membrane cleaning |
| RAS-DeOx | SS | Measure suspended solids concentration |
| 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 concentration |

**Table 8. Placement of online 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 |
| 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 at *net* values (using average data from the full operation cycle, permeation and relaxation = 11 minutes). The gross values are higher than the net, however the net corresponds better to the average operation. All values for the parameters described below are given as net values in this report.

**Flux**: Flowrate per membrane area, unit L/(m²·h). The flux is describing the load on the membranes. Flux is calculated as permeate flow divided by membrane area.

**TMP**: Transmembrane pressure, unit mbar. The difference in pressure before and after the membranes, this can be compared to filter resistance if TMP can vary. TMP is the driving force for transportation through the membrane. TMP is measured using online pressure transmitter in the membrane tank and on the permeate pipe.

**Permeability**: Flux per TMP, unit L/(m²·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

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

## 5.3.2 Sludge quality

In addition to the parameters analysed at the external and internal laboratory, and 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), colloidal TOC (cTOC) 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) and colloidal TOC (cTOC)

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

### Trash content

The method for defining the Trash content is described in detail in a previous report (Andersson et al., 2017). In short, the sludge is filtered through screens with different slot width and the amount of trash captured in the screens is measured. This analysis was carried out 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 screen size of 2 mm, should not exceed 2 mg/L (according to the membrane supplier).

## 5.3.3 Anaerobic digestion

### Digestion efficiency (degradation rate)

The digestion efficiency [% of $VS_{in}$] was calculated based on the mass flow of VS [kg VS/d] into the digester (BS) and out of the digester (RS) using the equation below:

$$Degradation\ efficiency = {}^{(VS_{RS} - VS_{BS})}/_{VS_{BS}}$$

### Specific biogas and methane production

The specific biogas production [$m^3$/kg $VS_{deg}$] was calculated using the daily biogas production [$m^3$/d] and dividing it by the amount of degraded VS per day which was calculated as the difference between VS into the digester ($VS_{BS}$) and out of the digester ($VS_{RS}$). The specific methane production was calculated in the same way, but the daily biogas production was multiplied with the percentage of methane in the gas.

### Theoretical biogas production

Periodically there has been problems with the accuracy of the gas flow meters. Therefore, the theoretical biogas and production was calculated as a reference to control and sometimes replace the measured values. Theoretical biogas production was calculated by multiplying the amount of degraded VS per day with a factor. For biogas the factor was 0.8 $m^3$/kg $VS_{deg}$. The value was found in literature and calibrated to values from previous years operation.

# 6 Results and discussion

## 6.1 Primary treatment

| Trial | J | F | M | A | M | J | J | A | S | O | N | D |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Inlet cooling | | | | | | | | | | | | |
| Increased inflow | | | | | | | | | | | | |
| Imitation of first phase operation SFA | | | | | | | | | | | | |

### 6.1.1 Inlet screen

Different punch hole screens, from 2 mm to 6 mm, has been used on the inlet flow. In 2017 and throughout 2018 a 6 mm size was used, and inlet suspended solids concentrations corresponded well to data from Henriksdal WWTP. The course 6 mm screen contributed to an increase in primary sludge production which was required to meet the design for the sludge pilot. However, it also caused frequent clogging of the inlet heat exchanger. To be able to cool the inlet temperature, the 3 mm screen was put back in place from December 2018 to the beginning of April 2019. As the problems with clogging of heat exchangers continued it was decided to reinstall the 6 mm screen and relocate the heat exchangers to the nitrate recirculation stream in the biological treatment. Figure 11 shows that the SS-concentration increased slightly after installing the 6 mm screen and the SS-concentration to the pilot was similar as the concentration to Henriksdal. The large variation during 2019 was partly due to SVOA emptying digesters during spring and summer to a point near the inlet pumps to the pilot.

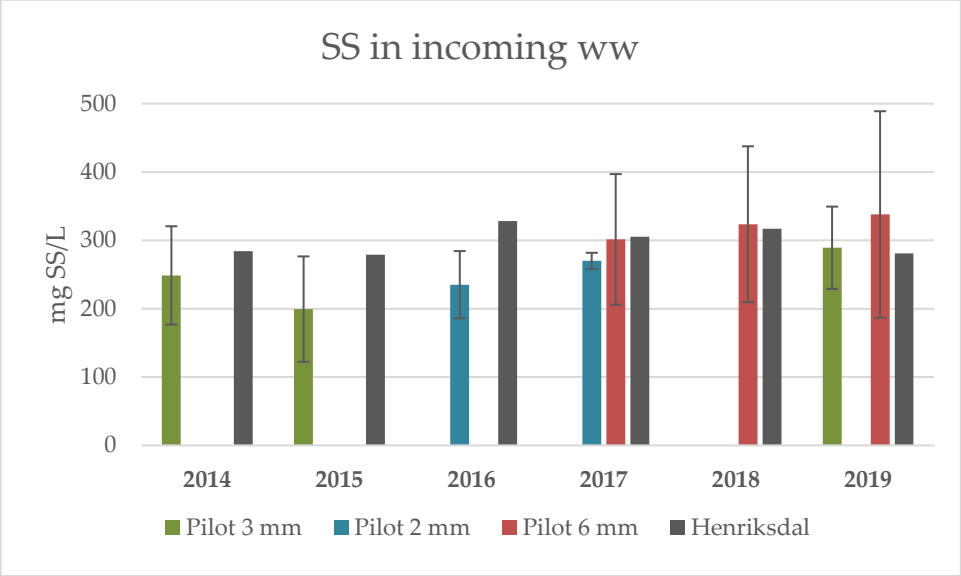



**Figure 11. Incoming SS-concentration to the pilot line after passing through 2, 3 or 6 mm punch hole screen and to Henriksdal (about half of the flow not sieved, half sieved through 3 mm step-screen).**

## 6.1.2 Efficiency of primary settler

The primary settling volume (3.3 $m^3$), and capacity, is smaller than it should be (design value 4.5 $m^3$) resulting in poor reduction, compared to Henriksdal WWTP, and insufficient primary sludge production. The reduction rate over the primary settler is showed Table 9 and Figure 12 below.

**Table 9. Reduction over the primary settler and primary sludge production for the pilot compared to Henriksdal WWTP 2019 data and the future design for Henriksdal WWTP.**

| Parameter | 2019 | 2018 | 2017 | 2019: Henriksdal WWTP[a] | Design future Henriksdal |
|---|---|---|---|---|---|
| Reduction of SS (%) | 36% | 35% | 37% | 52% | 60% |
| Reduction of $BOD_7$ (%) | 27% | 25% | 30% | 51% | 50% |
| Reduction of TP (%) | 12% | 10% | 14% | 38% | 40% |
| Reduction of TN (%) | 4% | 1% | 4% | 9% | 10% |
| Reduction of TOC (%) | 20% | 18% | 17% | 36% | - |
| PS-production (kg/d) | 18.5 | 16.1 | 13.1 | 62 200[b] | 115 000 |

a) At Henriksdal during 2019 trials with enhanced pre-precipitation using $Al^{3+}$ during high flows were conducted in half of the primary clarifiers. The strategy was permanently implemented in December.

b) Based on uncertain TS-measurements due to installation of new thickening unit. Probably underestimated.

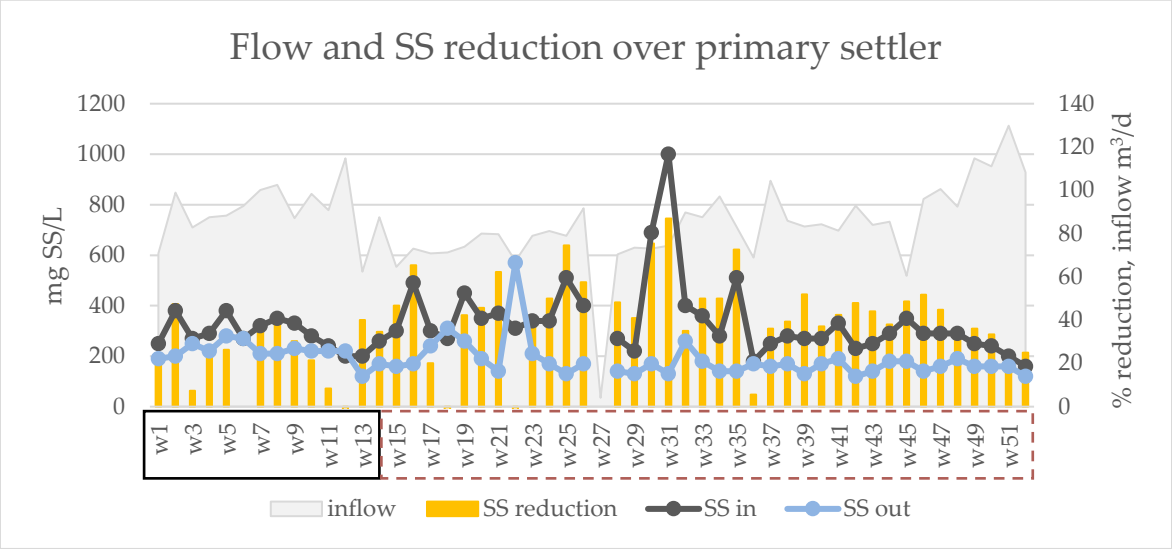



**Figure 12. Reduction of SS over primary settler with 3 mm screen (black box, w1-w13) and with 6 mm screen (red dashed box w14-w52).**

### 6.1.3 Screen and sieve – effect on trash content

During 2019 the pilot line was operated with first a 3 mm and then a 6 mm punch hole inlet screen (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. Results are presented in Table 10.

The results from previous years have been consistent when the 6 mm + 0.6 mm screen/sieve configuration was implemented. In addition, visual inspection of the membrane cassettes in 2018 show very little build-up of trash indicating that the measured values are good. During 2019 only three analyses of trash content were conducted, and the results varied more than before.

**Table 10. Trash content in waste activated sludge (WAS) with various sieve-configurations. According to membrane supplier the trash content at 2 mm sieve should not exceed 2 mg/L.**

| Sieves and hole size | Dates | No of analyses | 1 mm sieve mg/L | 2 mm sieve mg/L |
|---|---|---|---|---|
| 3 mm screen at inlet pump | Dec 2013 | 2 | 11.6 ± 5.4 | 1.0 ± 0.7 |
| 2 mm fine screen at inlet pump | Nov 2016 – Feb 2017 | 4 | 6.4 ± 2.4 | 1.1 ± 0.7 |
| 6 mm screen at inlet pump and 0.6 mm fine sieve before biology | Mar 2017 → Dec 2017 | 8 | 4.1 ± 3.8 | 0.6 ± 0.3 |
| 6 mm screen at inlet pump and 0.6 mm fine sieve before biology | May 2017 → Dec 2017 | 6 | 2.2 ± 1.6 | 0.6 ± 0.3 |
| 6 mm screen at inlet pump and 0.6 mm fine sieve before biology | Jan 2018 → Nov 2018 | 6 | 2.0 ± 1.1 | 0.6 ± 0.5 |
| 3 mm screen at inlet pump and 0.6 mm fine sieve before biology | Dec 2018 – Mar 2019 | 1 | 2.3 | 2.0 |
| 6 mm screen at inlet pump and 0.6 mm fine sieve before biology | Apr 2019 – Dec 2019 | 2 | 5.1 ± 5.4 | 1.5 ± 2.0 |

### 6.1.4 Pre-treated wastewater

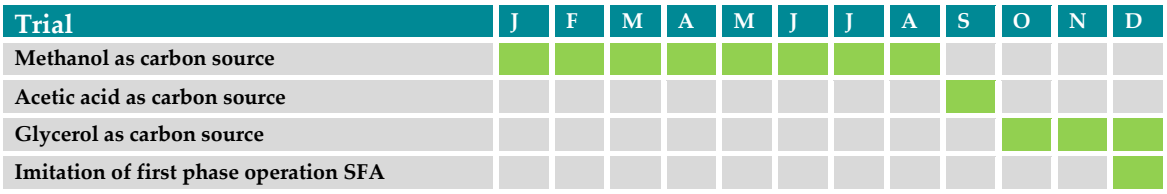
The quality of the pre-treated wastewater (PTW) is presented in Table 11. The concentrations measured in the pilot were significantly higher than the corresponding concentrations measured in the Henriksdal WWTP for SS, $BOD_7$, TN and TP. This is mainly due to the poor performance of the primary settler and was also observed last year. These concentrations were also higher than the design values given for the future plant, which on the other hand are relatively low since it is a weighted average between Henriksdal WWTP and Bromma WWTP, and Bromma WWTP has a diluted inflow.

The concentrations in pre-treated wastewater will affect the biological treatment, including WAS production (and thereby SRT and the amount of phosphorus assimilated in sludge), pre-denitrification capacity and the need for simultaneous precipitation. The difference in concentrations between the pilot and the future Henriksdal design thereby will affect comparison of the evaluation parameters related to these aspects. 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 design. This is due to enhanced biological phosphorus removal (EBPR) which is described in chapter 6.3.1.

**Table 11. Data on PTW from the pilot compared to data from Henriksdal 2018 and the design data for the future Henriksdal WWTP.**

| Parameter | Value Pilot 2019 | Value Henriksdal 2019 | Design future Henriksdal | Design H-dal/ Value Pilot |
|---|---|---|---|---|
| *Pre-treated wastewater (PTW) – into biological treatment* | | | | |
| SS (mg/L) | 191 ± 71 | 135 | 113 | 0.6 |
| $BOD_7$ (mg/L) | 207 ± 54 | 112 | 108 | 0.5 |
| TN (mg/L) | 45 ± 7 | 38 | 33 | 0.7 |
| TP (mg/L) | 5.3 ± 0.9 | 3.7 | 3.0 | 0.6 |
| Fe (mg/L) | 7.9 ± 2.8 | 10.9 | 12 | 1.6 |
| Alkalinity (mg/L) | 280 | 254 | | |
| $BOD_7$/TN (mg/mg) | 4.6 | 2.9 | 3.3 | 0.8 |

# 6.2 Nitrogen removal




Nitrogen concentrations in the incoming water to the biological treatment (PTW, primary treated water) and in the effluent are presented in Table 12. On average the effluent total nitrogen concentration was 4.3 mg/L. Three out of 51 weekly composite samples were above the target concentration of 6 mg N/L. The reduction of total nitrogen (measured in primary treated water) including reject water was 90.2%.

**Table 12. Nitrogen concentrations in primary treated water (PTW) and effluent during 2019.**

| Parameter | Limit | Average | Min | Max | No. of weekly samples |
|---|---|---|---|---|---|
| TN PTW (mg/L) | - | 45 | 25 | 58 | 52 |
| TN EFF (mg/L) | 6 | 4.3 | 2.1 | 8.2 | 51 |

Effluent nitrogen concentrations as weekly composite samples are presented in Figure 13. During most of the year nitrogen removal has been satisfactory with stable effluent concentrations below 6 mg/L as also shown previous years. In November (week 45-47) problems with aeration in BR3-BR5 led to reduced nitrification causing effluent total nitrogen to exceed 6 mg/L. However, reduced nitrification capacity around this time of the year has been observed every year.

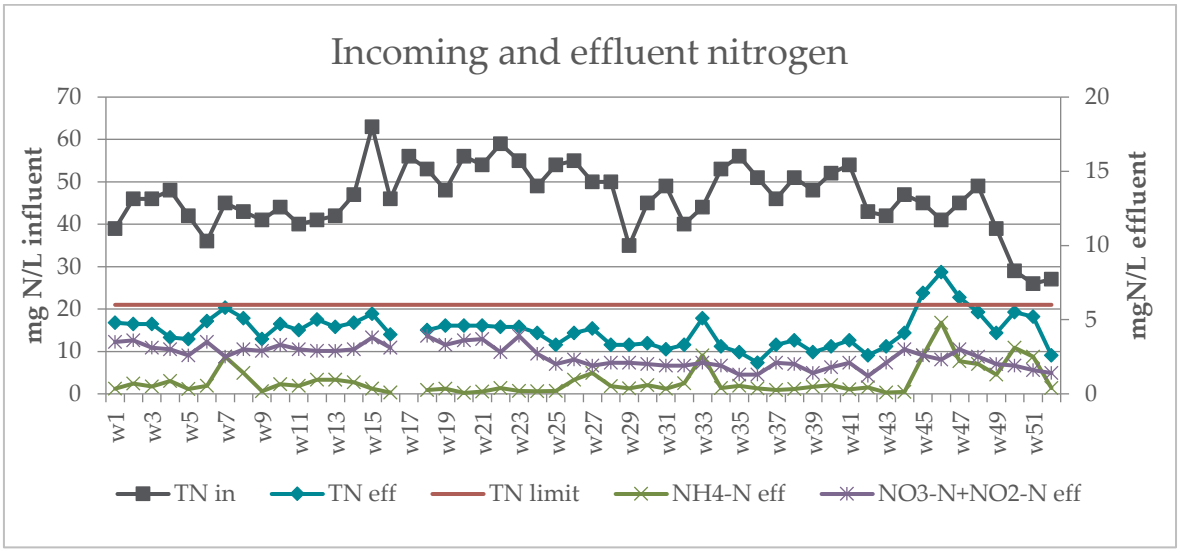



**Figure 13. Incoming and effluent nitrogen concentrations from analysis of weekly composite samples. Limit for effluent total nitrogen was set to 6 mg N/L. Effluent data is missing from week 17.**

In Table 13, key values for the nitrogen removal are presented for the MBR pilot and compared to the SFA design. The amount of removed total nitrogen was the same as 2018 and somewhat higher compared to the design. Different external carbon sources have been tested during this year, methanol dosage was used for the majority of the year and the average consumption per day was the same as 2018, 0.8 kg COD/d. Details about the other external carbon sources can be found in section 6.2.2 Denitrification.

The nitrogen removal rate presented in Table 13 was similar between the pilot and the SFA design. The large differences in system configuration makes it impossible to conduct a fair comparison between aeration to biological treatment. One reason for a higher value in the pilot is that the basins in the pilot were about 4 times less deep than the design. The aeration of the biology is slightly lower 2019, compared to previous years, 2017 and 2018, when average aeration was 54 $m^3/h$ and 52.4 $m^3/h$, respectively.

**Table 13. Comparison of parameters related to the nitrogen removal between operational data from the pilot and the design of future Henriksdal.**

| Parameter | Unit | Value Pilot | Design future Henriksdal | Value pilot/ scaled design future Henriksdal[i] |
|---|---|---|---|---|
| Removed nitrogen (including reject water) | kg N/d | 3.6 | 21 000 | 114% |
| Nitrogen removal rate | g N/kg VSS, d | 17.9 | 17.6 | 102% |
| Aerated sludge age (including membrane tanks) | d | 7.0 | 9.4[ii] | 74% |
| Aeration biology (activated sludge) | $Nm^3/h$ | 50.1[iii] | 68 000 | 494% |
| Specific oxygen demand (SOTR) | kg O2/d | 58[iv] | 240 000 | 162% |
| Consumption of external carbon | kg COD/d | 0.8[v] | 12 000 | 45% |

[i] The value of SFA-design divided by the scale factor 6 700

[ii] Assumed that ¾ of all membrane tanks are in operation as a yearly average.

[iii] $m^3/h$ not $Nm^3/h$

[iv] 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 $O_2/Nm^3$, m.

[v] Only including methanol (week 1 to week 34).

### 6.2.1 Nitrification

Nitrification worked satisfactory most of the year with low effluent concentrations except for shorter periods and a peak in effluent ammonium towards the end of the year. When comparing the aeration of the biology with aeration of the membrane tanks (Figure 14) it can be observed that the membranes most of the time were operated at the lower aeration level (Leap-Lo corresponding to 14 $m^3/h$ each, 28 $m^3/h$ in total) with only a few peaks above 28 $m^3/h$. The aeration of the biology varied as weekly average between 28 and 85 $m^3/h$ and was 50 $m^3/h$ as yearly average.

