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

-pilot Henriksdal 2040, results from 2020

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ivl

In cooperation with

STOCKHOLM VATTEN OCH AVFALL

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Preface

This report presents work performed during 2020, within the long-term pilot study trials of municipal wastewater treatment with Membrane Bioreactors (MBR) and sludge treatment including thermophilic and mesophilic digestion. The study is carried out in cooperation between IVL Swedish Environmental Research Institute and Stockholm Vatten och Avfall (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.

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, 5 and 6 the reports are in English, see Andersson et al. (2019; 2020; 2021).

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Summary

Henriksdal wastewater treatment plant (WWTP) in Stockholm is currently being extended and rebuilt for increased capacity (from 0.8 to 1.6 million PE) and enhanced treatment efficiency (6 mg TN/L, 0.20 mg TP/L, 5 mg BOD_7/L). The reconstruction includes 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. It also includes extended pretreatment and a new treatment step for thickening of primary sludge. Digestion of thick sludge (~6% TS) will be done at thermophilic conditions, unlike today's mesophilic operation, with high organic load and relatively short retention time.

To increase the knowledge of MBRs in Nordic conditions, Stockholm Vatten och Avfall (SVOA) and IVL Swedish Environmental Research Institute have conducted long-term MBR studies in pilot scale at the R&D facility Hammarby Sjöstadsverk, located on the premises of the Henriksdal WWTP. The MBR-pilot was taken into operation in 2014 and was reconstructed to its current configuration in 2016. In 2017 the MBR pilot was supplemented with a sludge treatment line in order to study different aspects of sludge digestion.

During 2020, the MBR-pilot was operated at a fixed inflow of 4.4 m^3/h, which is 37% higher than the design average flow, with glycerol as external carbon source for post-denitrification, and aluminum (PAX) instead of Ferric (PIX) as complement to Ferrous ($FeSO_4$) for phosphorous precipitation. This was done to test the operational strategy for the first MBR-line in Henriksdal WWTP. The average effluent concentration of nitrogen and phosphorus was 3.9 mg TN/L and 0.10 mg TP/L, respectively, which means that the effluent requirements were met also this year. To achieve this, 13.7 g Fe^{2+}/m^3, 0.5 g Al^{3+}/m^3 and a glycerol dose equivalent to 16.7 g COD/m^3 wastewater was required. The low consumption of phosphorus precipitation chemicals, 0.95 mole metal per mole of phosphorus removed, was mainly due to a high enhanced biological phosphorus removal (EBPR) activity. In 2020 the phosphorous release rates (lab test which indirectly quantifies EBPR activity) went up to above 10 g PO_4-P/kg VSS,h, which is higher than all previous years and indicate very good EBPR-activity. The iron and aluminum content in the activated sludge was 4.1 and 0.5%, respectively. Two master thesis projects on EBPR were conducted to learn more about the process. Results showed that most P-release took place in the pre-denitrification zone in the presence of nitrate. During spring, ethanol was used instead of glycerol to see if EBPR was boosted, but no effect on the EBPR was observed.

The consumption of glycerol (external carbon source) was similar to values obtained for methanol in terms of COD-dose (g/m^3 ww), which was also confirmed by denitrification rate lab tests. This indicates that glycerol can be a good temporary carbon source at Henriksdal WWTP until the methanol tanks are constructed. However, the glycerol dosed is not a pure product and potential long-term effects on the concentration of colloidal total organic carbon (cTOC) and indirectly membrane performance has to be evaluated further. Ethanol had a higher denitrification rate (3.1-4.7 mg NO_3-N/g VSS, h) compared to glycerol (2.4 mg NO_3-N/g VSS, h) and a lower specific COD consumption; 1.7-4.4 g COD/g NO_3-$N_{denitrified}$ compared to 6.6 g COD/g NO_3-$N_{denitrifed}$ for glycerol in the lab tests. Nevertheless, when dosed in the pilot the consumption was almost identical to the glycerol dose, 16.2 g COD/m^3 while the specific dose was higher than for glycerol (2.9 g COD/g NO_3-$N_{denitrified}$ for ethanol and 1.8 g COD/g NO_3-$N_{denitrified}$ for glycerol).

Average total sludge age during 2020 was 17.5 days and average aerated sludge age was 6.5 days. Nitrification was complete with ammonia concentrations below 2 mg/L at all times except week 11, week 36 and week 50-51. Both week 11 and 38 the aeration was deliberately reduced to minimize foaming. In week 50-51 the nitrification was inhibited by hypochlorite in combination with cold temperature after recover cleaning was performed.

Like previous years, the membranes in membrane tank 1 (MT1) was cleaned with oxalic acid and the membranes in MT2 with citric acid. Both membranes were also cleaned with hypochlorite. The membranes

were operated with an annular average net flux of approximately 22 lmh at an average TMP of 79 mbar. The resulting permeability was 325 lmh/bar. In 2019, the cleaning scheme for maintenance cleaning (MC) with acid was optimized to reach a minimum consumption of 50% of the recommended cleaning scheme for citric acid and 25% for oxalic acid. In 2020, focus was put on reducing the use of sodium hypochlorite, which is important since the hypochlorite supposedly wears most on membranes and cause them to age and also may cause emissions of chlorine gas and formation of toxic chlorinated organic compounds. During the first 43 weeks, the aeration of the membranes was kept in Leap-Lo to facilitate evaluation of results from the chemical cleaning trials. Acid cleaning was operated according to the optimized scheme most of the year. For 24 weeks, MC with hypochlorite was turned off completely in both membrane tanks (normally done twice per week in each MT). During the first 10 weeks the hypochlorite MC was replaced with water backpulses. The permeability remained stable during this period. Thereafter both hypochlorite and water backpulses were turned off and only acid MC was done. The permeability then decreased from almost 400 lmh/bar to below 175 lmh/bar (MT1 to as low as 121 lmh/bar). To minimize hypochlorite consumption without affecting the membrane function, a filtration resistance-based method for hypochlorite MC was then tested during the last 10 weeks of the year (after restoring permeability). During this period, 4 hypochlorite MC:s in MT1 and 1 in MT2 were carried out. This corresponds to a hypochlorite consumption that is 8 times lower than the recommended cleaning scheme (2 times/week in each MT).

Recovery cleanings (RC) of the membranes were performed twice in 2020 (March-April and December) 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 rich gas emission from the tank was performed in March. The results showed that emissions occurred during a much longer period than expected and extended measurement will therefore be conducted during next year.

Two experiments were conducted in the sludge pilot in 2020; a thermophilic and a mesophilic hydraulic retention time (HRT) crash test where the HRT was gradually lowered until the anaerobic digestion processes collapsed from microbial imbalance and subsequent acidification. Due to recurrent operational problems with the thickener, it was bypassed throughout the experiments. The first trial started in July 2019 at thermophilic conditions with HRT 9 days. The initial organic loading rate (OLR) was low, around 2 kg VS/m^3,d but increased with decreasing HRT. It was difficult to provoke a crash and the digester still performed relatively well after 7 weeks at HRT 4 days and OLR >4 kg VS/m^3,d. Since it was not technically possible to further lower the HRT, glycerol and cooking oil was added in increasing amounts to the digester until the digester finally crashed (pH dropped below 6, marginal biogas production) at an OLR of almost 14 kg VS/m^3,d. A similar trial at mesophilic conditions started in June 2020 with HRT 10 days and an OLR of 1.5-2 kg VS/m^3,d. By the end of 2020 the digester had been operated at HRT 4 days and OLR >3 kg VS/m^3,d for a couple of weeks without any decrease in pH or reduction in specific methane production compared to normal values. After 5 days of glycerol dosing the process crashed on the 15th of January 2021 with an OLR of 5.3 kg VS/m^3,d. While both thermophilic and mesophilic digestion showed stable performance down to 4 days HRT, thermophilic operation managed a higher OLR before crashing and also showed a better overall performance (degradation rate, methane production) at low HRTs.

The overall resource consumption in the pilot showed that the optimization of phosphorus precipitation and membrane cleaning chemicals resulted in a significantly lower dosing than design values for the future Henriksdal WWTP. The consumption of external carbon source was similar to the design, despite the lower set point for nitrate in the pilot. The airflow demand, on the other hand, was significantly higher in the pilot, which largely is explained by the configuration/smaller depth in the tanks.

Sammanfattning

Henriksdals reningsverk i Stockholm byggs nu ut och om för ökad kapacitet (från 0,8 till 1,6 miljoner PE) och för förbättrad reningsgrad (6 mg TN/L, 0,20 mg TP/L, 5 mg BOD_7/L). Projektet inkluderar uppgradering av den befintliga konventionella aktivslamprocessen till en ny membranbioreaktorprocess (MBR) med mer än 1,6 m^2 installerad membranyta. Det inkluderar även utökad förbehandling och ett nytt steg för primärslamförtjockning. Termofil rötning av tjockt slam (~6% TS) vid hög organisk belastning och relativt låg uppehållstid kommer ersätta dagens mesofila rötning.

För att öka kunskapen om MBR-teknik i nordiskt klimat har Stockholm Vatten och Avfall (SVOA) och IVL Svenska Miljöinstitutet genomfört långtidsstudier på en membranprocess i pilotskala på F&U-anläggningen Hammarby Sjöstadsverk, som ligger i anslutning till Henriksdals reningsverk. MBR-piloten togs i drift 2014 och byggdes om till sin nuvarande utformning under 2016. År 2017 kompletterades MBR-piloten med en slamlinje för att kunna studera olika aspekter av slamrötning.

Under 2020 drevs MBR-piloten med fast inflöde på 4,4 m^3/h, vilket är 37% högre än designflödet, med glycerol som kolkälla för efterdenitrifikation och aluminium (PAX) istället för trevärt järn (PIX) som komplement till järnsulfaten för fosforfällning. Detta för att testa den driftstrategi som kommer tillämpas under de första årens drift av fullskalelinjen på Henriksdals reningsverk. Utsläppsvärden för fosfor och kväve var som årsmedel 0,10 mg TP/l och 3,9 mg TN/l vilket innebär att de framtida reningskraven underskreds även detta år. För att åstadkomma detta doserades i snitt 13,7 g Fe^{2+}/m^3 (doserades innan försedimenteringen), 0,5 g Al^{3+}/m^3 och glycerol motsvarande 16,7 g COD/m^3 avloppsvatten.

Den låga förbrukningen av fällningskemikalie för fosforrening, 0,95 mol metall per mol avskild fosfor, berodde till stor del på en hög biologisk fosforreningsaktivitet (bio-P), trots avsaknad av en anaerob zon. Under 2020 ökade fosforsläppshastigheten (indirekt mått på bio-P aktivitet) till 10 g PO_4-P/kg VSS, h vilket är högre än tidigare år och indikerar mycket väl utvecklad bio-P. Halten av järn och aluminium i slam var låg, 4,1% respektive 0,5%. Två examensarbeten utfördes inom ramen för projektet för att studera bio-P processen. Resultaten visade bland annat att fosforsläppet skedde i fördenitrifikationszonen i närvaro av nitrat. Under våren testades etanol som extern kolkälla för efterdenitrifikation istället för glycerol för att se om det förstärkte bio-P aktiviteten, men ingen effekt kunde ses.

Förbrukningen av glycerol (extern kolkälla) var likvärdig med resultaten från tidigare försök med metanol, sett till COD-dos (g/m^3 avloppsvatten), något som också bekräftades med denitrifikationstester på labbet. Resultaten visar att glycerol kan vara en fullvärdig ersättare för metanol på Henriksdals reningsverk fram till dess metanolanläggningen byggts och driftsatts. Glycerolen är dock inte en ren produkt och potentiella långtidseffekter på mängden kolloidalt organiskt material (cTOC) i processen och deras effekt på membrandriften måste utvärderas mer ingående. Etanol uppvisade en högre denitrifikationshastighet vid labbtester (3,1-4,7 mg NO_3-N/g VSS, h) jämfört med glycerol (2,4 mg NO_3-N/g VSS, h) och även en lägre specifik COD-förbrukning, 1,7-4,4 g COD/g NO_3-$N_{denitrifierat}$ jämfört med 6,6 g COD/g NO_3-$N_{denitrifierat}$ för glycerol. Trots detta var förbrukningen av etanol när den doserades i piloten 16,2 g COD/m^3 vilket är nästan identiskt med glyceroldoseringen samtidigt som den specifika dosen var lite högre än den för glycerol (2,9 g COD/g NO_3-$N_{denitrifierat}$ för etanol och 1,8 g COD/g NO_3-$N_{denitrifierat}$ för glycerol).

Slamåldern i processen år 2020 var i snitt 17,5 dagar och den luftade slamåldern 6,5 dagar. Fullständig nitrifikation med utgående ammoniumhalter på under 2 mg/l uppnåddes alla veckor utom vecka 11, 36 och 50-51. Under både vecka 11 och 38 minskades luftflödet medvetet i försök att minska flyslambildning. Under vecka 50-51 inhiberades nitrifikationen av hypoklorit i kombination med kallt vatten i samband med återställningsrengörning (RC), något som aldrig tidigare inträffat.

Liksom tidigare år tvättades membranen i membrantank 1 (MT1) med oxalsyra och MT2 med citronsyra. Båda membranen rengjordes även med hypoklorit. Membranen kördes med ett årsmedelflux på ungefär 22 lmh vid ett TMP-medel på 79 mbar. Detta resulterade i en permeabilitet på ungefär 325 lmh/bar (årsmedelvärde). Under 2019 optimerades syraförbrukningen vid underhållsrengöring (MC) till motsvarande 50% av specificerad mängd citronsyra och 25% av specificerad mängd oxalsyra. Under 2020 låg fokus på att minska förbrukningen av hypoklorit. Detta är viktigt eftersom hypoklorit är den tvättkemikalie som sliter mest på membranen och får dem att åldras samtidigt som den potentiellt kan ge upphov till utsläpp av klorgas och giftiga klororganiska föreningar. Under årets första 43 veckor låstes membranluftningen i Leap-Lo läge för att underlätta utvärderingen av hypoklorittvätten. Samtidigt kördes syratvätten enligt det optimerade schemat från året innan. Under 24 veckor användes ingen hypoklorit alls i piloten (jämför med normalt 2 tvättar per vecka och membran). De första 10 veckorna, då hypoklorit MC ersattes med vatten, höll sig permeabiliteten på en jämn nivå. När sedan backpulser med både hypoklorit och vatten stängdes av och enbart MC med syra användes, sjönk permeabiliteten från ca 400 lmh/bar till under 175 lmh/bar (i MT1 till så låt som 121 lmh/bar) över 13 veckor. Slutsatsen var att antalet hypoklorittvättar kan minskas rejält men inte helt uteslutas. För att ta reda på exakt hur ofta hypoklorittvättar måste genomföras för att upprätthålla membranens drift och funktion så infördes en motståndsbaserad styrning av hypoklorit MC som testades under årets sista 10 veckor (efter att permeabiliteten återställts). Under denna period genomfördes 4 MC med hypoklorit i MT1 och en i MT2. Jämfört med mängden hypoklorit som skulle åtgå vid standardrengöring (2 ggr/vecka i varje MT) var förbrukningen med denna metod 8 ggr mindre.

Återställningsrengöring (RC) genomfördes vid två tillfällen under 2020, en i mars/april och en i december, genom att blötlägga membranen först i hypoklorit och sen i syra. Under RC med hypoklorit i mars genomfördes mätningar av klorhaltiga gaser i frånluften. Resultaten visade att klorhaltiga gaser förekom under en längre tid än förväntat varför utökade mätningar kommer utföras vid RC nästa år.

Två experiment genomfördes i slampiloten under 2020, ett termofilt och ett mesofilt kraschtest där uppehållstiden (HRT) i rötkammaren minskades stegvis tills rötningsprocessen kollapsade till följd av surrötning. På grund av återkommande driftproblem med förtjockaren tidigare år beslutades att förbileda förtjockningen under försöken. Första försöket startade i juli 2019 med termofil rötning vid 9 dagars uppehållstid. Den organiska belastningen (OLR) var då låg, omkring 2 kg VS/m^3,d men ökade sedan med minskande HRT. Det var svårt att framkalla en krasch och rötningen fungerade fortfarande relativt bra efter 7 veckor vid 4 dagars HRT och en OLR på >4 kg VS/m^3, d. Eftersom det inte var tekniskt möjligt att sänka uppehållstiden i rötkammaren mer så tillsattes istället glycerol och matolja i ökande mängder för att på så sätt öka OLR. När rötkammaren till slut kraschade (indikerades av pH<6 och marginell gasproduktion) var OLR nästan 14 kg VS/m^3, d.

Ett likadant försök genomfördes sedan vid mesofila förhållanden. Vid starten i juni 2020 var HRT 10 dagar och OLR 1,5-2 kg VS/m^3,d. Vid slutet av 2020 hade rötkammaren drivits vid HRT 4 dagar och OLR>3 kg VS/m^3, d under några veckor utan att pH minskat eller den specifika metanproduktionen förändrats jämfört med värden vid normal drift. Efter tillsats av glycerol under 5 dagar kraschade reaktorn den 15 januari 2021 vid ett OLR på 5,3 kg VS/m^3, h. Trots att både termofil och mesofil rötning uppvisade stabil drift ner till 4 dagar HRT, kunde den termofila processen klara en högre OLR innan kraschen samtidigt som den gav bättre resultat i form av nedbryningsgrad, metanproduktion mm vid låga HRT.

Resursförbrukningen i piloten visade att optimering av fosforfällning och membranrengöring resulterade i betydligt lägre relativ förbrukning än designvärden för framtida Henriksdals reningsverk. Förbrukningen av extern kolkälla vad snarlik designvärdet trots lägre börvärdet för utgående nitrat i piloten. Luftningsbehovet å andra sidan var betydligt högre i piloten är i designen för framtida Henriksdal vilket troligtvis förklaras av utformningen av reaktorerna med relativt grunda tankar.

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 on membranes (sometimes referred to as scaling).
GS	Grab sample
Hepta	Iron(II)sulfate heptahydrate
H-dal	Short for Henriksdal WWTP
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)
PA	Pre-aeration
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/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
WWTP	Wastewater Treatment Plant

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

1 Introduction

This report presents the results from year 2020 (project year 7), of the pilot scale trials with membrane biological treatment of municipal wastewater with Membrane BioReactor (MBR, operated since 2014) and associated sludge treatment (operated since 2018). The project is 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 unique process configuration is combined with membrane filtration to reach both a high degree of purification and operational stability. Project years 2014-2019 have been presented in separate reports earlier.

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.

No major alterations to the configurations have been made in 2020, and thus chapter 2 and 3 are similar as in the previous report. However, the flowrates and load to the pilot for 2020 are summarized in section 3.3. Also, the method section is similar as described in the previous reports.

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 the plant 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 concentrations of organic matter and nutrients 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 unique MBR-process containing 1.6 million m^2 of membrane area. The first MBR-line, out of seven, will be taken into operation in 2021 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 treatment of the flow bypassing the biology at high flows. In order to increase the capacity and quality of the sludge 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 wastewater treatment plants. 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 is currently Seine Aval in France (commissioned in 2016). The plant, with a design average inflow of 2.6 m^3/s (www.thembrsite.com, 2021-01-05), is also the largest MBR installation using SUEZ's ZeeWeed ultrafiltration (UF) hollow-fiber membranes as used in this project.

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

- High seasonal variations in wastewater temperature and inflow, affecting both the membrane performance and 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 resource consumption.

There are MBR-plants in the USA, eg. *Broad Run* and *King William County* in Virginia, *Ruidoso* in New Mexico and *Cauley Creek* and *Yellow River* in Georgia, that reach very low effluent nutrient concentrations, 0.05-0.10 mg TP/L and below 6 mg TN/L, without final polishing steps (Pellegrin et al., 2015). Phosphorus removal at these plants is achieved by a combination of enhanced 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. Standard protocols for air scouring and chemical cleaning of the membranes exist. However, each plant is unique, and the cleaning schedule can and should be optimized for the local conditions to save resources.

The future doubled treatment capacity at Henriksdal WWTP will also affect the sludge treatment. The load on the digesters is expected to double but the digester volume was, at the time of the experiments, not planned to be expanded (during 2021 it was decided to construct one more digester, increasing the total volume with 20%). Consequently, digestion must be performed with higher organic load and shorter hydraulic retention time. To manage this, not only the waste activated sludge but also the primary 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 and IVL 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, the project team decided to supplement the MBR-pilot with a sludge treatment line in order to study the future digestion process.

3 Description of the pilot plant

The pilot plant is designed to be a small copy of the future Henriksdal WWTP plant, scale 1: 6,700. The incoming wastewater is pumped from the Henriksdal inlet with a mean flow of around 3.2 m^3/h. Primary treatment comprise of a fine screen (6 mm), pre-aeration, a primary settler and a fine sieve (2 mm). The biological treatment consists of 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 not to disturb the pre-denitrification. The sludge treatment consists of mechanical thickening, anaerobic digestion, and mechanical 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.

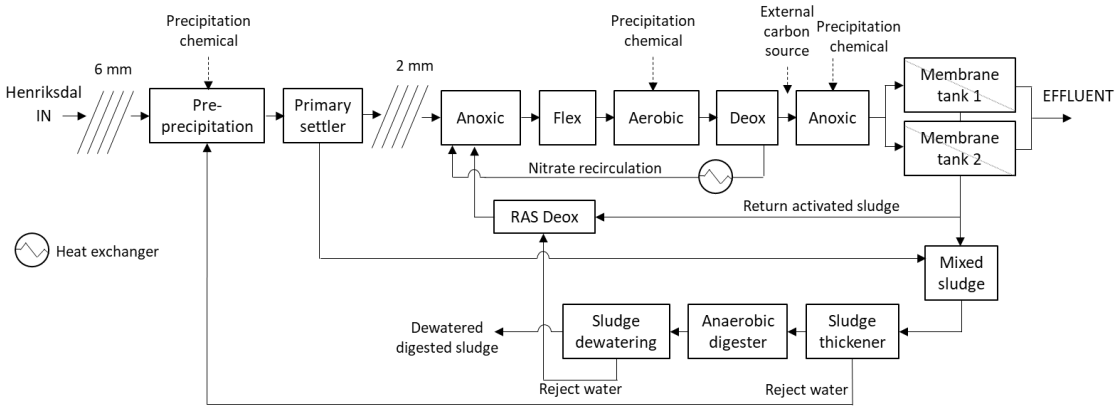


Figure 1. Flow scheme of the pilot WWTP.

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 pilot WWTP compared to the future Henriksdal WWTP.

Tank	Pilot (m^3)	Future H-dal (m^3)	Scale factor H-dal/Pilot	Specification
Pre-treatment				
PA (sand trap)	0.7	2 460	-	Pre-aeration. Dosing point 1 Fe^{2+}.
SED	3.3	30 000	9 200	Primary settler. Withdrawal of primary sludge.
Membrane bioreactor (MBR)				
BR1	4.8	33 500	7 000	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.
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+} or Al^{3+}.
MT1	1.45	9 750	6 700	Membrane tank. Aerated.
MT2	1.45	9 750	6 700	Membrane tank. Aerated.

Tank	Pilot (m^3)	Future H-dal (m^3)	Scale factor H-dal/Pilot	Specification
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

3.1.1 Incoming wastewater

Incoming wastewater to the pilot plant is pumped from the Danviken channel, one of five inlet tunnels to Henriksdal WWTP plant. The pilot influent contains a 10-20% higher concentration of organic matter (measured as BOD_7) than the combined average inflow to the Henriksdal WWTP. It also has about 60% higher BOD_7-concentration than the inlet to 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. However, flow variations in the pilot inflow are proportional to the current actual inflow variations to 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 resulting incoming load on the pilot plant is higher than the corresponding design load for the Henriksdal WWTP, year 2040, see Table 2. In 2020, the inlet concentrations of nitrogen and phosphorous to the pilot were lower than normal from March onward. This resulted in a yearly average concentration of total nitrogen 17% lower than 2019 and a total phosphorus concentration 26% lower than 2019. Also, at Henriksdal WWTP the inflow was lower than previous years, resulting in a load that was around 12% lower for nitrogen, 18% lower for phosphorus and 17% lower for BOD. This was presumably caused by the corona pandemic, leading to more people working from home and therefore fewer working people in the office areas dominating central Stockholm that are serviced by Henriksdal WWTP.

In addition, incoming wastewater to the pilot has a higher temperature than incoming wastewater to Henriksdal. Previous years, the incoming wastewater was during some periods cooled in heat exchangers to compensate for this. However, due to continuous problems with clogging, cooling was during 2020 limited to the nitrate recirculation (flow from BR5 to BR1). The processes have been tested during cold inlet temperatures previous years. The heat exchanger was controlled to maintain a temperature in MT1 corresponding to 1°C higher than inlet to Henriksdal, since temperature normally increases from inlet to biology in Henriksdal with about 1°C. The temperatures in the incoming wastewater to Henriksdal and to the pilot are presented in Figure 2. On average the temperature of the inlet water to the pilot was 19.0°C, which is 2.2°C higher than the influent wastewater to Henriksdal (16.8°C). The daily average temperature in the pilot

inlet varied between 12.8°C and 25.0°C. With cooling on the nitrate recirculation using the temperature in MT1, process temperature could be controlled to match the temperatures in Henriksdal WWTP +1°C. On average the temperature in MT1 was 17.8°C.

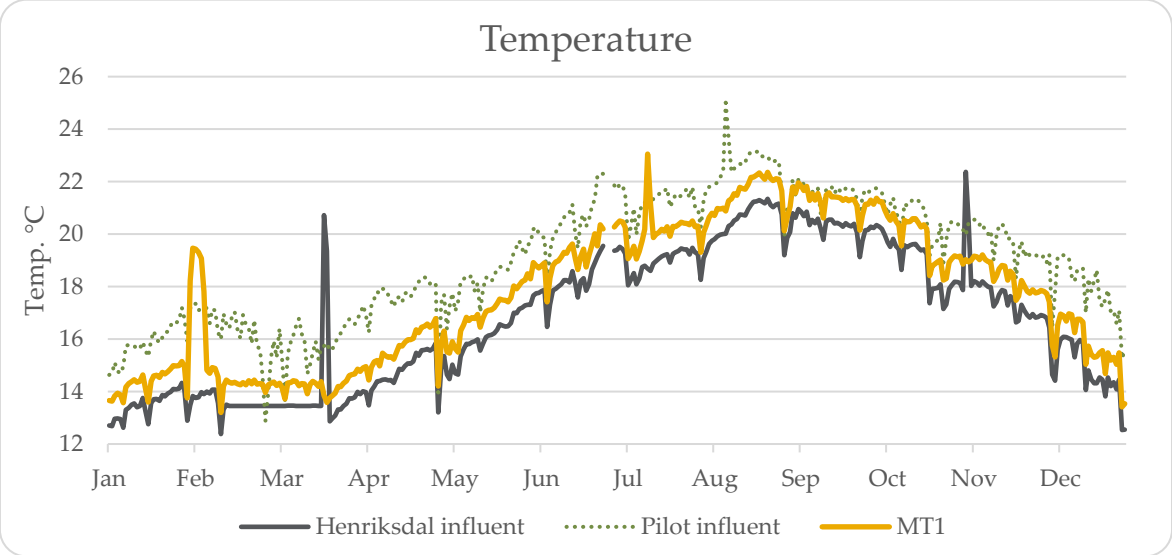


Figure 2. Influent temperatures to the MBR pilot (dotted line) and Henriksdal WWTP (black line) together with temperature in the Membrane tank 1 (MT1) during 2020.

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 3. The small hole size of the drum sieve was chosen to enable the study of clogging tendencies. In June 2020, the 0.6 mm fine sieve before the biology was changed to a 2 mm punch hole drum sieve, which is the same hole size as in future Henriksdal WWTP, to study tendencies of accumulation of trash in the biological sludge.