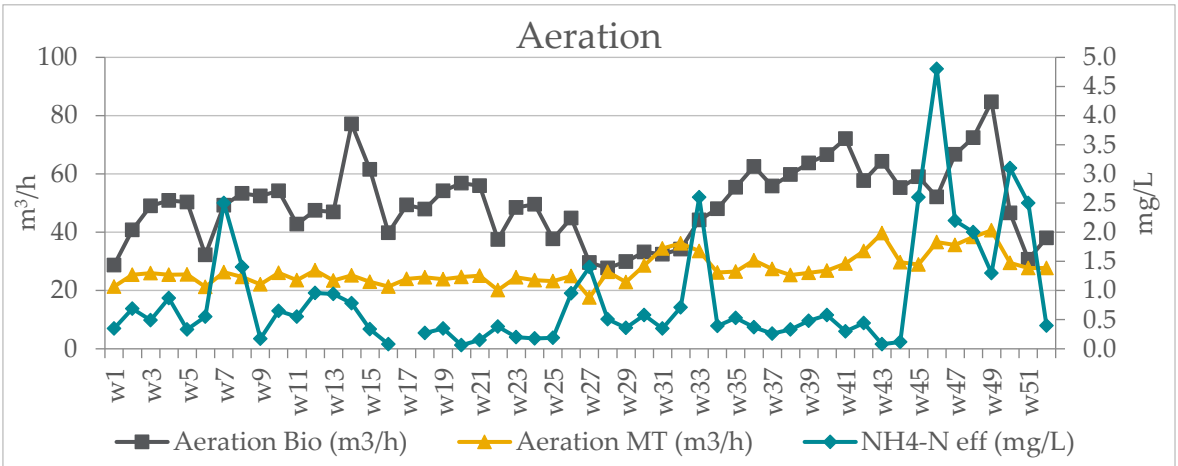



**Figure 14. Aeration need in biology and membrane tanks (MT) together with effluent $NH_4$-N.**

On average, the aeration of the membranes accounted for 36% of the total aeration, same as previous years (2017 and 2018). This is a significant reduction compared to the first year (2016) when 54% of total aeration was used for the membranes.

As previous years, problems with foaming and unreliable DO sensor readings in the foamy sludge have made aeration control difficult at times. BR4 and BR5ox was aerated to keep a manually selected DO setpoint between 2 and 3 mg/L throughout the year. During the colder season (beginning and end of the year) BR3 was continuously aerated. In periods, from week 17, intermittent aeration of BR3 was activated. The trigger for aeration of BR3 was first ammonium concentrations in BR2 above 6 mg/L. This was changed (in week 43) to flow-based control, where aeration was on only when flowrates to the treatment was above 2.5 $m^3/h$ to prevent excessive aeration at night when inflow was low.

The initial aim for the pilot was to operate at a total sludge age of 25 days, to evaluate the performance at design conditions. 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. This year, in week 21, a feedback controller was implemented to automatically adjust the WAS flowrate in order to keep a setpoint of 8 000 mg SS/L in BR4.

The resulting total and aerated (including membrane tanks) sludge age is presented in Figure 15 together with the WAS flowrate. The sludge age was higher during the summer period although this is typically a period when a lower sludge age would be required for the nitrification. This is a result of lowering the WAS-pumping to maintain 8 000 mg SS/L in the biological treatment. The calculated total sludge age was on average 19 days which is lower than SFA design (25 days), however 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 calculated aerated sludge age (including membrane tanks) was on average 7 days.

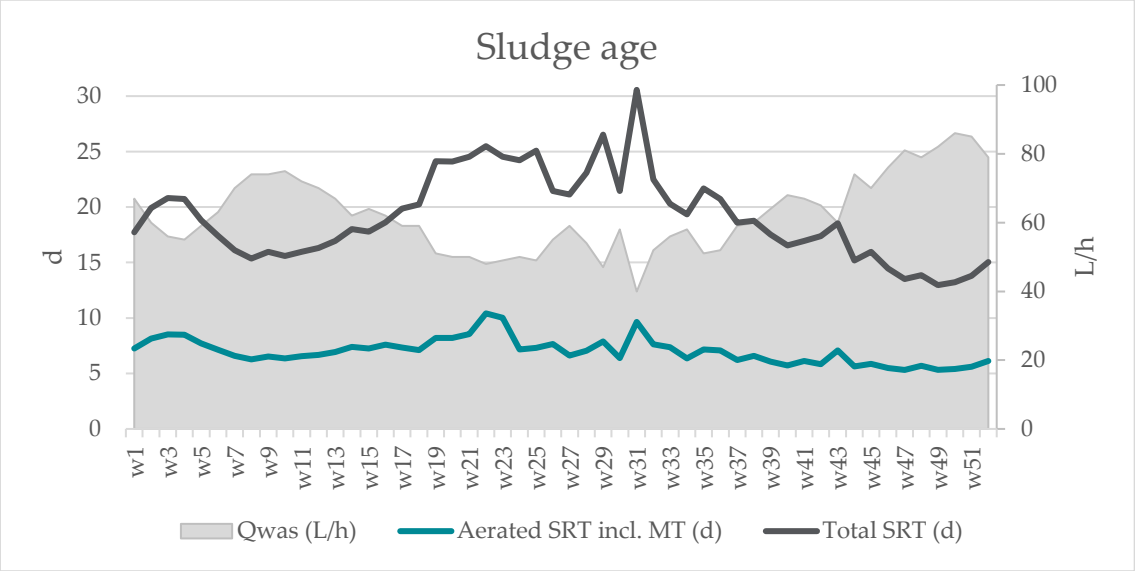



**Figure 15. Total and aerated (incl. MTs) sludge age (moving average one month back in time) together with WAS flowrate.**

## 6.2.2 Denitrification

Both pre and post denitrification is utilized in the pilot. For pre-denitrification (BR1, BR2 and sometimes BR3) the nitrate recirculation flow rate (from BR5deox to BR1) has been flow proportional to the inflow using a factor of 3 x Qin, except for a short period (3 days) when the heat exchangers were clogged and nitrate recirculation was stopped. A maximum flow of about 13 $m^3/h$ have been used as this corresponds to the maximal flow in the full scale. The concentration of $NO_3$-N in the recirculation flow was 2.8 – 6.0 mg/L. In addition, nitrate was recirculated with the RAS (from RAS-deox to BR1) with a flow corresponding to 4 x Qin. The nitrate concentration in the RAS was 1.3 - 3.9 mg $NO_3$-N/l.

For the post denitrification (BR6) different external carbon sources has previously been used. This year methanol was used until week 35 when acetic acid was tested for two weeks and after that, glycerol was used for the remainder of the year. The carbon source has been added to a point in the piping between BR5 and BR6 and the dosage have been controlled by online nitrate concentration.

Due to sensor problems the signal used has been alternating between nitrate sensor in BR6 and nitrate sensor in the effluent. The sensor readings from BR6 were preferred as this gives a faster control strategy. However, due to foaming and sensor failure, the effluent nitrate sensor was used at times. The nitrate setpoint was varied from 4 mg $NO_3$-N/L to 1 mg $NO_3$-N/L. To avoid overdosing carbon source, the pump used have in periods been operated intermittently in order to give time for the nitrate concentration to change after a period of dosing.

An overview of the external carbon source addition is given in Figure 16. All three carbon sources managed to control effluent nitrate and keep concentrations below 3 mg $NO_3$-N/L. The need for external carbon source varied more during the period with glycerol compared to methanol. The short test with acetic acid showed an increased need for COD compared to methanol. A more

detailed evaluation of the acetic acid is presented in the next section (6.2.3 Comparison of different external carbon sources).

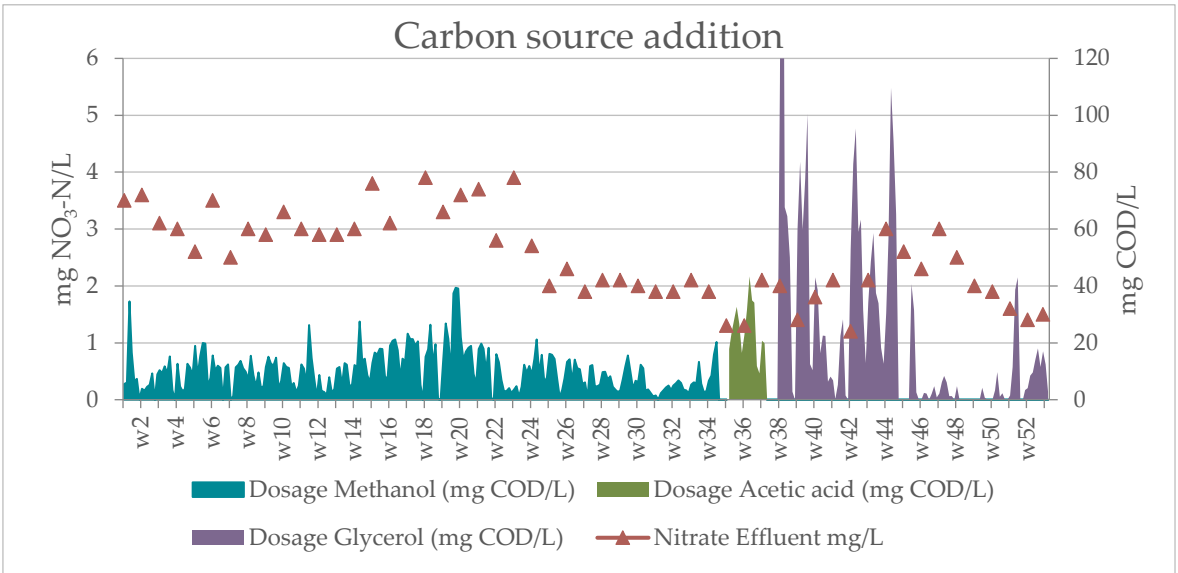



**Figure 16. Dosage of methanol, acetic acid and glycerol as daily average values and effluent nitrate analysed in weekly composite samples.**

The online sensors for nitrate in BR5deox and BR6 are helpful in the daily operation to monitor the denitrification, however it is difficult to use the data for estimation of how much nitrogen is denitrified in the post denitrification as the sensors tend to drift and the error in the reading is close to the difference in measured concentration. The weekly average concentration of nitrate, as measured by the sensors, in BR5deox, BR6 and in the effluent is presented in Figure 17. The concentrations in BR5deox varied between 3 and 6 mg $NO_3$-N/L while the concentration in BR6 varied between 0.2 and 4.7 mg $NO_3$-N/L. Week 25 to 43 the $NO_3$-N concentration in BR6 was much lower than the effluent readings. Although effluent nitrate can be higher than in BR6 due to nitrification in the membranes, the difference is more likely due to sensor problems in BR6.

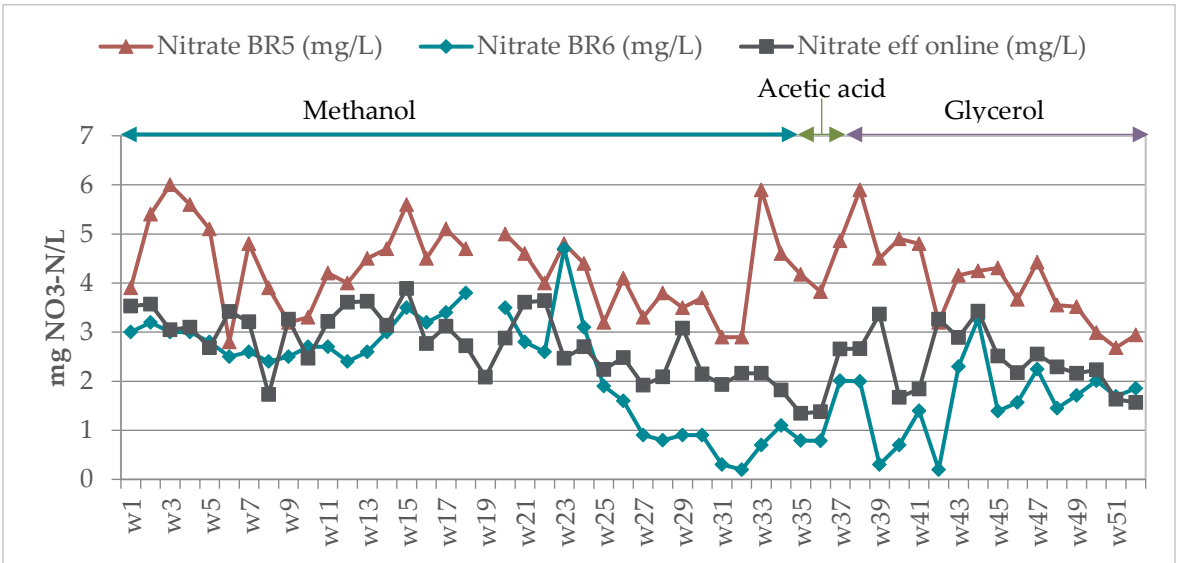



**Figure 17. Online nitrate concentrations as weekly average.**

## 6.2.3 Comparison of different external carbon sources

The nitrogen treatment process design for future Henriksdal WWTP is based on methanol as external carbon source for post denitrification. Methanol was chosen since it is a well-known, well-functioning carbon source that is available in large quantities to low cost. It has been used to boost pre-denitrification at Bromma WWTP for many years. However, in order to decrease the carbon dioxide footprint of the two WWT-plants SVOA is searching the market for more environmentally friendly options for the future. Methanol is also flammable, which is a risk to the work environment, at the same time as it requires ATEX approved installations and special permits. It would thereby be preferred to find a carbon source that is non-flammable. Furthermore, the carbon source storage facility for the full-scale treatment in Henriksdal will not be in operation during the first phases of the SFA-project and therefore a temporary installation will be needed at Henriksdal during a few years. With this background, different carbon sources have been tested within the pH2040 project, both in lab scale batch tests and in the pilot line.

A summary of the carbon sources tested so far is presented in Table 14.

**Table 14. Some properties of the tested carbon sources.**

| | Acetic acid (68-80%) | Brenntaplus | Glycerol | Methanol (fossile) |
|---|---|---|---|---|
| Produced from | Biprod. food industry | Non fossil | Bioprod. biodiesel production | Natural gas |
| COD-content (g/l) | 1 070[3] | 1 000[4] | 1 470[1] | 1 200[1] |
| Specific COD consumption (g COD/g N-red) | 6.1[2] | - | 4.2[2] | 4.8[2] |
| Max den. growth rate at 20°C ($d^{-1}$) | - | - | 2.0-3.4 | 1.0-1.8 |
| Specific sludge prod. (g SS/g COD) | - | - | 0.3 | 0.4 |
| Bulk density ($kg/m^3$) | 1 060[3] | 1 270[4] | | 790 |
| ATEX- classified | no | no | no | yes |
| Freezes/ Boils (°C) | -10/118[3] | -15/130[4] | -38/290[5] | -98/65[1] |
| Dynamic viscosity at 21°C (mPa×s) | 1.06[3] | 13.5[4] | 1 485[1] | 0.59[1] |
| Corrosive | yes | no | no | no |
| pH | 2.5[3] | 8.7[4] | 7[5] | - |
| Contaminants | no | metals and phosphorus | metals | no |

[1]Jansson (2017), [2]USEPA (2013), [3]Safety Data Sheet, acetic acid, Helm, [4]Productblad Brenntaplus VP1 (2018), [5]Ingfeldt (2020).

## Pilot tests

Since 2016 four different carbon sources have been tested in the pilot; Brenntaplus, Methanol, Acectic acid and Glycerol. Although the operational conditions in the treatment line have varied over time an attempt to compare the specific consumption of carbon source for post denitrification have been made. Based on the online data of COD dosage and nitrogen removal (over the post denitrification zone, g N-red, and over the whole biological treatment, g N), the specific consumption of carbon source as g COD/g N or g COD/g N-red was calculated. Results are shown in Figure 18. In general, the specific COD-consumption was lower than the theoretical COD-consumption for denitrification, which is 2.86 g COD/g N-red (Metcalf & Eddy, 2014). Normally the meassured specific COD-consumption is higher than the theoretical value since some of the added carbon is utilized for microbial cell growth (sludge production). Normal values for methanol is 4.8 g COD/g N-red, for acetate and acetic acid it is 6.1 g COD/g N-red and for MicoC (glycerol based product ≈70%, commonly used in England and USA) it is 6.4 g COD/g N-red (USEPA, 2013). In the case of the pilot trials, the calculated specific COD-consumption for the different carbon sources also include denitrification utilizing internally produced carbon from hydrolysis, something that occures all the time, also during periods with no addition of external carbon. The numbers presented should therefore not be compared to litterature values but can still be used to evaluate the relative performance from the different carbon sources.

As can be seen in Figure 18 the specific methanol COD-consumption was much lower during 2018 and 2019 compared to when it was first introduced in 2017. According to litterature the denitrifying bacterial community need time, up to several months, to adapt to methanol before the carbon source can be efficiently used, something that is not seen for other carbon sources (Carlsson & Hallin, 2000). This might be the reason for the higher specific methanol COD conumption in 2017. Brenntaplus, acetic acid and glycerol all showed similar consumption of COD per g nitrogen removed in the post denitrification zone (values were: Brenntaplus 1.72, Acetic acid 1.86 and glycerol 2.04) which was clearly lower than the highest value for methanol (2.53 g COD/g N-red), but much higher than the lowest (0.93 g COD/g N-red in 2018).

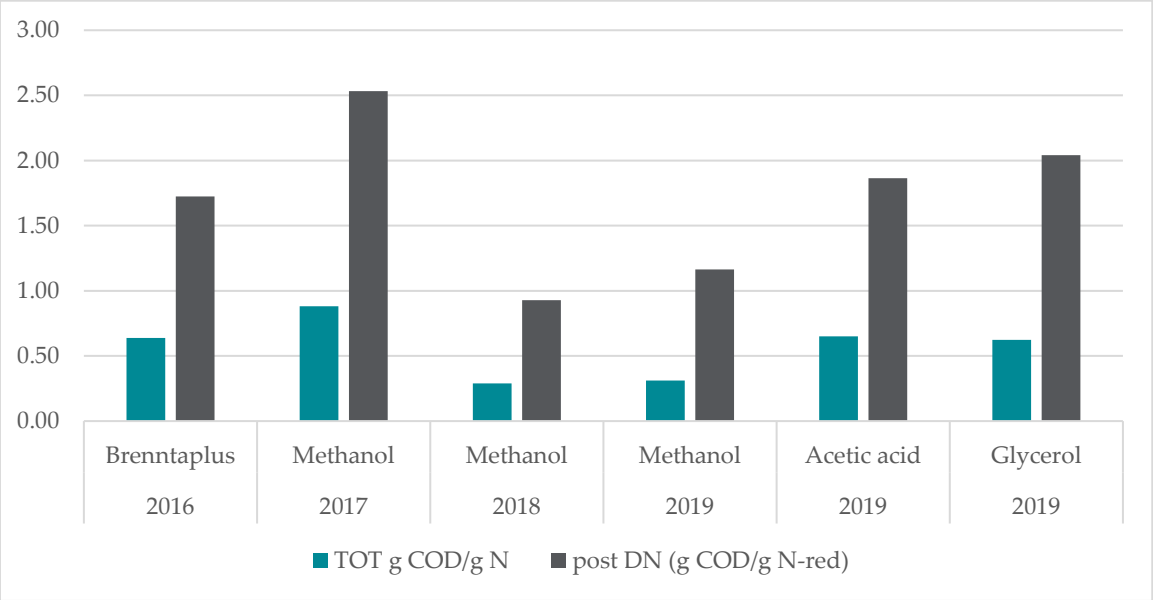



**Figure 18. Comparison of different external carbon sources used in the pilot 2016 to 2019.**

When looking at some of the operating conditions during the trials it is clear that the acetic acid trial stands out, being a very short trial with high temperature and low effluent nitrate concentration (due to low nitrate set-point for dosing), see Table 15.

**Table 15. Carbon source dosage and nitrogen removal for trials using different carbon sources.**

| | | Weeks | kg COD/d | kg $NO_3$-N red in post DN/d | Total kg N-red /d | Temperature (°C) | Effluent $NO_3$-N (mg/L) |
|---|---|---|---|---|---|---|---|
| 2016 | Brenntaplus | 19 | 1.75 | 1.0 | 2.7 | 16.9 | 3.1 |
| 2017 | Methanol | 44 | 2.42 | 1.0 | 2.7 | 17.3 | 2.9 |
| 2018 | Methanol | 25 | 0.80 | 0.9 | 2.7 | 17.7 | 3.3 |
| 2019 | Methanol | 34 | 0.85 | 0.7 | 2.7 | 18.2 | 2.8 |
| 2019 | Acetic acid | 2 | 2.13 | 1.1 | 3.3 | 21.9 | 1.3 |
| 2019 | Glycerol | 15 | 1.92 | 0.9 | 3.1 | 18.6 | 2.1 |

## Lab tests

In order to further compare and follow any adaption of the sludge to a new carbon source denitrification batch tests have been carried out. The methodology used for the denitrification batch tests is described by van Loosdrecht et al. (2016) and is the same as used previously in the project. Temperature compensated denitrification rates for different external carbon sources using sludge from Henriksdal (no external carbon source) and Bromma (adapted to methanol) as references in the batch tests are presented in Figure 19.

For methanol, denitrification rate during 3 tests on MBR sludge in May 2019 varied between 2.0 and 2.4 mg $NO_3$-N/ g VSS, h. For the Bromma ASP sludge the rate was slightly higher which suggests that this sludge was more adapted to methanol. The reference MBR-sludge (no added carbon source) varied between 0.7 and 1.7 mg $NO_3$-N/ g VSS, h.

For acetic acid, the test results were different than expected. The pH in the lab reactor decreased from around 7.0 to about 5.5 when acetic acid was added to the reactor. The low pH resulted in inhibited denitrification and much lower denitrification rates compared to methanol. When pH was adjusted in the acetic acid reactor (test on May 20$^{th}$) the denitrification rate increased from 1.3 to 3.5 mg $NO_3$-N/ g VSS, h which still was lower than expected, however, higher than with methanol. Common values for denitrification rate for acetic acid is 5.9 mg $NO_3$-N/g VSS, h. Lower rates have been observed in literature, for example Elefsionitis & Li (2016) measured 1.46 mg $NO_3$-N/ g VSS, h at pH 6.5 using acetic acid. Glass & Silverstein (1999) also observed a significant reduction in denitrification rate at pH 6.5 to 7 compared to at pH 7.5 to 9. In the same publication, higher denitrification rate was reported for acetate in an acetate-acclimated biomass compared to a biomass used to methanol. It can be assumed that there is a certain adjustment time for the biomass to a new carbon source and that the denitrification rate for acetic acid would have been higher if the biomass were used to acetic acid.

In September glycerol was tested and showed a denitrification rate of 2.1-2.4 mg $NO_3$-N/g VSS, h.

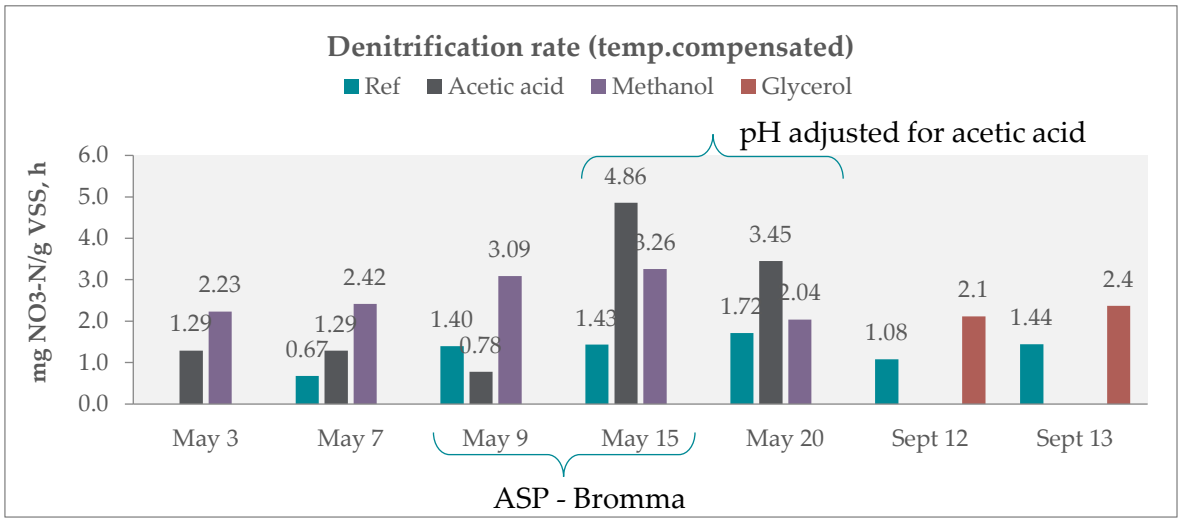



**Figure 19. Denitrification rate during laboratory denitrification batch tests using different carbon sources.**

The specific COD-consumption during denitrification tests varied (Figure 20) and showed correlation to how denitrification was affected by pH. For acetic acid the COD-consumption was lower in the first tests compared to the last, when pH was adjusted. The COD-consumption for the test on May 20th was high even in the reference reactor which suggests that there was a significant amount of easily available COD present in the sludge already before the addition of external carbon source. The specific COD-consumption in the batch test were in general higher than the results from the pilot-scale trials which was expected.

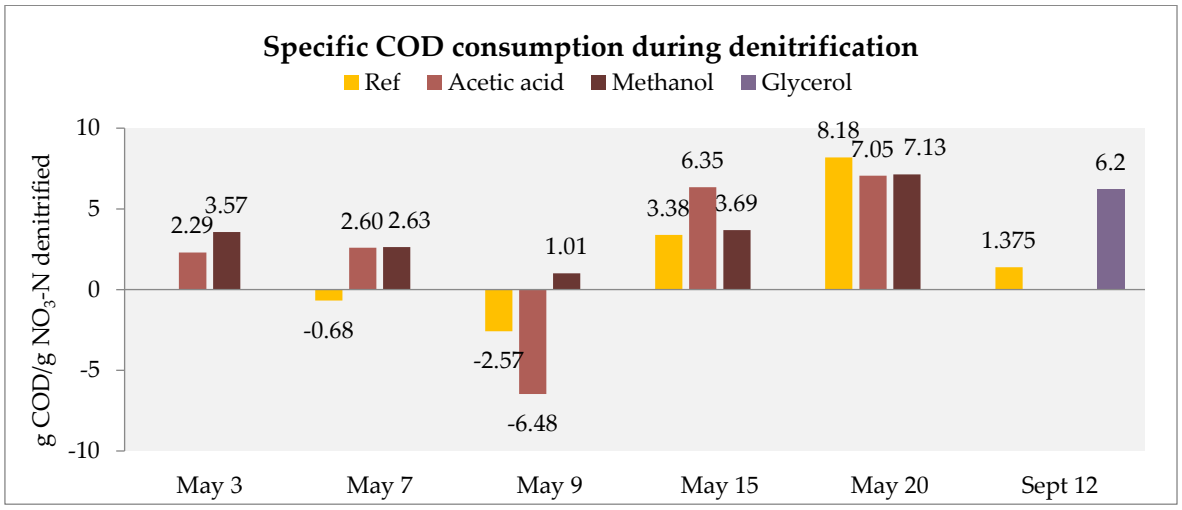



**Figure 20. Specific COD consumption during laboratory denitrification batch test in laboratory tests using different carbon sources.**

Batch denitrification test previously performed using pilot sludge have shown a specific COD-consumption for methanol of 1.0-2.7 g COD/g N-red with increased ratio over time as the sludge adapted to methanol. The tests 2019 (excluding deviating results) showed further increase to about 2.6 – 3.7 g COD/g N-red, which is closer to the reference values of 4.8 g COD/g N-red. This could be an indication of increased acclimatization to methanol based on the assumption that the

hydrolysis products, and not methanol, were utilized as electron acceptor for denitrification in the initial tests resulting in a low COD-consumption whereas methanol was utilized, and thus consumed to a greater extent, in the later tests.

The initial specific COD-consumption when using glycerol was 6.2 g COD/g N-red which was similar to the results for pH adjusted acetic acid and similar to reference values for glycerol (USEPA 2013). Future tests will be conducted to monitor any adaption of the sludge.

An important observation during the denitrification tests with acetic acid was a significant release of phosphorus. No such release was observed for methanol or in the reference test. The release was not related to the pH, as it was observed also in the pH adjusted test with acetic acid. The increase in phosphorus was observed immediately as the acetic acid was introduced and continued to increase reaching a final concentration between 2 and 10 mg P/L. The phosphate concentration in the acetic acid was analysed but considered neglectable.

The phosphate release most likely occurs due to the enhanced biological phosphorus removal (EBPR) activity detected in the pilot line. During anaerobic (or anoxic) conditions and in presence of VFA, phosphate is released from poly-P granules in the cells of certain bacteria (more about this in chapter 6.3.1). Acetic acid belongs to the VFA group and most likely induced a phosphorous release.

During the denitrification tests a much higher dose of carbon source is added to the reactor compared the actual dose in pilot or full scale. The effect on pH and phosphate release at realistic (average and maximum) doses of acetic acid was also tested in the lab reactors. The tests showed an immediate increase of phosphate in the reactor to above 1 mg P/L at average dose and to 3 mg P/L using the maximum dose (Figure 21). After the initial release, the phosphate concentration showed a decreasing trend during the first 20 to 40 minutes followed by stable values for the remainder of the test. A second test was carried out comparing sludge from BR5 and RAS with max dose of acetic acid and more frequent sampling in the first minutes of the test. The results showed an increase in phosphate concentration during the first 2 minutes (Figure 22) followed by a rapid decrease.

Considering the strict effluent requirements for phosphorus, the acetic acid was therefore considered inappropriate as external carbon source for post denitrification. The other carbon sources, which are not VFAs, did not cause a P-release.

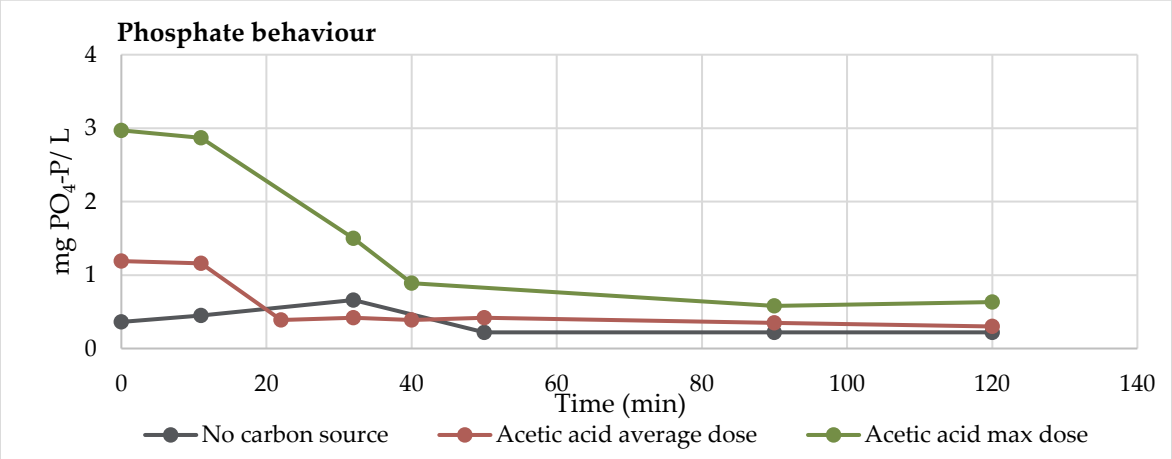



**Figure 21. Phosphate concentrations during denitrification tests with acetic acid as carbon source. Average dose was 0.07 ml acetic acid to 5 L sludge. Max dose was 0.15 ml acetic acid to 5 L of sludge.**

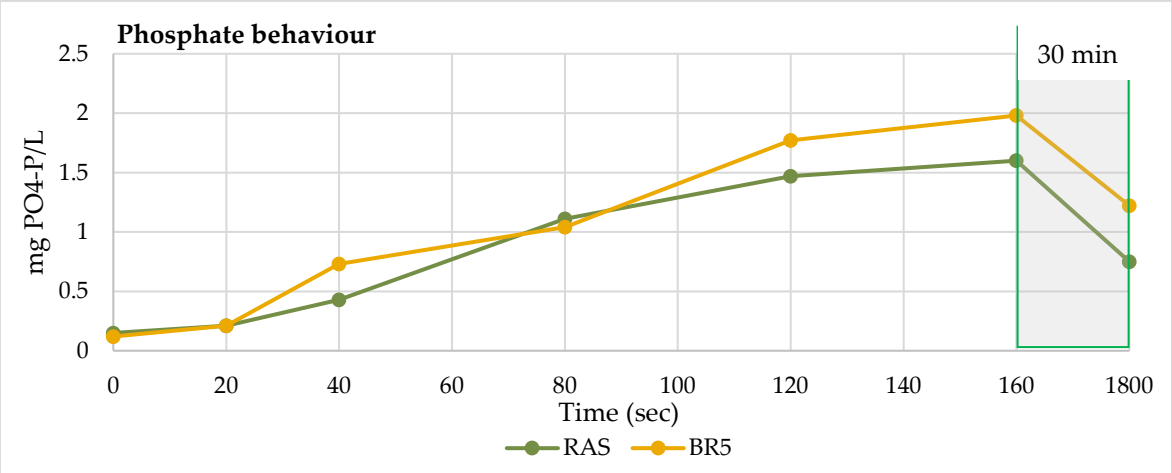



**Figure 22. Phosphate concentrations during denitrification tests with acetic acid as carbon source, dosage was 0.15 ml acetic acid to 5 L of sludge from BR5 and RAS.**

## 6.3 Phosphorus removal

| Trial | J | F | M | A | M | J | J | A | S | O | N | D |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Effluent phosphorus target 0.10 mg P/L | | | | | | | | | | | | |
| Strategy for increased PIX dosage | | | | | | | | | | | | |
| Effluent phosphorus analyser in use | | | | | | | | | | | | |

The goal of reaching stable effluent phosphorus concentrations below 0.15 mg P/L had been achieved previous years by using a control strategy with dosage of ferrous sulphate and ferric chloride in three points in the process, where two points were controlled using feedback control from online effluent phosphate measurements. The dosage of ferric chloride was then very low (<2 mg Fe/L as weekly average). In week 4 2019 a change in precipitation strategy was made in order

to dose more ferric chloride than before to be able to study any effects on the membrane performance. A new target of 0.10 mg P/L was set from week 14.

The phosphorous concentrations in and out from the biological treatment is presented in Table 16 and Figure 23 below. The yearly average effluent total phosphorus corresponded exactly to the new target of 0.10 mg P/L.

**Table 16. Phosphorus concentrations in primary treated water (PTW) and effluent during 2019.**

| Parameter | Limit | Average | Min | Max | Nr of weekly samples |
|---|---|---|---|---|---|
| TP PTW (mg/L) | - | 5.5 | 2.6 | 7.4 | 52 |
| TP EFF (mg/L) | 0.20 | 0.10 | 0.028 | 0.22 | 51 |

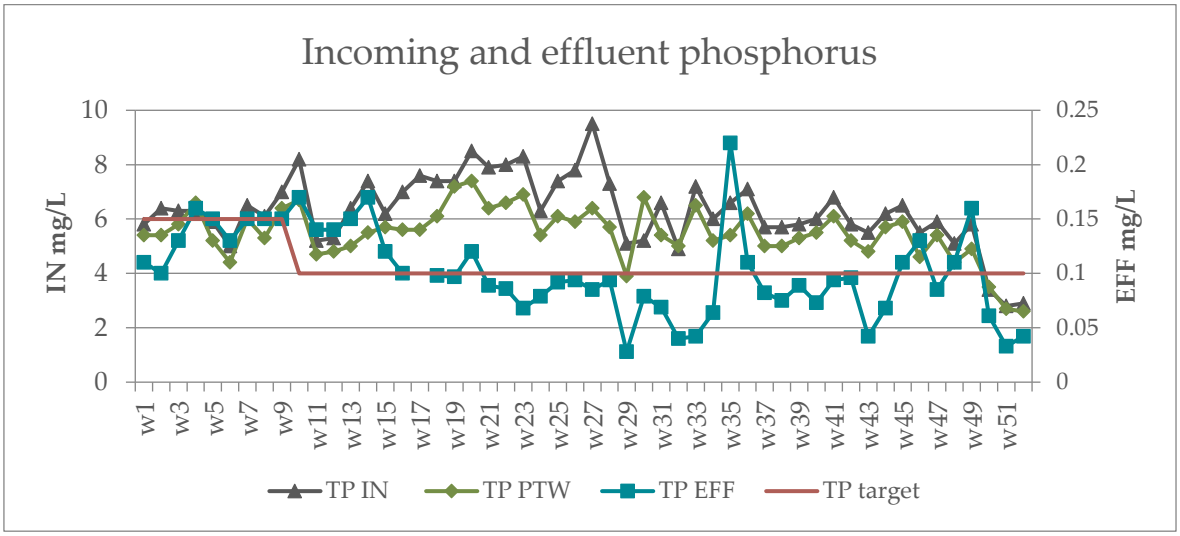



**Figure 23. Influent and effluent total phosphorus analysed on weekly composite samples. Effluent data is missing for week 17.**

Key parameters for the phosphorus removal, both for the pilot and for the SFA design, are presented in Table 17. The phosphorus load in the pilot was much higher than the phosphorus load which the SFA design was based on. This resulted in higher phosphorus removal in the biology of the pilot in comparison to the SFA design. Despite the high phosphorus removal, iron consumption in the pilot was lower in the pilot compared to the SFA design. Although the consumption of iron for precipitation of phosphorus was one of the most uncertain parameters in the SFA design, the difference is supported by the observed EBPR in the pilot. In the SFA design a yearly average dosage of 20 mg Fe/m$^3$ treated water was assumed (sum of the three dosing points). In the pilot, the average dosage 2019 was 13.7 mg Fe/L.

**Table 17. Comparison of operational data from the pilot with data for the SFA design, yearly average values.**

| Parameter | Unit | Value pilot 2019 | SFA design | Value pilot/scaled SFA design[i] |
|---|---|---|---|---|
| Phosphorous load influent | kg P/d | 0.53 | 2 594 | 137% |
| Phosphorous load PTW | kg P/d | 0.46 | 1 580 | 196% |

| | | | | |
|---|---|---|---|---|
| Phosphorus load reject water | kg P/d | 0.01 | 480 | 17% |
| Total phosphorous load biology | kg P/d | 0.47 | 2 060 | 154% |
| Phosphorus load effluent | kg P/d | 0.0087 | 80 | 73% |
| Phosphorus removed in biology | kg P/d | 0.45 | 1 980 | 154% |
| Iron consumption (PS+BR4+BR6) | kg Fe/d | 1.16 | 10 000 | 78% |
| Iron consumption per removed phosphorus | mole Fe/ mole P | 1.4 | 2.8 | 50% |
| Phosphorus in sludge | % of SS | 3.5 | 5.4[ii)] | - |
| Iron in sludge | % of SS | 7.6 | - | - |
| VSS in sludge | % of SS | 74 | 66 | - |

i) SFA divided by 6 700.

ii) No EBPR. Mainly chemically bound phosphorous.

The total amount of iron dosed is presented as daily average values in Figure 24. The base dose of $Fe^{2+}$ was added to the pre-aeration which was controlled flow proportionally to a dose of 5-12 mg Fe/L (manually adjusted based on effluent phosphate). A supplementary dosage of $Fe^{2+}$ was added to the aerated part of the biological treatment (BR4). This dose was controlled using a slow (time constant 1 h) online feedback control from effluent phosphate. The weekly average of the dose to BR4 varied in the range 0-13.5 mg Fe/L. A third and final polishing dose using $Fe^{3+}$ was added in BR6 (just prior to the membrane tanks). This dose was initially added during shorter peaks in effluent phosphate. In week 4 the dosing of $Fe^{3+}$ was changed to be proportional to the dose in BR4 in order to use the $Fe^{3+}$ more often than before.

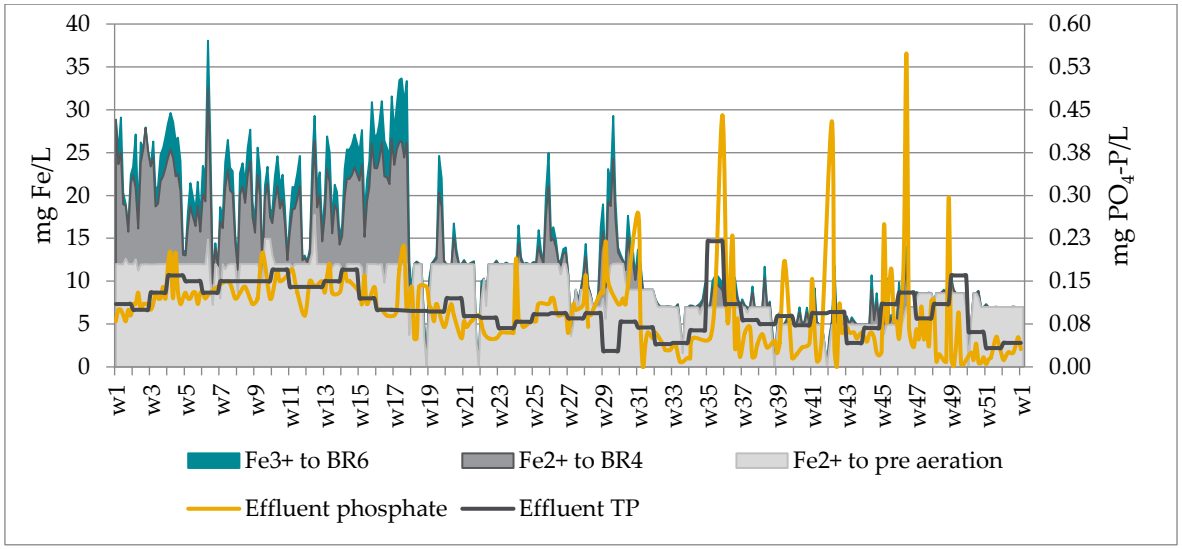



**Figure 24. Iron dosage (the sum of three dosing points) as daily average, effluent phosphate analysed in daily composite samples and effluent total phosphorus in weekly composite samples.**

A drastic reduction in iron dosage can be observed during week 17 (Figure 24). In the beginning of week 17 the phosphate setpoint for the controller of dosage to BR4 and BR6 were lowered from 0.1 mg/L to 0.08 mg/L resulting in the highest dosage of iron to the biology for a couple of days. In week 18, the inflow was heavily diluted by rain causing low phosphorus concentrations in the water treatment and effluent, which significantly reduced the dosage of precipitation chemicals.