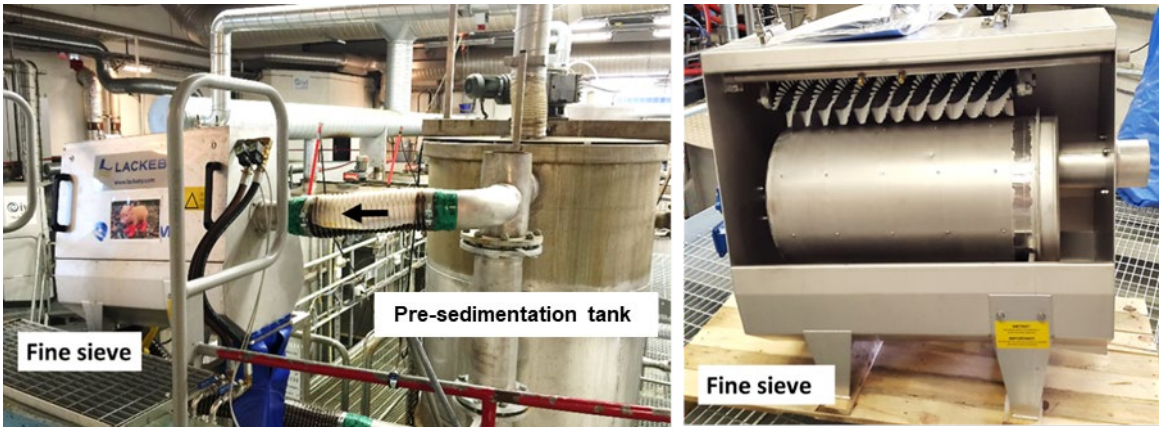


Figure 3. Photo of the fine sieve installation.

3.1.3 Biological treatment

The biological treatment consisted of six identical biological reactor tanks, BR1-6, see Figure 4. 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 addition of primarily glycerol as an external carbon source. The oxygen-rich return activated sludge (RAS) flow ($4\times Q_{in}$) 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 4. 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 compared to the full-scale design. 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. One of the questions for the future Henriksdal WWTP is how to control the RAS-pumps to maintain a steady level in the RAS-channel and to avoid flooding or creation of waves through the plant. This could not be tested in the pilot since the RAS-Deox volume is much larger (HRT ~ 10 min) and the water level did not change as fast with changes in inflow. 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 5, 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 5c, the membranes were not completely tensioned between the top and bottom, so that the air bubbles cause the fibers to move and thus more easily remove sludge stuck on the membrane fibers.

The two membrane cassettes in the pilot were operated in parallel but independently of each other to enable comparisons of different operational strategies.

Figure 5. 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 (206 m^2) corresponds to the design membrane surface installed in six treatment lines (out of seven in total) in the full-scale plant. There are two reasons for this. First, the setup corresponds to two standard design pilot cassettes from the manufacturer. Secondly, the design max flow rate to the biological treatment, according to the full-scale design of the future Henriksdal WWTP, could be treated even if one of the seven treatment lines are out of operation.

In the future Henriksdal WWTP, each treatment line (a total of seven) will have 12 membrane tanks of which each 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 average flow rate (see Table 2) 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. This mode of operation was chosen since it did not affect the concentration of sludge or pollutants around the membranes but allowed operators to control the membrane operation by adjusting the flux.

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 lower 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 6.

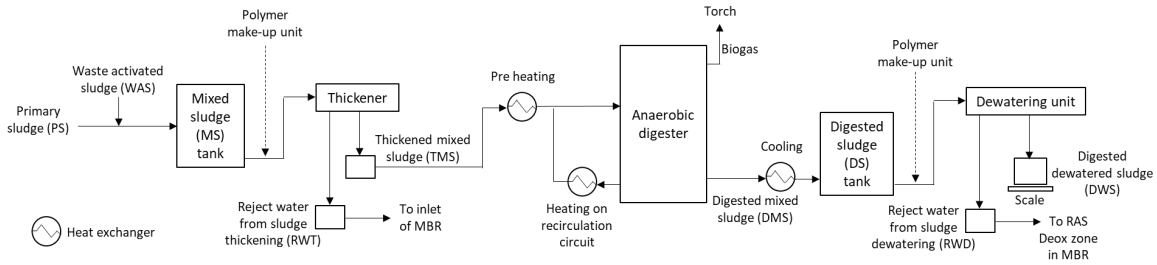


Figure 6. Process set-up for the sludge treatment line.

3.2.1 Thickening

Primary sludge and waste activated sludge was intermittently pumped to the mixed sludge tank. In the standard set-up, mixed sludge was then pumped to a rotating drum sieve thickener with addition of polymer and from there into the digester. In 2020, however, the thickener was bypassed (see chapter 6.7 for motivation).

3.2.2 Digestion

The digester is cylindrical with a base area of 2.54 m^2 and a variable sludge level. A volume of 5.7 m^3 corresponds to full digester capacity in the future Henriksdal WWTP (scale 1:6,700). During 2020 the volume was kept between 3.4 and 5.5 m^3 (instead of 5.7 m^3) in order to reach lower HRT (down to 4 days). The sludge, which is fed continuously into the recirculation circuit, 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).

During 2020, the trial with decreasing retention time at thermophilic condition, which was initiated in 2019, was finalized. A new trial with decreasing retention time at mesophilic conditions started. Special attention was given the operation at short HRT, with a master thesis student performing additional monitoring and evaluating the mesophilic period.

In the future Henriksdal WWTP design, fat from grease traps at restaurants and industrial by-products like glycerol will be co-digested with sludge. However, no external organic material was fed to the pilot digester, except during the last weeks of the trials with thermophilic and mesophilic digestion at 4 days HRT, when grease and glycerol was added to simulate a high organic loading rate (OLR) until the process crashed.

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 7.

Reject water from dewatering of digested sludge was collected in a tank for continuous pumping into the RAS-Deox zone in the wastewater treatment line. Due to operational problems the reject water was not continuously returned to the process.

Mid-November to mid-December, the dewatering unit was not in operation, but since one of the trials required reject water addition, reject water was collected at Henriksdal WWTP and dosed into the water line.

Figure 7. Photo of the dewatering equipment in the pilot.

3.3 Flow rate and load

Mean values for flow rates and loads of the pilot wastewater and sludge treatment lines during 2020 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 2020 was 4.37 m^3/h which is higher than 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 2020	Design Pilot	Design future H-dal	Value Pilot/ Design H-dal[i]
Flowrates					
Average influent flowrate, Q_{in}	m^3/h	4.37	3.16	20 880	0.021%
Design average flowrate, Q_{dim}	m^3/h		3.32	21 960	0.015%
Max flowrate	m^3/h	5.5	5.44	36 000	0.015%
Min flowrate	m^3/h	1.8	1.8	11 600	0.016%
Nitrate recirculation flowrate	m^3/h	5.1-13.1	3.8-13.3	-	-
Nitrate recirculation flowrate	× Q_{in}	2.5	1.2-4.2[ii]	0-4	-
RAS flowrate	m^3/h	4.1-23.3	3.6-19	-	-
RAS flowrate	× Q_{in}	4.2	1.1-5.9[ii]	4 (3-5)	100%
Temperatures					
Temperature influent	°C	19.0	-	-	-
Temperature biology	°C	17.8	-	-	-
Incoming load					
BOD_7 influent	mg/L	195	206[iii]	216	90%
SS influent	mg/L	221	201[iii]	280	79%
TN influent	mg/L	39	44[iii]	37	105%
TP influent	mg/L	4.7	5.7[iii]	4.9	96%
Primary settler (SED)					
BOD_7 reduction over SED	%	21	46	50[iv]	42%
SS-reduction over SED	%	33	60	60[iv]	55%
TN reduction over SED	%	0	10	10[iv]	-
TP reduction over SED	%	7	40	40[iv]	17.5%
BOD_7 PTW	mg/L	159	112	108	147%
SS PTW	mg/L	145	80	112	130%
TN PTW	mg/L	39	40	33	118%
TP PTW	mg/L	4,4	3.4	3.0	147%
SS removed over SED	kg SS/d	7.6	13.3[v]	89 300	0.009%
Primary sludge production	kg SS/d	21.7	17.2[v]	115 000	0.019%
VS-concentration PS	% of TS	88%	77%	77%	114%
Biological treatment					
BOD_7-load PTW (at average flowrate)	kg BOD_7/d	16.0	8.6	57 500	0.028%
Specific WAS-production [vi]	kg SS/kg BOD_7	0.99	1.02	1.02	97%

Parameter	Unit	Value Pilot 2020	Design Pilot	Design future H-dal	Value Pilot/ Design H-dal[i]
WAS production, average	kg SS/d	15.9	8.8	58 600	0.027%
VSS-concentration WAS	% of TSS	80%	64%	64%	125%
SS in biological tanks	mg/L	7 600	8 000	8 000	95%
SS in membrane tanks	mg/L	9 900	10 000	10 000	99%
Total sludge age	d	17.5	32.0	31.2	56%
Membrane tanks					
Installed membrane area (gross)	m^2	206	206	1 600 000	0.013%
Permeate recirculation	m^3/h	0.03-0.9	0.05-2	-	-
Net flux average (at average T)	$l/m^2,h$	22.0	17.9	20.9	105%
Net flux max	$l/m^2.h$	25.3	30.8	30	84%
Permeate pumping max	m^3/h	7.0	12.4	62 250	0.011%
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	139%
Specific air demand at Leap-Hi [vii]	$Nm^3/h, m^2$	0.252	0.252	0.196	129%

[i] Value pilot divided by Design future H-dal. Value 100% for complete compliance.
[ii] Based on average flowrate 3.2 m^3/h.
[iii] Design based on data from 2015.
[iv] Measured at Fe-dosage ca 10 g/m^3 in FL/sand trap.
[v] Calculated based on incoming load/scaled from future H-dal design with factor 6,700.
[vi] Excluding external carbon source. Calculated from process data for Values Pilot 2020. 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 from the thermophilic (w.1-17) and the mesophilic trial (w. 25-53)

Parameter	Unit	Value pilot w.1-17	Value pilot w.25-53	Design future H-dal
Sludge into digester (Thickener bypassed in the pilot. Thickened sludge in design Henriksdal)				
Flow into digester (TMS/MS)	L/h	31.4	27.0	118 000
TS-concentration TMS/MS	%	2.4%	1.9%	6.0%
TS-load TMS/MS	kg TS/d	18.0	12.4	172 000
VS-load TMS/MS	kg VS/d	13.1	9.8	124 000
VS-load EOM	kg VS/d	1.7[a]	0	44 000
Digestion				
Digester temperature	°C	55	37	55
Rector volume	m^3	4.1	4.9	38 000
Hydraulic Retention time, HRT	d	5	8	13[b]
Specific VS-load (OLR)	$kg\ VS/m^3,d$	4.3	2.0	3.3[b]
Organic degradation rate (ODR)	% of VS_{in}	52%	40%	42%[b]
VFA/Alkalinity	mg CH_3COO^- eq/mg $CaCO_3$	0.12	0.07	-
Out of digester				
Flow Digested mixed sludge, DMS	L/h	31.4	27.0	123 000
TS-concentration DMS	%	1.2%	1.1%	3.9
VS-concentration DMS	% of TS	77%	75%	60%
TS-load DMS	kg TS/d	8.8	7.4	124 000
VS-load DMS	kg VS/d	6.8	5.5	74 000

Parameter	Unit	Value pilot w.1-17	Value pilot w.25-53	Design future H-dal
Specific biogas production	$m^3/kg\ VS_{deg.}$	0.67	0.88	1.0
Specific biogas production	$m^3/kg\ VS_{in.}$	0.35	0.35	0.42
Flow biogas	m^3/d	4.5	3.1	52 000[b]
Methane content biogas	%	57%	59%	65%
TS-concentration DDMS[c]	%	26%	21%	30%
SS-concentration reject RWD	mg/L	1 810	652	900
NH_4-N in reject water	mg/L	298	150	1 500
Polymer consumption dewatering[c]	g/kg TS	14.5	11.5	6-10

a) Only feed to digester at the end of the crash test during 3 weeks. Averaged out over the whole period.
b) Numbers excluding addition of external organic material in order to be comparable to data from the pilot.
c) Different methods of dewatering. Not comparable.

3.4 Chemicals

During 2020, glycerol and ethanol were used as external carbon source (only one at a time) in the post-denitrification zone. Phosphorus was precipitated using ferrous(II)sulphate at one dosing point and aluminium(III)chloride in another dosing 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

Glycerol was tested in the pilot during 2020 since it is a potential temporary carbon source for the full-scale process at Henriksdal WWTP. The full-scale process is designed to use methanol as external carbon source, but the methanol storage facility will be constructer in a later phase of the project and therefore a temporary installation will be needed at Henriksdal during a few years. Scandinavian Biogas, the company responsible for upgrading and selling the produced biogas at Henriksdal WWTP, have storage tanks for glycerol on site and use glycerol to increase the biogas production in Henriksdal digesters when the biogas demand is high. The facts that tanks are already available, and that glycerol is not flammable, and thus doesn't require ATEX approved installations and special permits, makes it an attractive candidate for post-denitrification. However, the efficiency of glycerol as a carbon source was not known and therefore it was tested in the pilot. Glycerol was collected in 25 L containers from Scandinavian Biogas's storage. The measured COD concentration was about 850-900 g COD/L.

Ethanol was used during a couple of months for a special trial when a master student evaluated the effect of carbon sources on the enhanced biological phosphorus removal (EBPR) activity. The ethanol was delivered as 96% solution in 25 L containers with a concentration of about 1 600 g COD/L. As the available pump for dosing was too large, the ethanol solution was diluted 1:1 based on volume before added to the process.

The external carbon sources that have been tested previously in the pilot are:

- Sodium acetate (2014-2015)
- Brenntaplus (2015-2016)
- Methanol (2017-2019)
- Acetic acid (2019)
- Glycerol (2019)

The dosing point of external carbon source was 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 aluminium(III)chloride. Previous years, a combination of iron(II)sulfate and iron(III)chloride (PIX 111) was used but during 2020 the iron(III)chloride was replaced with aluminium(III)chloride.

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 36 and 70 g/L. For the batches used in the experiment, the iron content was determined by density measurement for each batch.

Aluminium chloride (PAX-XL60 from Kemira) was delivered as a solution with at concentration of 7.5% Al by weight as specified by the supplier. An aluminium concentration of 98.3 g Al/L has been used for dose calculations.

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 2020, 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 35 and 75 g Cl_2/ L during the year. Due to trials with reduced use of sodium hypochlorite in 2020, the storage time and potential degradation has been greater than before.

Citric acid solution was delivered with a concentration of 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 dewatering of digested sludge Flopam EM 640 HIB (SNF) was used during 2020 for both the mesophilic and thermophilic sludge. Previous years other polymers have been tested.

Polymer was delivered in solution and prepared to desired 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 2020

An overview of the experimental plan year 2020 is presented in Table 4 and in more detail in later chapters of the report.

The complete history of the pilot project includes initial pilot trials with flat sheet membranes and a standard MBR process configuration in 2014-2015 with the purpose to get to know MBR technology in general. Early 2016 the pilot was reconstructed to resemble the design of the future Henriksdal WWTP that had been set by then. 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 objectives reached, the overall goals for 2018 were 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 use of 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.

In 2020 the "how low can we go" theme continued with low setpoints for effluent phosphorus and nitrogen, low chemical consumption for membrane cleaning and low retention time in the digester.

Table 4. Experimental plan of year 2020.

Trial	Month of the year											
	J	F	M	A	M	J	J	A	S	O	N	D
Increased inflow	■	■	■	■	■	■	■	■	■	■	■	■
Imitation of first phase operation of full-scale	■	■	■	■	■	■	■	■	■	■	■	■
Cooling in the biology	■	■	■	■	■	■	■	■	■	■	■	■
Glycerol as carbon source	■	■	■			■	■	■	■	■	■	■
Ethanol as carbon source				■	■							
Reject water recirculation to pre-aeration	■	■										
Reject water recirculation to RAS-deox										■	■	
Effluent phosphorous target 0.10 mg TP/L	■	■	■	■	■	■	■	■	■	■	■	■
Aluminium chloride as precipitation chemical		■	■	■	■	■	■	■	■	■	■	■
In-depth study of EBPR	■	■	■	■	■	■						
Minimising membrane scouring air use	■	■	■	■	■	■	■	■	■	■	■	■
Oxalic acid and citric acid comparison	■	■	■	■	■	■	■	■	■	■	■	■
Minimising membrane cleaning chemical use	■	■	■	■	■	■	■	■	■	■	■	■
Recovery cleaning			■	■								■
Offgas measurement			■									
Extended membrane operational cycle	■	■	■	■	■	■	■	■	■	■	■	■
How low can we go - thermophilic	■	■	■	■								
Emptying and re-seeding the digester					■	■						
How low can we go - mesophilic							■	■	■	■	■	■
Sludge dewatering in operation	■	■	■			■		■	■	■	■	
Study on indicators of process failure								■	■	■	■	■

Minimizing the resource consumption, especially for membrane cleaning was continuously in the spotlight this year. Further reductions in acid cleaning chemical consumption, both oxalic and citric have been evaluated. In addition, the membranes were operated without sodium hypochlorite cleaning from 24th of March to 9th of September (169 days). Recovery cleaning of the membranes was carried out twice, first in March to April and later in December. Chlorine gas emissions from the process was measured during soaking of the membranes in sodium hypochlorite.

During most of the year, the production goal for phosphorus and nitrogen was lowered according to the theme of the year. 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, imply that the effluent phosphorous must be even lower than the concentration limit of 0.20 mg TP/L. To test the capacity of the process, the target concentration in the pilot operation was lowered to 0.10 mg TP/L.

At the same time, the inflow was increased in order to mimic the first years of operation of the full-scale MBR-line at Henriksdal WWTP, with a high, fixed inflow (37% higher than design average and 26% higher than 2019). The difference from previous similar trials was that glycerol was used as carbon source, while the previous version of this trial was carried out without external carbon source. The reason for this was that glycerol will be temporarily utilized as carbon source in the first MBR-line at Henriksdal WWTP for a few years until the methanol storage facilities have been constructed. It was therefore interesting to test its denitrification potential and possible effects on the overall treatment and membrane performance. In addition aluminum(III)chloride (PAX) was used in combination with ferrous sulphate (Hepta) for phosphorus precipitation, something that was not tested before. A strategy with three precipitation dosing points (a flow proportional dose of Hepta added at the inlet and two dosages controlled by phosphate feedback control as simultaneous precipitation using Hepta and ferric chloride (PIX) in BR4 and BR6, respectively) has been shown successful in maintaining stable low effluent phosphorous concentrations. But, as the PIX dosing equipment will not be ready to take into operation during the start of the first full-scale treatment line, PAX (which is currently used for storm water treatment at Henriksdal WWTP) will be used in the full scale initially. To study the capacity and any potential differences, PIX was replaced with PAX in the pilot in March 2020. To get a higher dosage of PAX, the dosage of Hepta in the biology was stopped. In addition to evaluating the phosphorus removal using PAX, the plan included monitoring potential effects of the acids used for membrane cleaning and possible effects on the sludge treatment from the aluminium product.

Digestion was tested at very low HRT at both thermophilic and mesophilic conditions to find out at what point the process starts to fail. The trial is important since the digestion retention time will be jeopardized in near future when digesters will be reconstructed and in a more distant future with high load and high dependency on mechanical thickeners to maintain the retention time.

Separate trials in 2020 included the evaluation of enhanced biological phosphorus removal within two master thesis projects. As part of one of the projects, another external carbon source, ethanol, was used in March. In the autumn, another master thesis project was performed focusing on indicators of digester failure while lowering the retention time.

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 8.

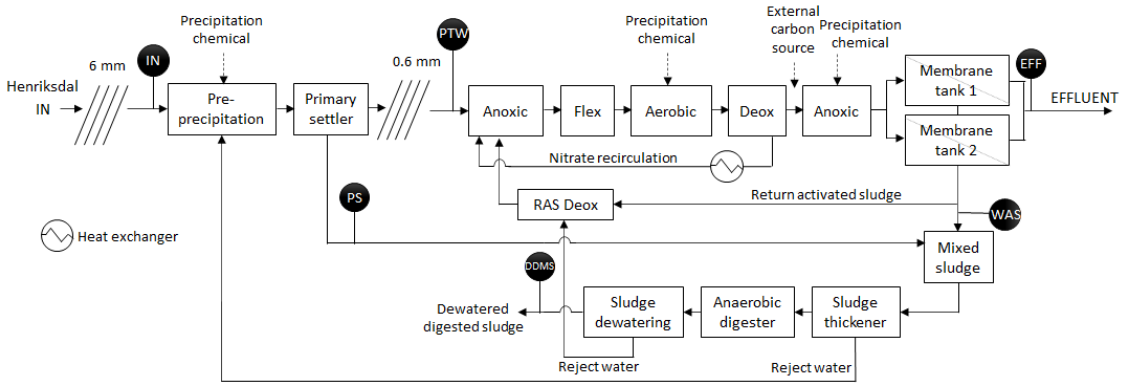


Figure 8. Sampling points in the pilot process marked as black circles (SLUDGE1 and SLUDGE2 sampling points not included in the figure).

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 point and sample type. 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 mixed 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 three months, 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). Parameters

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

Sampling point	Parameters														
	TOC	COD	BOD_7	TP	PO_4-P	SS	VSS	cTOC	NH_4-N	NO_3-N + NO_2-N	TN	Alkalinity	Fe (digested)	Al (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	1
Reject water mixed sludge thickening[1]						1	1								
Reject water digested sludge dewatering[1]			1	1	1	1	1		1		1		1		
Mixed sludge (MS)[2]		1													
Digested mixed sludge (DMS)[2]		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**	**4**	**3**	**7**	**4**	**6**	**3**	**1**	**4**	**3**	**4**	**2**	**8**	**1**	**1**

[1]Samples were sent when thickener/dewatering unit was in operation.
[2]Sampling started September 2 (w. 36) 2020.

In addition to the external analyses, analyses were also performed internally at IVL's laboratory at Hammarby Sjöstadsverk for daily operation. Water 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.

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

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 analysed onsite 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 the MBR pilot.

Placement	Parameter	Function
General	Flowrate (water)	Measure all recirculation flows
IN	Temperature	Measure the incoming wastewater temperature
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 into aerated part of biological treatment. Sometimes used for control of aeration.
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
BR5ox	DO	Controlling Dissolved Oxygen
BR5ox	Flowrate (air)	Measure air flow
BR5deox	NO_3-N	Measure nitrate concentration, monitor function of post-denitrification
BR5deox	NH_4-N	Measure ammonium after aerated part of biological treatment, monitor function of nitrification. Sometimes used for control of aeration.
BR5deox	DO	Monitor Dissolved Oxygen
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
EFF	NH4-N	Measure effluent ammonium 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
AD	Radar	Monitor foam in 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. The membranes are operated in cycles with 15 minutes of permeate withdrawal and 1-minute relaxation (extended operational cycle, standard operational cycle is 10+1 minutes). The membrane performance parameters can be calculated as *gross* values (using only data from the 15 minutes of actual permeate withdrawal) or as *net* values (using average data from the full operation cycle, permeation and relaxation = 16 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 and measured using online pressure transmitters 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 $T = Temperature;\ \theta = 1.025\ if\ T \geq 20\ °C; and\ \theta = 1.033\ if\ T < 20\ °C$

5.3.2 Sludge quality

In addition to the parameters analysed at the external and internal laboratory as 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 was 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). The concentration of colloidal TOC (cTOC) should be less than 10 mg/L (membrane supplier).

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 widths 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

Organic degradation rate (ODR)

The organic degradation rate (% of VS_{in}) was calculated based on the mass flow of VS (kg VS/d) into the digester (mixed sludge, MS) and out of the digester (digested sludge, DMS) using the equation below:

$$ODR = \frac{(VS_{MS} - VS_{DMS})}{VS_{MS}}$$

Specific biogas and methane production

The specific biogas production was calculated in two ways, either based on the amount of VS that was fed to the digester, VS_{in}, or the amount that was degraded in the digester, VS_{deg}. The values that are primarily evaluated in this report is m^3/kg VS_{in}, which was calculated using the daily biogas production (m^3/d) divided by the amount of VS fed to the digester per day (VS_{MS}). 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 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 factor was based on results from previous trials and corresponded well with values foud in literature.

6 Results and discussion

6.1 Primary treatment

Trial	J	F	M	A	M	J	J	A	S	O	N	D
37% increased inflow												
Imitation of first phase operation of full-scale												

6.1.1 Inlet screen

Different punch hole screens, from 2 mm to 6 mm, have been used for the initial screening since the start of the project. Since 2017 a 6 mm size was been used to increase the primary sludge production which was required to meet the design load for the sludge pilot. As Figure 9 indicates, inlet suspended solids concentrations corresponded, after that change, well to data from Henriksdal WWTP, especially during 2017 and 2018. During 2019, there were frequent maintenance work done in the Henriksdal digesters, which include that the sludge is emptied into the inlet channel where the inlet pumps to Hammarby Sjöstadsvek are located. This resulted in higher suspended solids concentrations in the pilot inlet compared to Henriksdal (and previous years). Figure 9 shows that the inlet SS-concentration to the pilot decreased in 2020 although the 6 mm screen was used. The main contributing factor was considered to be the corona pandemic which resulted in less people in the central parts of Stockholm, from where the wastewater to the pilot originated (Danviken inlet).

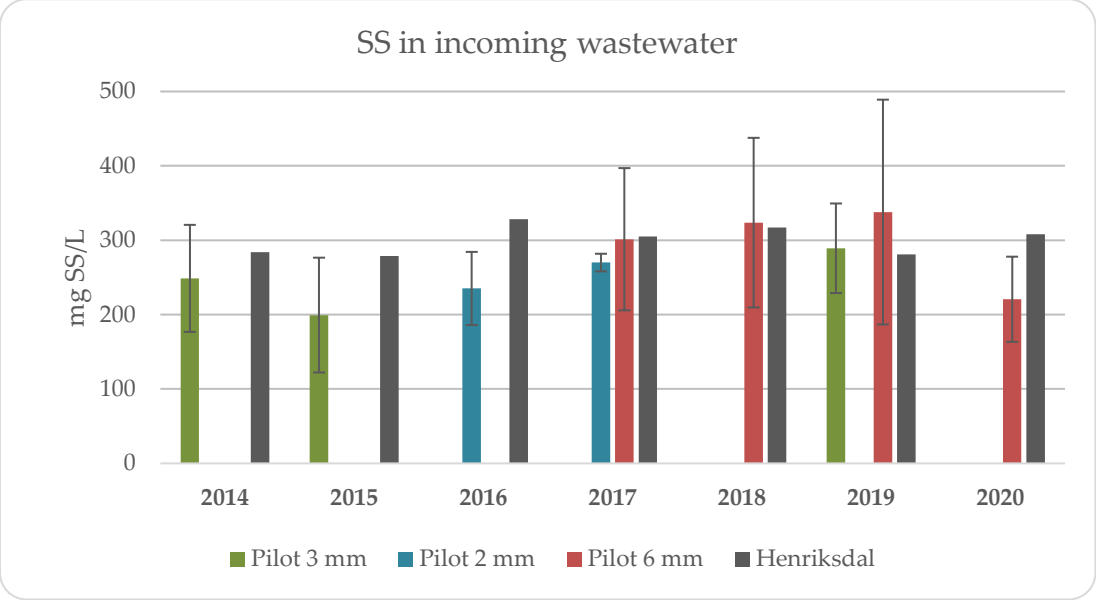


Figure 9. Incoming SS-concentration to the pilot after passing through 2-, 3- or 6-mm punch hole screens and to Henriksdal (3-mm step-screen).