Although the load returned to normal after week 18, the need for precipitation chemicals remained lower than before.

The iron content of the sludge decreased from 10-12% in the beginning of 2019 to about 3% at the end of the year (Figure 25). The phosphorus content in the sludge was more even throughout the year, around 3-4% of TSS.

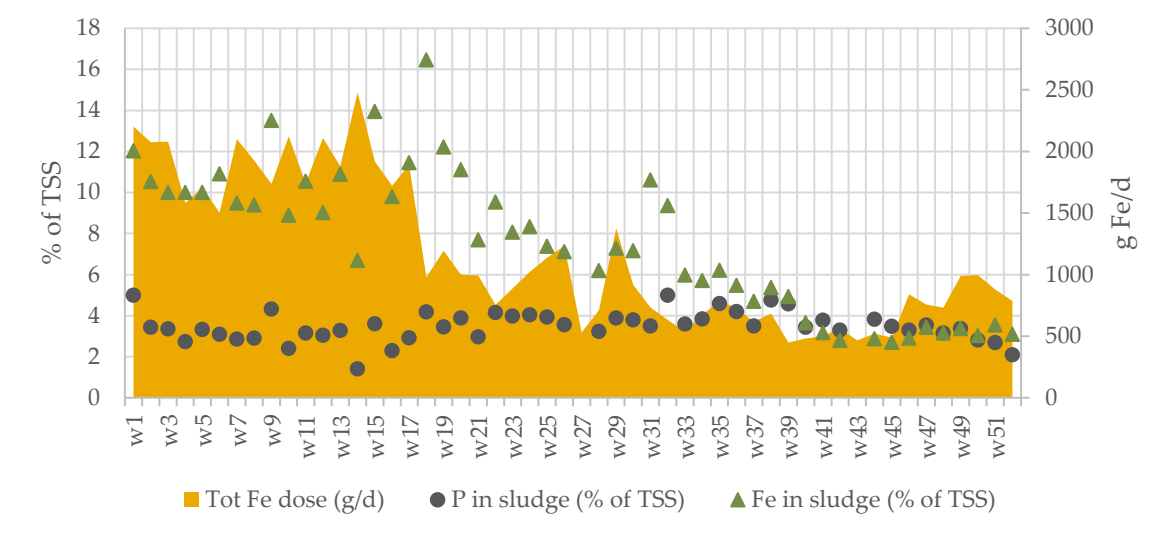



**Figure 25. Iron and phosphorous in sludge.**

## 6.3.1 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, an 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, which was also confirmed by monthly phosphate release tests (as described in Tykesson & la Cour Jansen, 2005) during 2018.

Results from the P-release tests from 2018-2019 are shown in Figure 26.

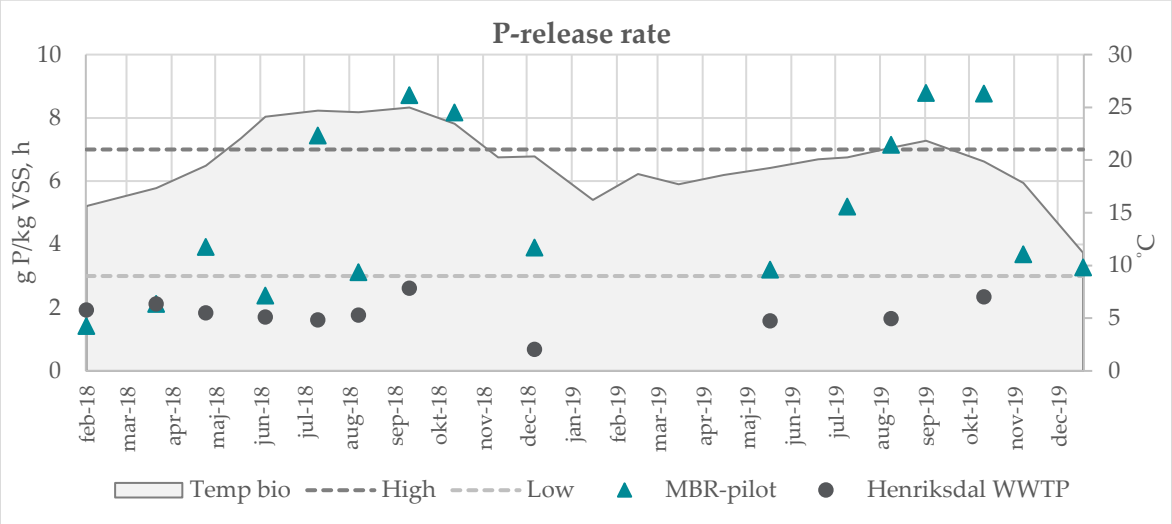



**Figure 26. EBPR activity in the MBR-pilot and Henriksdal WWTP (CAS). Dotted lines show limits for EBPR-activity (Janssen *et al.*, 2002), below the lower line = poor EBPR, above the higher line = high EBPR. Temp bio is the temperature in the treatment process.**

The activated sludge from the MBR-pilot showed high phosphate release rates in July to October, when the temperature in the treatment process is warm, for both 2018 and 2019. P-release rates above 8 g P/kg VSS, h was measured on several occasions, which is high enough to indicate a stable EBPR process but not as high as in well-established Swedish EBPR-plants like Öresundsverket, 12,5 g P/kg VSS,h (Tykesson et al., 2005) and Duvbacken WWTP, 20 g P/kg VSS,h (Salmonsson et al., 2017). In winter the phosphate release rate tends to decrease to almost as low levels as in the reference sludge from Henriksdal WWTP (CAS with no EBPR).

According to design guidelines for EBPR (Metcalf & Eddy, 2014) the following is required to achieve bio P:

(1) a strictly anaerobic zone with access to VFA (>8 g/g P) and 0.5-1.0 h hydraulic retention time followed by an aerated zone,
(2) a COD:P ratio in the inflow of >60:1,
(3) low load of TP from digested sludge reject water.

The MBR-pilot do not have an anaerobic zone in the water line, however, the DO and nitrate is often low, <0.2 mg $O_2$/l and <3 mg $NO_3$-N/l, in the anoxic pre and post denitrification zones, which both are followed by aerated zones. The hydraulic retention time in the pre denitrification zone is around 10 min, and in the post denitrification zone around 20 min, which is lower than the required minimum of 30 min. VFA is not analysed regularly in the process. A measurement campaign in July 2017 showed concentrations of acetic acid of 20 ± 3 mg/L (all other VFA:s was under detection limit) in the inlet to the biology, which is in the normal range for municipal wastewater, giving a VFA:TP ratio of approximately 4. The COD:TP ratio in the inlet to the biological treatment was 76 as a yearly average over 2019 which provides the necessary requirements for VFA production through hydrolysis and high WAS-production for removal of P. The TP and $PO_4$-P in the reject water from sludge dewatering was relatively high during 2019 - 122 mg/L and 33 mg/L respectively. The values can be compared to the concentration in the reject

water at Henriksdal WWTP (no bio-P) which were 9.9 mg TP/l and 1.6 mg $PO_4$-P/l. There was a correlation between high bio-P activity (P-release rate), low Fe concentration in WAS and phosphate concentration in the reject water. The TP load from the reject water was around 10% of TP in the inlet to the biology.

In Figure 27 the influence of iron dosage on the bio-P and P-removal is visualised. Combinations of EBPR and chemical precipitation have been shown to work previously at Swedish WWTPs (Jansen et al., 2009). The phosphorus content in sludge was rather similar over time, around 3% of SS in WAS. The iron content in sludge varied with time and in general, a higher P-release rate was observed when the iron content in sludge was low. Figure 28 shows the relationship between iron in sludge, iron dose and P-release rate. There is no clear correlation between the iron content in sludge and the P-release rate (if data from only 2018 is plotted a moderate correlation can be seen) but there is a moderate correlation between the iron dose and the P-release rate. This could be due to the iron dose being controlled by the effluent $PO_4$-P concentration. So, iron is dosed when bio-P activity goes down (thus, the decrease in bio-P activity is not induced by the increased iron dose). Instead the bio-P activity seems to be depending on the water temperature and perhaps the seasonal changes in sludge quality and composition that occurs at most Swedish WWTPs. A more detailed study on the EBPR in the MBR-pilot will be conducted in 2020.

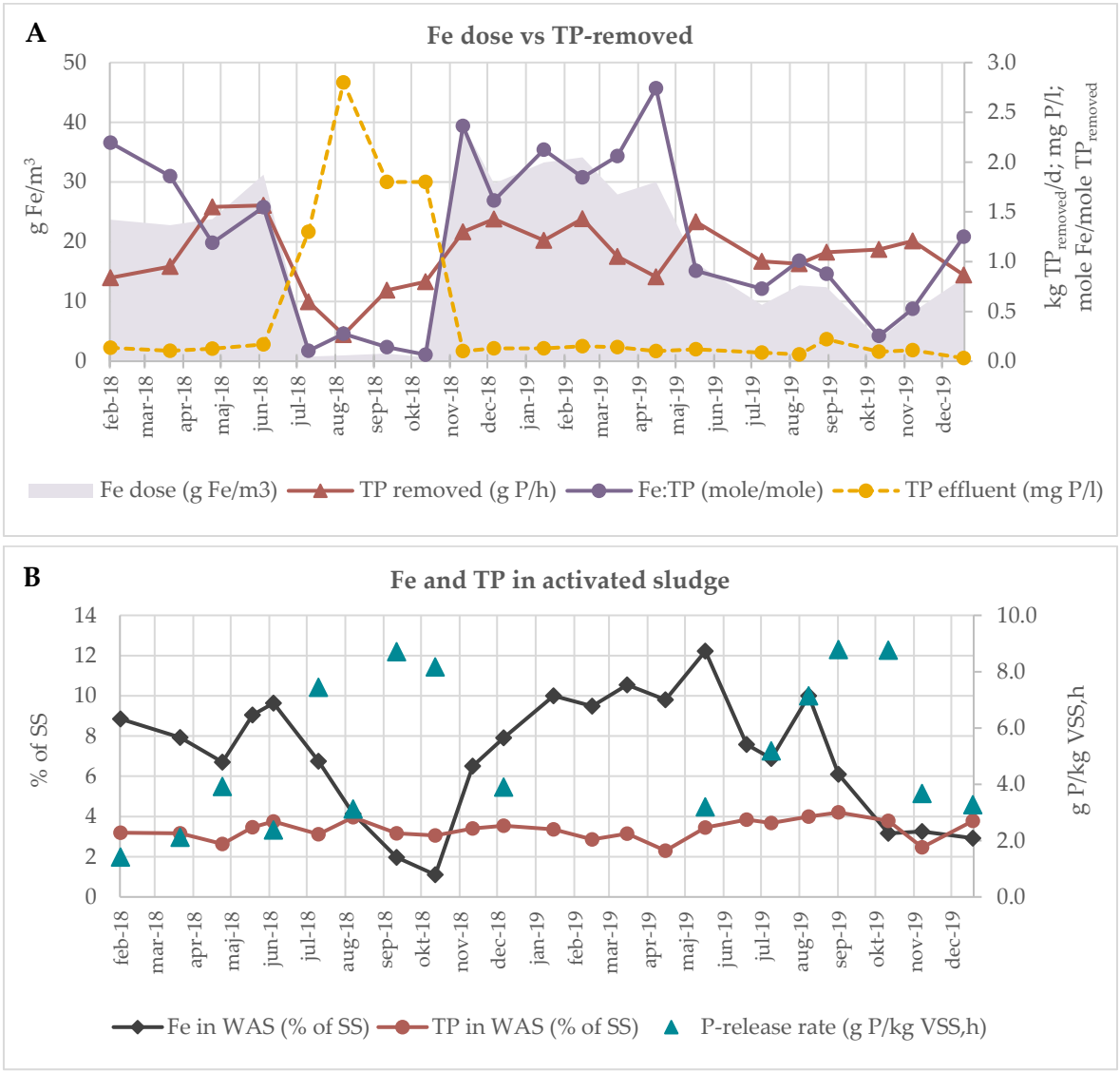



**Figure 27. A) Fe-dose in activated sludge and TP removed in the process. B) Fe and TP in waste activated sludge (WAS) together with P-release rate.**

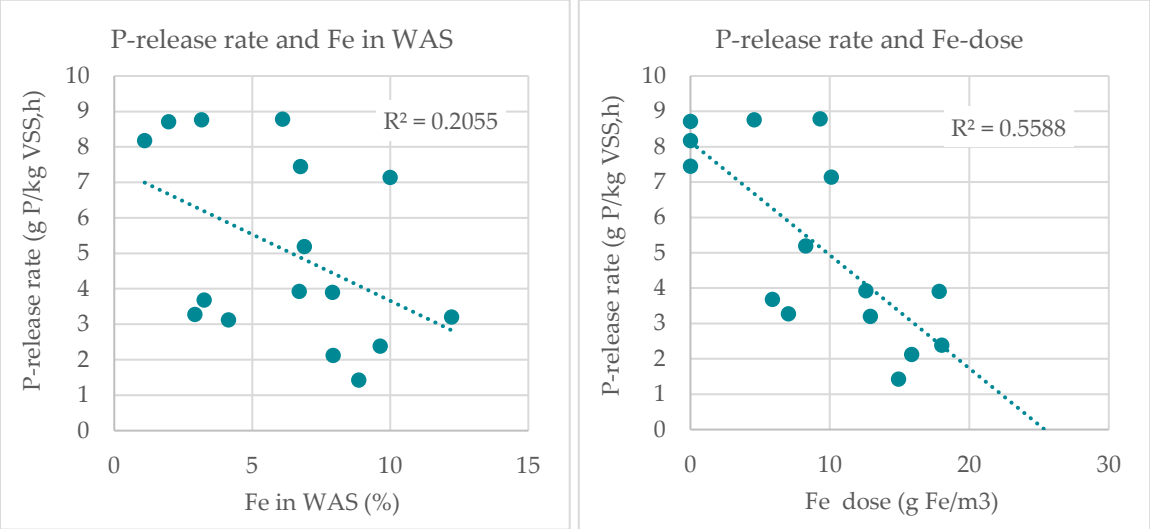



**Figure 28. The relationship between iron in waste activated sludge (WAS) and P-release rate (left) and iron dose and P-release rate (right). Data from 2018 and 2019.**

## 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 2019 and no specific measures have been taken to achieve a higher BOD reduction.

## 6.5 Membrane performance

| Trial | J | F | M | A | M | J | J | A | S | O | N | D |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Minimising membrane scouring air use | | | | | | | | | | | | |
| Oxalic acid and citric acid comparison | | | | | | | | | | | | |
| Minimising membrane cleaning chemical use | | | | | | | | | | | | |
| Recovery cleaning and sampling | | | | | | | | | | | | |
| Trial without membrane relaxation | | | | | | | | | | | | |
| Extended membrane operational cycle | | | | | | | | | | | | |

The standard membrane operation is 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 should be recycled to maintain design conditions. 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 10 % of withdrawn permeate during 2019.

This year many changes have been made regarding the membrane operation.

| | |
|---|---|
| Recovery cleaning with sodium hypochlorite | Week 11 |
| Recovery cleaning acids (oxalic and citric) | Week 13 |
| Synchronised MC with hypo and acid | Week 18 continuing |
| Test with no aeration of the membranes for one cycle (10 minutes of permeate production without aeration) | Week 28 |
| Operation without relaxation (capacity increased by 10%) | Week 29-34 |
| Operation with prolonged permeation cycle (15 min instead of 10 min, capacity increased by 3%) | Week 34 continuing |
| Operation with forced Leap-Lo (low membrane aeration) | Week 50 continuing |

### 6.5.1 Permeability

Permeability above 200 L/(m²·h·bar) is considered good according to the supplier. As can be seen in Figure 29, the permeability was above 200 L/(m²·h·bar) throughout the year for both membranes. A decreasing trend can be seen for both membranes.

In the beginning of the year, week 6, it was noted that the flow meters on the permeate lines from both membranes showed faulty values. Flow meters were switched in between the lines to see the difference. Permeability calculations increased with 100 L/(m²·h·bar) for MT2 and decreased with the same for MT1 after switching the flow meters. The flow meters showed damage on the inside, likely caused by a leaking valve on the sodium hypochlorite tube. In week 8 both flow sensors were replaced with new flow meters with sodium hypochlorite resistant liner.

The permeability was affected by the different operational changes and tests for the membranes, especially the different cleaning strategies. When acid maintenance cleanings were scheduled on the same day as the maintenance cleaning with sodium hypochlorite (starting week 18), a clear saw toothed patterned was observed in permeability (see Figure 29).

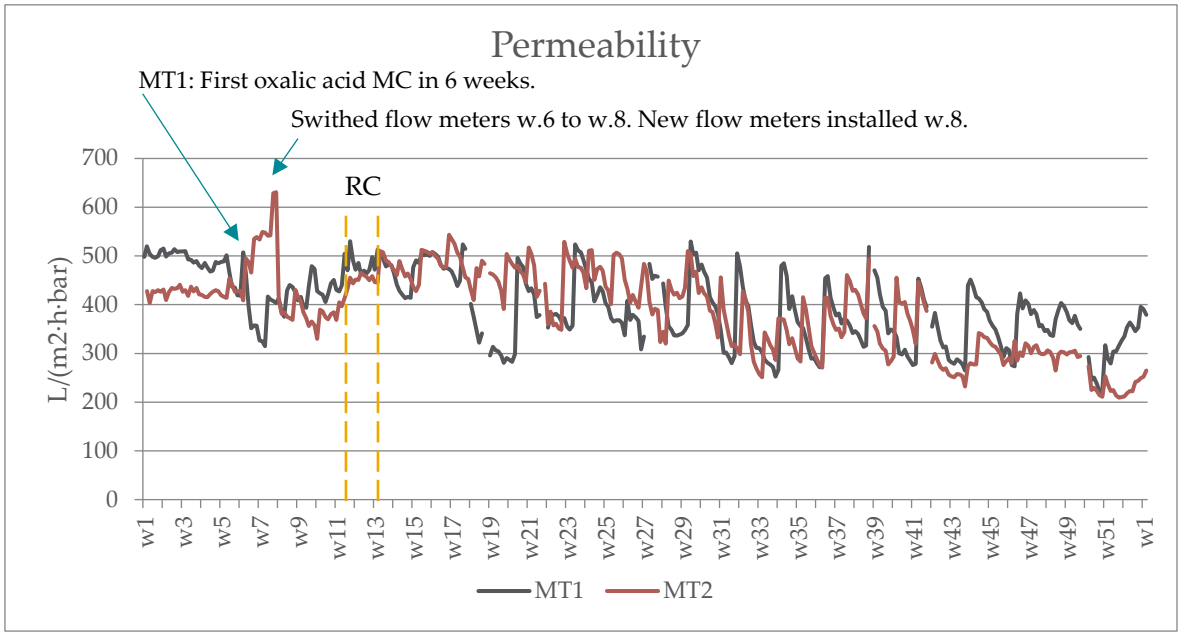



**Figure 29. Permeability (temperature compensated) for membrane 1 (MT1) and 2 (MT2) during project year 6 (2019). RC was performed in week 11 (hypo) and week 13 (acids).**

### 6.5.2 Flux and TMP

Fluxes for the two membranes are presented in Figure 30. Normally the membranes were operated with net flux around 20 to 25 L/(m²·h).

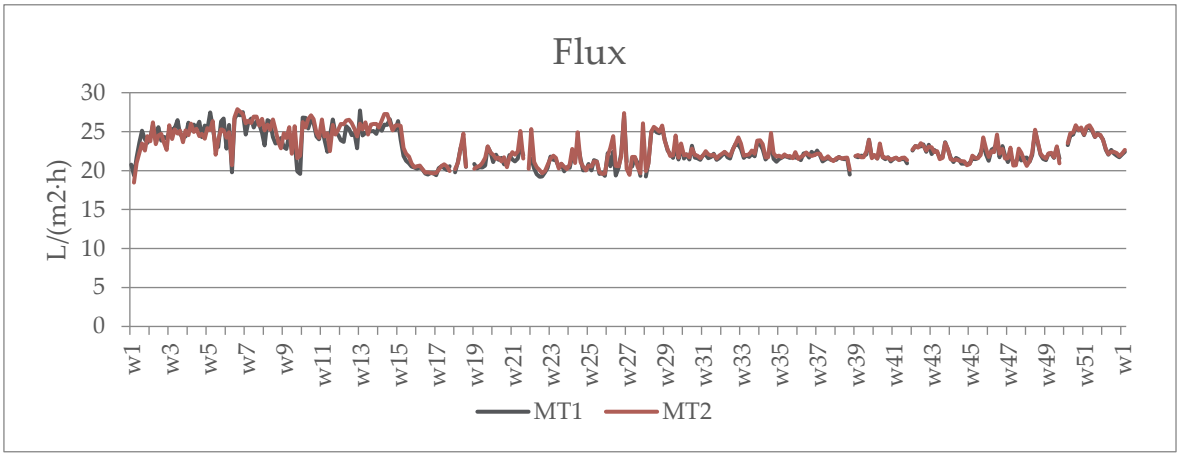



**Figure 30. Net flux during 2019.**

The transmembrane pressure (TMP) is presented in Figure 31. Based on daily average data, TMP varied between 36 and 148 mbar for MT1 and between 39 and 156 mbar for MT2 during 2019. An increasing trend in TMP towards the end of the year can be observed. The abrupt changes are often related to maintenance cleaning events.

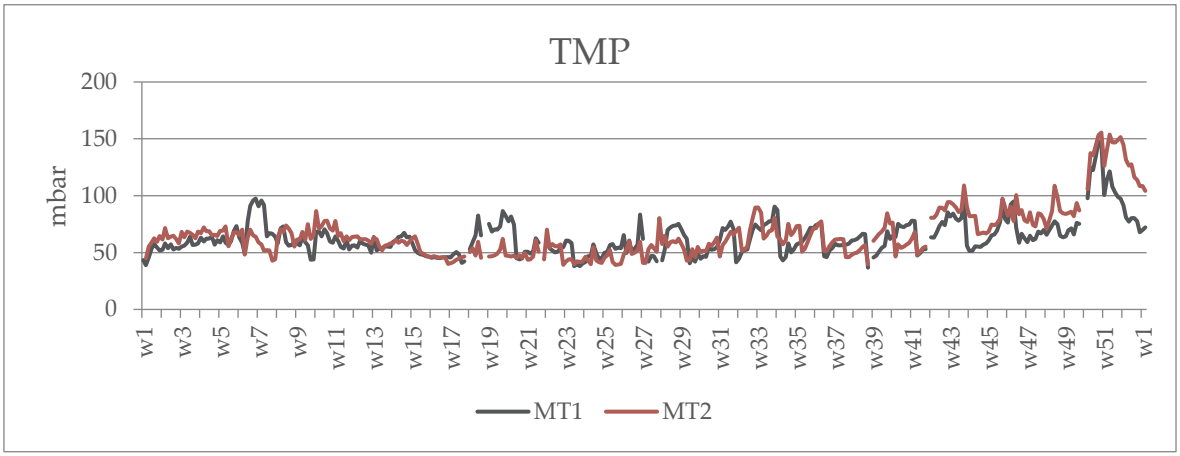



**Figure 31. Net TMP during 2019.**

## 6.5.3 Membrane cleaning

The membranes were cleaned with sodium hypochlorite and citric or oxalic acid. MT1 was cleaned with sodium hypochlorite and oxalic acid and MT2 was cleaned with sodium hypochlorite and citric 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 membrane tank still in operation, the influent flow set-point was set to half of the current value, although never lower than 1.8 $m^3/h$, during MC.

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. Standard cleaning procedure included nine back pulses, 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 $L/(m^2 \cdot h)$ 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.

In 2017 attempts of reducing the chemicals used for maintenance cleaning started. The time of the initial back pulse was reduced from 5 minutes to 2 minutes and later the number of back pulses were reduced from 9 (incl. the first longer one) to 7 in total. In 2018 further reduction of oxalic acid usage was done by increasing the interval in-between cleaning events.

The operational settings have been divided into separate trial periods. An overview of the trials regarding acid MC are presented in Table 18.

**Table 18. Overview of trials with reduced acid use for membrane cleaning.**

| Trial | Start | Description |
|---|---|---|
| T1 | Sept 2017 | Citric vs Oxalic - 7 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 |
| T6 | Feb 2019 | Trial reduced nr of BP and 100% longer time in-between oxalic acid cleanings. |
| T7 | May 2019 | Acid MC is carried out same night as Hypo MC, every 4th hypo for MT1 and every 2nd hypo for MT2 |
| T8 | Aug 2019 | Oxalic acid pumping reduced to 80% chemical flow during pumping. |
| T9 | Oct 2019 | Citric cleaning with reduced chemicals as well as Oxalic |
| T10 | Oct 2019 | Oxalic acid 80% chemical flow, 7 BP, every 4th Hypo MC<br>Citric acid 100% chemical flow, 7 BP, every 4th Hypo MC |

The amount of chemicals used normalized to the initial settings (back pulse duration 2 minutes + 8 x 30 seconds carried out after 345 $m^3$ permeate produced) are presented in Figure 32 together with the permeability.

Trial 5 (T5) started in October 2018 where MT1 was operated for 53 days without oxalic acid cleaning. One cleaning was carried out on the 17th of December 2018 and then followed by 50 days without cleaning. Trial 6 (T6) started in week 6 with half the frequency of oxalic acid MC. As the

flow meters used for permeability calculations were damaged from sodium hypochlorite and changed in week 8, data for the beginning of this year is uncertain. With the new flow meters, data was more reliable and in can be observed that, although less than half of the specified oxalic acid was used performance of MT1 was slightly better than MT2.

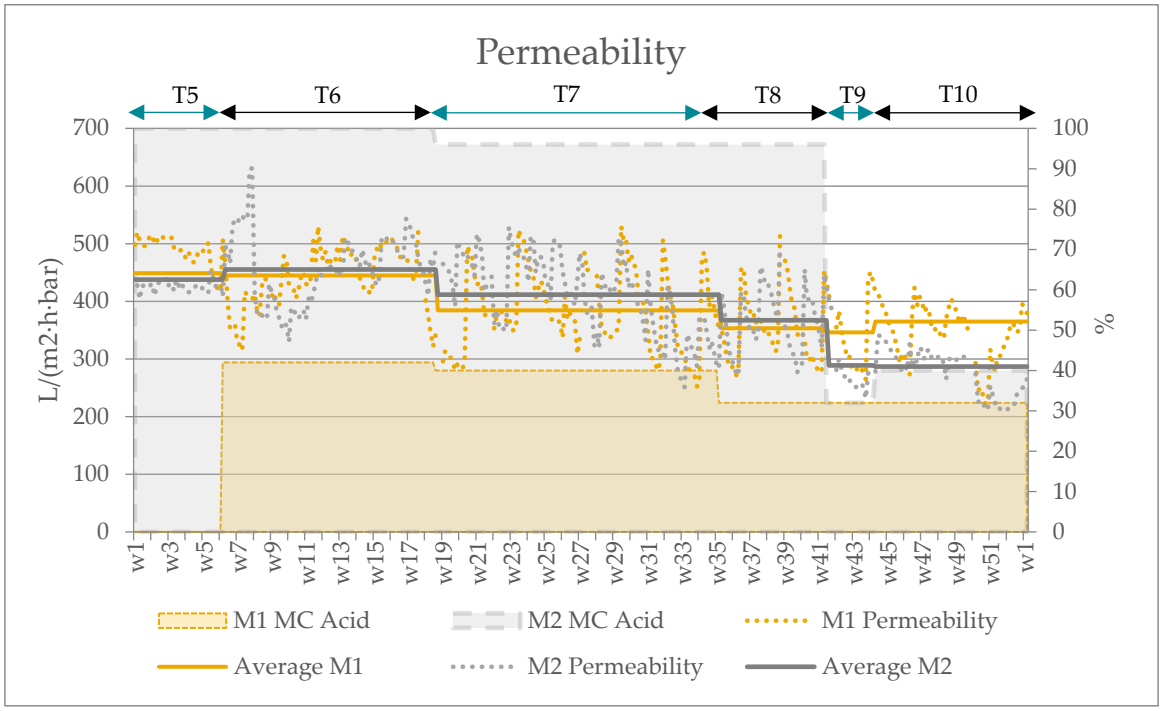



**Figure 32. Amount of acid used for MC, normalized to back pulse duration of 2 minutes + 8 x 30 seconds carried out with interval of 345 m³ of permeate produced. M1 was cleaned with oxalic acid, M2 was cleaned with citric acid. T5 – T10 are trial periods. T5 started in October 2018 (one oxalic MC for MT1 was carried out on the 17th of December 2018).**

Starting with trial 7 (T7) MC with acids were synchronised with the hypo MC, taking place 5 hours apart instead of separate times. Citric MC was carried out after every other hypo MC and oxalic MC after every 4th hypo MC. This had a major effect on the variation of permeability as it increased after acid MC and slowly decreased until the next acid MC, compared to previous years when permeability was much more even. As the permeability for MT1 still was comparable with MT2 it was decided to reduce the amount of oxalic acid further by decreasing the chemical flowrate to 80 % of standard flow rate (T8). This resulted in a 20 % lower concentration in the back pulsed solution. After about a month with this setting it was time to start reducing the citric acid for comparison. Initially (T9) the same settings were used for citric acid as was set for oxalic acid in T8. However, after a rapid decrease in permeability for MT2 the citric acid concentration in the solution back pulsed was increased to the standard concentration (T10). From week 44 and for the remainder of this year MC with acid was carried out after every 4th hypo MC for both MTs'. As citric acid consumption was reduced, the permeability for MT2 decreased compared to MT1. For T10 the average permeability for MT1 was 345 L/(m²·h·bar) and for MT2 277 L/(m²·h·bar).

Throughout 2019 maintenance cleaning with sodium hypochlorite has been carried out with interval according to supplier, but with reduced back pulses; 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 membrane supplier, 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. Last year (2018), RCs were carried out in May and this year, the RCs were carried out in March although permeability was good (around 400 L/(m$^2$·h·bar)). Last time, the membrane cassettes were lifted for inspection and membrane fibers were sent for analysis of type of fouling. It was then concluded that the membranes were in good condition, the foulant before recovery cleaning contained mainly iron, some organic material and trace amounts of calcium phosphate. This time, focus was on measuring the chlorine gas and chloramines emitted to air when the membranes were soaked in sodium hypochlorite solution. The main objective of the measurement campaign was to study if membrane cleaning with hypochlorite will result in high levels of these gases emitted, which can be hazardous from a working environment perspective. Since the ventilation from the membrane tanks, in the full scale plant, will be connected to a technical tunnel where different types of equipment will be installed, and since chlorine gas is very corrosive and can damage this equipment, it is from this perspective as well, interesting to investigate emitted levels.

Since the previous RCs about 13 700 m$^3$ permeate was produced by MT1 and 13 800 m$^3$ permeate by MT2. This corresponds to about 133 and 134 m$^3$ per m$^2$ membrane area for MT1 and MT2, respectively.

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 19 together with the amount of chemicals used and conditions at start and end of the soaking.

**Table 19. Results from RC.**

| Date | Membrane tank | Chemical | Amount | Measurements in tank at the start of soak | Soaking time | Measurements in tank at end of soaking |
|---|---|---|---|---|---|---|
| 2019-03-13 to 2019-03-14 | MT1 | Sodium hypochlorite (83.5 g/L) | 15.7 L | pH 9.5<br>$Cl_2$ 240 mg/L | 20.5 h | pH 8.1<br>$Cl_2$ 165 mg/L |
| 2019-03-14 to 2019-03-15 | MT2 | Sodium hypochlorite (91* g/L) | 15* L | pH 9.9<br>$Cl_2$ 310 mg/L | 19.5 h | pH 8.2<br>$Cl_2$ 250 mg/L |
| 2019-03-25 to 2019-03-26 | MT1 | Oxalic acid (8%) | 22.3 L | pH 2.5 | 19.5 h | pH 2.5 |

| | | | | | | |
|---|---|---|---|---|---|---|
| 2019-03-26 to 2019-03-27 | MT2 | Citric acid (51%) | 5.3 L | pH 2.7 | 20 h | pH 2.7 |

*problems with low level in hypo tank after 5 back pulses, 5 L 12% sodium hypochlorite was added to tank. Volume and concentration are estimated.

When comparing the effect on permeability there was no great difference between oxalic and citric acid (Figure 33). After the acid RCs the permeability was almost the same for the two membrane tanks, around 500 L/(m²·h·bar).

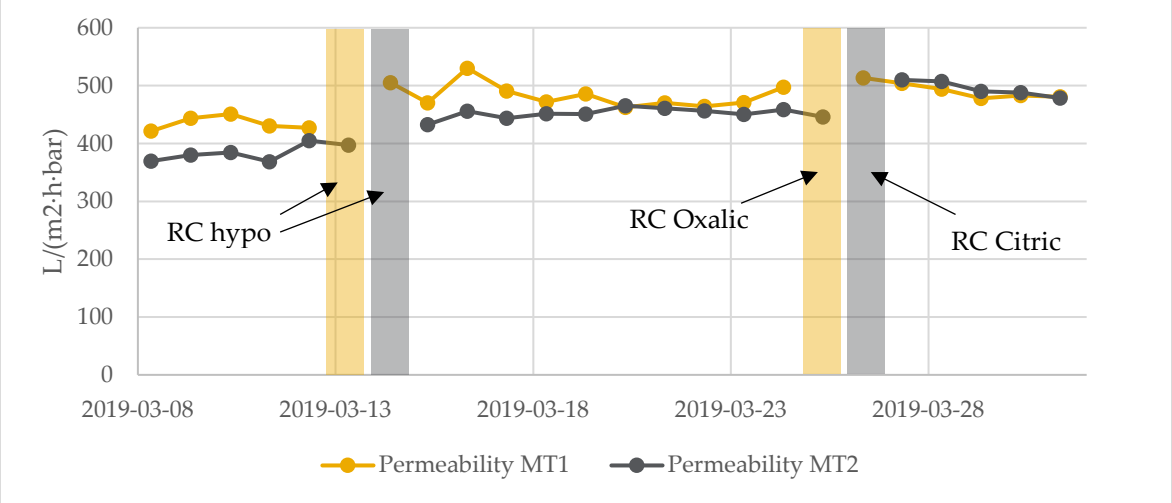



**Figure 33. Permeability before and after recovery cleaning (RC) with sodium hypochlorite (hypo) and acids (oxalic acid and citric acid for MT1 and MT2, respectively).**

Chlorine gas and chloramines were measured with passive samplers installed about half a meter above the water surface of each membrane tank. The tank was not sealed but operated under negative pressure thanks to an air exhaust inside the tank. The samplers started measuring about one hour after the sodium hypochlorite solution was introduced into the system. Filling of the tank took about 45 minutes and once filled the aeration was turned on for 5 minutes to mix the solution. Air sampling started after filling and mixing was finished. One reference sample was also collected where no cleaning was conducted.

Chloramine samples (samplers supplied by the Department of Public Health and Clinical Medicine, Umeå University) and chlorine samples (SKC 225-9006 samplers) were then collected for 3 hours, with a sample flow of about 1 l/min, using SKC AirChek TOUCH air sampling pumps. Sample flows were measured using a DryCal DC-Lite Primary flow meter. Chloramine samples were then analysed using ion chromatography, by Department of Public Health and Clinical Medicine, Umeå University. Chlorine samples were analysed using ion chromatography by the Department of Occupational and Environmental Medicine, Örebro University Hospital.

**Table 20. Results from trichloramine and chlorine gas measurements in the membrane tank ventilation.**

| Sample | Trichloramine (mg/m³) | Chlorine gas ($Cl_2$/mg) |
|---|---|---|
| Reference sample (no cleaning) | < detection limit | <0.005 |
| RC MT1 | 1.23 | 0.018 |

| | | |
|---|---|---|
| RC MT2 | 0.04 | <0.005 |

As can be seen in the results presented in Table 20 the emitted levels of both Trichloramine and Chlorine gas was much higher during recovery cleaning of membrane tank 1 compared to membrane tank 2. This was surprising as the measured dissolved chlorine concentration was higher during cleaning of membrane tank 2 both at the start and end of soaking. During cleaning of membrane tank 2, the levels of Trichloramine and Chlorine gas were comparable to the levels from the reference sample. It should be noted that the World Health Organisation reference limit for exposure is 0.5 mg/m$^3$ which was largely exceeded during cleaning of membrane tank 1.

Due to the large difference in the measurements from the two different membrane tanks, it was decided to conduct a more detailed measurement in the next recovery cleaning. In the next measurement, sampling will start directly when the soaking solution is pumped into the membrane tanks, and not after the tanks are filled. In the next measurement, also several samplers will be installed and operating in different time intervals, to study if there are specific time intervals where more gas is being emitted than others.

### 6.5.4 Membrane aeration

The membranes are aerated by a coarse bubble aeration system mounted at the bottom of the membrane tanks. The membrane aeration flow is operated at one out of two levels, Leap-Hi (26 m$^3$/h) or Leap-Lo (14 m$^3$/h). An algorithm, provided by the membrane supplier, is used to select which of them to be used. The control strategy is called fouling control. According to the membrane supplier, the aeration system operates very close to the minimum air flow rate at Leap-Lo. If the air flow rate would be reduced below the Leap-Lo level, the mechanical aeration equipment would not function as intended.

#### Fouling control

The fouling control algorithm caused the higher aeration level (Leap-Hi) to be used during the latter part of 2019 (Figure 34). Both membranes needed the higher aeration at times, however often not at the same time. MT2 was in Leap-Hi more often than MT1.

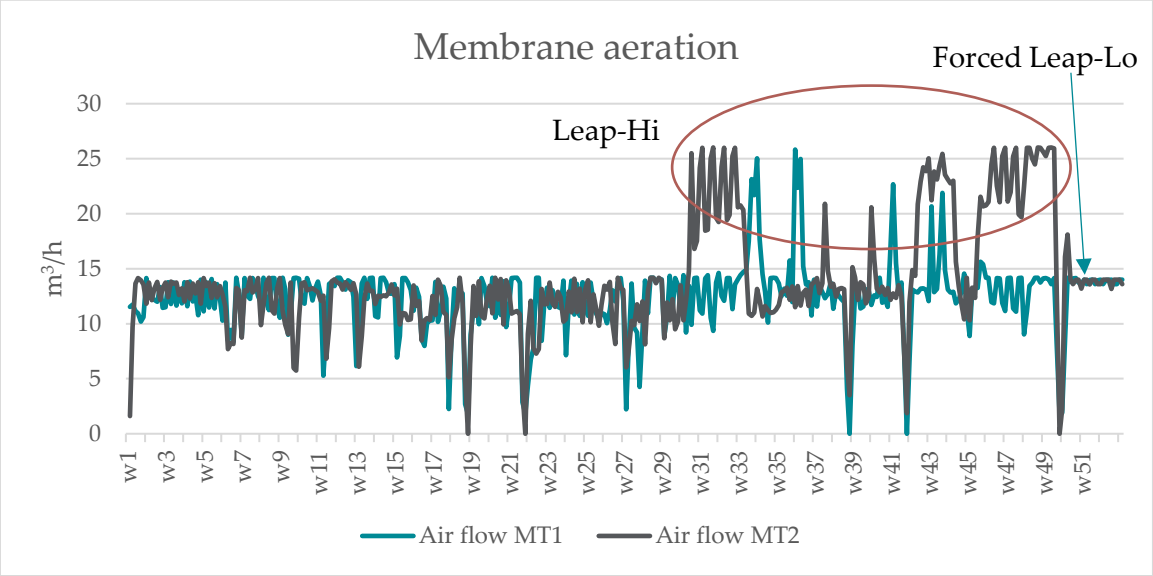



**Figure 34. Membrane aeration as daily average during 2019. Values above 14 m³/h correspond to usage of Leap-Hi. This is indicated with an ellipse in the figure.**

In week 50 the membrane aeration was forced to Leap-Lo for both membranes to evaluate membrane permeability without usage of Leap-Hi during 2020.