6.1.2 Efficiency of primary settler

The primary settling volume in the pilot (3.3 m^3) and its capacity, is smaller than it should be (design value 4.5 m^3). This results in a poor reduction of solids and organic material compared to Henriksdal WWTP, and insufficient primary sludge production. The reduction rate over the primary settler is showed Table 9 and Figure 10 below. During 2020, two additional aspects affected the efficiency of the primary settler. First, the inflow to the pilot was high, 4.4 m^3/h compared to the design average flow of 3.2 m^3/h. This resulted in a shorter reaction time for phosphorous coagulation as well as shorter settling times and thus even lower reduction compared to previous years, but higher total primary sludge production. Second, the corona pandemic affected the composition of the influent wastewater resulting in lower inlet concentrations.

Table 9. Reduction over the primary settler and primary sludge production in the pilot compared to Henriksdal WWTP 2020 and the future design of Henriksdal WWTP.

Parameter	2020	2019	2018	2017	Henriksdal WWTP 2020[a]	Design future Henriksdal
Inflow (m^3/h)	4.37	3.56	3.50	3.40	12 400	22 000
PE (1 PE = 70g BOD_7/d)	292	352	318	314	850 000	1 600 000
Reduction of SS	32%	36%	35%	37%	56%	60%
Reduction of BOD_7	19%	27%	25%	30%	54%	50%
Reduction of TP	8%	12%	10%	14%	33%	40%
Reduction of TN	0%	4%	1%	4%	2%[b]	10%
Reduction of TOC	10%	20%	18%	17%	37%	-
PS-production (kg TS/d)	21.7	18.5	16.1	13.1	61 000[c]	115 000
PS-production (g TS/PE/d)	74	53	51	42	73	72

a) At high flows Enhanced Pre-Precipitation using Al^{3+} is used for half of the primary clarifiers.
b) New sampling point for the Sickla inlet (SIN), no longer including the reject water from digested sludge dewatering.
c) Based on uncertain TS-measurements.

Higher concentrations of solids, organic material and nitrogen after the primary settler than in incoming wastewater have been observed at certain times, mainly after w45 (Figure 10). No obvious cause could be identified, but the installation of new samplers during this period may partly be responsible. This aspect will be further investigated.

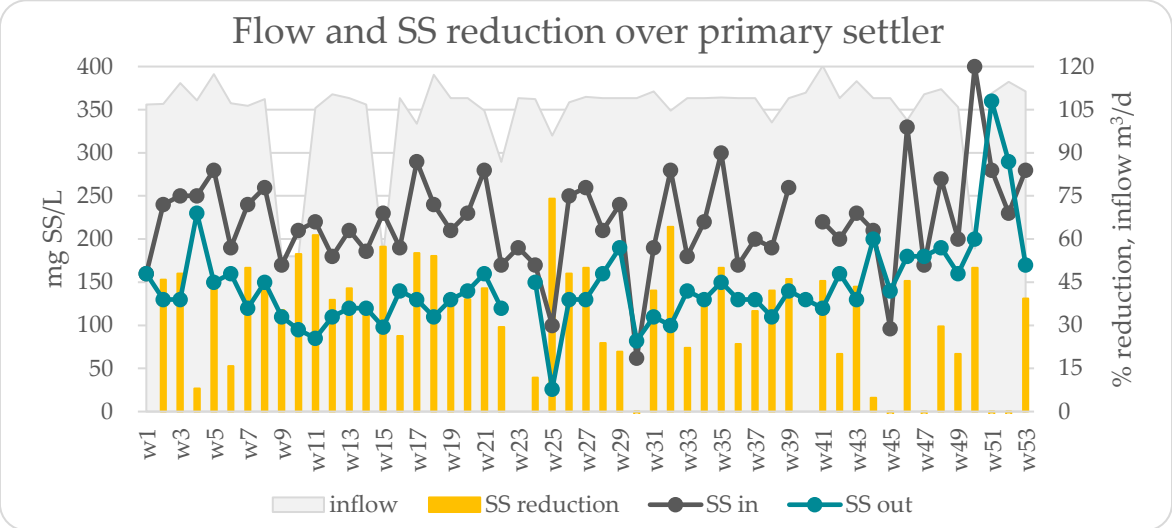


Figure 10. Reduction of SS over the primary settler after a 6 mm screen.

6.1.3 Screen and sieve – effect on trash content

During 2020 the pilot line was operated with a 6 mm punch hole inlet screen (see 6.1.1 Inlet screen) and at first a 0.6 mm, and later a 2 mm, mesh fine sieve before the biological treatment. Over the years, different screen-sieve configurations have been tested. In order to monitor the amount of particles, fibres and hair that accumulate 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 showed very little build-up of trash, indicating that the measured values are accurate. On June 24th, 2020, the 0.6 mm fine sieve before the biological treatment was replaced with a 2 mm fine sieve. The 0.6 mm sieve was initially selected (instead of the recommended 2 mm) to provide further protection of the membranes and to evaluate the amount of screenings removed by the fine sieve. Not much material was removed by the sieve and the trash-content in the sludge was low which led to the conclusion that larger hole size, which is preferable from a hydraulic point of view, can be used. In July the 0.6 mm drum was changed to one with 2 mm holes, which is the size that will be used in the full-scale plant. Surprisingly, the measurements from 2020 indicate less amount of trash content in the sludge although both screen (6 mm) and sieve (2 mm) are the largest sizes tested.

Table 10. Trash content in waste activated sludge (WAS) with various sieve-configurations. According to the membrane supplier, the trash content with a 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 mm screen at inlet pump and 0.6 mm fine sieve before biology	Jan 2020 – June 2020	1	0.4	0.2
6 mm screen at inlet pump and 2 mm fine sieve before biology	July 2020 – Dec 2020	3	1.4 ± 0.5	0.3 ± 0.2

6.1.4 Pre-treated wastewater

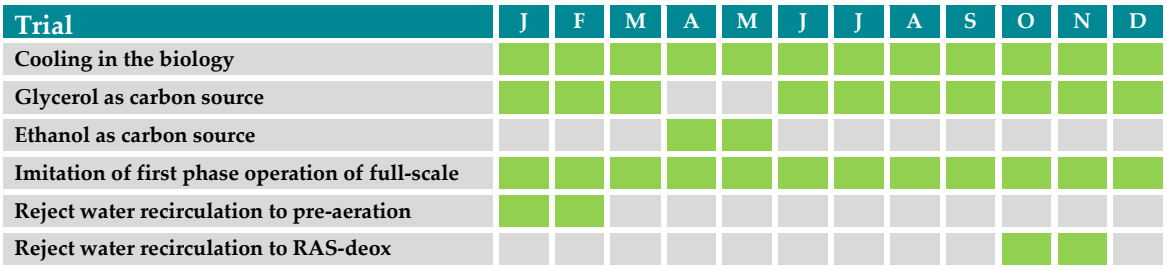
The quality of the pre-treated wastewater (PTW) is presented in Table 11. The concentrations measured in the pilot were significantly lower compared to previous years, however still higher than the corresponding concentrations measured in Henriksdal WWTP for SS, BOD_7 and TP. This is mainly due to the poor performance of the primary settler and was also observed previous years. The concentrations of SS, BOD_7 and TP in pre-treated water were also higher in the pilot than the design values given for the future plant, which though are relatively low since they are determined a weighted average between Henriksdal WWTP and Bromma WWTP, and Bromma WWTP has a diluted inflow.

The concentrations of nutrients and organic matter 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.

Table 11. Data on PTW from the pilot compared to data from Henriksdal 2020 and the design data for the future Henriksdal WWTP.

Parameter	Value Pilot 2020	Value Henriksdal 2020	Design future Henriksdal	Value pilot/ Design future H-dal
Pre-treated wastewater (PTW) – into biological treatment				
SS (mg/L)	145 ± 50	135	113	128%
BOD_7 (mg/L)	159 ± 35	99	108	147%
TN (mg/L)	39 ± 5	39	33	118%
TP (mg/L)	4.2 ± 0.8	3.3	3.0	140%
Fe (mg/L)	5.8 ± 1.9	10.4	12	48%
Alkalinity (mg/L)	251	250	250	100%
BOD_7/TN (mg/mg)	4.1	2.5	3.3	124%

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 3.9 mg/L. Four out of 53 weekly composite samples were above the target concentration of 6 mg N/L. The reduction of total nitrogen over the biological treatment step was 90%.

Table 12. Nitrogen concentrations in primary treated water (PTW) and effluent during 2020.

Parameter	Limit	Average	Min	Max	No. of weekly samples
TN PTW (mg/L)	-	39	28	52	53
TN EFF (mg/L)	6	3.9	1.7	11.0	53

Effluent nitrogen concentrations as weekly composite samples are presented in Figure 11. During most of the year, nitrogen removal has been satisfactory with stable effluent concentrations below 6 mg/L, as also shown previous years. In December (week 50-51), problems with the nitrification occurred after recovery cleaning of the membranes with sodium hypochlorite. The resulted in that the effluent TN reached 11 mg N/L. This has never been observed before. However, reduced nitrification capacity around that time of the year has been observed every year. Nitrification though quickly recovered and by the end of the year effluent ammonium was below 0.2 mg N/L.

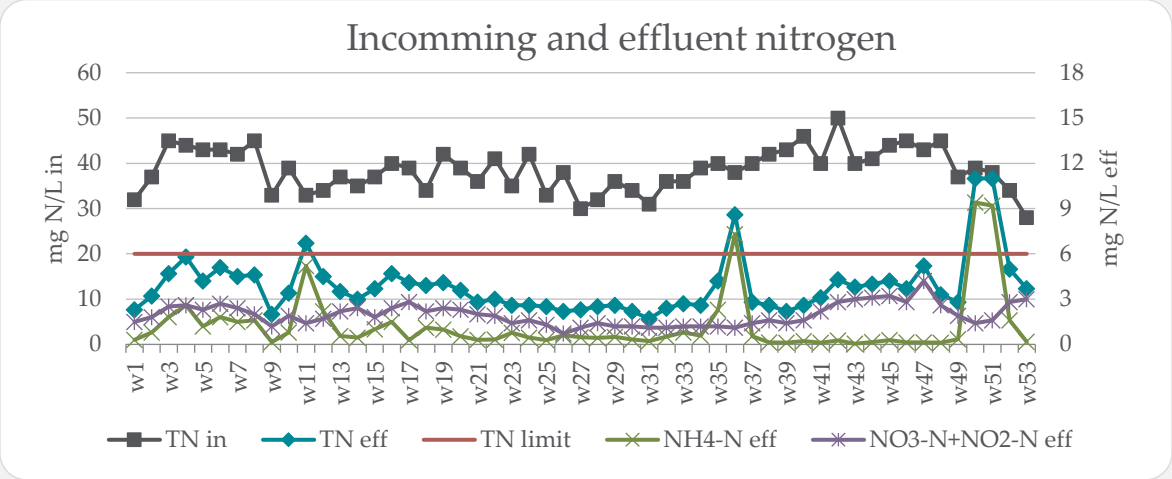


Figure 11. Incoming and effluent nitrogen concentrations from analysis of weekly composite samples during 2020. Limit for effluent total nitrogen was set to 6 mg N/L.

In Table 13, key values for the nitrogen removal are presented for the MBR pilot and compared to the future Henriksdal design. The amount of removed total nitrogen during 2020 (3.7 kg N/d) was the slightly higher compared to 2018 and 2019 (3.6 kg N/d) and somewhat higher compared to the full-scale design. Different external carbon sources have been tested during this year. Glycerol dosage was used for the majority of the year and the average consumption per day was 1.57 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 in the pilot and the future Henriksdal design.

The airflow to the pilot was much larger than the estimated for the future Henriksdal WWTP (Table 13). One reason for the big differences in airflow is that the depth in the basins of the pilot were one fourth of the depth in Henriksdal. The aeration of the biology in the pilot was also much higher in 2020 (72 m^3/h), compared to previous years, when average aeration has been around 50-54 m^3/h (more about this in chapter 6.2.1).

Consumption of external carbon source was more than double this year using glycerol compared to last year with methanol, which can partly be explained by the high inflow and thereby higher nitrogen load.

Table 13. Comparison of parameters related to the nitrogen removal between operational data from the pilot and the design for future Henriksdal (2040).

Parameter	Unit	Value Pilot	Future H-dal design	Value pilot/ scaled future H-dal design[i]
Removed nitrogen (incl. reject water)	kg N/d	3.7	21 000	118%
Nitrogen removal rate	g N/kg VSS, d	17.65	17.6	100%
Aerated sludge age (incl. MTs)	d	6.5	9.4[ii]	69%
Aeration biology (activated sludge)	Nm^3/h	72.3[iii]	68 000	712%
Specific oxygen demand (SOTR)	kg O_2/d	83[iv]	240 000	232%
Consumption of external carbon	kg COD/d	1.7[v]	12 000	97%

[i] The value of future H-dal 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 of ammonium except for shorter periods and three peaks in effluent ammonium (w.11, w.36 and w.50-51). When comparing the aeration of the biology with aeration of the membrane tanks (Figure 12) it can be observed that the airflow to the biology was more than double the airflow to the membranes 47 out of 53 weeks. The membranes were operated at the lower aeration level (Leap-Lo corresponding to 14 m³/h each, 28 m³/h in total) most of the time, with only a few peaks above 28 m³/h. The airflow to the biology varied as weekly average between 31.5 and 127 m³/h and was 72 m³/h as yearly average.

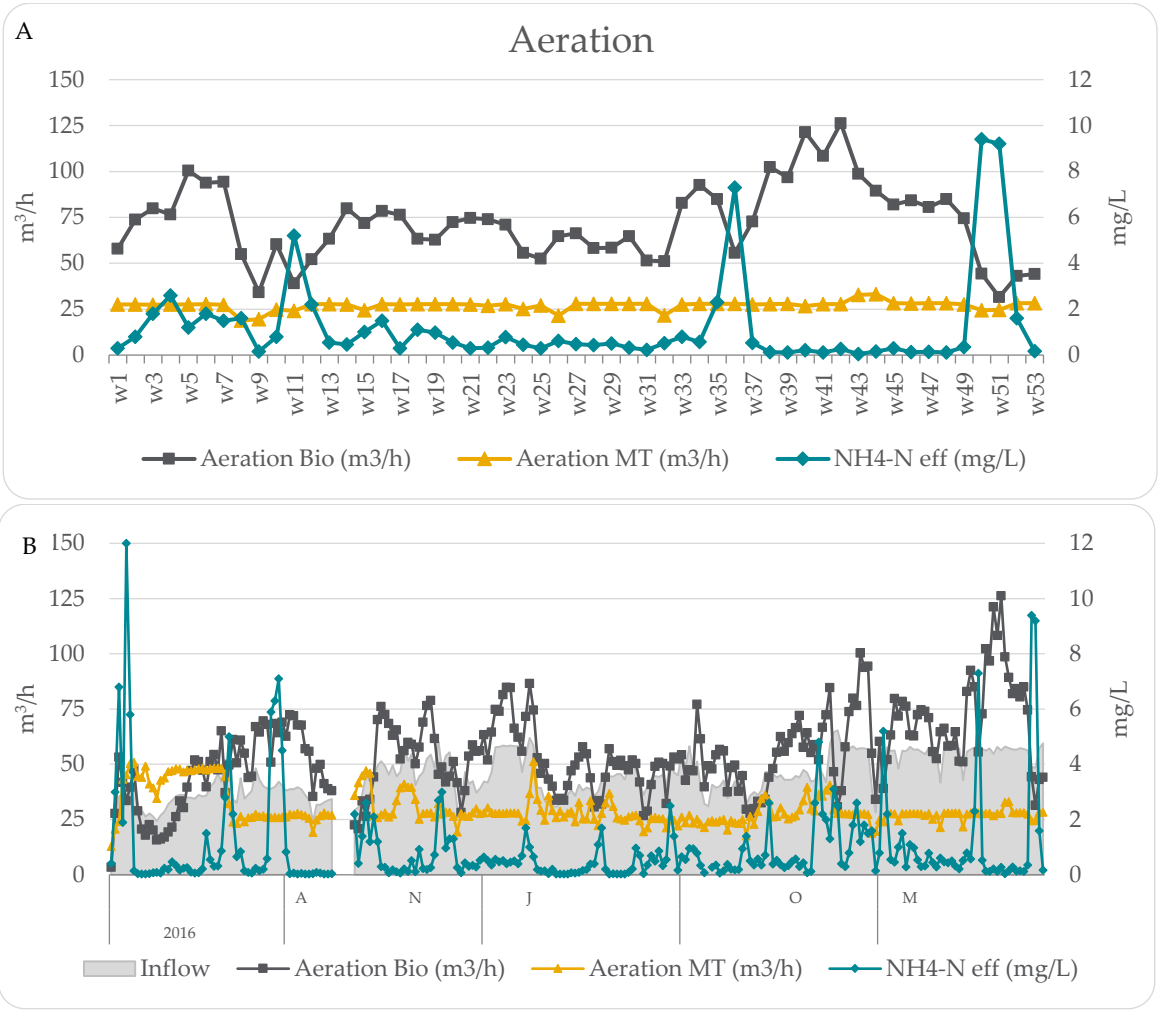


Figure 12. A) Aeration need in biology and membrane tanks (MT) together with effluent NH_4-N during 2020. B) Aeration, influent flowrate and effluent NH_4-N from 2016.

Table 14 shows the annular average airflows and DO levels in the aerated tanks year 2016-2020 together with the calculated theoretical airflow demand. As can be seen, the airflow year 2020 was higher than previous years. To some extent this can be explained by the high inflow resulting in a higher load of nitrogen and BOD on the biology. The inflow 2020 was 37% higher than design average flow and 26% higher than 2019. Nevertheless, the load of TN to the biology only increased by 12% and the BOD load was the same in 2020 as

in 2019. The airflow, on the other hand, was 49% higher in 2020 than in 2019. When calculating the airflow demand, using standardized design calculations, it can be seen in that the calculated value is within 10% difference from the measured value year 2016-2019 (Table 14). In 2020, however, the deviation is 38%, which clearly indicates some type of problems with the aerators, such as fouling or broken membranes, problems with the airflow meters or, less likely, some other type of unknown interference with aeration. The influence from the increased dose of external carbon source compared to previous years did not affect the airflow demand since it was consumed in the post denitrification zone, or in case of overdosing, in the membrane tanks. The possible effect of the BOD load from external carbon on the airflow demand was also investigated by adding the external carbon source to the incoming BOD load to the biology and recalculate the airflow demand. The result show that the airflow demand increased with 4% if all the external carbon went to the aerated zones, which would still not explain the 38% deviation. Because of the exceptionally high measured airflow, it was decided to exchange the aerator membranes in the beginning of 2021.

Table 14. Airflow and DO in the different zones year 2016-2020. Calculated airflow demand is based on the load of BOD and TN in PTW, as well as DO and temperature in the biological reactors for each year.

Year	Inflow (m^3/h)	Airflow BR3 (m^3/h)	DO BR3 (mg/L)	Airflow BR4 (m^3/h)	DO BR4 (mg/L)	Airflow BR5 (m^3/h)	DO BR5 (mg/L)	Airflow Tot (m^3/h)	Calc. Airflow Tot (m^3/h)
2020	4.38[f]	23.2	1.42	38.7	2.71	10.5	1.67	72.5	52.6
2019	3.48[d]	18.3	1.45	16.2	2.64	13.9	2.00	48.5	51.8
2018	3.52[d/f]	17.3	1.75	19.1	1.98	15.9	1.60	52.3	47.6
2017	3.47[d/f]	20.9	1.91	20.4	2.89	10.0	2.19	51.3	51.4
2016	2.84[d]	6.0	0.78	20.6	2.74	11.2	3.70	37.9	42.2

f - fixed flow
d -dynamic flow

The first period with incomplete nitrification in week 11 was related to low DO setpoints (< 1 mg/L) in the treatment line, especially BR3. This was done as a measure to reduce aeration to limit the overflow of sludge due to foaming. The same problems with foaming sludge occurred in week 35 and again the aeration was reduced, and effluent ammonium increased.

In week 50, directly after startup from recovery cleaning (RC) with sodium hypochlorite, nitrification was significantly affected and the ammonium concentration in the treatment line and effluent increased (Figure 13). After a couple of weeks, the process had recovered, and effluent ammonium was low again. This was the 12th recovery cleaning since 2016 and never before has the nitrogen removal been affected. No clear differences in the cleaning or neutralization procedure were observed and no other explanation has been found.

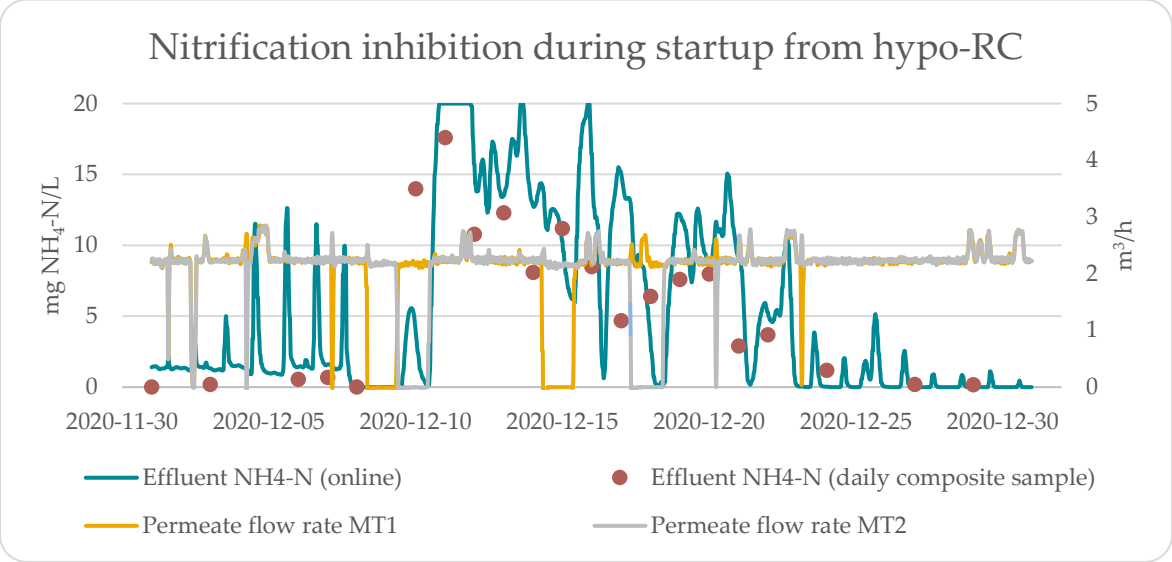


Figure 13. Nitrification inhibition after startup from recovery cleaning (RC) with sodium hypochlorite.

On average, the aeration of the membranes accounted for 28% of the total aeration, previous years (2017 to 2019) this number was about 36%. However, the lower value for 2020 is mainly caused by an increase in aeration of the biology. Membrane aeration has been significantly reduced compared to the first year (2016) when 54% of the total aeration was used for the membranes and the average membrane aeration was 42 m^3/h.

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 1 and 4 mg/L throughout the year. Aeration of BR3 was using a flow-based control, where aeration was on (with DO setpoint 2.5 mg/L) only when flowrates to the treatment was above 2.5 m^3/h to prevent excessive aeration at night when inflow was low. Due to problems with foaming, aeration of BR3 was manually operated from week 11 to week 23. From week 23 aeration was again automatically controlled, this time based on effluent ammonium concentration. Aeration of BR3 to DO setpoint (varying between 1.0 - 2.5 mg/L) was triggered if effluent ammonium was above target concentration (varying between 0.2 to 1.0 mg NH_4-N/L).

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 WAS outtake was determined based on the MLSS concentration in the line and the sludge age was not actively controlled and has only been monitored.

Waste activated sludge (WAS) flowrate has been automatically controlled using a feedback controller to keep the suspended solids concentration in BR4 at setpoint 8 000 mg SS/L. The resulting total and aerated (including membrane tanks) sludge age is presented in Figure 14 together with the WAS flowrate. The peaks in sludge age around week 12 and week 28 are the 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 17.5 days, which is lower than future Henriksdal design (25 days). This is partially due to the higher load of SS and BOD to the biology compared to the Henriksdal design. However, the calculations of sludge age in the pilot 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 6.5 days.

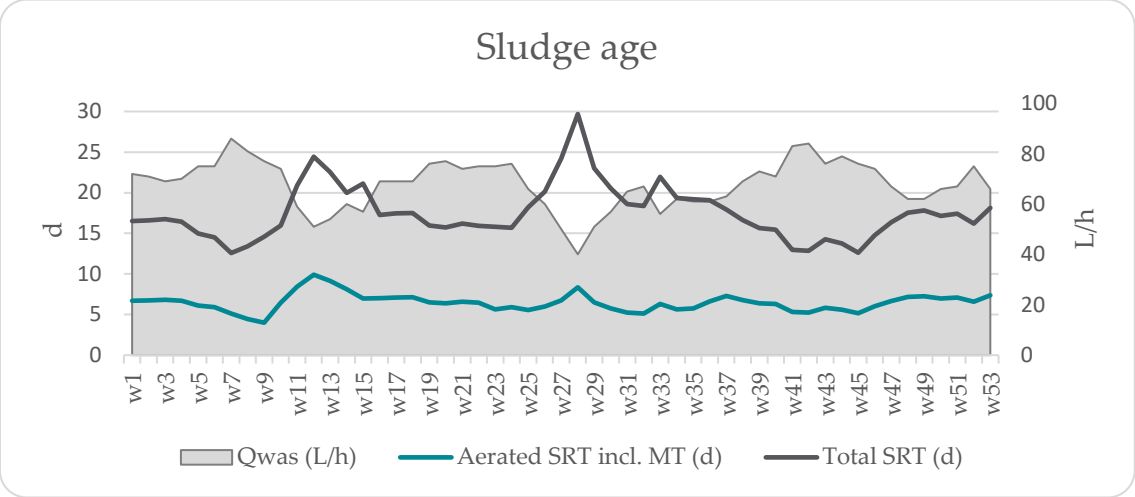


Figure 14. Total and aerated (incl. MTs) sludge age (moving average one month back in time) together with WAS flowrate.

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 in January/February (6 days) when the pump was out of operation and a month-long trial using 1 x Qin (September). 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 1.4 – 6.2 mg/L (average 4.2 mg/L) according to the online sensor. 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 0.8 - 4.9 mg NO_3-N/L with 2 mg/l as yearly average value (based on effluent online sensor).

For the post denitrification (BR6) different external carbon sources have previously been used. This year glycerol was used, except for a 10-week trial (w.14 to w.23), when ethanol was tested. The reason for testing ethanol was mainly for evaluation of the enhanced biological phosphorus removal (EBPR), as ethanol is considered a suitable carbon source for EBPR. The external carbon source has been added to a point in the piping between BR5 and BR6 and the dosage has been controlled by the online nitrate concentration.

Due to sensor problems, the signal used for dosage control has been alternating between the nitrate sensor in BR6 and the nitrate sensor in the effluent. The sensor readings from BR6 were preferred as these provide a faster control strategy. However, due to foaming and sensor failure, the effluent nitrate sensor was used from time to time. The nitrate setpoint was 2 mg NO_3-N/L until June when glycerol was again used and the setpoint was lowered to 1 NO_3-N/L. From w.41 onward the setpoint was increased to 3 mg NO_3-N/L.

An overview of the external carbon source addition is given in Figure 15. The effluent nitrate concentration was possible to control and keep below 3 mg NO_3-N/L with both glycerol and ethanol. The 10-week long test with ethanol showed a similar consumption of COD as for glycerol. A more detailed evaluation of the denitrification potential of ethanol is presented in the next section (6.2.3 Comparison of different external carbon sources). During week 36 the nitrate recirculation was decreased from 3 x Qin to 1x Qin in order to induce an increased carbon source dosage. The objective with this test was to evaluate any relation between glycerol dosage and cTOC. It is clear from Figure 15 that the lower nitrate recirculation resulted in an increased carbon dosage but not lower effluent nitrate, as expected. In week 41 the nitrate recirculation was readjusted to 3 x Qin and carbon source dosage was stopped until w.48 due to specific trials.