### 6.5.5 Experiment without membrane aeration

**Background and Objective:**
As the membrane permeability in the pilot most of the time is high, 300-600 L/(m²·h·bar), and the aeration often is in the lower aeration mode (Leap-Lo mode 80-90% of the time) it is believed that the membranes could perform well with even less aeration.

One idea to reduce the aeration is to aerate intermittently. In order to gain knowledge about the TMP increase when aeration is off, it was decided to try operating the membranes without aeration for one permeation cycle (10 minutes) and monitor the TMP.

**Method:**
Aeration was turned off at the start of the permeation cycle and started again at the beginning of the relaxation. The next permeation cycle was operated with normal aeration and relaxation followed by a 2-minute long back pulse at flux approximately 20 L/m²,h.

The procedure was carried out first for membrane tank 1 and then for membrane tank 2.

**Results:**
The way TMP increased during the permeation cycle without aeration was very similar for both membrane tanks (MTs), see Figure 35. When aeration was turned off it took approximately 3.5 to 5 minutes before a rapid exponential increase in TMP could be observed. After 10 minutes of permeation without aeration, TMP had increased from about 55-60 mbar to 172 mbar for MT1 and 187 mbar for MT2. Relaxation for 1 minute with aeration on did not help to reduce TMP.

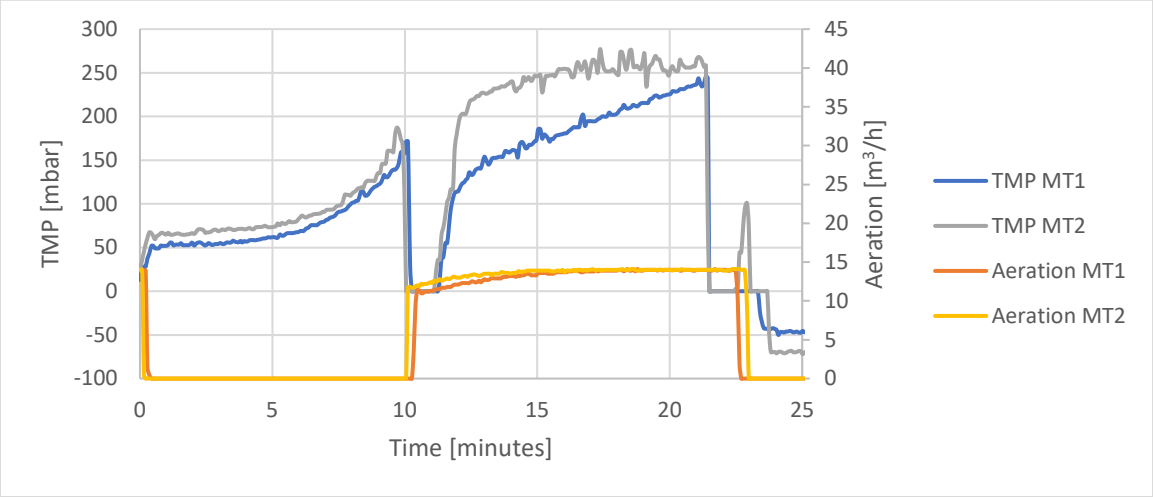



**Figure 35. TMP and aeration for MT1 and MT2 during the experiment without aeration for 10 minutes of permeation.**

TMP continued to increase during the second permeation cycle although aeration was on. However, TMP was reduced almost to the initial value after the 2 minute back pulse carried out before the third permeation cycle started, see Figure 36 for MT1 and Figure 37 for MT2.

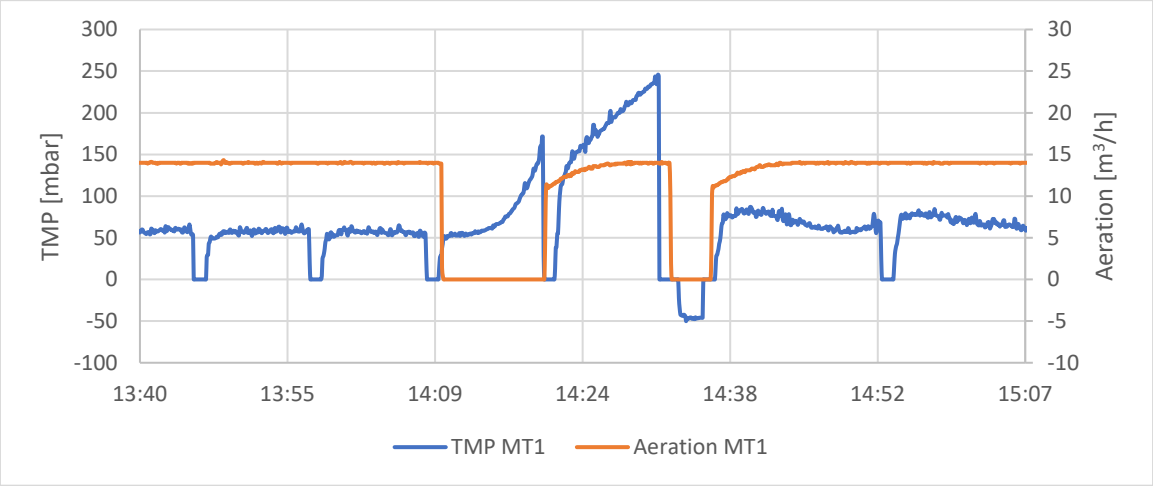



**Figure 36. TMP and aeration of MT1.**

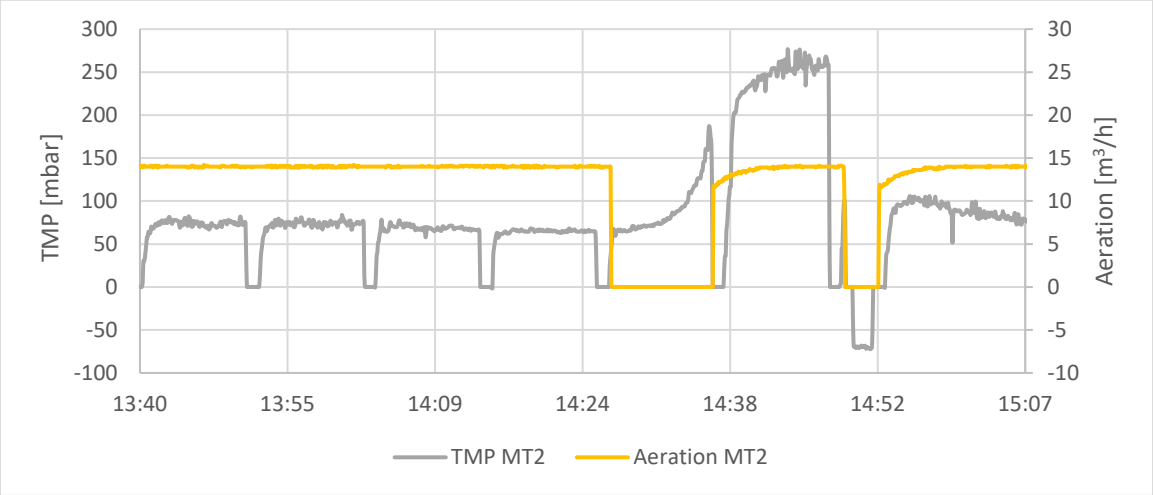



**Figure 37. TMP and aeration of MT2.**

Permeability during the experiment is presented in Figure 38. After operating without aeration during 10 minutes of permeation followed by one normal operational cycle permeability was at its lowest, just below 100 L/(m²·h·bar). Permeability was recovered to above 300 L/(m²·h·bar) after back pulsing with permeate for 2 minutes.

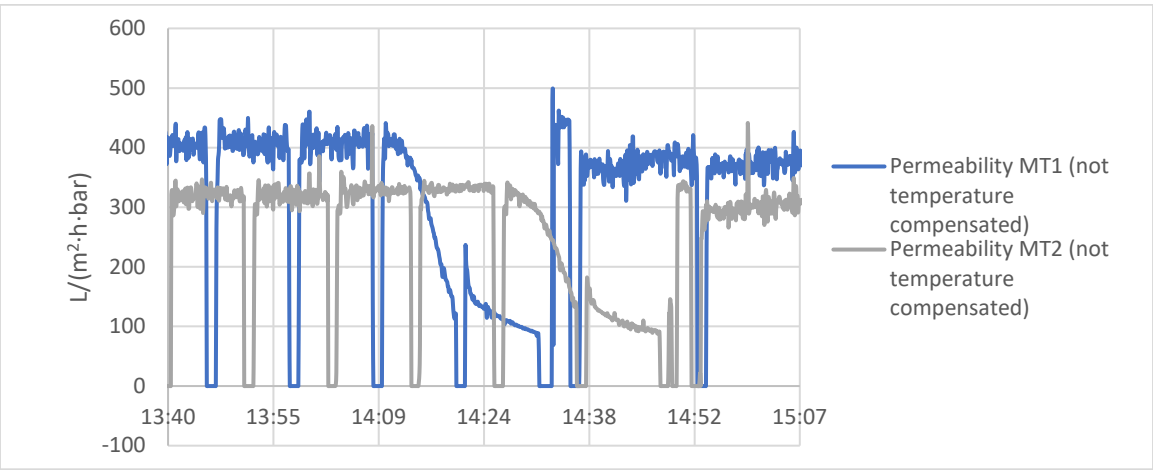



**Figure 38. Permeability during the experiment for MT1 and MT2.**

From this experiment it can be concluded that the membrane fouling created during the permeation cycle without aeration was not removed by just starting the aeration again. With a back pulse for 2 minutes permeability was increased from 100 L/(m²·h·bar) to above 300 L/(m²·h·bar) which was similar as before the trial.

As the lost permeability was mostly recovered by a 2-minute back pulse an alternative operational cycle with reduced aeration and back pulse instead of relaxation could be possible for future tests.

### 6.5.6 Experiment without relaxation

During the above presented experiment without membrane aeration, it was concluded that relaxation had much less impact on restoring permeability compared to a back pulse. Before switching to back pulse instead of relaxation, an estimate on how often the back pulses would be needed was sought. In order to determine performance when permeation was prolonged it was decided to test operation of the membranes without relaxation for a longer period.

The no-relaxation-test started on the 9th of July 2019 and permeation was on continuously (except during maintenance cleaning) for six weeks (w.29-w.34) without noticing any negative impact on membrane permeability and the trial was then ended.

As no clear decrease in permeability was observed within a time frame reasonable to be used as new settings for permeation cycle, it was decided (in agreement with membrane supplier) to extend the permeation cycle from 10 minutes of permeation + 1 minute of relaxation to 15 minutes of permeation + 1 minute of relaxation. The advantage of a longer permeation cycle is less down-time when the membrane tank is aerated but no permeate is being produced. This will in a full scale save energy as less trains will need to be in operation to produce the same amount of permeate. Without any relaxation the permeate withdrawal increased with 9%. The new settings with 15 min permeation and 1 min relaxation would increase the permeate production from 91% of the time to 94% of the time, i.e. 3%. One limitation in the pilot was that the extra permeate withdrawal needed more permeate recirculation, especially during low flows, to maintain the desired flux.

## 6.6 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 21 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 data from the Henriksdal WWTP (annular average 2019) and design data for the future Henriksdal WWTP according to SFA.

The big differences in sludge production between the pilot and Henriksdal WWTP is caused by the configuration of the primary treatment step and the relatively higher SS and BOD load in the pilot, which was described in previous reports together with its effect on the following treatment steps (Andersson et al., 2019; Andersson et al., 2020).

**Table 21. Sludge data from the pilot year 2019 compared to data from the Henriksdal WWTP 2019 and design data for the future Henriksdal WWTP.**

| Parameter | Pilot data 2019 | Henriksdal WWTP data 2019 | Design future Henriksdal |
|---|---|---|---|
| WAS production (kg SS/d) | 13.3 | 22 700 | 59 000 |
| Part of total sludge production (%) | 42% | 27% | 34% |
| VSS in WAS (% of SS) | 74 | 68 | 63 |
| Fe in WAS (% of SS) | 7.7 | 11 | - |
| PS-production (kg TS/d) | 18.5 | 62 200* | 117 000 |
| Part of total sludge production (%) | 58% | 73% | 66% |
| VS in PS (% of TS) | 88 | 78 | 80 |
| Total sludge production (kg TS/d) | 31.8 | 84 700 | 176 000 |
| Total sludge age, SRTtot (d) | 19 | 14 | 28 |
| Aerated sludge age, SRTox (d)** | 7.0*** | 6.0 | 7**** |
| SVI jan-jun (mL/g) | 138 | 167 | - |
| SVI jul-dec (mL/g) | 219 | 126 | - |

*based on uncertain TS measurements.

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

***including membrane tanks, without membrane tanks SRTox = 5.2 d

****excluding membrane tanks, yearly average.

The sludge quality in terms of its filterability is an important factor for the membrane operation. Poor sludge quality will cause more rapid reversible and irreversible fouling, higher TMP and lower permeability which causes higher energy and chemical consumption for cleaning and eventually leads to a faster ageing of the membranes. The filterability of sludge is believed to depend on a number of factors such as microbial composition, extracellular polymeric substances (EPS), soluble microbial products (SMP), biopolymer clusters (BPC), fine colloidal particles and inorganic foulants such as cat- and anions in wastewater and chemical precipitants (Iorhemen et al., 2016; Du et al., 2020).

It has been debated how to measure and monitor the filterability of MBR-sludge in a representable way and several more or less complex methods have been proposed (Gkotsis & Zouboulis 2019). In this project, time-to-filter, TTF (described in chapter 5.3.2) has been used since it was suggested by the membrane supplier. The fouling propensity of membranes can proposedly also be monitored by analysis of colloidal TOC (cTOC) which are particles in the size range 0.04-1.2 µm (chapter 5.3.2, Fan et al. 2006). Analysis of cTOC was also adopted in this project.

A summary of sludge properties analysed in this project that might affect the membrane performance are shown in Figure 39 and Figure 40 below, together with temperature compensated permeability. Sludge volume index (SVI), normalized TTF, colloidal TOC (cTOC) increase in the second half of the year as is the TMP. At the same time, the iron dose was reduced and the iron content in sludge was decreased. The cTOC value should be below 10 mg/L to insure minimal fouling (according to the membrane manufacturer). In 2019 the cTOC was above 10 mg/L on 11 out of 52 weeks.

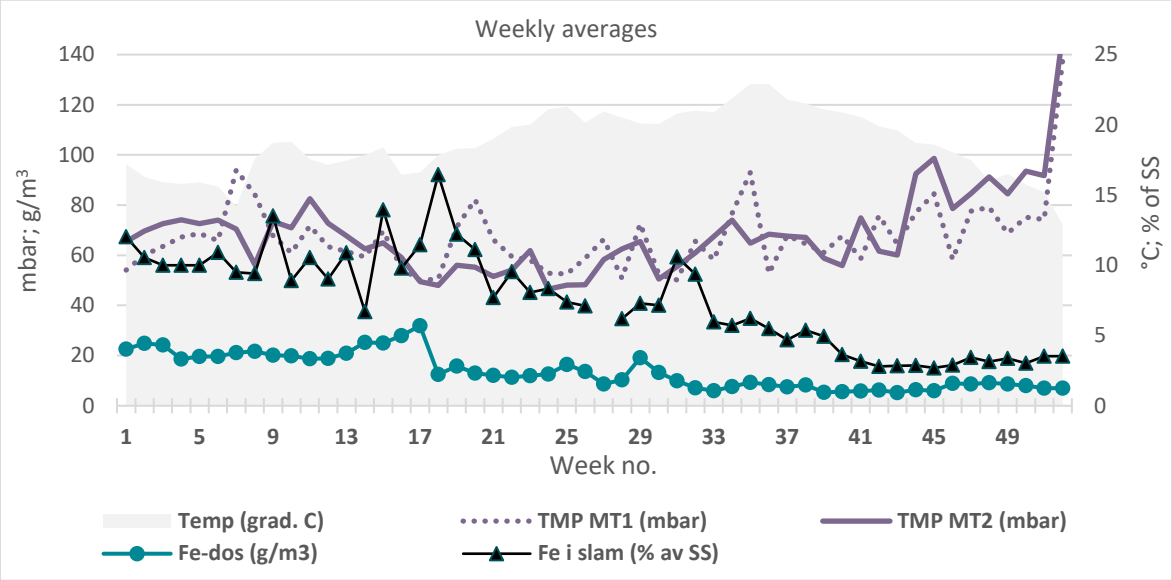



Figure 39. TMP, temperature, Fe-dose and Fe content in sludge over the year.

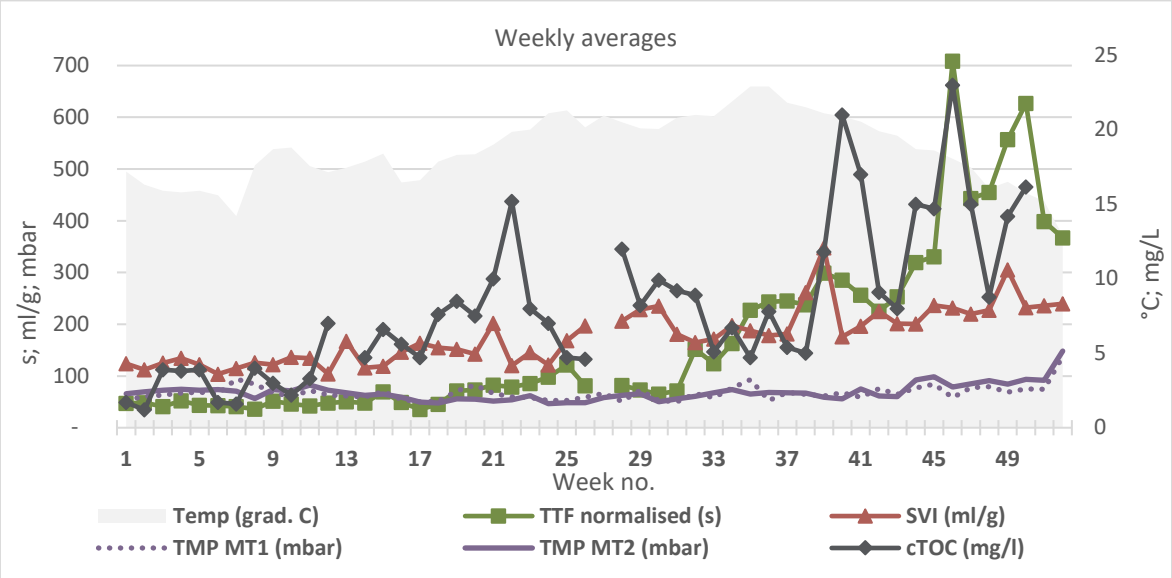



Figure 40. TMP, temperature, TTF (normalized to SS-concentration), SVI and cTOC over the year.

It is hard to find a single sludge property that has a clear correlation to the TMP or permeability of the membranes. The only example where a clear correlation was seen was between the iron content in sludge (Fe as % of SS) and the time to filter (TTF), see Figure 41. The correlation to the actual iron dose (which is not a sludge property), however, was as strong. Zhang and colleagues (2015) showed that addition of ferrous ($Fe^{2+}$), but not ferric ($Fe^{3+}$), had a positive effect on the filterability of MBR-sludge. The iron could theoretically bind the fine colloidal particles to flocs and thereby increase the filterability of the sludge. There was, however, no correlation between iron dose or iron in sludge and cTOC concentration, indicating some other mechanism increasing the filterability in sludge with high iron content.

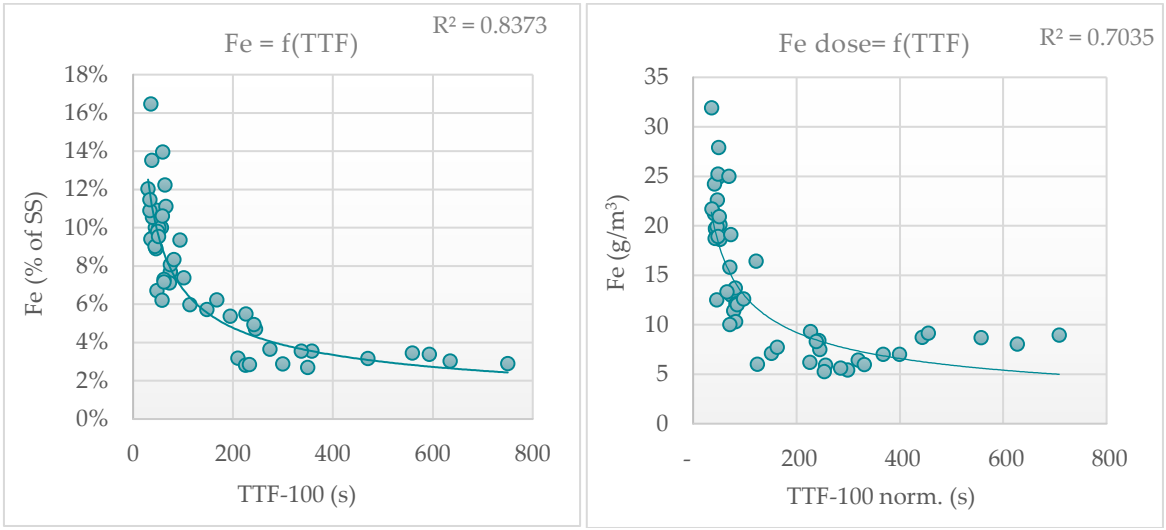



**Figure 41. The correlation between chemical precipitant in sludge (left) and dose (right) and normalised TTF-100.**

In Table 22 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. A more efficient precipitation strategy combined with the previously described enhanced biological phosphorus removal, EBPR, activity caused the Fe/P ratio to decrease over the first years. Stricter effluent goals for phosphorus in the effluent then caused the ratio to increase.

**Table 22. WAS composition (annual average) in the pilot over 6 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) |
|---|---|---|---|---|---|
| **2019** | 9 932 | 7.6 | 3.5 | 75 | 1.5 |
| ***n*** | *50* | *50* | *50* | *50* | *50* |
| **2018** | 8 480 | 6.4 | 3.3 | 77 | 1.1 |
| ***n*** | *50* | *50* | *50* | *50* | *50* |
| **2017** | 9 632 | 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* |

Data on metals and some organic micro pollutants in dewatered digested sludge have been collected over the years and was analysed and compared to data from Henriksdal WWTP in a separate report (Nähri et al., 2020). A summary of the evaluation will be published in next year´s report from pH2040.

## 6.7 Sludge pilot

| Trial | J | F | M | A | M | J | J | A | S | O | N | D |
|---|---|---|---|---|---|---|---|---|---|---|---|---|
| Mesophilic operation of the digester | | | | | | | | | | | | |
| Transition to thermophilic digestion | | | | | | | | | | | | |
| Thermophilic operation of the digester | | | | | | | | | | | | |
| Sludge thickener in operation | | | | | | | | | | | | |
| Trial with decreasing digester HRT | | | | | | | | | | | | |
| Dewatering in operation | | | | | | | | | | | | |

The sludge pilot was taken into operation in September 2017, comprising thickening of mixed sludge (PS and WAS), digestion and dewatering of digested sludge. During 2018 the sludge pilot operation was discontinuous as the pilot experienced problems with clogging of pipes and instrument failures.

During 2019 two trials were performed in the sludge pilot line; transition from mesophilic to thermophilic digestion during spring and an HRT reduction test for the digester during autumn. The HRT reduction test was not finalized within 2019.

The year 2019 started with a mesophilic reference period followed by the transition to thermophilic conditions by increasing the temperature in the digester. After the transition the digester was operated at thermophilic conditions with high load followed by thermophilic conditions with short HRT. In the middle of June 2019, when starting the trial with short retention time in the digester, the sludge thickener was by-passed to secure adequate load into the digester.

### 6.7.1 Feed characteristics

As shown in Figure 7, the Mixed sludge tank receives Primary Sludge (PS) and Waste Activated Sludge (WAS) from the MBR-pilot line. The digester was fed with mixed sludge (MS) from the MS tank. The proportion of PS to WAS, in terms of TS weight, was approximately 50/50 before week 24 and was adjusted to 60/40 after week 24 by discharging a fraction of the WAS. The total and volatile solids in the MS changed during the year following Figure 42.

Until week 27, the sludge was being thickened to around 5%TS, but several problems with the thickened sludge pump and pipes led to interruptions in the digester's feeding. Thus, for the second semester, the thickener was bypassed to ensure a continuous feed and stable HRT (Figure 42).

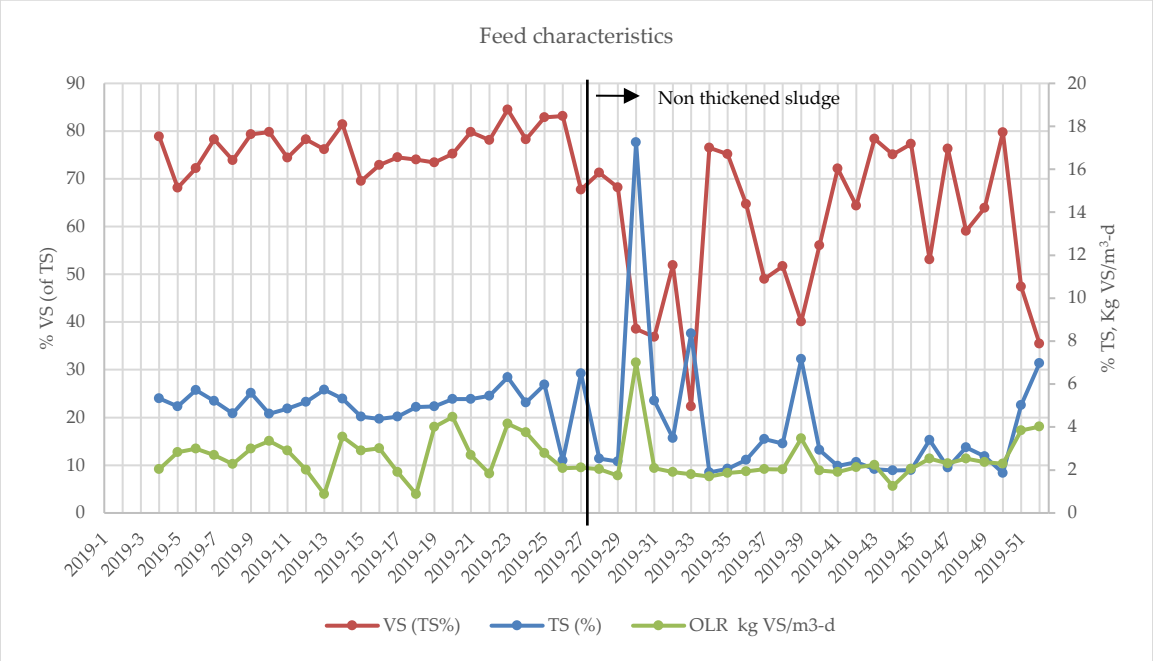



**Figure 42. Feed characteristics 2019.**

After the thickener was taken out of operation, sand and grit in the feed began to be noticed. These substances are considered non-degradable, inorganic compounds, and their presence affects the measured TS values. Most likely, the sedimentation of sand and grit in the BS-tank brings along organic material as well since the measured VS mass flows increase (although VS as a percentage of TS decreases). Sampling for TS and VS analyses were taken as grab samples and in periods with random sand and grit accumulation, the use of these data to calculate weekly average OLR values gives strange results, which was observed in week 32 and 39 of Figure 42. Many key numbers on digestion performance, which will be discussed in the following chapters, are based on VS% destruction and OLR and will thus be affected by this.

## 6.7.2 Mesophilic operation

From week 1 to 12, during the months before the mesophilic-thermophilic transition, the process was being operated in mesophilic conditions with the aim to obtain a stable process.

## 6.7.3 Mesophilic – Thermophilic transition

The need to handle more sludge at Henriksdal WWTP in the future led to the decision to implement thermophilic digestion. Thermophilic digesters (55°C) can be operated at lower retention times than mesophilic digesters (37°C) without decreasing the biogas production. However, it is known that the transition of an anaerobic digester (AD) from mesophilic to thermophilic temperatures could pose problems to the microbial process and operation as well as causing odor problems. This trial aimed to increase the temperature while monitoring and evaluating the pilot's digester's performance. A study performed at Ryaverket (Bittlingmayer, 2017) suggested that a fast increase in temperature is the best strategy for the transition. However,

the heating system at Henriksdal WWTP limits the speed with which the sludge can be heated. The maximum heating capacity in the digesters at Henriksdal will be +2°C per day. Thus, the same speed of heating was chosen for the pilot trial.

## Method

The temperature was increased from 37°C to 55°C in a time period of 18 days. Since the microbial community shifts in the period from 47 to 50 degrees (Schnürer & Jarvis, 2017), the trial was divided into two parts; 37-47°C and 47-55°C.

Part 1:

- Time frame: 4 days (18th March – 22nd March)
- Temperature increase: 1°C on the 18th and then 2°C until the 22nd.
- Feeding: continuous
- OLR: 2.8 Kg VS m³/d

Part 2:

- Time frame: 4 days (25h March – 29th March)
- Temperature increase: 3°C on the 25th, 1°C -on the 27th and 3 on the 29th.
- Feeding: intermittent
- OLR: 1.4 Kg VS m³/d

The feed was stopped during the weekend between part 1 and part 2 (23rd and 24th), and then several times during the second part of the trial to avoid VFA accumulation and due to problems with the thickener. These variations are shown in Figure 43, and the average HRT for the entire trial was 25 days.

After the temperature reached 55°C, daily sampling and analyses for process monitoring was performed until the 17th of May. Thereafter sampling and analyses interval went back to twice per week.

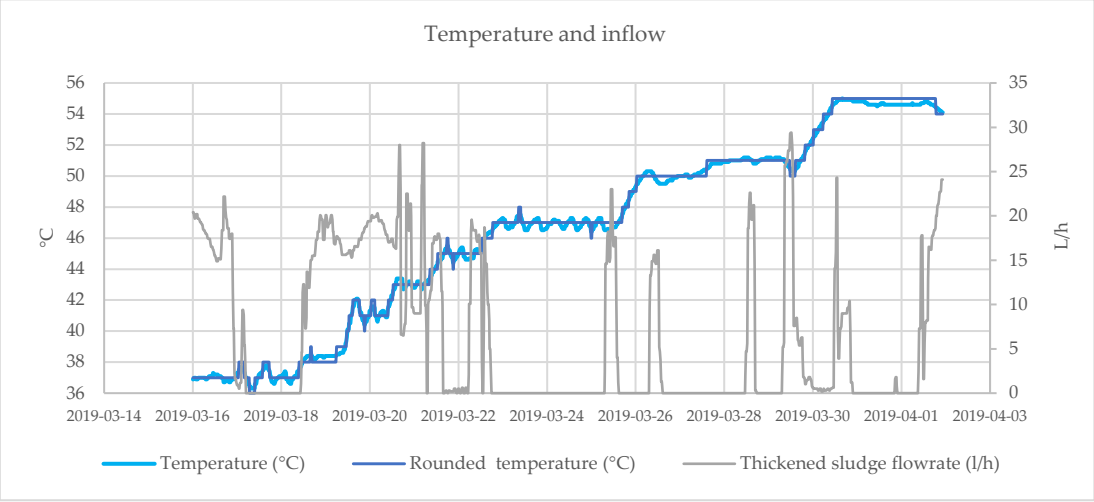



**Figure 43. Temperature and inflow during transition from mesophilic to thermophilic operation.**

## Results

The transition trial was performed as master's project by Joakim Gustavsson and the results will also be published as a master thesis report. Operational problems took place during the transition.

The thickener clogged several times, affecting the inflow to the digester, especially after reaching thermophilic conditions. In addition, the heat exchanger had some problems maintaining 55°C during the first few days (Figure 43). Despite these operational problems, the transition was carried out successfully.

Figure 44 shows the follow-up parameters of the anaerobic digester. Between the black dotted lines is the specific transition period, where all the changes occurred. Even if it is known that the major changes occur between 47 and 50 degrees, VFA accumulation and pH decreasing started a few days before. On the 22nd, when the digester reached 44°C, VFA increased from 180 to 780 mg $CH_3COOH$/L, and pH dropped 0,1.

As expected, it was not until the digester overturned 47 degrees on the 25th, and the digester was fed again after two days feeding stop, that significant changes started to be noticed. During 2 days after the 25th, the alkalinity in the digester decreased (from 4 300 to 3 547 mg $CaCO_3$/L), followed by the peak in VFA concentration (2 373 mg $CH_3COOH$/L), and the effect on the pH value, which reached its minimum of the trial, 7.2. Furthermore, during this critical period there was an increase in ammonia (both $NH^3$-N and $NH_4$-N) and $H_2S$ but both were far from the inhibitory levels. The $H_2S$ content peaked with 16 ppm (10 000 ppm is inhibitory level). Total ammonia nitrogen peaked at 1 300 mg/L (around 1 700 mg/L is considered inhibitory level).

When the digester reached over 50 °C, alkalinity considerably increased (from 4000 to 6000) and VFA was consumed by methanogens, as it can be seen in the small increase in gas production and methane concentration that took place on the 27th (Figure 45), and the decrease in VFA concentration.

The digester stabilized after the temperature stopped oscillating (heat exchanger problems) on the 3rd of April and continued with stable operation until the 24th of April, when there was a stop on the inflow due to operational issues. The pH increased for a few days until the problem was fixed, to then decreased to under 7.3.

During the stable period from the 4th until the 25th of April, the digester showed good results in terms of gas production and methane content. The average gas production for this period was 5 $m^3$/d (0.4 $Nm^3$/kg VS) and 60% methane content (Figure 45).

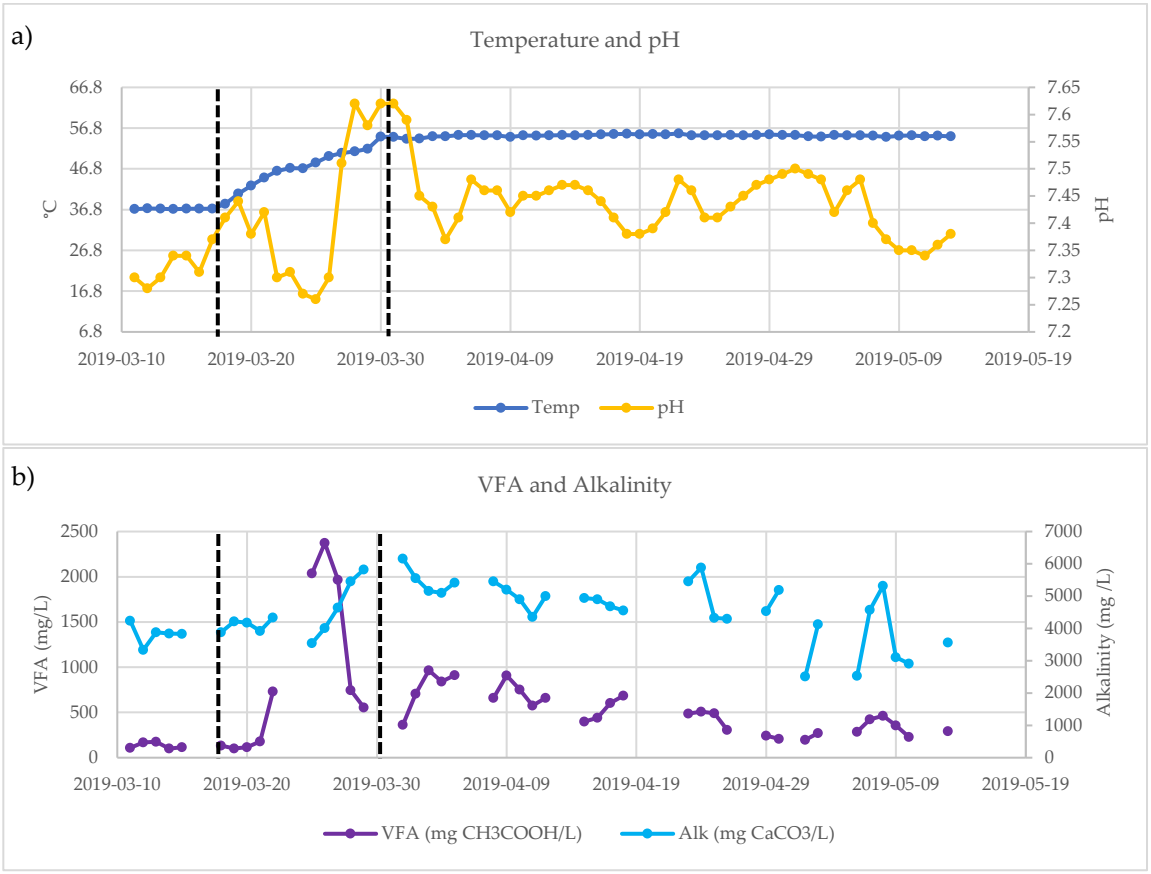



**Figure 44. Changes in the digester during transition. a) Temperature and pH and b) VFA and alkalinity.**

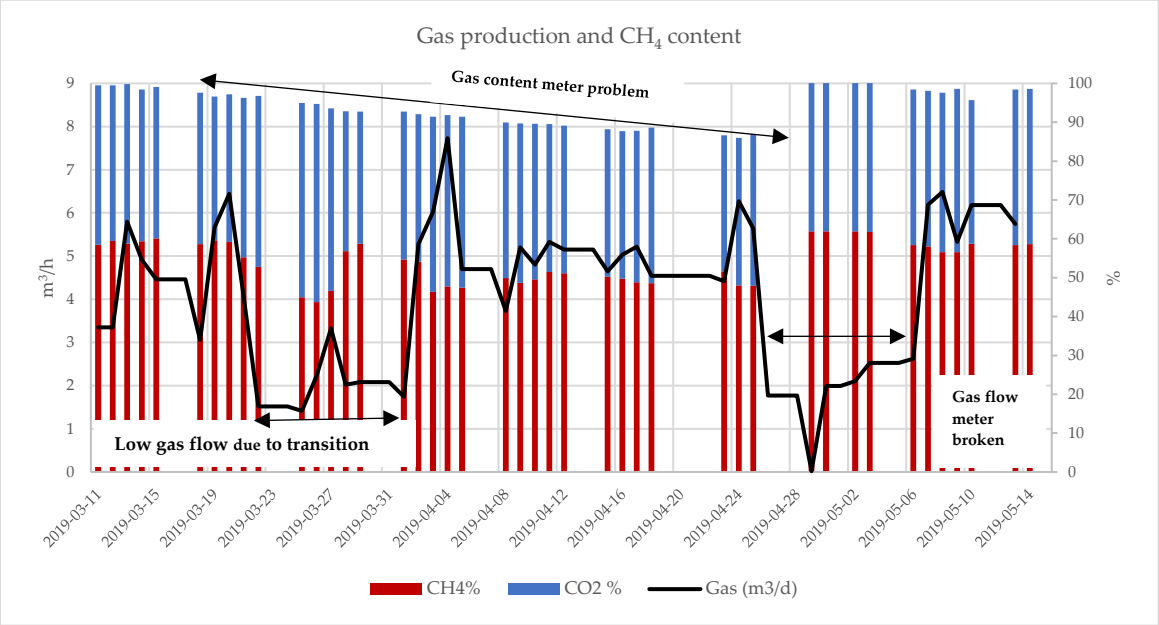



**Figure 45. Gas production, $CH_4$ and $CO_2$ content during the mesophilic to thermophilic trial.**

One of the main concerns of the future full-scale transition at Henriksdal WWTP, is the potentially strong smell when VFA or ammonia are accumulated, which might disturb the people living nearby. However, no strong odor was detected in the digester's surroundings during the transition.

The gas flow meter had some issues after the 26th of April. It could have been showing inaccurate values before, but no clear indications were seen. Furthermore, the gas content meter also posed some issues, the total gas content ($CO_2$% + $CH_4$%) started to decline after the second week of the trial (Figure 45). Initially, it was thought that this could be an accumulation of an unknown gas, but it was later found that the gas meter was measuring lower $CH_4$ levels. After it was sent for service, the values came back to normal by the end of April. The problems with quantification and qualitative analyses of the gas made the evaluation of results more complicated.

Another trial with transition from mesophilic to thermophilic digestion is planned to be done in 2021.