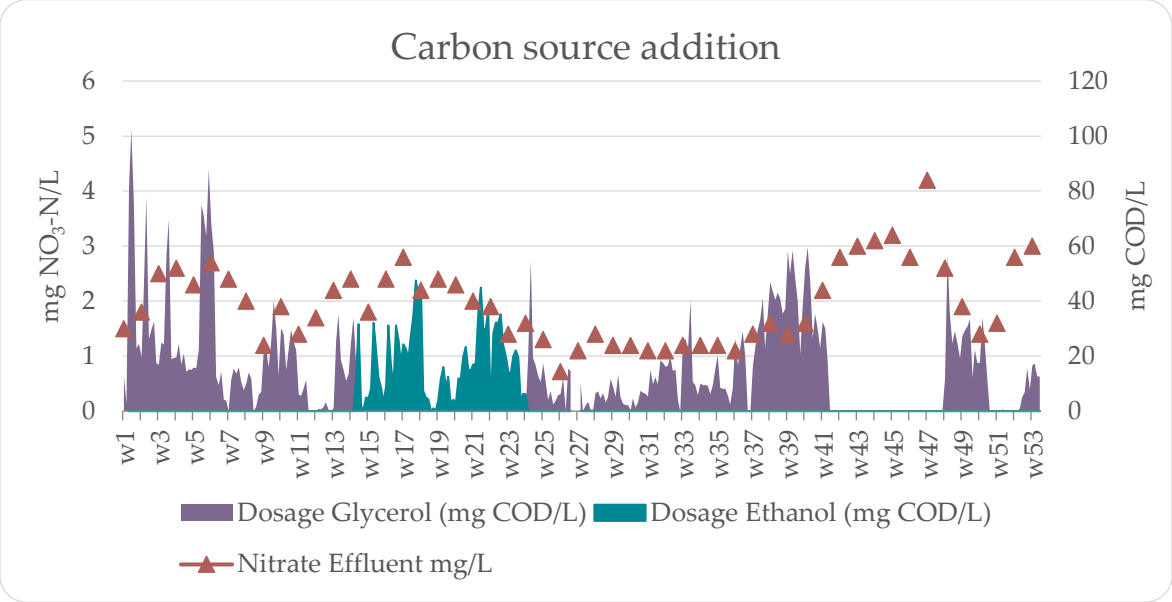


Figure 15. Dosage of glycerol and ethanol (mg COD/L incoming wastewater) as daily average values and effluent nitrate analysed in weekly composite samples.

The online sensors for nitrate in BR5deox and BR6 were helpful in the daily operation to monitor the denitrification. However, it was difficult to use the data for estimation of how much nitrogen was denitrified in the post-denitrification as the sensors tended to drift and the error in the reading was 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 16. The concentrations in BR5deox varied between 1.4 and 6.2 mg NO_3-N/L while the concentration in BR6 varied between 0.6 and 3.7 mg NO_3-N/L. The NO_3-N concentration in BR6 was similar to the effluent reading, sometimes higher and sometimes lower. Although effluent nitrate can be higher than in BR6 due to nitrification in the membranes, the difference is more likely due to sensor inaccuracy in BR6. Note that week 42 to 49, when there was no external carbon source addition, effluent nitrate was higher than normal, and higher than nitrate readings in BR6. This is also explained by faulty sensor readings in BR6.

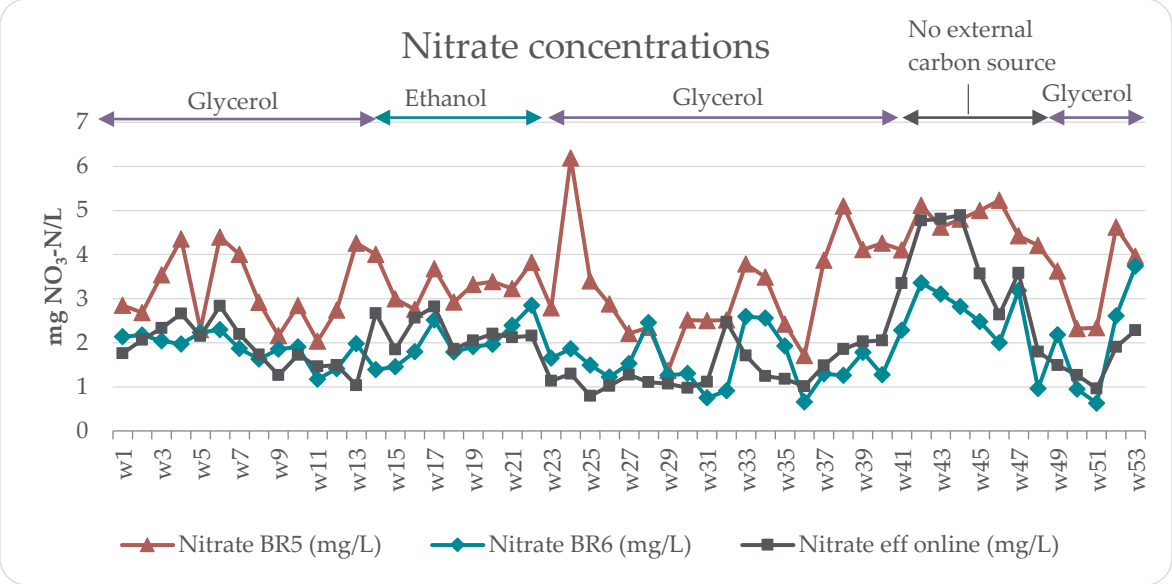


Figure 16. Online nitrate concentrations as weekly average.

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 at low cost. The carbon source storage facility for the full-scale treatment in Henriksdal will not be in finished during the first phases of operation of the full-scale plant and therefore a temporary installation will be needed at Henriksdal during a few years. With this background, glycerol has been tested as an alternative external carbon source in the pilot during 2020. A shorter test (10 weeks) was also done using ethanol as carbon source as part of evaluation of enhanced biological phosphorus removal.

Pilot results

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 has been made. Based on the online data of COD dosage and nitrogen removal (NO_3-N removal over the post-denitrification zone is labelled "g N-red", and TN-removal over the whole biological treatment is labelled "g N"), the specific consumption of carbon source as g COD/g N or g COD/g N-red was calculated. Results for glycerol and ethanol from 2020 are shown in Figure 17 and Table 15, previous results using other external carbon sources can be found in Andersson et al. (2021).

The theoretical COD-consumption for denitrification 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) and in cases when all oxygen is not removed in the deox zone some of the COD might also be oxidized. 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 17, the specific COD consumption was higher for ethanol compared to glycerol when only considering the nitrgen removed in the post-denitrification (2.91 compared to 1.84 g COD/g N-red). When relating the COD consumption to the total nitrogen removal the consumptions were similar for the two carbon sources.

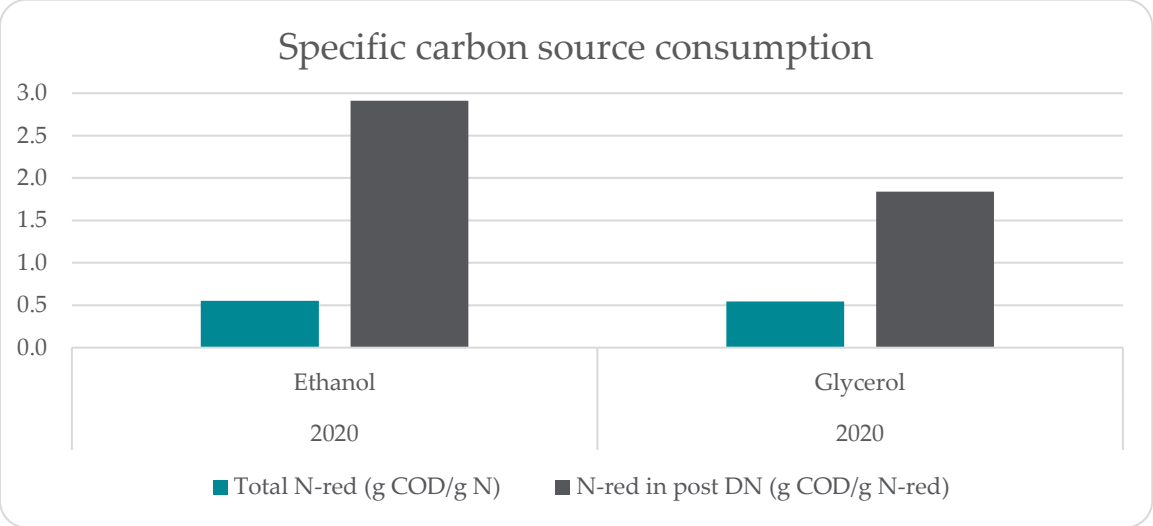


Figure 17. Comparison of COD consumption for ethanol and glycerol used in the pilot during 2020.

When looking at some of the operating conditions during the trials, both effluent nitrate and total nitrogen removal was lower when using glycerol (see Table 15). This is explained by a higher load (3.6 kg TN/d) entering the biological treatment in the period when ethanol was used compared to the period with glycerol (when the load was 3.2 kg TN/d).

A comparison between glycerol and ethanol in the pilot during February to April 2020 was done by Roberts (2020) and this evaluation also found the consumption of the two carbon sources to be comparable.

Table 15. Carbon source dosage and nitrogen removal for trials using ethanol and glycerol during 2020.

	Carbon source	Weeks	kg COD/d	kg NO_3-N red in post DN/d	Total kg N-red /d	Temperature biology (°C)	Effluent NO_3-N (mg/L)
2020	Ethanol	9	1.70	0.6	3.3	15.6	2.1
2020	Glycerol	22*	1.57	0.9	2.9	19.6	1.5

* w.41 to 48 excluded (dosage was paused).

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 (normalized to 20°C) for ethanol and glycerol are presented in Figure 18.

For glycerol, the denitrification rate in March 2020 was 2.4 mg NO_3-N/ g VSS, h. This is similar as in September 2019 when glycerol was tested and showed a denitrification rate of 2.1-2.4 mg NO_3-N/g VSS, h (Andersson et al., 2021).

For ethanol, the denitrification rate increased from 3.1 to 4.7 mg NO_3-N/ g VSS, h as the sludge adapted to ethanol. The first denitrification test (March 2020) was carried out before dosage of ethanol had started in the pilot and the second test (May 2020) was carried out after two months using ethanol in the pilot.

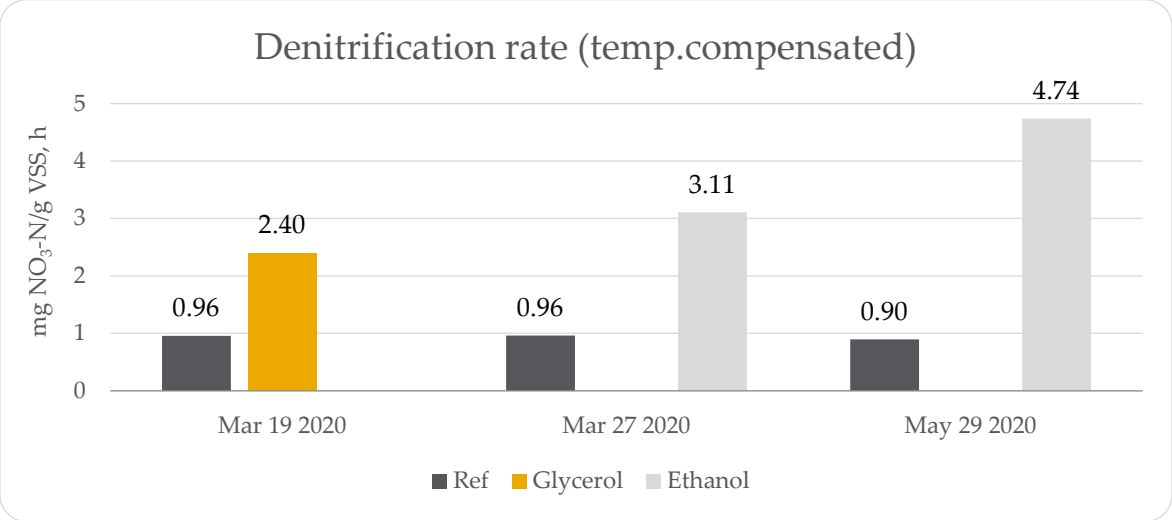


Figure 18. Denitrification rate during laboratory denitrification batch tests using glycerol and ethanol as carbon source.

The specific COD consumption during denitrification tests varied (Figure 19), especially for the reference test without external carbon source, and showed correlation to how denitrification was affected by pH. For methanol the COD-consumption was between 2.1 and 2.6 g COD/g NO_3-N denitrified. The COD consumption for glycerol was much higher, 6.6 g COD/g NO_3-N denitrified which is slightly higher than previous test in September 2019 when COD consumption was 6.2 g COD/g NO_3-N denitrified, which in turn is similar to reference values for glycerol (USEPA 2013). For ethanol, the COD consumption increased from first to second test when the sludge had adapted to ethanol. Similar results have been obtained for methanol in previous batch tests.

The specific COD-consumption in the lab test was significantly higher than in the pilot. This is most likely due to the intermittent dosing of external carbon source in the pilot that is controlled by the effluent nitrate concentration. The incoming nitrogen peaks in the morning and afternoon but goes down during nighttime. Thus, dosing of external carbon source is only active at daytime. However, at nighttime when no external carbon source is dosed, some post denitrification still takes place using organic carbon produced from hydrolysis of activated sludge. So, when calculating the specific COD-consumption by summarizing the dose of external organic carbon and comparing it to the amount of nitrogen denitrified in the post-denitrification zone, the ratio will be lower than the actual value.

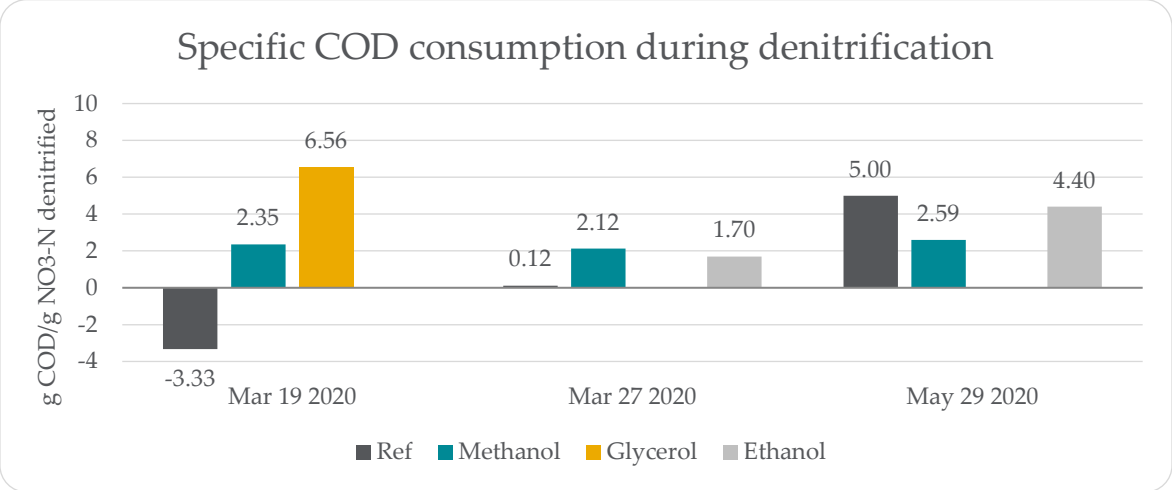


Figure 19. Specific COD consumption during laboratory denitrification batch test in laboratory tests using different carbon sources.

6.3 Phosphorus removal

Trial	J	F	M	A	M	J	J	A	S	O	N	D
Effluent phosphorus target 0.10 mg TP/L												
Aluminum chloride as precipitation chemical												
In-depth study of EBPR												

The goal of reaching a stable effluent phosphorus concentration below 0.15 mg P/L as monthly and annular averages was achieved previous years by using a control strategy with dosage of ferrous sulphate and ferric chloride in three points in the process; one flow proportional dose prior to pre-sedimentation and two dosage points in the biology (ferrous sulphate to BR4 and ferric chloride to BR6) controlled using feedback control from online effluent phosphate measurements. During 2019 the effluent target was lowered to 0.10 mg TP/L which was achieved using the same precipitation strategy but using lower phosphate setpoints. This year (2020) the goal of 0.10 mg TP/L was met using aluminum chloride (PAX, in operation from week 7) instead of ferric chloride (PIX) in the third dosing point. This was tested because the first MBR-line in Henriksdal will use precipitation chemicals from the existing storm water treatment system during the first years of operation as the PIX-tanks will not be built until a later stage of the project. The second dosage point of ferrous sulphate to the biology was taken out of operation to increase the aluminum chloride dosage and enable a better evaluation of potential effects on membranes and sludge quality of using an aluminum product in the process.

The phosphorous concentrations in and out from the biological treatment is presented in Table 16 and Figure 20 below. The yearly average effluent total phosphorus was 0.05 mg TP/L, the lowest since the project started, and well below the target of 0.10 mg TP/L. It should be noted that the inlet total phosphorus was also around 18% lower than normal. Total phosphorus in primary treated water has previous years varied between 5.3 and 5.5 mg TP/L as yearly average. In Henriksdal WWTP, the TP in the inflow was around 16% lower than previous years.

Table 16. Phosphorus concentrations in primary treated water (PTW) and effluent during 2020.

Parameter	Limit	Average	Min	Max	Nr of weekly samples
TP PTW (mg/L)	-	4.4	3.0	5.8	53
TP EFF (mg/L)	0.20	0.05	0.023	0.28	52

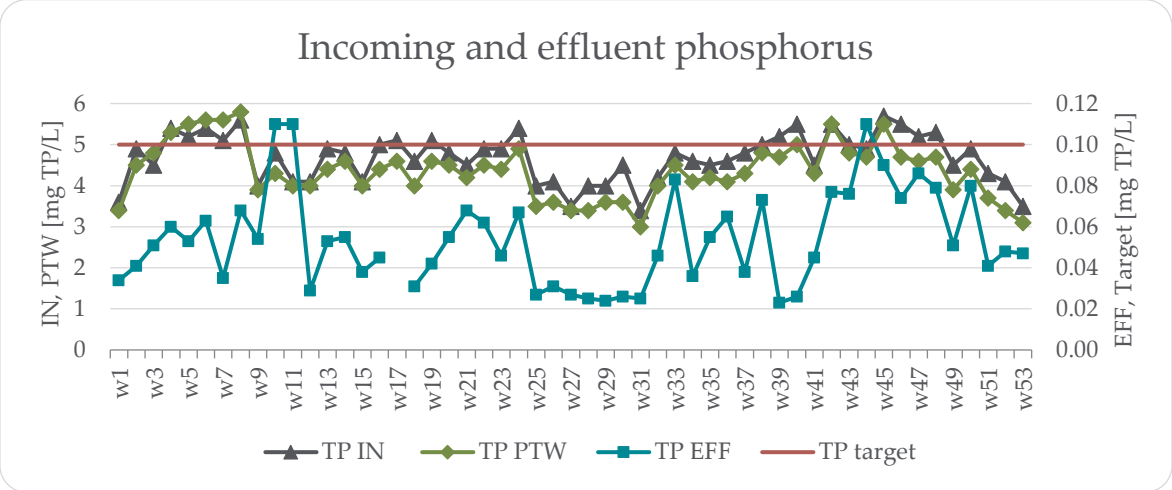


Figure 20. Influent and effluent total phosphorus analysed in weekly composite samples during 2020. Effluent data from week 17 is considered a measurement error and was excluded from calculations.

Key parameters for the phosphorus removal, both for the pilot and for the future Henriksdal design, are presented in Table 17. The phosphorus load in the pilot was much higher than the phosphorus load which the future Henriksdal design is based on. This resulted in higher phosphorus removal in the biology of the pilot in comparison to the future Henriksdal design. Despite the high phosphorus removal, the relative metal consumption in the pilot was lower in the pilot compared to the future Henriksdal design. Although the consumption of iron for precipitation of phosphorus was one of the most uncertain parameters in the future Henriksdal design, the difference is supported by the observed EBPR in the pilot. In the future Henriksdal 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 2020 was 7.1 mg Me/L (6.6 mg Fe/L + 0.5 mg Al/L).

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

Parameter	Unit	Value pilot 2020	Future H-dal design	Value pilot/scaled Future H-dal design[a]
Phosphorous load influent	kg P/d	0.49	2 594	127%
Phosphorous load PTW	kg P/d	0.46	1 580	194%
Phosphorus load reject water	kg P/d	0.01	480	16%
Total phosphorous load biology	kg P/d	0.47	2 060	153%
Phosphorus load effluent	kg P/d	0.0055	80	46%
Phosphorus removed in biology	kg P/d	0.46	1 980	157%
Iron consumption	kg Fe/d	0.7	10 000	47%
Aluminium consumption	kg Al/d	0.05	0	-
Metal consumption (Fe+Al)	kg Me/d	0.75	10 000	50%
Metal consumption per removed phosphorus	mole Me/ mole P	0.95	2.80	34%
Phosphorus in sludge	% of SS	3.3	5.4[b]	-
Iron in sludge	% of SS	4.1	-	-
Aluminium in sludge	% of SS	0.5	-	
VSS in sludge	% of SS	80	66	-

a) Future H-dal design divided by 6 700.
b) No EBPR. Mainly chemically bound phosphorous.

As can be seen in Table 17, the phosphorus fraction of the activated sludge (% of SS) was lower in the pilot than in the future Henriksdal design although the relative amount of phosphorus that was removed is larger. This can be explained by the high WAS production (high microbial cell growth) in the pilot, caused by high incoming load of SS and BOD. A high sludge production means that more phosphorus is bound in the sludge which gives a high removal of phosphorus per day even though the percentage in sludge is low.

The total amount of iron dosed is presented as daily average values in Figure 21. The base dose of Fe^{2+} was added to the pre-aeration which was controlled flow proportionally to a dose of 5-10 mg Fe/L (manually adjusted based on effluent phosphate). A supplementary dosage of Al^{3+} was added in BR6 (just prior to the membrane tanks). This dose was controlled by effluent phosphate feedback control and only added during shorter peaks in effluent phosphate. Although a higher aluminium dosage was intended, it was not needed due to very low effluent phosphate concentrations. An even lower dosage of Fe^{2+} to the pre-aeration would have risked the primary sludge production, and indirectly the ongoing trials in the sludge pilot, and was therefore not tested as an alternative to increase the aluminium dosage. The yearly average dosage of aluminium chloride was only 0.5 mg Al/L, and the highest weekly dosage was 3.8 mg Al/L.

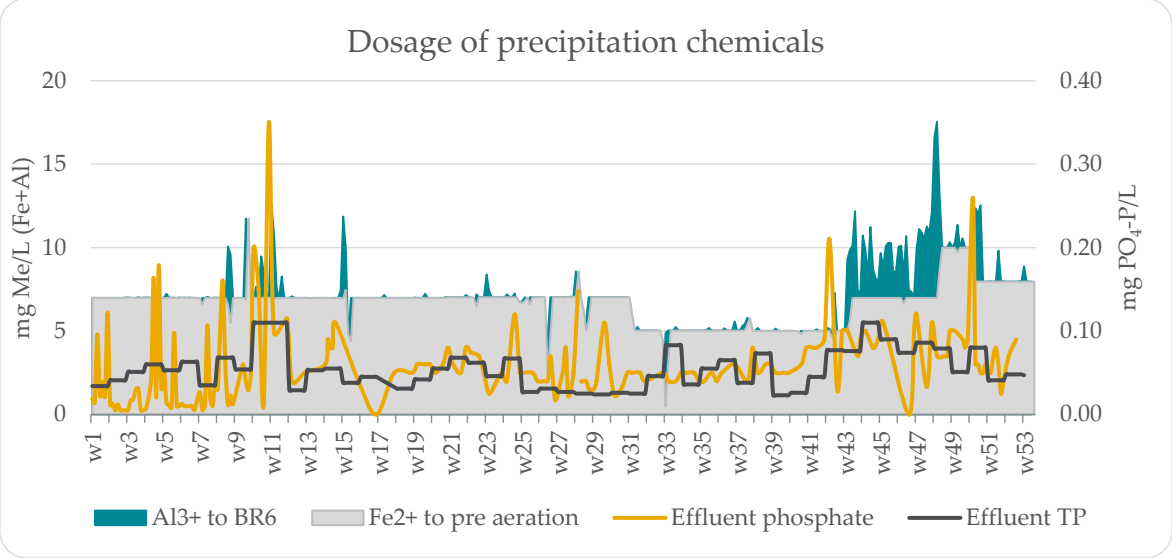


Figure 21. Metal dosage (the sum of Fe^{2+} to pre-aeration and Al^{3+} to BR6) as daily average, effluent phosphate analysed in daily composite samples and effluent total phosphorus in weekly composite samples. Note that the iron and aluminium are presented as a sum of added metals (g Me/d).

The iron, aluminium and phosphorus content of the sludge was monitored during 2020. The aluminium content in the sludge was mostly below 0.5% of TSS but increased to a maximum of 1.7% of TSS in week 48 after some weeks with slightly higher dosage of aluminium chloride (Figure 22). The iron content of the sludge, which varied between 2% and 7% of TSS, did not clearly correlate to iron dosage. The reason for this is most likely because ferrous creates an iron hydroxide buffer in the sludge (which has a sludge age of around 20 days) that delays its effect since phosphate can bind to the iron hydroxide at a later time point. Al^{3+} and Fe^{3+} rapidly forms stable molecules with phosphate ($AlPO_4$, $FePO_4$). This difference complicates the evaluation. The phosphorus content in the sludge was even throughout the year, around 3-4% of TSS, which is similar as previous years.

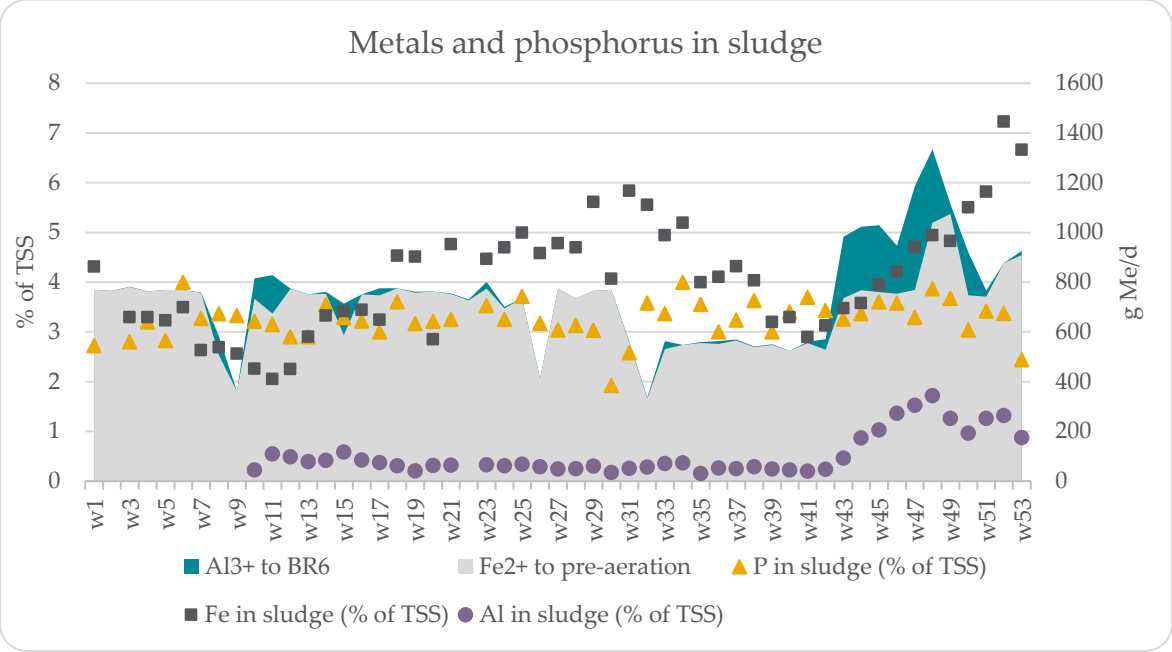


Figure 22. Iron, aluminium and phosphorous in sludge (weekly data). Note that the iron and aluminium are presented as a sum of added metals (g Me/d). The highest dose was in week 48 when 1334 g Me/d was added. Out of this, the majority was iron (1039 g Fe/d) and the remainder was aluminium (295 g Al/d).

Enhanced biological phosphorus removal (EBPR)

The treatment 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 from the pilot 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. This was also confirmed by monthly phosphate release tests (as described in Tykesson & la Cour Jansen, 2005) initiated in 2018. Further, the data on the molar ratio of iron to removed phosphorus from 2020, presented in Table 17, confirms that EBPR occurred in the pilot.

Results from the P-release tests from 2018-2020 are shown in Figure 23.

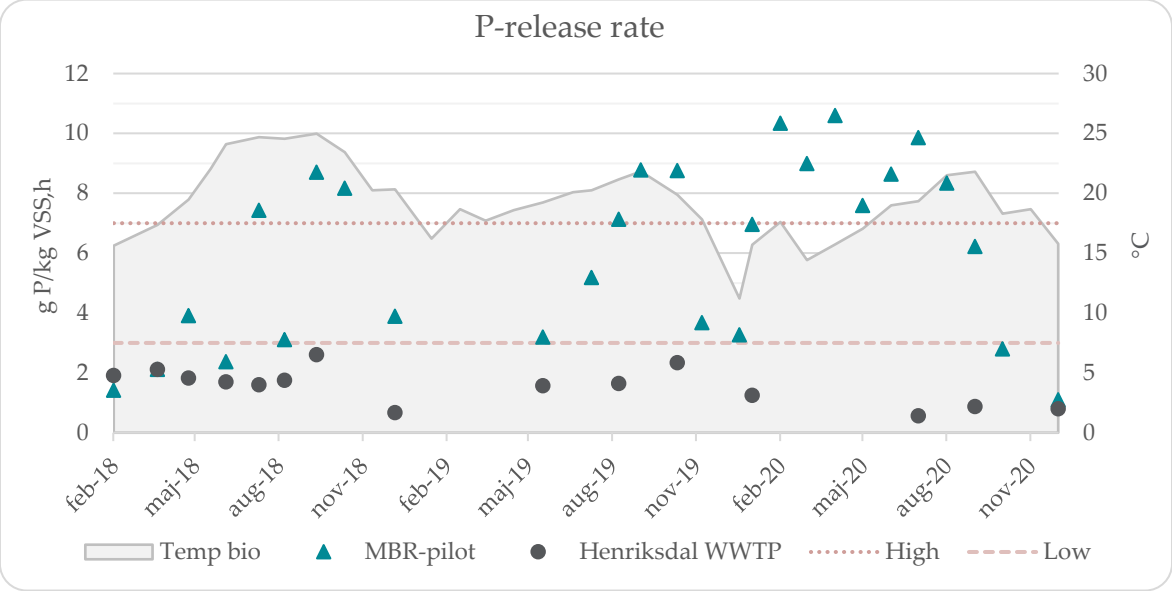


Figure 23. EBPR activity in the MBR-pilot and Henriksdal WWTP (CAS). Dotted lines show limits for EBPR-activity (Janssen et al., 2002), below the lower line = poor EBPR, above the higher line = high EBPR. Temp bio is the temperature in the treatment process.

During 2020, the EBPR activity was higher than previous years, with values on or above 7 g P/kg VSS, h from January to August. The P- release rates reached top values above 10 g P/kg VSS, h, which is almost in line with the well-established Swedish EBPR-plant Öresundsverket, 12,5 g P/kg VSS,h (Tykesson et al., 2005), although still significantly lower than 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 EBPR:

(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 does not contain an anaerobic zone in the water line. However, the DO and nitrate levels are often low with <0.2 mg O_2/L and <3 mg NO_3-N/L, respectively, in the anoxic pre- and post-denitrification zones. These zones are both followed by aerated zones, but the hydraulic retention time in the anoxic zones is shorter than 30 min. The COD:TP ratio in the inlet to the biological treatment (PTW) averaged at 81:1 in 2020, which should be enough for EBPR. In addition, external carbon source is added to the post-denitrification zone. The reject water from sludge dewatering contained 99 mg TP/L and 60 mg PO_4-P/L (yearly average), which is significantly higher than at Henriksdal WWTP (15 mg TP/L and 0.7 mg PO_4-P/L), indicating P-release in the digesters.

In Figure 24 a and b the influence of iron and aluminum dosage on the bio-P and P-removal is visualised. Combinations of EBPR and chemical precipitation have been shown to work well at Swedish WWTPs (Jansen et al., 2009). The phosphorus content in sludge was rather constant over time, between 3 and 4% of SS in WAS. The iron content in sludge was low during 2020 (2-5% of SS) and the modest dosing of PAX due to EBPR resulted in low levels of aluminum in the sludge (0-2% of SS). When EBPR activity was high, the dose of coagulant decreased, since it is controlled by the PO_4-analyser in the effluent. Therefore, a correlation between

Fe/Al in sludge, coagulant dosing and EBPR activity could be seen. It is, however, hard to determine the exact reason for the drop in EBPR activity from September to December, due to the combined effect of various parameters.

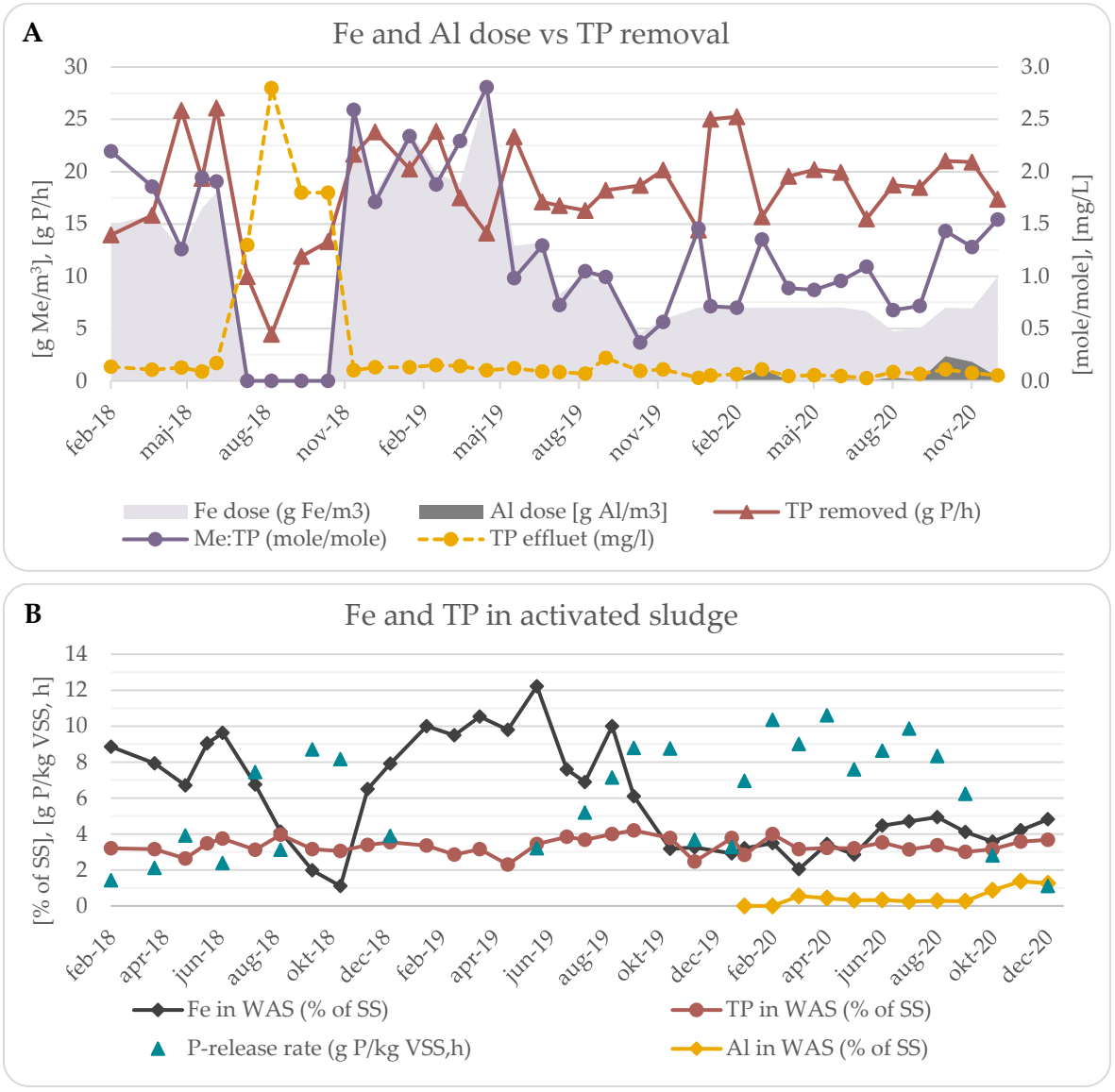


Figure 24. A) Fe- and Al-dose in activated sludge process and TP removed in the process. Me = Metal. B) Fe, Al and TP in waste activated sludge (WAS) together with P-release rate.

During spring 2020, two MSc thesis projects performed in cooperation with KTH, by Ross Roberts and Benjamin Fridh, studied the EBPR process in depth (Roberts, 2020; Fridh, 2020). It was concluded, based on phosphate profiling and Neisser staining (polyphosphate granules in poly-P accumulating organisms are stained blue and visualization in microscope according to method described in Fridh (2020), that the P-release mainly took place in the pre-denitrification zone (Figure 25 and Figure 26). The Neisser staining also indicated some release in the post-denitrification zone while this could not be seen in the PO_4-profilings. In a previous PO_4-profiling performed on one single occasion 7/6 2018, the largest P-release was seen in the post-denitrification zone (Andersson et al., 2020). The external carbon source used at the time was methanol and the nitrate concentration in and out of the post-denitrification zone was low at the time of sampling, 2.8 mg/L

and 0.8 mg/L respectively. Corresponding concentrations during the sampling period 2020 were around 4 mg/L and 2 mg/L while the nitrate concentration in the pre-denitrification zone was around 0.5 mg/L. Although the studies by Roberts and Fridh both pointed to the pre-denitrification zone for P-release, it is possible that both the pre- and post-denitrification zones act as the "anaerobic" zone in the EBPR-process depending on other factors such as nitrate concentration, available VFA, recirculation of oxygen from RAS-deox and deox zone etc.

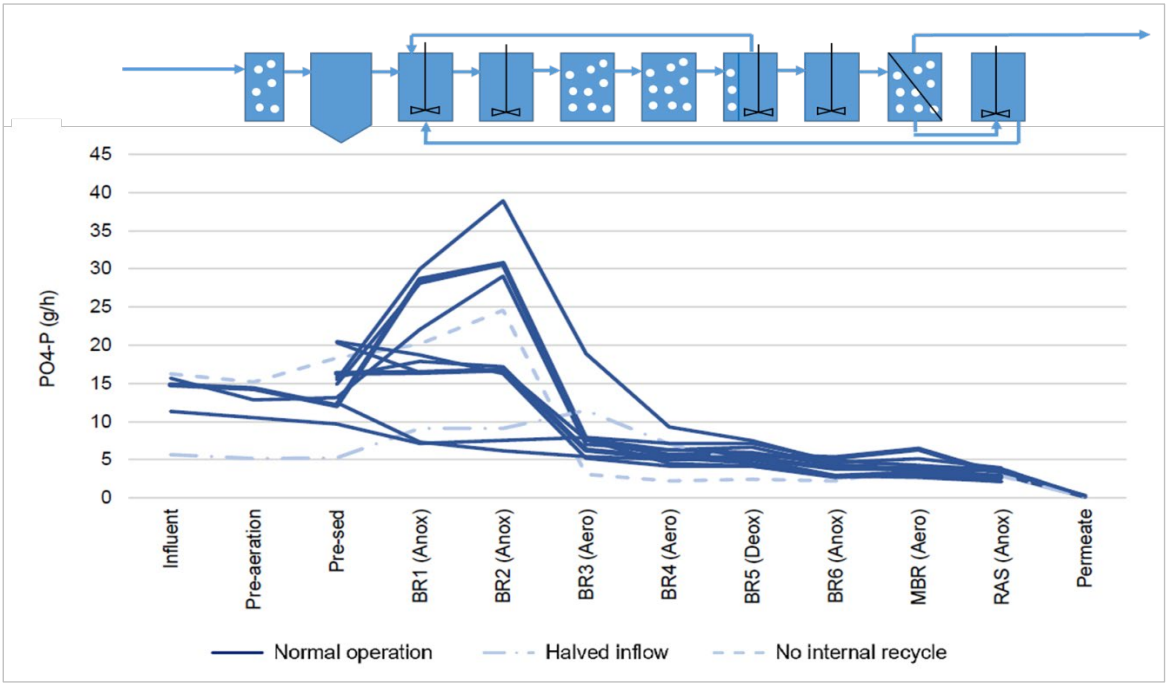
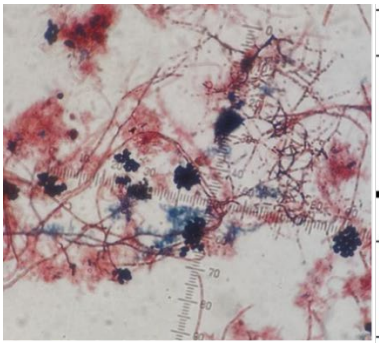


Figure 25. Phosphate profile in the water line. Y-axis show the mass flow of phosphate as grams PO_4-P per hour. Since the total flow is different in different zones the concentrations cannot be compared. From Roberts 2020.



Bioreactor	BR2	BR3	BR4	BR5	BR6
Count	123	126	252	205	153
	114	127	234	230	220
	156	122	268	239	215
Average	131	125	251	225	196
Average Concentration (Cluster ml-1)	1.31×10^6	1.25×10^6	2.51×10^6	2.25×10^6	19.6×10^6

Figure 26. Left, a micrograph of sludge from ox zone where poly-P granules are stained with Neisser´s blue and microorganisms are counterstained with Neutral red. Right, the number of poly-P clusters in different zones. From Fridh (2020).

P-release tests in the presence of nitrate showed P-release rates in the same range as the reference without nitrate when acetate was added in excess, dose of 400 mg/L acetate as soluble COD. However, if the addition of acetate was limited (20-40 mg/L sCOD) in the presence of nitrate, no P-release was measured. Results are shown in Table 18. The conclusion was that P-release in the MBR-pilot is achieved in the presence of nitrate

and that sufficiently high VFA-concentrations is provided by the incoming wastewater or by hydrolysis in the pre-denitrification tanks.

Table 18. P-release rates in presence of nitrate, with different acetate addition (Roberts 2020).

Carbon source	Dose mg COD/L	Temp. °C	pH	Start sCOD mgCOD/L	Start VFA mg H_3COO^-/L	Start NO_3-N mg/L	VSSg /L	Max PO_4-P mg/L	1st P-release rate mg P/g VSS·h
Acetate	400	19.9	7.8	459	326	26.8	5.9	56	8.0 (anox)
Acetate	400	20.0	7.4	450	324	<0.5	5.9	60	9.0
Acetate	400	20.2	7.9	384	333	39.0	6.2	54	7.8 (anox)
Acetate	400	20.3	7.3	380	332	0.7	6.2	60	8.7
Acetate	20	20.4	7.1	84	<50	29.4	5.8	0.9	0.1(anox)
Acetate	20	20.3	7.4	84	<50	3.9	5.8	7.6	0.1
Acetate	40	20.3	7.3	96	<50	27.3	5.8	3.2	0.3 (anox)
-	0	20.4	7.2	86	<50	4.1	5.7	7.8	0.1

The effect of using different external carbon sources as substrate for the P-release test and as a carbon source for the post-denitrification in the MBR-pilot was also studied. Acetate gave the highest P-release rates with an average of 8.6 g P/kg VSS, h. A VFA-slurry, produced by fermentation of primary sludge and food waste, followed at 6.3 g P/kg VSS,h (Owusu-Agyeman et al. 2020). Brenntaplus, a mixture of proteins, sugars and alcohols, gave a P-release rate of 5.0 g P/kg VSS, h. Ethanol, methanol and glycerol resulted in P-release rates insignificantly higher than the negative reference (no carbon addition), 0.1-1.3 g P/kg VSS, h compared to 0.1 g P/kg VSS, h. The external carbon source added to the post-denitrification zone was changed from glycerol to ethanol for a few months to see if it affected the EBPR. However, no effect could be detected in the pilot line or when performing P-release tests in the lab. To improve the EBPR-activity in the line, a VFA-based carbon source could be used instead of glycerol or methanol, either in the pre- or post-denitrification zone.

In addition, the influence of temperature on the P-release rates was investigated in the range of 15-25°C with the conclusion that the initial P-release rate was higher at higher temperatures whereas the highest overall P-release was achieved at lower temperatures.

During spring 2020 samples were also collected regularly for DNA-analysis of the PAO community. DNA-extraction and PCR were carried out but because of the COVID-19 situation, the samples could not be sequenced during this study. This is planned in the future.

Details about the two studies and additional results are presented in the MSc-thesis reports (Roberts 2020; Fridh 2020).

6.4 BOD reduction

Analysis on BOD_7 from daily composite samples have, since the 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 2020 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												
Offgas measurement												
Extended membrane operational cycle												

During 2020 the membranes were operated in extended cycles with 15 minutes of permeation, followed by 1 minute of relaxation (standard is 10 + 1 minute). 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. This control strategy means that the membrane flux will be controlled (but varying with inflow), and the transmembrane pressure (TMP) will be allowed to vary.

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

During 2020 the inflow to the pilot was higher than average and permeate recycling, which normally is used as the membranes area is slightly too large, was therefore not applied.

Table 18 list the main events and experiments regarding the membrane operation during 2020.

Table 19. Main events and experiments regarding the membrane operation.

Event	Time period (2020)
Recovery cleaning (RC) with sodium hypochlorite	Week 11 and 50
Recovery cleaning (RC) with acids (oxalic and citric)	Week 15 and 51
Maintenance cleaning (MC): Hypo replaced with water	Week 13 to 23
No hypo MC	Week 24 to 37
Hypo MC initiated based on fouling control algorithm	Week 42 continuing
Operation with prolonged permeation cycle (15 min instead of 10 min, capacity increased by 3%)	Throughout 2020
Operation with forced Leap-Lo (low membrane aeration)	Week 1 to 43

6.5.1 Permeability

Generally, a permeability above 200 L/(m²·h·bar) is considered as good according to the supplier. As can be seen in Figure 27, the permeability was above 200 L/(m²·h·bar) throughout most of the year 2020 for both membranes. The permeability was greatly affected by the different cleaning strategies tested this year, see details in section 6.5.3 Membrane cleaning.

Recovery cleaning was carried out twice during 2020 with a clear increase in permeability after cleaning. Prior to the first RCs, permeability was higher for MT1 (cleaned with oxalic acid) compared to MT2 (cleaned with citric acid). After the first RCs, both membranes had similar permeability. From week 15 a decreasing trend can be seen for both membranes and the lowest permeability was obtained in week 37 after 13 weeks of

operation without use of sodium hypochlorite for maintenance cleaning. To restore permeability, the hypo cleanings were started again in week 37. The maintenance cleanings using acid did have a small positive effect on permeability, which can be seen as a biweekly increase in permeability (Figure 27).

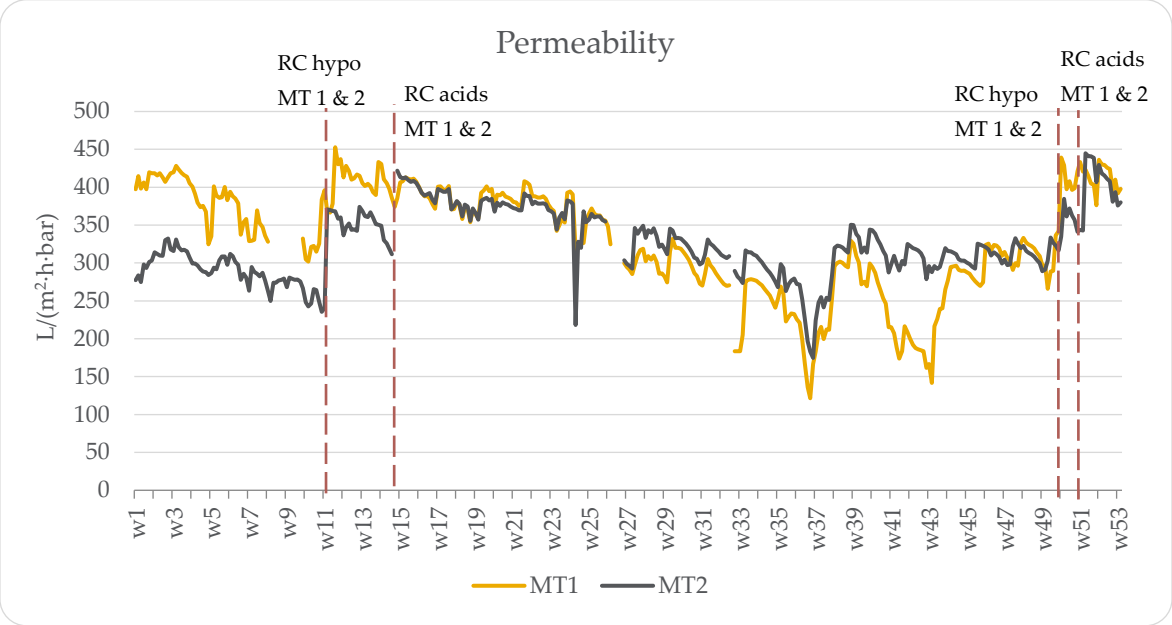


Figure 27. Permeability (temperature compensated) for membrane 1 (MT1) and 2 (MT2) during project year 7 (2020). Recovery cleanings (RCs) were carried out with hypo in week 11 and week 50 and with acids in week 15 and 51 for both membranes. MT1 was cleaned with oxalic acid, MT2 was cleaned with citric acid.

During 2020, the membrane aeration was forced to the lower air flow rate (see section 6.5.4 Membrane aeration). No negative effects on membrane permeability were observed. The decrease in permeability could be related to aeration, but the reduced cleanings are considered the major factor affecting membrane permeability.

There was a rapid decrease in permeability for MT1 from week 39 to week 43, but no clear explanation to this has been found. Previously, the general trend has been that MT1 has performed better than MT2 although it has been cleaned using less chemicals. In week 43 the oxalic acid settings were changed to increase the amount of acid used during each MC (from 1040 mg/L to 1300 mg/L which is the design concentration).

During 2020, aluminium chloride was dosed in BR6. The highest dosage was from week 43 to 48 when 2-4 mg Al/L inflow was added. No negative effect on the permeability was observed when adding aluminium chloride.

6.5.2 Flux and TMP

Fluxes for the two membranes are presented in Figure 28. Normally the membranes were operated with net flux around 21 to 25 L/(m²·h). Design net flux for future Henriksdal is 20.9 L/(m²·h), and design max net flux is 30 L/(m²·h). If the flux was lower than design, one of the membranes would enter standby mode. During 2020 the flux was controlled proportional to the inflow. This year the inflow to the pilot was kept constant at 4.54 m³/h and was increased to 5.5 m³/h (design max flow) during rain events. The even flux with shorter peaks reflects the inflow flowrates. As the inflow was constantly higher than design, membranes rarely entered standby mode (with exception for downtime).

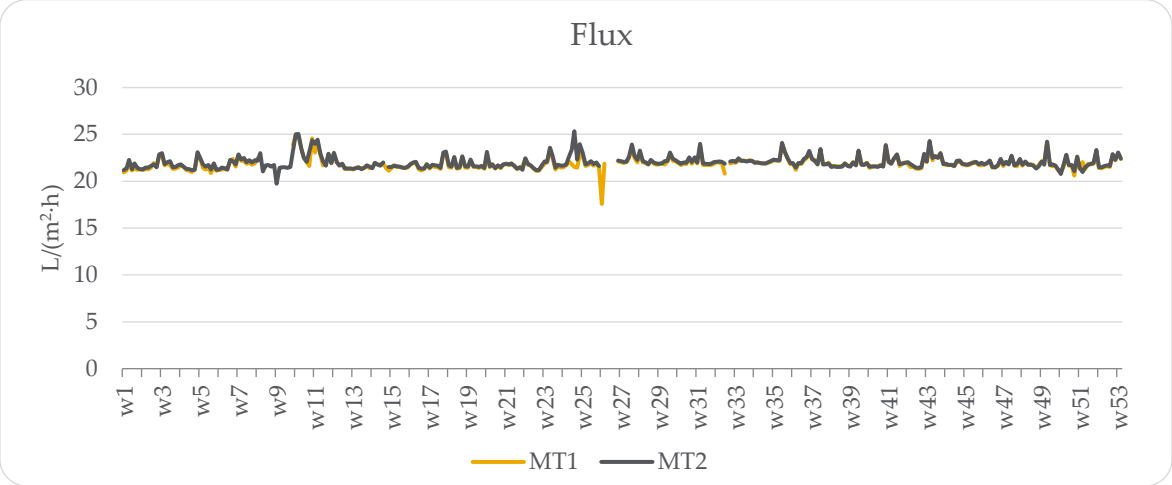


Figure 28. Net flux during 2020.

The transmembrane pressure (TMP) is presented in Figure 29. Based on daily average data, net TMP varied between 51 and 177 mbar for MT1 and between 56 and 128 mbar for MT2 during 2020. TMP is reduced after recovery cleaning (RC) with hypo (w11 and w50). A smaller decrease in TMP can be observed for MT2 in week 15 when RC with citric acid was carried out. The peaks in TMP around week 37 is related to the trial with only acid MCs which was ended in week 37 and TMP decreased after MC with hypo. MT1 show higher TMP than MT2 from week 39 with a peak week 43. During this period the membranes were operated with different amount of acids for MCs and membrane aeration was forced to the lower air flow (Leap-Lo) until week 43. The higher TMP for MT1 resulted in lower permeability and extra MCs were manually initiated in week 43. Other abrupt changes in TMP are often related to the normal maintenance cleaning events.

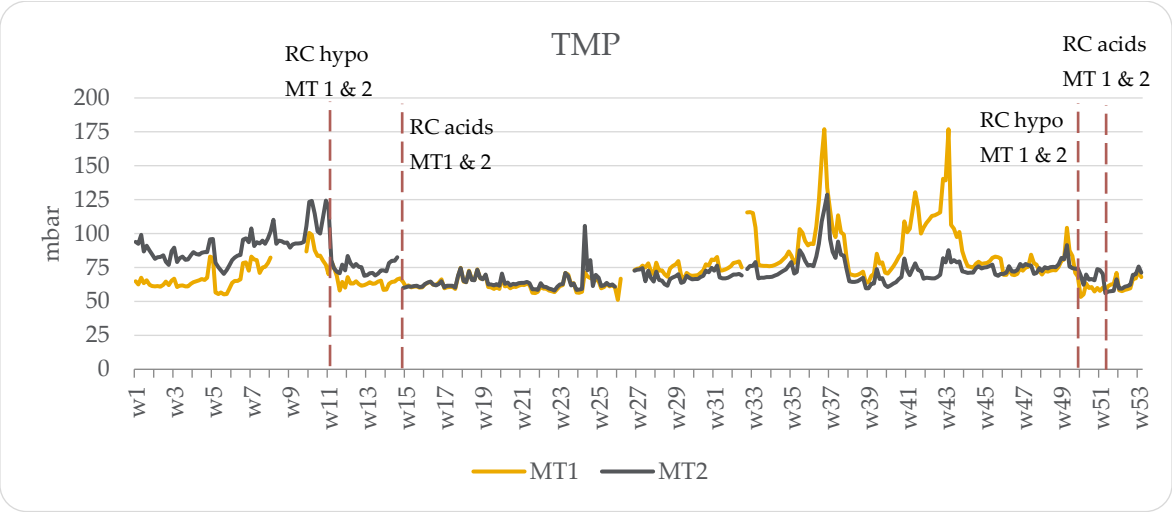


Figure 29. Net TMP during 2020. Recovery cleanings (RCs) were carried out with hypo in week 11 and week 50 and with acids in week 15 and 51 for both membranes. MT1 was cleaned with oxalic acid, MT2 was cleaned with citric acid.