## 6.7.4 Thermophilic operation

After the pilot digester transition from mesophilic (37°C) to thermophilic anaerobic digestion (55°C), a period of 10 weeks of stabile operation was carried out to simulate the full scale operation with a HRT of 12 days and organic loading rate of 3.5 kg VS/m³,d. To achieve this with a mixed sludge of 5.9% TS the digester volume had to be increased from 4 to 5 m³.

During this this period the problems with the thickener continued and the volume was step wise increased.

## 6.7.5 Thermophilic HRT reduction

During reconstruction of the digesters at Henriksdal, the operating volume will be reduced which will cause a reduction in the hydraulic retention time (HRT) in order to manage all the sludge produced. Lower retention times increase the risk to wash out methanogens because of their long generation time, and the lack of time for hydrolyzing the substrates. This trial aimed to decrease the retention time until a crash in the pilot's digester performance occurred. Either by inhibiting the methanogens activity leading to accumulation of VFA or by the inability of bacteria to perform hydrolysis. The results would add information to the decision-making during the reconstruction of the sludge line at Henriksdal WWTP where a stable operation and good quality gas production must be ensured.

### Method

In this trial it was important to have a controlled and steady HRT. Because of the major problems related to the operation of the thickener it was decided to bypass the unit during this trial. Thus, the digester will be operated with a low organic loading rate (OLR). Although this was not the preferred mode of operation it was preferable to big fluctuations in HRT.

The pilot digester was continuously fed with non-thickened mixed sludge and operated at thermophilic conditions (55°C) for 22 weeks. During this period, the digester's HRT was decreased from 9 to 6 days, with a 1-day interval (Table 23). Each HRT was operated for at least 5 retention times unless operational problems occurred, in which case the time was extended.

**Table 23. Trial dates.**

| *Date* | **Week experiment** | **HRT (days)** |
|---|---|---|
| *Jul – Aug 2019* | 1 - 6 | 9 |
| *Aug –Oct 2019* | 7 - 14 | 8 |
| *Oct – Dec 2019* | 15 - 22 | 7 |
| *Dec 2019 - ongoing* | 23 - … | 6 |

For this trial, the mixed sludge had a TS between 2% and 5% (av. 2.8%) and VS as percentage of the TS between 40% and 89% (av. 66%) (Chapter 6.7.1, Figure 42). Furthermore, one sample every week was frozen for further DNA extraction to determine the microbial population changes during the experiment. DNA analyses will be done after the trial is completed in 2020 and results are not presented here.

### Results

During 2019, four different retention times were tested as it is shown in Table 24. By the end of the year, with a retention time of 6 days, the digester was still operating, and gas production was stable. Furthermore, no major acidification occurred in the digester due to the HRT reduction (Figure 47). Some operational problems affected the results, especially by the end of the year.

Figure 46 shows the organic loading rates and VS degradation rates throughout the 2019 test. These values are calculated based on lab results for TS and VS in the sludge, which could sometimes be affected by the presence of sand and other inert (non-degradable) solids. After sorting out the data,

the period with an HRT of 7 days was the one with less presence of sand and grit in the sludge and therefore, the one with more accurate values of VS destruction.

According to literature, an acceptable degradation rate is >40%, during the trial in 2019 the average VS degradation rate was 45.95%. However, as shown in table the HRT with lowest %VS destruction was unfortunately, 7 days. However, the total solids in the influent to the AD were the lowest of the 2019 trial and it must be considered that there was a stop in the AD operation in weeks 44 to 47, in the middle of the trial.

**Table 24. VS degradation and OLR by HRT.**

| *HRT* *d* | IN TS% | IN VS% | OUT TS% | OUT VS% | Inflow AD L/h | OLR kg VS/m$^3$d$^{-1}$ | VS degradation % of VSin |
|---|---|---|---|---|---|---|---|
| *9* | 3.75 | 50.11 | 1.96 | 68.10 | 22.00 | 1.78 | 52.56 |
| *8* | 3.18 | 60.67 | 1.17 | 72.76 | 25.00 | 1.92 | 46.30 |
| *7* | 2.45 | 68.44 | 1.24 | 76.93 | 27.65 | 2.18 | 33.33 |
| *6* | 4.62 | 54.19 | 1.01 | 74.94 | 33.17 | 3.38 | 59.61 |

The temperature is a determining factor in the metabolic activity of bacteria, transfer and solubility of gases, and chemical reactions. Changes in temperature affect methanogens more than they affect other organisms, like the fermentative, which is why VFA accumulation occurs. This can clearly be seen in Figure 46 and Figure 47, where the operational problems with the heat exchanger in weeks 29, 32, 38, from 43 to 48 and the last two weeks of 2019, resulted in VFA/alk ratio increases and pH decreases.

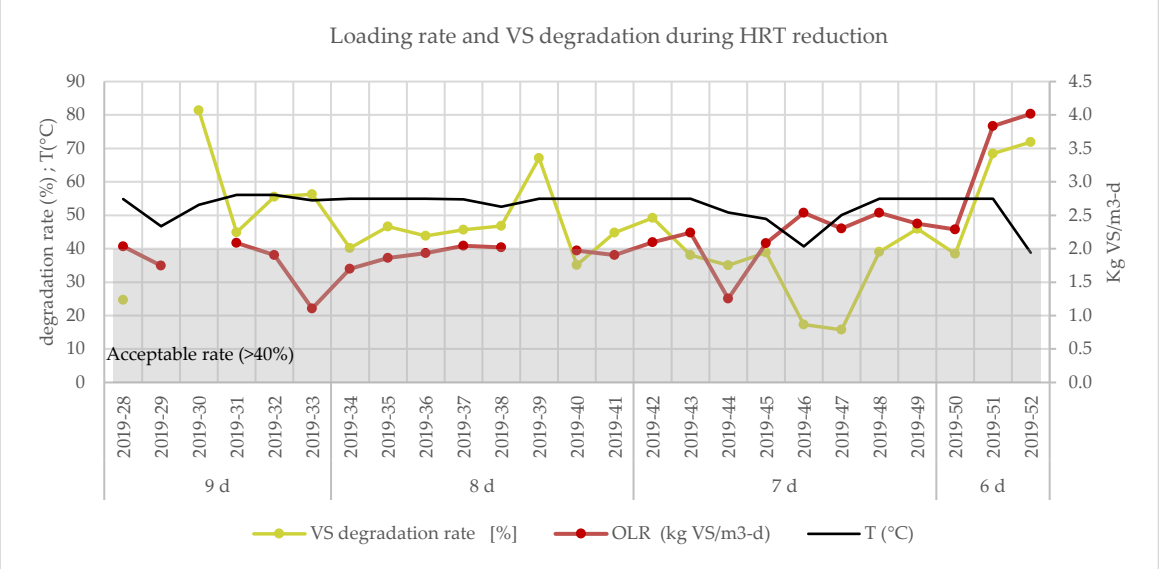



**Figure 46 OLR and VS degradation weekly average values (HRT reduction trial).**

pH was closely monitored during the trial as it determines the digester's stability. The pH showed a reduction every time the HRT was reduced, having the greatest drop at the beginning of the experiment when HRT changed from 12 to 9 days. These changes in the pH are dependent on the ratio between VFAs and alkalinity buffer. Ammonium was also measured as it is part of the buffer

system in the digester and has a high impact on the pH. Furthermore, it can be seen that all the operational problems can be reflected in the pH changes (Figure 44).

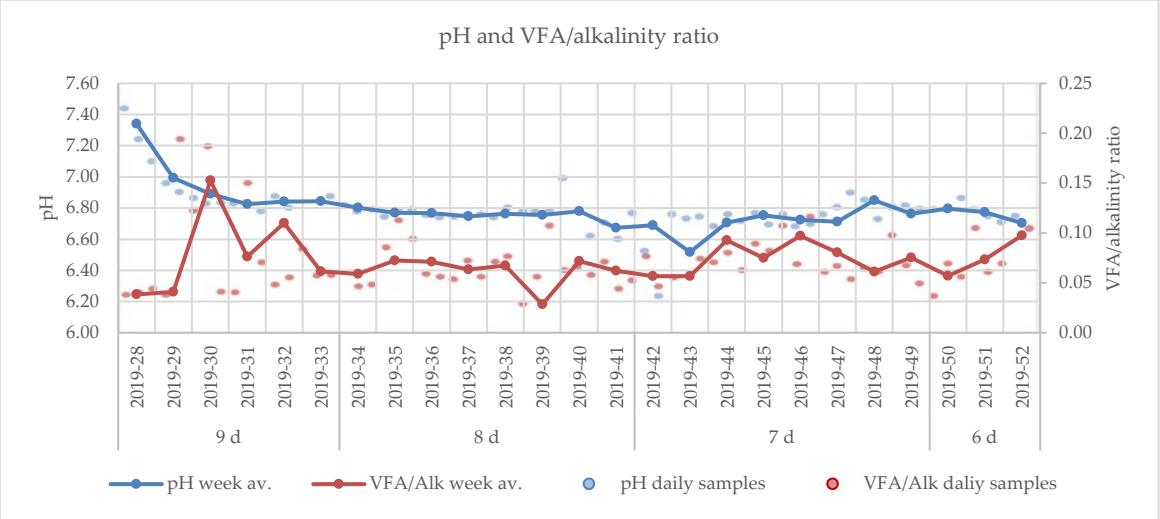



**Figure 47 pH and VFA/alkalinity ratio weekly average values together with daily samples (HRT reduction trial).**

The inaccurate TS and VS results (discussed in Chapter 1.1) not just affected the VS degradation calculations, but also the theoretical gas production calculations which are based on the degraded amount of VS, like it is shown in Figure 48. False peaks in the calculated values (especially weeks 30, 39, 51 and 52) show the problem with this calculation method when the inflow contains large amounts of sand (see chapter 6.7.1).

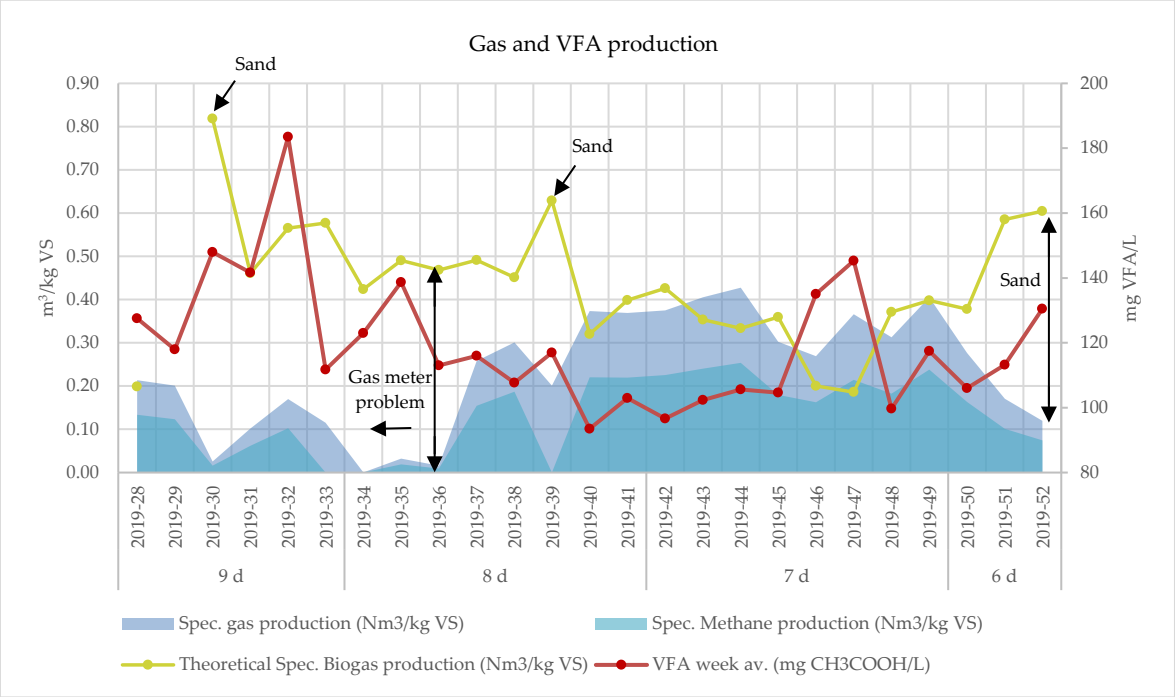



**Figure 48 Specific gas production measured and theoretical (HRT reduction trial)**

The gas flow meter was not showing accurate values between weeks 30 and 36. The issue was not detected for some weeks because it was thought that the AD was actually having problems and it was close to its' process crash. A control against the calculated theoretical values made it clear that the gas flow meter was broken.

Furthermore, the dewaterability of the sludge was seen to be affected by the reduction in the retention time. The changed sludge properties caused clogging of the effluent pipes from the polymer mixing tank to the dewatering unit and therefore no dewatering was done. The reason why it stopped working was not investigated further. The dewatering unit was shut down from week 51 until the end of the winter break.

During the last two weeks of the 2019, the temperature dropped to 18°C causing accumulation of VFA. The pH decreased to 6.1 on the 27th of December. In order to avoid further acidification of the digester due to operational reasons, the feed to the digester was stopped for 10 days.

The digester is expected to recover, and the experiment will continue in 2020. The 6 days retention time period will be repeated once the digester is stable again.

The results from the full trial will be presented and discussed in next year's report.

## 6.8 Resource consumption

Resource consumption in the pilot for 2019 is summarized in Table 25. A comparison with the future Henriksdal design was made where possible (design values available). Pilot values contain great uncertainties due to problems with pumps, air in the tubes, difficulties in manually reading levels and degradation of some chemicals, to mention a few.

This year, three different external carbon sources were tested. Methanol, which have been tested previous years was used in the pilot for about 34 weeks in 2019. A short, 2 weeks, trial with acetic acid was conducted before switching to glycerol as external carbon source for the remainder of the year. Comparing the three, methanol and glycerol was more similar in consumption per day and per kg N removed. The acetic acid consumption was more than double compared to methanol and glycerol which could be partly explained by the trial conditions which differed from normal operation.

The daily iron consumption was slightly lower than the future Henriksdal design although the phosphorus load to the pilot was almost doubled compared to design. This is reflected in the mole Fe to mole P consumption which was only 50% in the pilot compared to the design. The effluent phosphate concentration was the same as the target concentration which was lower than design calculations. The low iron consumption is explained by the EBPR activity in the pilot. Compared to previous years the iron consumption has decreased as the EBPR activity has increased.

Despite the efforts made in reducing the amount of chemicals for MC, the daily amounts used are not much lower than the SFA design. One explanation is that the influent to the pilot (on which the interval for MC was based) was 13% higher during 2019 than the pilot design inflow. For the citric acid, as only one out of two membranes were cleaned with citric acid, the consumption was lower than the design value (37% of design for two membrane tanks).

**Table 25. Chemical consumption during 2019.**

| Resource | Unit | Value pilot | Future Hdal design | Value pilot/scaled future Hdal design |
|---|---|---|---|---|
| External carbon source (Methanol, 234 days) | kg COD/d | 0.80 | 12 000 | 45% |
| | g COD/g N* | 0.29 | - | - |
| External carbon source (Acetic acid, 15 days) | kg COD/d | 2.15 | - | - |
| | g COD/g N* | 0.66 | - | - |
| External carbon source (Glycerol, 107 days) | kg COD/d | 0.82 | - | - |
| | g COD/g N* | 0.27 | - | - |
| Iron (FV+BR4+BR6) | kg Fe/d | 1.16 | 10 000 | 78% |
| | mole Fe/mole P | 1.4 | 2.8 | 50% |
| Citric Acid (51%) | L/d | 0.077** | 1 400 | 37% |
| Sodium hypochlorite (12%) | L/d | 0.191*** | 1 200 | 107% |
| Oxalic acid (8%) | L/d | 0.24** | - | - |
| Aeration biology | $m^3/d$ | 1201 | - | - |
| Aeration MT | $m^3/d$ | 652 | - | - |

* N removed in total, from inlet to effluent.
** Number of MCs with each acid, multiplied with time settings and number of back pulses using design flowrate of chemical. Measured consumption for one RC preformed with each acid.
*** Number of MCs with hypo was multiplied with time settings and number of back pulses using design flowrate of chemical.

## 6.9 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.

### 6.9.1 Method

A total of four separate sampling campaigns are 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 were conducted in 2019, one in January 2019 and one in October 2019. The 2019 sampling campaigns also included samples from the Henriksdal WWTP CAS process, for comparison. Both the MBR pilot process and Henriksdal

WWTP are treating the same influent wastewater, and the samples from each respective site were taken at the same dates and times.

Both water phase and sludge phase samples were included in the sampling, 3 water phase sampling points (IN, PTW and EFF) and 3 sludge phase sampling points (PS, WAS and DDMS), see Figure 49. 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 hypochlorite which has been shown to be a source for different chlorinated compounds (Ma et al., 2013).

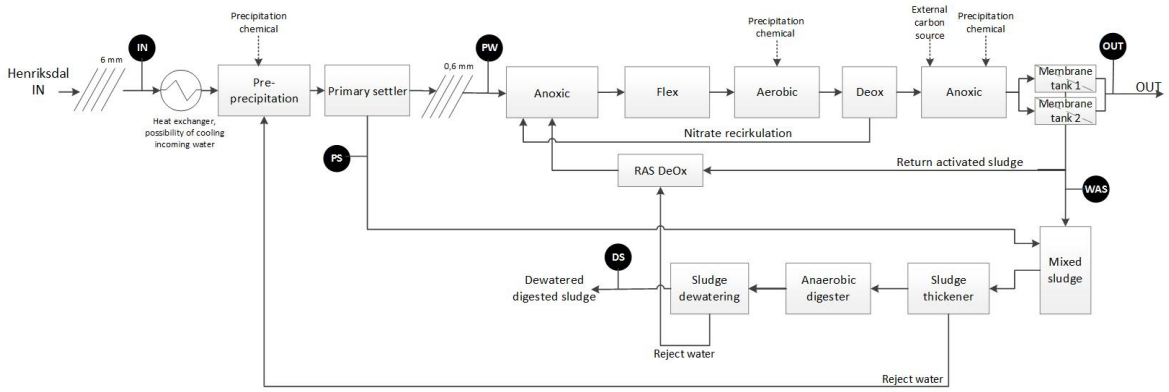



**Figure 49. Process scheme with sampling points (IN, PTW, EFF, PS, WAS, DDMS).**

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 26. 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 |

## 6.9.2 Results

The results from the 2017 sampling campaign was presented in the yearly report of 2017 (Andersson et al., 2019) and 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. The results of all four sampling are, at present date of publishing this report, being compiled and evaluated, and will be reported in a separate report to be published by the financer of this specific side project; the Swedish Water and Wastewater Association. This report will be published later in 2020.

# 7 Related publications

From the project five conference presentations were held in 2019 at three different international conferences:

Westling, K., Baresel, C., Andersson, S. L. & Narongin, M. (2019) *Membrane Bioreactors (MBR) in municipal WWTPs as turning point in wide-ranging water reuse?* IWA International Conference on Water Reclamation and Reuse, Berlin, Germany, 16-20 June 2019.

Andersson, S. L., Westling, K., Andersson S., Lindblom, E. & Taylor, S. (2019) *Evaluation of a resource efficient deoxygenation MBR zone using trials and simulation.* IWA Membrane Technology Conference & Exhibition for Water and Wastewater Treatment and Reuse. Toulouse, France, 23-27 June 2019.

Westling, K., Andersson, S. L., Andersson S. (2019) *Citric vs Oxalic Acid for Membrane Cleaning in MBR*. IWA Membrane Technology Conference & Exhibition for Water and Wastewater Treatment and Reuse. Toulouse, France, 23-27 June 2019.

Westling K., Andersson, S. L. & Andersson, S. (2019) *Citric vs Oxalic acid for membrane cleaning in MBR*. Nordic Wastewater Conference, NORDIWA, Helsiniki, Finland, 23-25 September 2019.

Westling, K., Baresel, C., Andersson, S., Wahlberg, C. & Närhi, K. (2019) *Micropollutants and Microplastics in MBR*. Nordic Wastewater Conference, NORDIWA, Helsiniki, Finland, 23-25 September 2019.

Two closely related publications (in Swedish) are to be published:

- Master thesis by Joakim Gustavsson on the digester transition from mesophilic to thermophilic conditions.
- SVU-Report on Micropollutants in wastewater treatment plants with MBR (the pilot) – a comparison to a conventional wastewater treatment plant (Henriksdal) and recipient effects (Närhi et al., 2020).

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