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 (MCs) were automatically carried out on a weekly basis. 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 backpulse was reduced from 5 minutes to 2 minutes and later the number of backpulses were reduced from 9 (incl. the first longer one) to 7 in total. In 2018, further reduction of oxalic acid usage was done by increasing the interval in-between cleaning events. In 2019 both oxalic and citric acid usage was optimised and a synchronization with the hypo MC was tested. During 2020 main focus has been on reducing the amount of hypo used for MC.

The operational settings have been divided into separate trial periods. An overview of the trials regarding the MCs are presented in Table 20.

Table 20. Overview of trials regarding the maintenance cleaning (MC) of the membranes.

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 acid cleaning with reduced chemicals, oxalic acid as previous period
T10	Oct 2019	Oxalic acid 80% chemical flow, 7 BP, every 4th Hypo MC Citric acid 100% chemical flow, 7 BP, every 4th Hypo MC

T11	Feb 2020	Oxalic acid 80% chemical flow, 7 BP, every 4th Hypo MC Citric acid 100% chemical flow, 9 BP, every 4th Hypo MC
T12	Mar 2020	Hypo MC replaced with water MC
T13	June 2020	Hypo MC excluded until permeability is below a low limit of <150 L/(m²·h·bar). Acid MCs carried out with half standard interval (once every other week)
T14	Sept 2020	Recovery period
T15	Oct 2020	Fouling-based trigger for hypo MC Acid MCs carried out with half standard interval (once every other week)

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 30 together with the permeability for the trials during 2020.

Trial 10 (T10) started in October 2019 where MC with acid was carried out after every 4th hypo MC for both MTs. This trial continued until February 2020. As citric acid consumption was reduced compared to the design, the permeability for MT2 decreased compared to MT1. For T10 the average permeability for MT1 was 345 L/(m²·h·bar) and 277 L/(m²·h·bar) for MT2. Based on these results, it was decided to return to design settings for citric acid MC (but still performed at half design interval) from February 2020 (T11). During T11, recovery cleaning (RC) of the membranes were carried out in preparation for the next trial.

In March 2020, the sodium hypochlorite was replaced with water, trial 12 (T12) to see the effect of the backpulse alone without addition of any chemicals. As permeability remained stable for more than two months with this approach it was tested to completely exclude the hypo MCs (T13) and monitor for how long the membranes could operate with only acid MCs before the permeability decreased below 150 L/(m²·h·bar). This trial started on June 9th (w24) and continued for three months (until September 9th, w37). The membranes were operated without hypo MC for 92 days and permeability decreased from maximum 394 L/(m²·h·bar) to minimum 121 L/(m²·h·bar) for MT and from 383 to 175 L/(m²·h·bar) for MT2. After a short recovery period with standard cleaning schedule (T14) a new trial started in October 2020 (T15) where the hypo MCs were initiated based in indication of membrane fouling. The same algorithm that controls the membrane aeration (called fouling control, see section 6.5.4 Membrane aeration) was used.

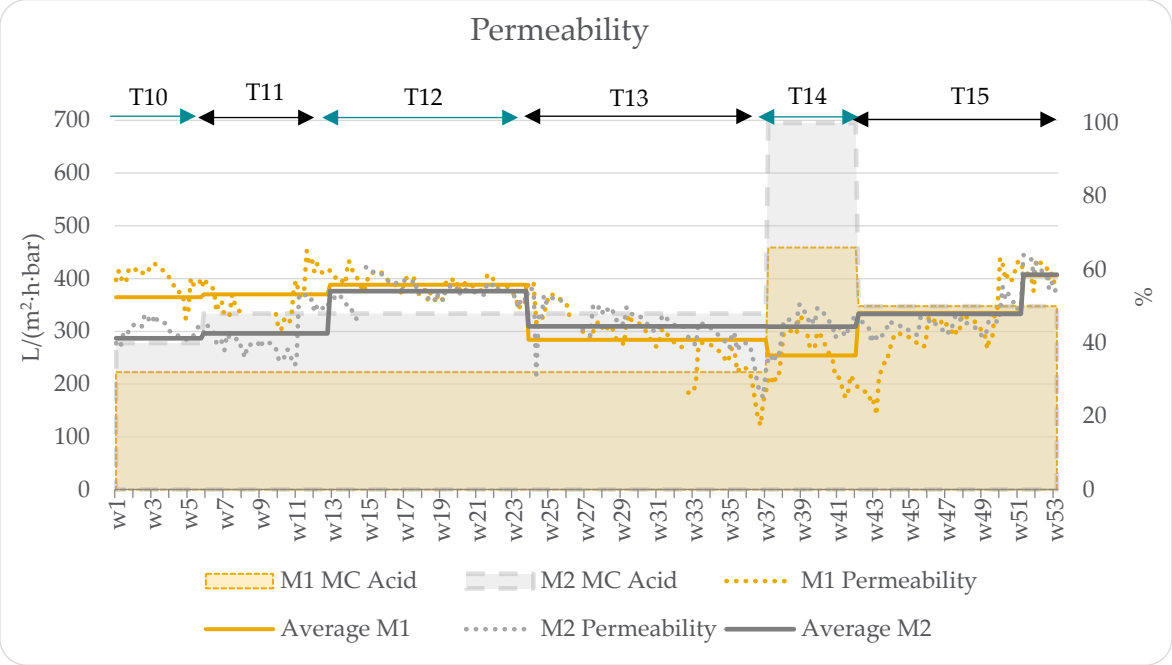


Figure 30. Membrane permeability and amount of acid used for MC, normalized to back pulse duration of 2 minutes + 8 x 30 seconds carried out with interval of 345 m³ of permeate produced. M1 was cleaned with oxalic acid, M2 was cleaned with citric acid. T10 – T15 are trial periods. T10 started in October 2019.

Maintenance cleaning with sodium hypochlorite has been carried out with interval according to supplier, but with reduced backpulses; 2 min initial pumping followed by 6 x 30 seconds until week 13. From week 13 to week 37 no hypo was used. In the five week long recovery period (T14), four MCs with hypo + acid were carried out for each MT. During T15 (week 42 to 52), when hypo was initiated by the fouling control algorithm, 4 hypo MCs were carried out for MT1 and 1 hypo MC for MT2 (all carried out in October). During November and December no hypo MCs were needed, but recovery cleaning was carried out by the end of the year increasing the permeability for both membranes. The trial with fouling-based control of hypo MCs will continue during 2021. The results so far indicate that hypo MC is required to maintain good permeability, but the intervals can be considerable reduced from standard interval of about 2 MCs with hypo per MT per week. Trial 13 (88 days), with no hypo MCs and half interval of acid MCs showed a slow reduction of permeability followed by more sudden drops in permeability (especially for MT1) but permeability was only below 200 L/(m²·h·bar) for 9 days (MT1) and 1 day (MT2) and easily recovered by initiating standard cleaning.

Recovery cleaning

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

According to the membrane supplier, RC should be carried out twice every year with both sodium hypochlorite and acid. However, as the permeability has been good, RC has normally only been performed once per year. Last year (2019), RCs were carried out in March and focus was on measuring the chlorine gas and chloramines emitted to surrounding air when the membranes were soaked in sodium hypochlorite. This year RC was performed in March/April and in December. RC with hypochlorite was done 1-4 weeks before RC with acid on both occasions.

When comparing the effect of the recovery cleanings on permeability, the RCs resulted in similar final permeability in both MTs both times (Figure 31). After the acid RCs the permeability was around 400-450

L/(m²·h·bar). Generally, oxalic acid RCs resulted in less improvement of permeability compared to the citric acid RCs. This can be explained by a higher permeability (and thus cleaner membranes) before the acid RC for MT1 compared to MT2.

For MT1, RC with hypo in March followed by MC with oxalic acid (carried out 15th of March), led to a total increase in permeability from 315 to 453 L/(m²·h·bar). The following RC with oxalic acid did not improve permeability more than an MC with oxalic acid (compare increase after oxalic MC on the 31st of March to oxalic RC on 7th of April). For MT2 the hypo RC in March plus a citric acid MC (carried out on the 24th of March) resulted in a total increase from 235 to 373 L/(m²·h·bar), however the citric acid MC contribution was small.

The decrease in permeability between the hypo RCs and the acid RCs in March/April indicate that a higher final permeability after acid RCs could be achieved if hypo and acid RCs were carried out closer in time. This was observed after the RCs carried out in December when permeability after acid RC reached a maximum of around 450 L/(m²·h·bar) during the week after acid RCs. However, this was not better compared to the result after hypo RC followed by acid MCs in March.

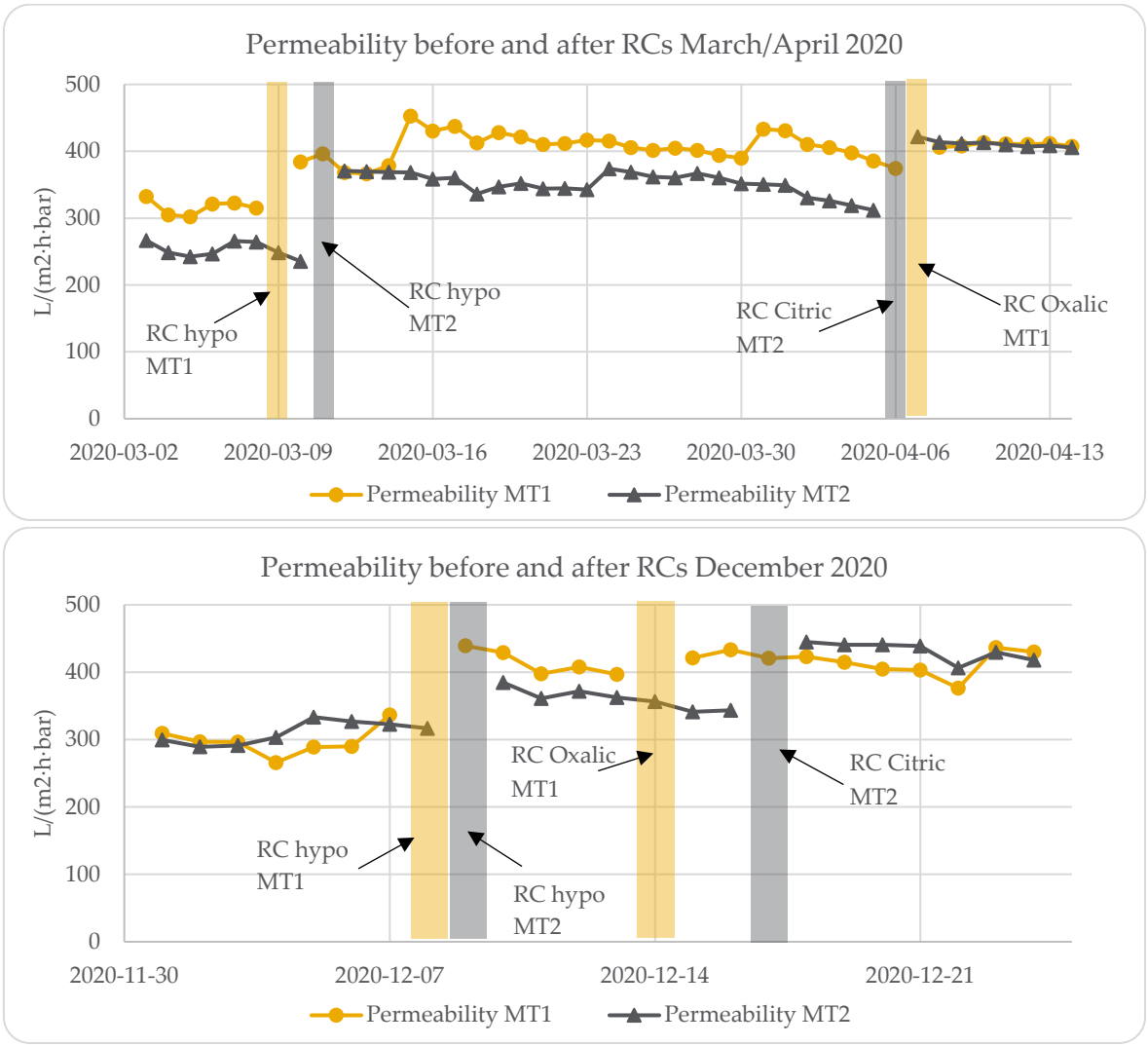


Figure 31. Permeability before and after recovery cleaning (RC) with sodium hypochlorite (hypo) and acids (oxalic acid and citric acid for MT1 and MT2, respectively) in March/April 2020 (upper) and December 2020 (lower).

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 designed to be 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.

By the end of 2019 and throughout most of 2020 the membrane aeration was forced to the Leap-Lo aeration flowrate (Figure 32). This was tested to evaluate the effects of low aeration on membrane permeability. Instead, the fouling control algorithm was used to trigger MC with hypo (see section 6.5.3 Membrane cleaning). Leap-Hi aeration was manually selected for MT1 during 6 days in October (w43) to improve

permeability. After this period, the fouling control algorithm was again activated to control membrane aeration. No automatic usage of Leap-Hi occurred during the remaining part of 2020.

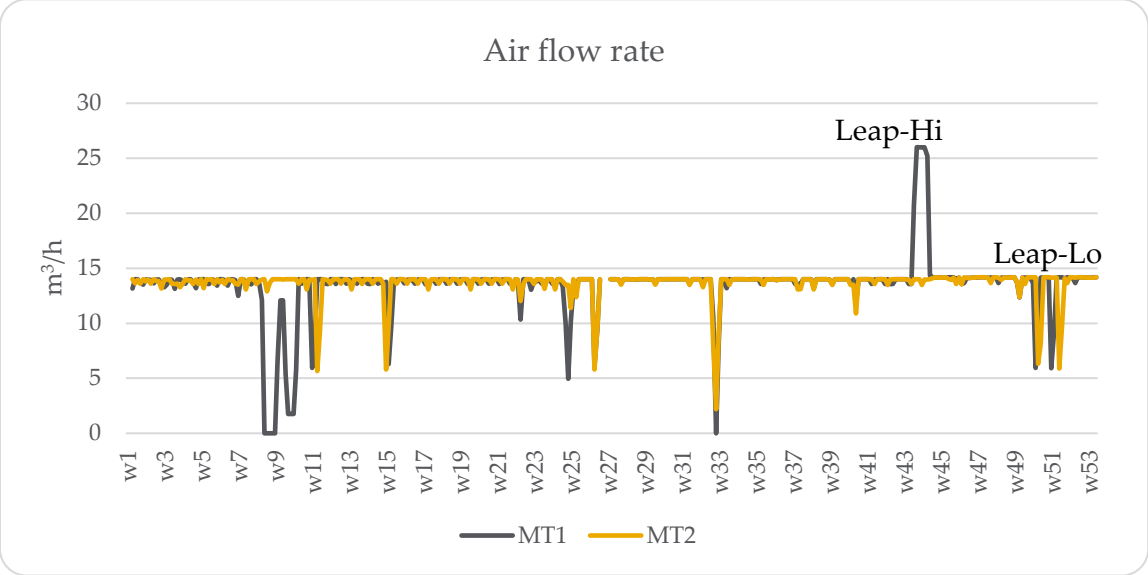


Figure 32. Membrane aeration as daily average during 2020. Leap-Hi corresponds to 26 m³/h and Leap-Lo corresponds to 14 m³/h.

6.5.5 cTOC and membrane performance

Two separate experiments focusing on factors effecting the collodial Total Organic Carbon (cTOC) were performed in 2020. One experiment investigated how carbon sources and reject water effect the cTOC. The second experiment, the effect of cTOC on the membrane performance in terms of transmembrane pressure (TMP). According to the membrane supplier SUEZ, the cTOC concentration should be kept <10 mg/L to not affect the membrane performance. However, it is not exactly known how much the variation in cTOC affects the membranes in terms of transmembrane pressure (TMP). To gain good knowledge of the cTOC and thus potential adverse effects on the membrane performance, the effect of additions that presumably affect the cTOC, such as external carbon source and reject water, needs to be understood.

Effects of carbon sources and reject water on cTOC

During 2019 and 2020, the cTOC concentration in the pilot increased up to 25 mg/L (Figure 33). It was observed that the cTOC concentration was higher from September 2019 until April 2020. During this time, two crucial changes in the operational parameters were made. First, the carbon source was changed from methanol to glycerol, and second, reject water from the sludge dewatering was added to the pre-aeration tank (Figure 33). When both addition of reject water was stopped and the carbon source was changed to ethanol (March 2020), there was a clear decrease in the cTOC concentration.

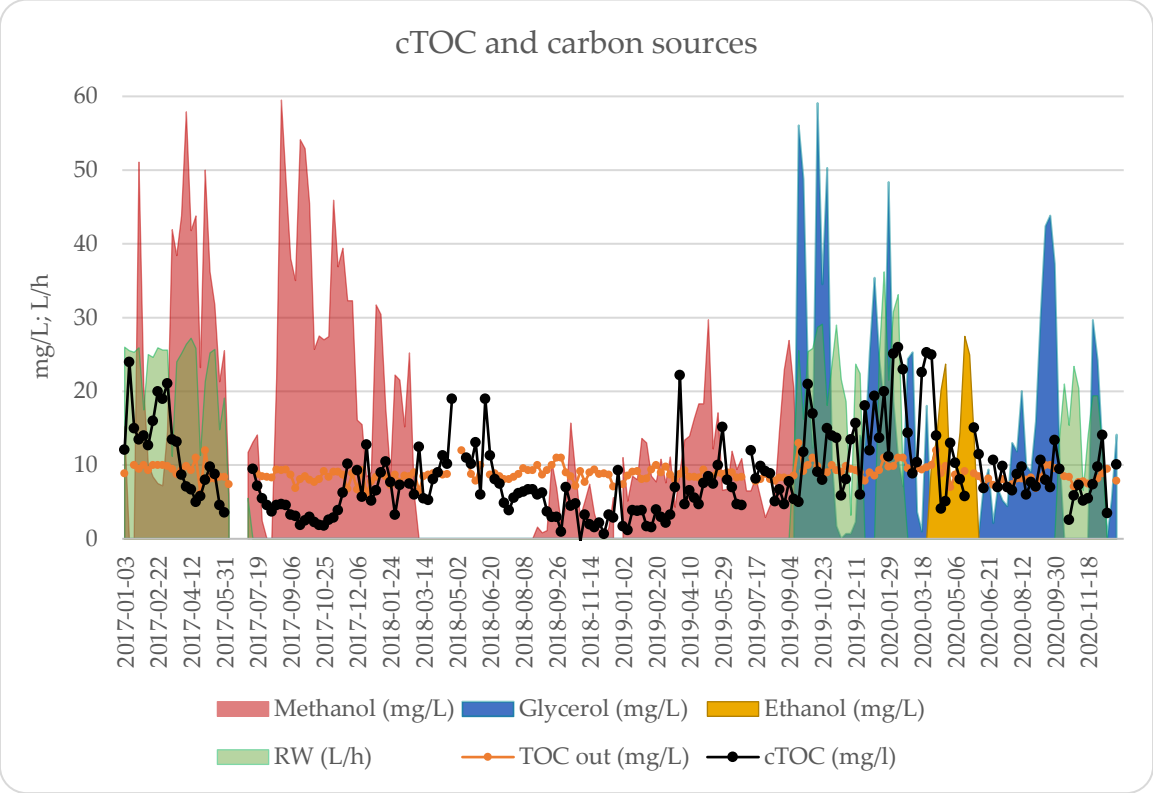


Figure 33. Summary of effluent TOC, dosing of external carbon sources and recycling of reject water from digested sludge dewatering (RW) during 2020.

In order to investigate if the increase in cTOC was related to the addition of glycerol or reject water, a first trial T1 without the addition of reject water (w24 to 41) and a second trial T2 without addition of external carbon source (w41 to 48) were carried out. After these trials (week 48 onwards), the process was operated with both, reject water and carbon source, with only two interruptions due to operational problems in the dewatering unit (trial 3, T3). A limitation is that it is unknown if the cTOC concentrations might increase by accumulation during longer periods of time, which may affect the outcomes of the trials. Furthermore, the values shown are weekly averages, which could minimize the impact of high and low doses on the real cTOC values.

Following the historical data and the results from the tests performed (Figure 34), no clear correlation between the addition of glycerol or reject water and the increase in the cTOC concentration could be found. During the first trial, where glycerol was the only carbon source and the reject water recirculation was off, the cTOC concentration exceeded 11 mg/L just one time, on the 9th of September. This, even at high glycerol addition (up to 44 mg/L), which demonstrates that the addition of glycerol alone, does not contribute to the cTOC increase.

During the second trial period, the cTOC concentration decreased from an average of 8.7 mg/L, in T1, to an average of 5.3 mg/L in trial 2 (Figure 34).

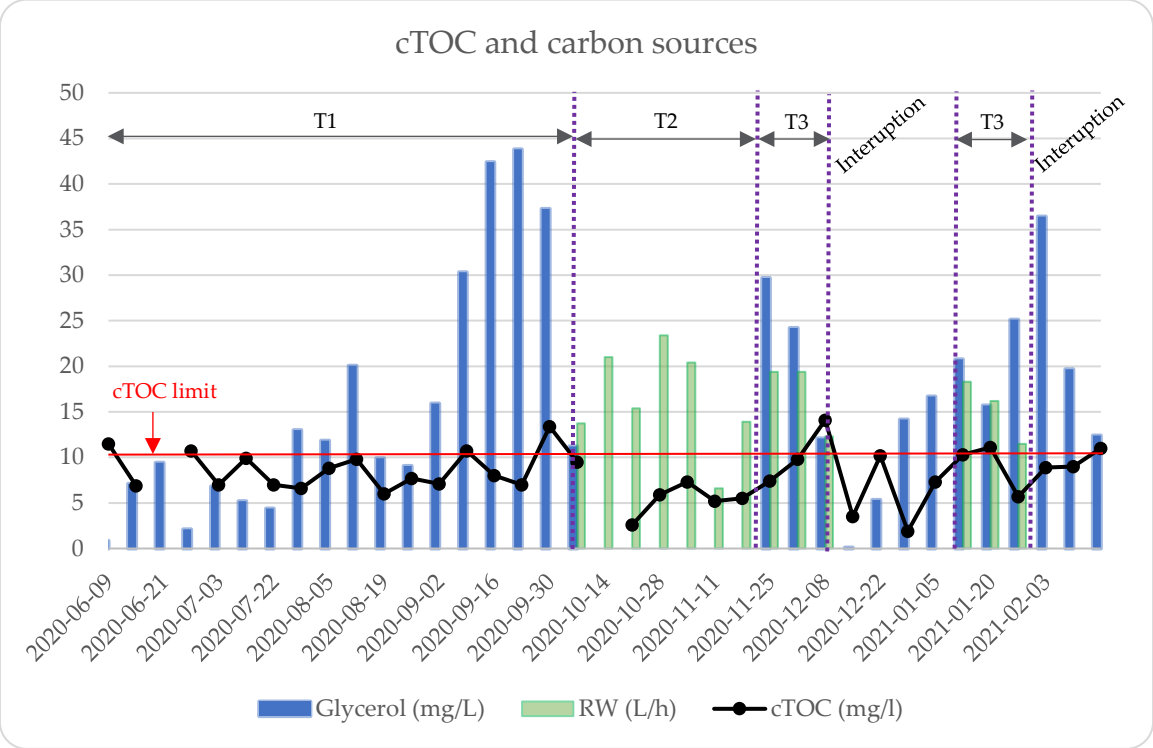


Figure 34. cTOC variations during the trial periods T1, T2, T3, and interruptions

In week 48, both addition of glycerol and reject water started. However, due to operational problems, the reject water could be recirculated only for 3 weeks. An increase in cTOC during this period can be observed in Figure 34; values increased from 5.5 mg/L to 14.1 mg/L in three weeks. The following four weeks had a stop in the reject water and a low but increasing glycerol dose with oscillating cTOC values. Finally, during the last three weeks, with addition of glycerol and reject water, there was an increase followed by a decrease in cTOC.

There is no clear increase in the cTOC concentration when glycerol or reject water were added separately to the process. However, it is suspected that the addition of both glycerol and reject water might affect the concentration of cTOC. This needs to be investigated over a more extended period of time with the addition of both and with shorter interruptions. Furthermore, if a new trial is performed, it is advised to take samples more than once per week.

Effect of cTOC on membrane performance

Generally, the two membrane cassettes perform differently due to their operational parameters, including use of cleaning chemicals and cleaning schedules, and therefore, the evaluation of the effect of cTOC on membrane performance is not possible with the available data. In order to investigate if there is any correlation between the membrane fouling (measured indirectly by permeability and TMP) in MT1 and MT2 respectively and cTOC, the two membranes had to be studied separately. The membrane permeates from the two MTs was therefore collected by two different samplers. By separating the permeate flows, the data collection allowed investigating the difference in the effluent TOC concentration and cTOC from the two membranes (oxalic and citric acid) and if the variations of cTOC affect the permeability. The colloidal particles measured by cTOC is thought to easily get stuck in the membrane pores and cause fouling, but it is not known if different cleaning chemicals restore this type of fouling to an equal extent. Since both MTs are fed with identical sludge with the same amount of colloidal particles, a lower value of TOC in the effluent (and subsequently a lower cTOC) from one of them

mean that more cTOC is deposited on the membrane surface of that MT. Weekly sludge samples and permeate samples were collected and analysed for TTF, TOC in, TOC out, and cTOC. The trial was performed fro 2.5 weeks between 2020-06-22 and 2020-07-10.

The results are presented in Table 21. Unfortunately, the collected data is insufficient to draw clear conclusions and cTOC concentrations were too low during the trial period (<10 mg/L) to observed clear effects. More data and perhaps a better analysis method is needed to evaluate the effect of cTOC on the performance of the membranes. It is further suggested that future trials include a measuring campaign with daily sampling to obtain results of TTF and cTOC. Furthermore, cTOC should be >10 mg/L during such a campaign if possible.

Table 21. Data from the trial where the effect of cTOC on membrane performance was investigated.

Date	Membrane	SS (mg/L)	TTF-100 (s)	TTF-100 Normalized (s)	TOC Filtrate (mg/L)	TOC Permeate (mg/L)	cTOC (mg/L)	TMP (mbar)	Permeability (lmh)
2020-06-22	MT1	9855	312.5	317	15	8.1	6.9	66.1	347.8
2020-06-22	MT2	9855	321	326	15	7.7	7.3	66.6	349.8
2020-07-01	MT1	9459	318.5	337	14	6.2	7.8	79.0	293.9
2020-07-01	MT2	9459	326	345	14	7.4	6.6	78.2	298.0
2020-07-06	MT1	8615	395.5	459	12	6.6	5.4	77.4	305.1
2020-07-06	MT2	8615	418.5	486	14	6.8	7.2	72.0	331.9
2020-07-08	MT1	9164	344	375	14	5.3	8.7	83.8	299.8
2020-07-08	MT2	9164	363.5	397	13	5.6	7.4	76.7	331.0
2020-07-10	MT1	10186	355.5	349	16	7.4	8.6	77.6	299.8
2020-07-10	MT2	10186	367.5	361	14	7.6	6.4	69.6	334.3

6.6 Offgas measurements during NaOCl RC

It is known that sodium hypochlorite used for water treatment gives rise to the formation of chloramines, which are emitted into the air (Thicket, 2002; Jacobs, 2007). Health effects ranges from increased problems for people suffering from asthma, to irritated airways, to chronic problems. There is no occupational limit value in Sweden, but the World Health Organisation (WHO) has suggested a reference value of 0.5 mg/m^3 for trichloramine (WHO, 2006). The Nordic Expert Group – a Nordic collaboration for production of criteria documents on chemicals for occupational exposure limits – has suggested a threshold limit value (time-weighted for 8-hour workday) of 0.1 mg/m^3 for trichloramine (Wastensson, 2019).

In March 2020, a RC was performed (one year since last RC) although permeability was good (around 250-300 L/(m2·h·bar)) to get more results on the chlorine gas and chloramines emitted to air during RC with hypo. From previous testing in March 2019, it had been determined that chlorine gas and chloramines were formed in the pilot during RC with sodium hypochlorite. For the RC in March 2020, a test was set up for sampling in a ventilation channel connected to the hood covering the membrane tank (Figure 35). In addition, the sampling procedure was further developed based on last years' results and supplemented with analysis of other volatile chlorinated organic compounds. Sampling was started at the same time as addition of sodium hypochlorite to the membrane tank was started. Filling of the tank took about 45 minutes and once filled the aeration was turned on for 5 minutes to mix the solution.

The air flow in the exhaust was measured using a vane anemometer (Testo 417) and a funnel (Testovent 417), before and after the sampling was done, and the flow was determined to be approximately 150 m^3/h.

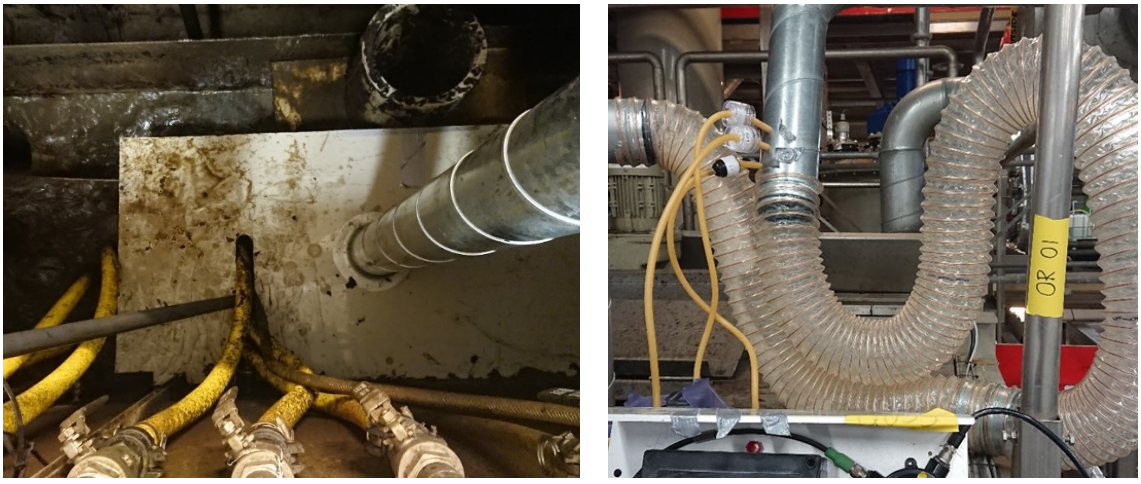


Figure 35. Left: Gas-sampling hood and ventilation channel placed over a membrane tank. Right: Ventilation channel and example of sampling setup for volatile chlorinated compounds.

Chloramine samples (samplers supplied by the Department of Public Health and Clinical Medicine, Umeå University) and chlorine samples (SKC 225-9006 samplers) were collected in series during a 7-hour time period after addition of hypochlorite was started (see Table 22). Sampling of volatile organic compounds (VOC) using Tenax samplers was added to the testing, to screen for possible formation of chlorinated organic compounds. The latter samples were only collected during the first 2 hours after addition of hypochlorite was started (see Table 22). Sampleswere collected using SKC AirChek TOUCH air sampling pumps.

Preferably sampling would have been done for a longer period of time than 7 hours, but the sampling technique had limitations due to the capacity of the samplers and consideration had to be taken to working day limitations.

Table 22. Sampling setup of chlorinated compounds during RC with hypo in March 2020.

Analyte	Sampling period	Flow rate (l/min)
Chloramine	0–0.5 h	2
	0.5–1 h	2
	1–1.5 h	2
	1.5–2 h	2
	2–3 h	2
	3–5 h	1
	5–7 h	1
Chlorine gas	0–1 h	1
	1–2 h	1
	2–4 h	1
	4–7 h	1
Chlorinated VOC	0–0.5 h	0.1
	0.5–1 h	0.1
	1–2 h	0.1

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 and VOC samplers were analysed by thermal desorption, gas chromatography and mass spectroscopy.

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.

As the membranes were operated with low maintenance cleaning during 2020, which resulted in lower permeability towards the end of the year, and to repeat the measurement campaign on gas emissions during RC with hypo, it was decided to do an additional set of RCs in December 2020. Unfortunately, the gas emission measurement campaign could not be conducted this time due to various practical reasons but is planned to be repeated in 2021.

The RCs were carried out first with sodium hypochlorite and then with acids (oxalic acid for MT1 and citric acid for MT2) some weeks later.

The schedule for cleanings can be seen in Table 23 together with the amount of chemicals used and conditions at start and end of the soaking. Higher starting values during 2020-12-09 to 2020-12-10 may be due to pipes being flushed from the previous RC and thus more recent sodium hypochlorite reaching the tank?

Table 23. 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
2020-03-09 to 2020-03-10	MT1	Sodium hypochlorite (66 g/L)	19.0 L	pH 9.3 Cl_2 895 mg/L	18.5 h	pH 8.5 Cl_2 490 mg/L
2020-03-11 to 2020-03-12	MT2	Sodium hypochlorite (74.5 g/L)	16.8 L	pH 9.2 Cl_2 730 mg/L	21 h	pH 8.0 Cl_2 425 mg/L
2020-04-06 to 2020-04-07	MT2	Citric acid (51%)	4.3 L	pH 2.8	20 h	pH 2.9
2020-04-07 to 2020-04-08	MT1	Oxalic acid (8%)	18.9 L	pH 2.3	19.5 h	pH 2.3
2020-12-08 to 2020-12-09	MT1	Sodium hypochlorite (53.5 g/L)	23.9 L	pH 9.4 Cl_2 600 mg/L	20.5 h	pH 8.2 Cl_2 430 mg/L
2020-12-09 to 2020-12-10	MT2	Sodium hypochlorite (53.5 g/L)	22.9 L	pH 9.4 Cl_2 755 mg/L	20.5 h	pH 8.0 Cl_2 375 mg/L
2020-12-14 to 2020-12-15	MT1	Oxalic acid (8%)	20.0 L	pH 2.2	21 h	pH 2.1
2020-12-17 to 2020-12-18	MT2	Citric acid (51%)	4.5 L	pH 2.8	19.5 h	pH 2.8

Results from the measurement of volatile chlorinated compounds are presented in Table 24 and Figure 36 to Figure 38. The VOC samples showed that the most abundant compound was by far chloroform. Other identified compounds were 1,1-dichloroethen, methylene chloride, carbon tetrachloride, dibromochloromethane and tetrachloroethene. The latter were detected only in trace amounts so only data for chloroform is presented.

Table 24. Total amounts and peak concentration of volatile chlorinated compounds emitted during recovery cleaning.

Analyte	Total emission (mg)	Timespan (h)	Peak concentration (mg/m³)	Peak occurrence (h)	Limit value (mg/m³)
Chloramine	6200	7	9.1	1-1.5	0.1(1)
Chlorine gas	900	7	1	1-4	1.5(2)
Chloroform	200	2	1	1-2	10(3)

(1) Suggested threshold limit value (time-weighted for 8-hour workday)
(2) Short-term limit value (15 minutes exposure)
(3) Threshold limit value (time-weighted for 8-hour workday)

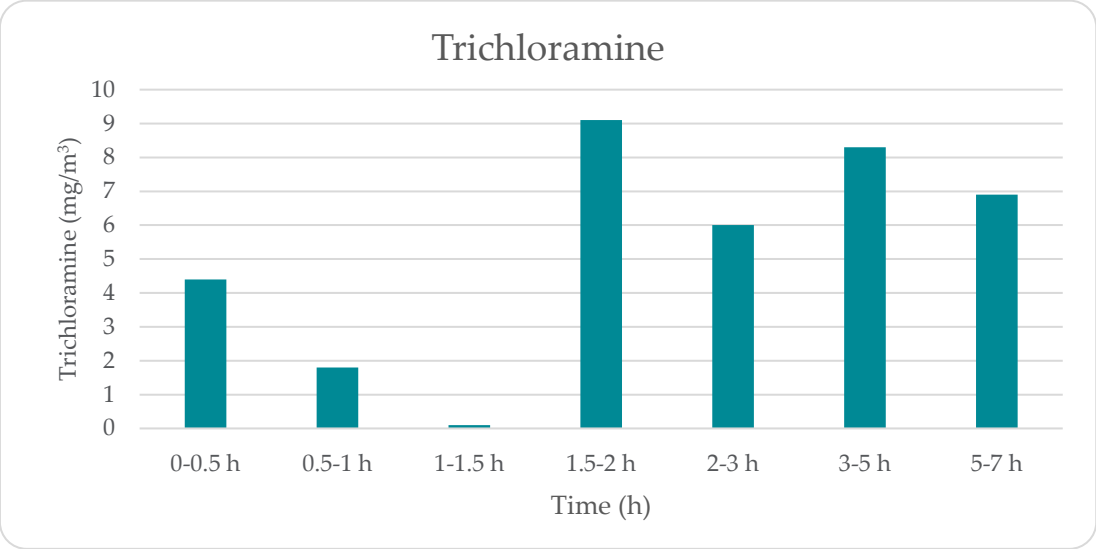


Figure 36. Emission of trichloramine during a 7-hour period, while cleaning wastewater treatment membranes with sodium hypochlorite.

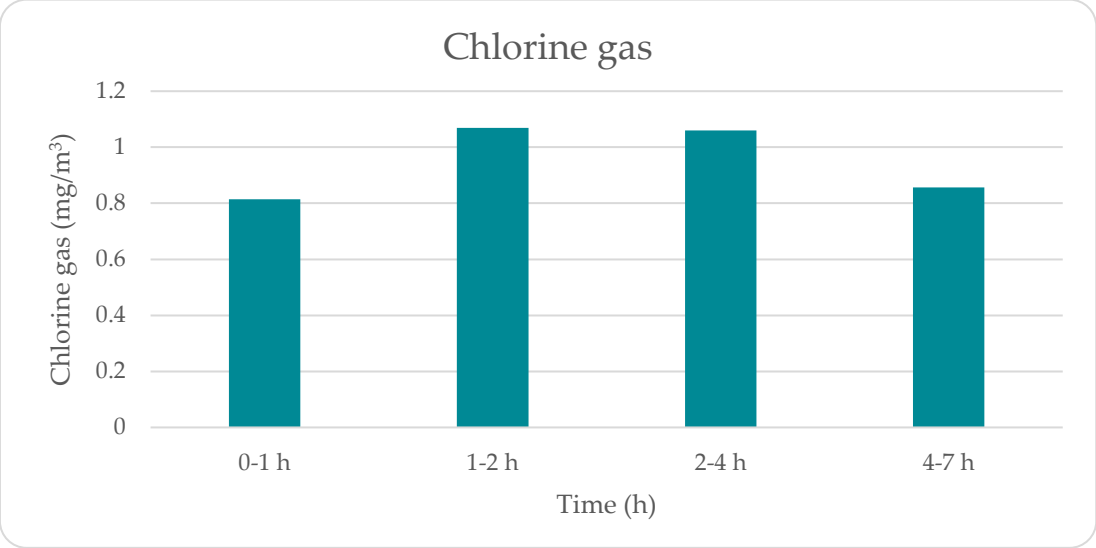


Figure 37. Emission of chlorine gas during a 7-hour period, while cleaning wastewater treatment membranes with sodium hypochlorite.

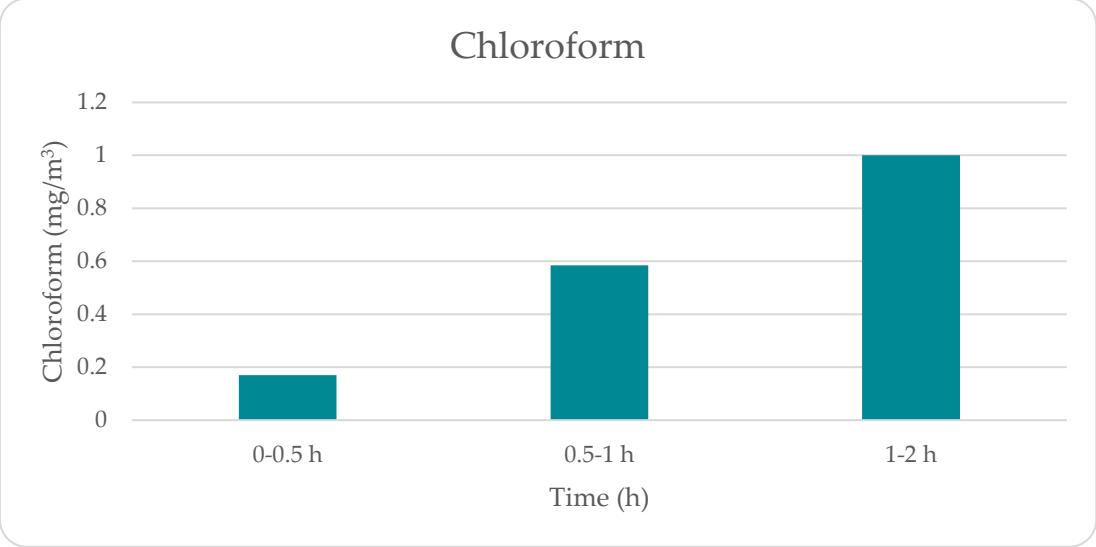


Figure 38. Emission of chloroform during a 2-hour period, while cleaning wastewater treatment membranes with sodium hypochlorite.

As can be seen in Figure 36 to Figure 38, the emission process was slower than expected, and the total amounts of compounds presented in Table 24 are an underestimation as the measuring period of 7 hours most likely was too short. The data works well to give a rough estimate of the amounts emitted, and the results can be used as a source for further calculations of occupational exposure levels that can occur, when scaling up to a full-size wastewater treatment plant. Even though sampling methods do not follow standards for measuring occupational exposure, the peak concentrations can be compared to the occupational limit values in Sweden (Arbetsmiljöverket, 2018), or in the case of trichloramine, recommended exposure limit as a reference. Chloramine peaked at 90 times the recommended limit, chlorine gas at 70% of the short-term exposure limit (15 min exposure) and chloroform at 10% of the occupational exposure limit (8-hour workday average).

As the intended longer measurement campaign planned for the December RCs was cancelled, it is the intention to do another measurement campaign of volatile chlorinated compounds during RC next year.

6.7 Sludge production and sludge properties

In Table 25, data from the pilot related to sludge production and sludge quality is compared to data from the Henriksdal WWTP (annual average 2020) and design data for the future Henriksdal WWTP according to the SFA project. The differences in sludge production between the pilot and Henriksdal WWTP are caused by the configuration of the primary treatment, which was described in previous reports together with the effect on the subsequent treatment steps (Andersson et al., 2019, 2020, 2021).

Table 25. Sludge data from the pilot year 2020 compared to data from the Henriksdal WWTP 2020 and design data for the future Henriksdal WWTP.

Parameter	Pilot data 2020	Henriksdal WWTP data 2020	Design future Henriksdal
WAS production (kg SS/d)	15.9	22 100	59 000
Part of total sludge production (%)	42%	27%	34%
VSS in WAS (% of SS)	80	68	63
Fe in WAS (% of SS)	4.1%	11	-
Al in WAS (% of SS)	0.5%	-	-
PS-production (kg TS/d)	21.7	61 000	117 000
Part of total sludge production (%)	58%	73%	66%
VS in PS (% of TS)	88	79	80
Total sludge production (kg TS/d)	37.6	83 100	176 000
Total sludge age, SRTtot (d)	17.5	14.5	28
Aerated sludge age, SRTox (d)*	6.5**	5.7	7***
SVI jan-jun (mL/g)	330	141	-
SVI jul-dec (mL/g)	346	117	-

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

**including membrane tanks, without membrane tanks SRTox = 4.8 d

***including membrane tanks, yearly average

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 normalized permeability. The coagulant dose (Fe^{2+} and Al^{3+}) was rather constant over the year until week 43 when the dose went up significantly. Consequently, the metal content in the sludge started to increase slowly. At the same time the permeability increases but that is most likely due to membrane cleaning and not the metal content in the sludge. In theory, high values on Time-to-filter, TTF (described in chapter 5.3.2), and colloidal TOC (cTOC) should correlate with a decrease in the permeability (Gkotsis and Zouboulis, 2019). This could not be seen in the process data. The cTOC value should be below 10 mg/L to ensure minimal fouling and values above 20 mg/l indicates poor sludge filterability (Gkotsis and Zouboulis, 2019). In 2020, the cTOC was above 10 mg/L during 26 out of 51 weeks and above 20 mg/L during 7 weeks, which is worse than previous years. TTF-100 should be lower than 100 s for high filterability while values above 300 s indicates poor filterability (Gkotsis and Zouboulis, 2019). In 2020 the TTF-100, normalized or not, was never under 100 s. TTF-100 and normalized TTF was under 200 s on 1 and 2 timepoints respectively and between 200-300 s on 10 and 11 timepoints out of 51. The highest values were 855 and 1015 s for TTF-100 and normalized TTF. In 2019 the values were below 100 s from week 1 to 32 when the TTF started to increase steadily. This years results indicate very poor filterability of the sludge. Still, the permeability was good throughout the year regardless of high TTF and cTOC and the fact that the membrane cleaning scheme was decreased (see chapter 6.5.3). In the beginning of the year, high values for both TTF and cTOC were observed. From week 17, both TTF and cTOC dropped and from week 41 onwards there was a drop in cTOC values but not in TTF. All in all, only a weak correlation between TTF and cTOC was found (R^2=0.38). Previous years a correlation between TTF-100 and Fe in sludge as well as Fe-dose could

be seen. This year no such correlation existed. This could be due to the high EBPR activity (chapter 6.3) and subsequent low dose of Fe^{2+} and Al^{3+} in the pilot.

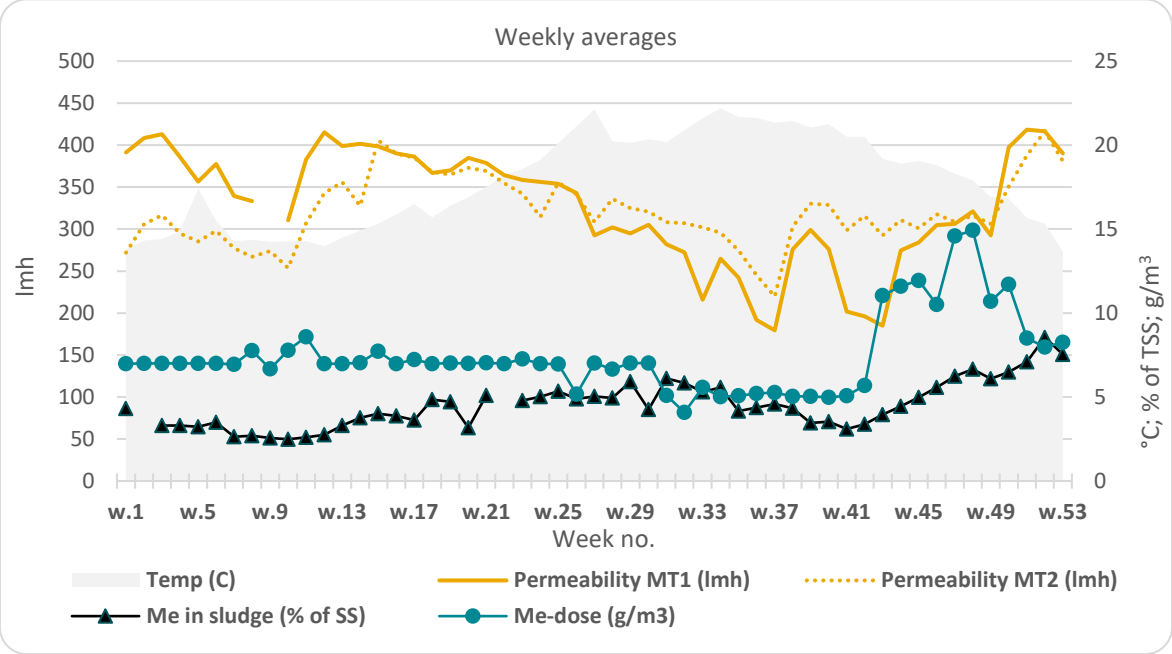


Figure 39. Normalised permeability, temperature, Metal (Me)-dose and Metal (Me) content in sludge over the year.

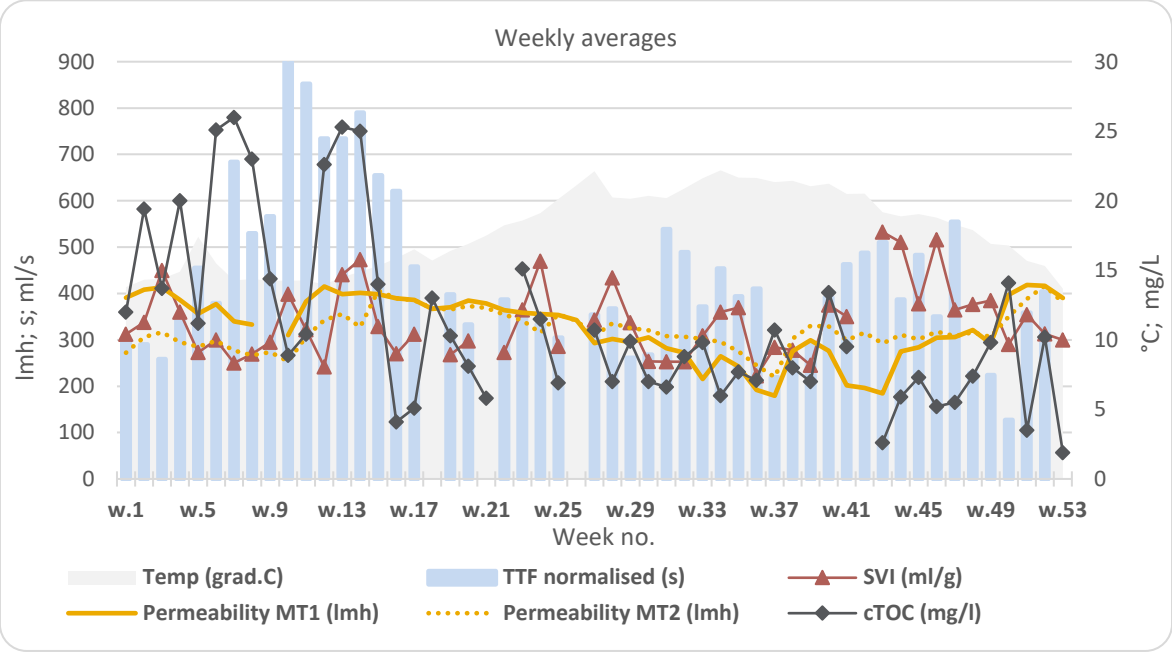


Figure 40. Normalized permeability, temperature, TTF-100 (normalized to a MLSS-concentration of 10 000 mg/L), SVI and cTOC over the year.

In Table 26 the TSS concentration in waste activated sludge (WAS) as well as the content of iron, aluminium, phosphorus and VSS are listed as annular averages for the years the pilot plant has been in operation. A more

efficient precipitation strategy combined with the previously described enhanced biological phosphorus removal, EBPR, caused the Me/P ratio to decrease over the first years. Stricter effluent goals for phosphorus in the effluent then caused the ratio to increase 2019. During 2020 the EBPR was higher than previous years which most likely is the reason for the low Me/P ratio in the sludge.

Table 26. WAS composition (annual average) in the pilot over 7 years of operation (Me = metal (Fe + Al), n = number of samples)

Year	TSS (mg/L)	Fe in sludge (% of TSS)	Al in sludge (% of TSS)	P in sludge (% of TSS)	VSS (% of SS)	Me/P in sludge (mole/mole)
2020	8 967	4.1	0.5	3.3	80	0.9
n	*51*	*51*	*43*	*51*	*51*	*51*
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., 2021). A short summary of the results and conclusions is given below.

Contradictory to the expectations, the concentrations of both metals and organic micro pollutants were in general lower in the sludge from the MBR-pilot than in sludge from Henriksdal WWTP. Since practically all particulate matter is retained by the membranes and accumulated in the sludge, the total area of removed particles is larger in an MBR and the removal of metals and organic materials that adsorb to those surfaces should therefore be bigger in an MBR than in a conventional activated sludge process. On the other hand, there is a dilution effect in the MBR-pilot sludge since the total amount of sludge produced per volume of treated wastewater was larger and the organic degradation rate was lower than in the full-scale plant. In Henriksdal an average of 0.16 kg TS/m^3 incoming wastewater was produced while the value for the pilot was 0.25 kg TS/m^3. This dilution effect seems to overshadow the effect of a larger particle surface. The fact that Henriksdal WWTP receives wastewater from three different tunnels (uptake areas) and the pilot only receives water from one of these can of curse also affect the results since the metal content in wastewater might differ between the different tunnels.

Comparing the quality of the dewatered digested sludge from thermophilic and mesophilic digestion showed that higher metal concentrations were found after thermophilic digestion but lower concentrations of organic micro pollutants. This indicates a higher degradation of organic micro pollutants at the higher thermophilic temperature.

Results from analysis of metals in dewatered digested sludge from the pilot and Henriksdal WWTP are found in Figure 41. Figure 42 shows the metal to phosphorus ratios for some of the metals that are most important for quality control before spreading sludge on farmland.

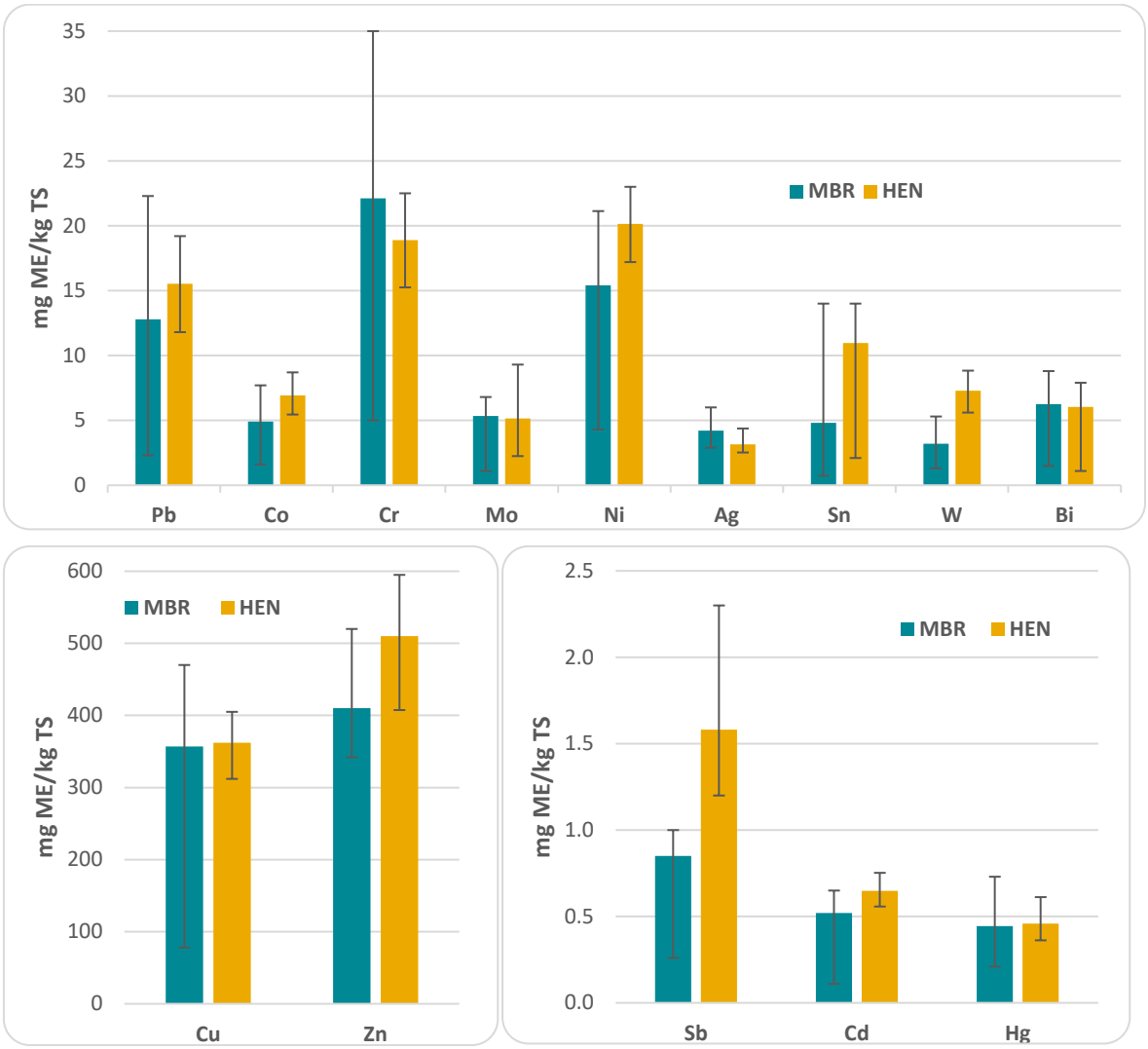


Figure 41. Average values of metal concentrations in dewatered digested sludge from Henriksdal WWTP (HEN) and the MBR pilot. The unit is mg ME/kg TS. Error bars show the lowest and highest analysis value.

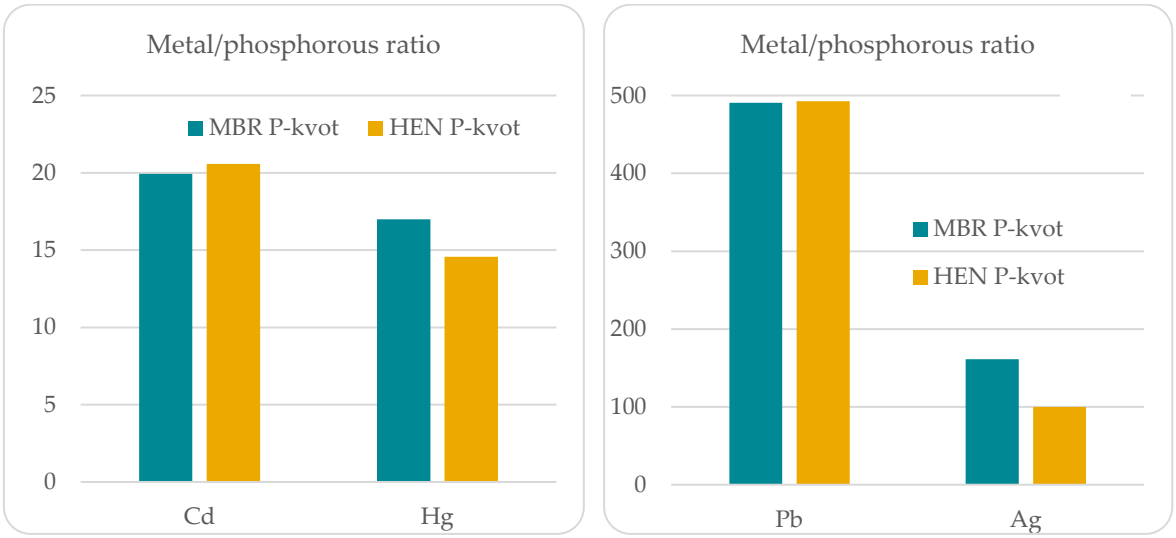


Figure 42. Metal/phosphorus ratio (mg/kg P) for a few important metals. HEN = Henriksdal WWTP.

6.8 Sludge pilot

Trial	J	F	M	A	M	J	J	A	S	O	N	D
How low can we go - thermophilic												
Emptying and inoculation of the reactor												
How low can we go - mesophilic												
Sludge dewatering in operation												
Study on indicators of process failure												

During 2020, the theme of the sludge pilot trials was "how low can we go". Two interconnected trials were performed, a thermophilic and a mesophilic crash test, where the goal was to see at how low HRT the digestion process could be operated before it failed. The thermophilic trial started in 2019 and the mesophilic trial was finalized in the beginning of 2021. Nevertheless, both trials are presented in this year's report. Since the operation of the thickener was very unreliable during previous years, it was decided to bypass the thickener and feed mixed sludge directly into the pilot during these trials since it was crucial to control the HRT. This resulted in an initially low organic loading rate (OLR) which increased as the HRT decreased.

6.8.1 Feed characteristics

As shown in Figure 6, the mixed sludge tank receives Primary Sludge (PS) and Waste Activated Sludge (WAS) from the MBR-pilot line. The digester was continuously fed with mixed sludge (MS) from the MS tank throughout the trials, bypassing the thickener. The proportion of PS to WAS, in terms of TS weight, was adjusted to 60/40 by discharging a fraction of the WAS to be more similar to the mixed sludge at Henriksdal WWTP. The total and volatile solids in the MS varied during the year which can be seen in Figure 43. Since the sludge was not thickened, the TS-concentration in the feed was low, generally around 1.5-3.0%, and the OLR was consequently also low, around 2 kg VS/m^3,d; although increasing with reduced HTR. The OLR at each retention time for the two trials can be seen in Figure 44. OLR was slightly lower during the mesophilic trials due to lower TS in the primary sludge.

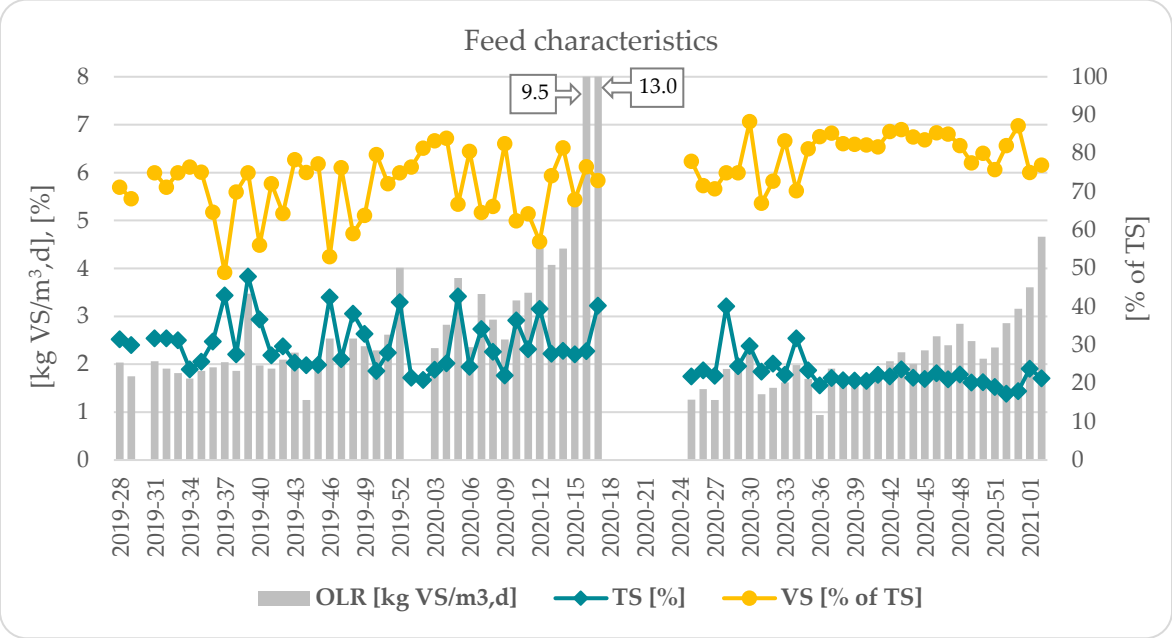


Figure 43. Feed characteristics in terms of TS and VS in mixed sludge and total organic loading rate (OLR) for the trial periods w.28 2019 to w.1 2021.

The MS-pump takes sludge from the bottom of the MS-tank and on several occasions during the trials, sand and grit from the primary clarifier accumulated in the bottom of the tank. Samples for TS and VS were collected from the pipe after the MS-pump. This accumulation of inorganic material resulted in unrepresentative values of TS and VS for some weeks. To compensate for this in Figure 44, a corrected value for TS and VS was used for five of the 73 weekly averages shown. The corrected value was obtained by multiplying the measured TS and VS and then dividing by a fictive VS-value of 75%, which is the average value for the period (example 15/8 2019 w. 33; measured TS=8.8%, VS=19.5%, corrected VS=75%, corrected TS=8.8x19.5/75=2.10%). For calculation of other key values reflecting digestion performance, such as VS% degradation and OLR, the original data was used since the amount of VS [kg/d] is not affected by the correction. In order to mitigate the problem of sand and grit, the mixed sludge tank was partially emptied from the bottom regularly during the second half of the year.

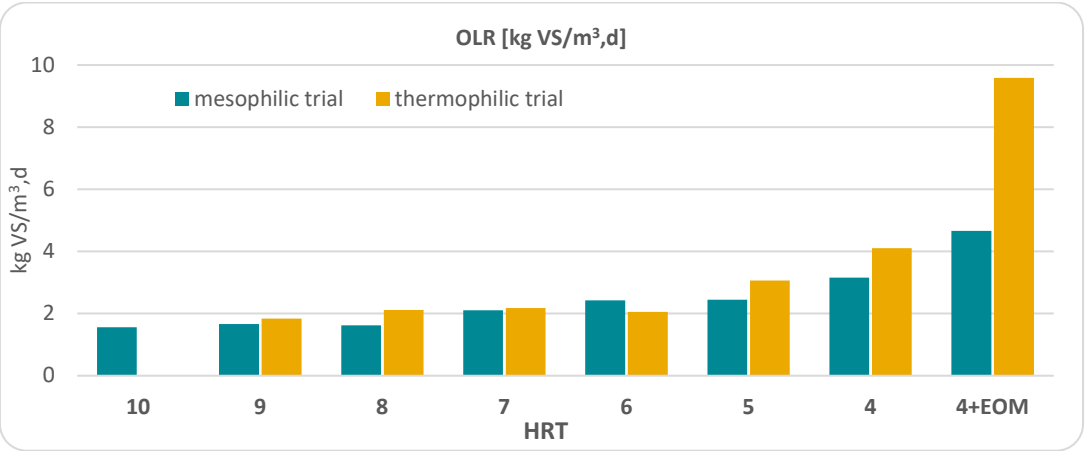


Figure 44. Organic loading rate (OLR) at each retention time (HRT) during the mesophilic and thermophilic trial.

From the end of September 2020, mixed sludge (MS) was analyzed for the content of carbohydrates, proteins fats and COD. The result can be seen in Table 27. The average COD-value of MS was 20 827 mg/L (±15%) which corresponded to 0.9 g COD/g TS or 1.2 g COD/g VS in the sludge.

Table 27. The average content of carbohydrates, proteins and fats in MS.

Component	% of VS	Standard deviation
Carbohydrates	45	±11
Proteins	37	±9
Fats	16	±6

Due to limitations in the digester configuration and pump capacity, the HRT could not be decreased below 4 d. After operation at 4 d HRT for at least 3 HRTs the OLR was increased by addition of external organic material (EOM). At Henriksdal WWTP fat separated from restaurant kitchen drains is added directly to the digesters together with glycerol (a byproduct from biodiesel production) which is added according to biogas demand and digester capacity by Scandinavian biogas Fuels. Therefore, kitchen oil and glycerol were added to the pilot digester week 15-17 2020 (Table 28) and glycerol only week 2 2021 (Table 29). The COD-content was different in the two different glycerol batches used, resulting in different organic loading at the same pump flow. Since it was tricky to pump oil during the end of the thermophilic trial, it was decided to only use glycerol at the end of the mesophilic trial.

Table 28. Addition of EOM during the thermophilic crash test in 2020.

Date	Flow Glycerol [L/d]	Org. Loading Glycerol [kg VS/d]	Flow Oil [L/d]	Org. Loading Oil [kg VS/d]
06-apr	0	0	2.5	2.3
07-apr	2.52	2.70	2.5	2.3
08-apr	5.07	5.43	5	4.6
09-apr	7.61	8.15	5	4.6
10-apr	7.61	8.15	0	0
11-apr	7.61	8.15	0	0
12-apr	7.61	8.15	0	0
13-apr	3.80	4.07	0	0
14-apr	8.54	9.15	5	4.6
15-apr	9.98	10.69	0	0
16-apr	15.00	16.07	5	4.6
17-apr	19.92	21.33	5	4.6
18-apr	19.92	21.33	5	4.6
19-apr	19.92	21.33	0	0
20-apr	24.96	26.73	0	0
21-apr	30.00	32.13	0	0
22-apr	31.68	33.93	0	0
23-apr	31.68	33.93	0	0
24-apr	36.68	39.28	0	0

Table 29. Addition of EOM during the mesophilic crash test in 2021.

Date	Flow Glycerol [L/d]	Org. Loading Glycerol [kg VS/d]	Flow Oil [L/d]	Org. Loading Oil [kg VS/d]
11-jan	2.5	1.61	0	0
12-jan	5	3.21	0	0
13-jan	7.5	4.82	0	0
14-jan	10	6.43	0	0
15-jan	12.5	8.03	0	0

6.8.2 How low can we go – background and set-up

During the reconstruction of the digesters at Henriksdal WWTP, the operating volume will be reduced for a few years, which will cause a reduction in the hydraulic retention time (HRT). In addition, the retention time year 2040 is expected to be around 12 days at normal operation which includes thickening to >6% TS. If thickening fail or some of the digesters need to be taken out of operation for longer periods, the HRT might be reduced to <8 days. Low retention times increase the risk of washing out methanogens because of their long generation time (slow growth), and the lack of time for hydrolysis of 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, good quality gas production and good properties of the digested sludge must be ensured.

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

The pilot digester was operated at thermophilic conditions (55°C) during the first trial and mesophilic conditions (37°C) during the second trial. Digestion of sludge at Henriksdal is expected to be performed at thermophilic conditions in the future since thermophilic operation is supposed to better cope with shorter retention times and higher organic loads (Metcalf & Eddy 2014; Henze et al. 2010). The thermophilic trial was performed to determine the critical HRT (which is equal to sludge retention time, SRT, in a continuous stirred-tank reactor, CSTR) for thermophilic digestion and assess the risks associated to the future full-scale operation. In order to investigate if thermophilic operation really is required in the future or if energy can be saved by continuing to operate mesophilic, it was decided to also perform a mesophilic trial at similar conditions as the thermophilic.

During each trial, the digester's HRT was decreased stepwise with one day HRT at the time (Table 30 and Table 31). Each HRT was planned to be in operation for 3-4 retention times each. If the sludge pilot experienced operational problems affecting the trial, the time of operation was extended. For the thermophilic trial, an initial retention time of 9 days was selected, and for the mesophilic trial 10 days, based on previous experiences from operation of the sludge pilot at stable conditions.

Table 30 and Table 31 summarize the implementation of the thermophilic and mesophilic trial respectively.

Table 30. Experimental set up for the thermophilic trial.

HRT [d]	Date [year-week]	Duration [×HRT]	OLR [kg VS/m³,d]	Digester V [m³]	Flow MS [L/h]
9	2019-28 to 2019-33	4.2	1.91	5.0	22
8	2019-34 to 2019-41	6.2	2.09	5.0	24
7	2019-42 to 2019-49	8	2.18	5.0	29
6	2019-50 to 2020-6	10.5	2.89	5.0	34
5	2020-7 to 2020-10	5.6	3.06	4.2	34
4	2020-11 to 2020-14	7	4.11	3.4	35
4 + EOM	2020-15 to 2020-17	5	9.98	3.4	35

Table 31. Experimental set up for the mesophilic trial.

HRT [d]	Date [year-week]	Duration [×HRT]	OLR [kg VS/m³,d]	Digester V [m³]	Flow MS [L/h]
10	2020-25 to 2020-29	3.5	1.56	5.0	20
9	2020-30 to 2020-37	6.2	1.66	5.0	22
8	2020-38 to 2020-41	3.5	1.63	5.0	24
7	2020-42 to 2020-44	3.0	2.11	5.0	29
6	2020-45 to 2020-47	3.5	2.42	5.0	34
5	2020-48 to 2020-51	5.6	2.45	4.6	36
4	2020-52* to 2020-01	5.3	3.21	3.4	35
4 + EOM	2021-02	1.8	4.66	3.4	35

*Year 2020 had 53 weeks.

Sampling and analysis

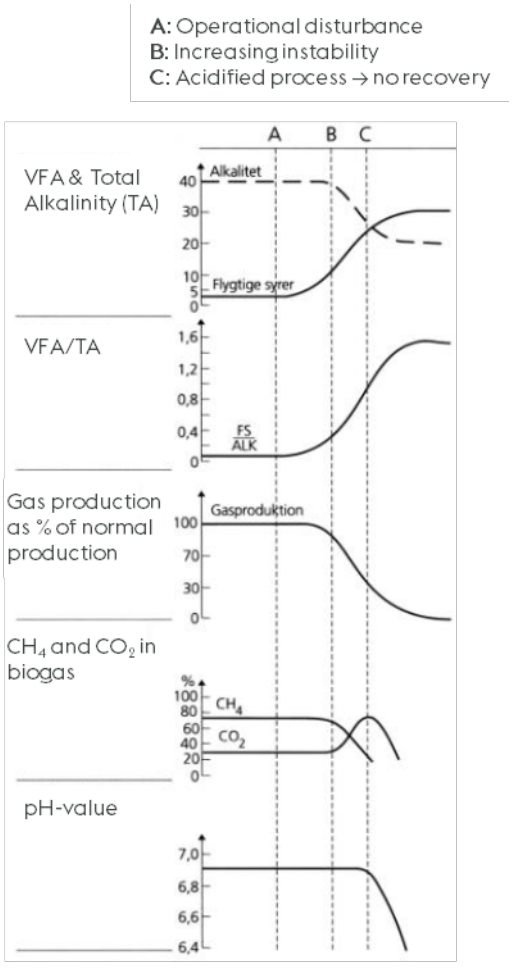
Besides the regular sampling and analysis program described in chapter 5.1, intensive sampling and analysis was done on a daily basis towards the end of each trial. In addition, a sludge sample from the digester was collected and frozen every week for DNA-analysis, which was done in cooperation with SLU. The aim was to determine how the microbial population changes during the experiments. At the time of writing, data from the DNA-analysis was not yet provided and results will instead be published in next year's report. During the mesophilic trial, autumn 2020, analysis of fat, protein and carbohydrates were done by Eurofins for three samples per HRT of mixed sludge (MS) and digested sludge (DS) respectively. At the end of each trial, when VFA increased over 150 mg/L, samples were taken daily and sent to KTH for GC analysis according to Owusu-Agyeman et al. (2020).

Previous studies and expectations

Methanogens grow slowly, with a generation time of 1-12 days depending on species and environmental conditions (Jarvis and Schnürer, 2009). Therefore, it is common to choose a retention time that is at least 12 days when designing anaerobic digesters. At higher retention times, a higher diversity of methanogens is most likely present. When the retention time decreases only the fastest growing methanogens, like the family of *Methanosaetaceaes*, will survive (Lee at al. 2011), which will make the process more vulnerable to disturbances. Studies using semi-continuous CSTR lab scale reactors have previously shown that process imbalances (foaming, VFA accumulation etc.) started to show at retention times below 9 days and that digestion could be performed down to 5 days retention time at both thermophilic and mesophilic conditions but with significantly decreased VS destruction efficiency (Nges and Liu 2010). At 3 days retention time both processes crashed due to accumulation of VFA. Lee et al. (2011) showed that mesophilic digestion could be maintained in lab-scale at 4 days retention time, although less efficient in terms of specific gas production, and that a shift in microbial population occurred at 5 and 4 days retention time. Both studies were performed at

higher OLRs than the study presented here. It is difficult to find relevant references on digestion of WWTP-sludge at low retention times in pilot- or full-scale. Since pilot trials usually experience more process disturbances than laboratory experiments, it was not expected that the conducted pilot test would be possible to run at such low retention times. Based on literature and experiences the expectations were set on a crash at 6 days HRT at thermophilic digestion and 7 days at mesophilic.

During autumn 2020, Jessica Sellin performed her MSc-thesis project studying parameters for process monitoring and the indicators for digestion process instability, both by going throw operational data from the thermophilic crash test and by following up the results from the ongoing mesophilic crash test (Sellin 2021). It was expected that changes in VFA and alkalinity, and especially the ratio between them (VFA/TA) would be the earliest indicators of process instability, followed by a drop in pH, as can be seen in Figure 45.



https://www.cok.dk/sites/cok.dk/files/2017-01/Drift%20af%20renseanlaeg%2C%20Teknik.pdf

Figure 45. Important process parameters and their expected behaviour when the process is approaching a crash (Henze et al., 2010).

6.8.3 Results thermophilic trial

An overview of the operation of the digester during the thermophilic trial can be seen in Figure 46. As can be seen, the thermophilic digestion process was operated rather stably down to 4 days retention time. After 28 days (7 x HRT) at 4 days retention time with an OLR of just above 4 kg VS/m³,d and no sign of instability; EOM was added in an attempt to crash the digester (since the HRT could not be lowered further). The OLR was increased over 20 days to 15.5 kg VS/m³,d on the 24th of April when it finally crashed.

It can also be seen in Figure 46 that the actual HRT increased slightly at the end of HRT 9, 6 and 5. This was caused by short stops in the inflow to allow the water level in the digester to decrease and thereby decrease the reactor volume. In week 44, the inlet pump feeding the digester broke on Friday afternoon and was not fixed until Monday morning the week after, which caused the HRT to rise from 7 to 11 days as a weekly average.

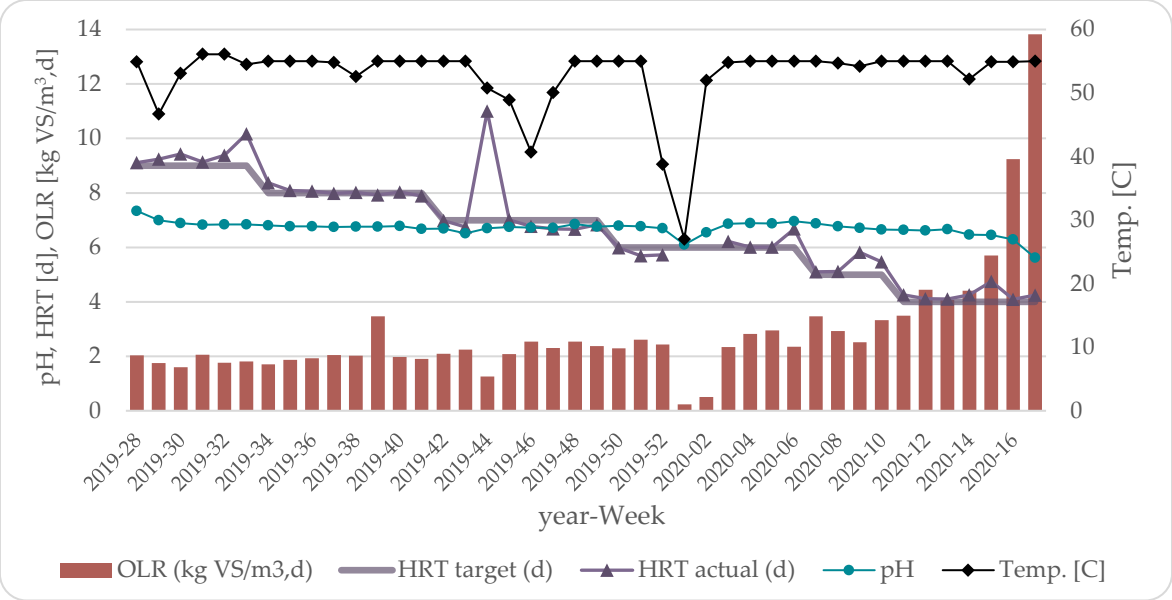


Figure 46. Operation during the thermophilic trial. Weekly averages.

During the thermophilic trial, problems with the heating system occurred three times, resulting in temperature drops (Figure 46). The largest temperature decrease occurred during 15 days at HRT 6 d, starting on Christmas eve 2019 when no operational staff worked, and no mechanics were available until after the holidays (see Figure 47). The temperature went as low as 18°C. The feeding was stopped on December 30 and resumed on January 9. The VFA increased to a maximum of 767 mg/L on January 2. In order to stop acidification and save the trial, 1.7 kg of bicarbonate mixed with water was fed to the digester on December 30 and <1 kg on January 2, resulting in normal VFA values from January 14 onward. Although the pH dropped to 6.11, the process recovered surprisingly well and after three retention times the "How low can we go" trial was resumed.

Data from times with process disturbances was not included in the evaluations.

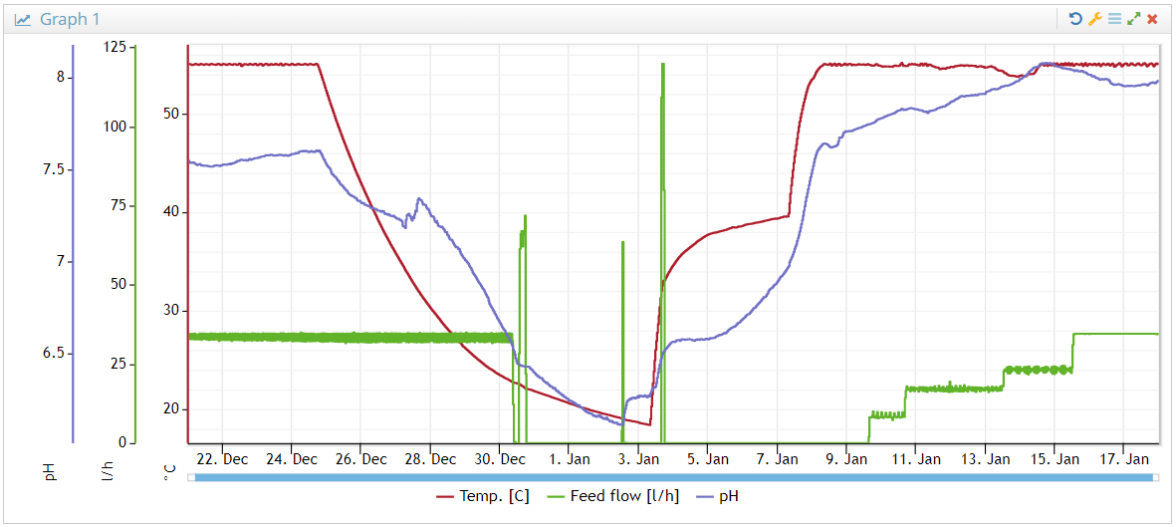


Figure 47. Online data on temperature, feed flow and pH from December 22, 2019 to January 17, 2020 when the heating system broke. Note that the online pH sensor shows a value that is about one unit higher than the actual value (Screenshot from aCurve, resolution 1 min).

Parameters often used to monitor digestion processes and more specifically the buffering system in the sludge, are shown in Figure 48 as an average value per retention time. pH decreased from 6.96 to 6.60 before addition of EOM while VFA decreased somewhat from 138 to 102 mg/L. At the same time, the total alkalinity (TA) went down from 2081 to 1458 mg/L and ammonium decreased from 374 to 268 mg/L, further lowering the buffer capacity. Since both VFA and TA decreased, the VFA/TA ratio was rather constant around 0.08. The average values for the period when EOM was added shows a steep increase in VFA and a steep decrease in ammonium, whereas the total alkalinity did not change much. These changes resulted in a VFA/TA ratio of 0.26 and a pH of 6.1. In conclusion, neither VFA and TA nor pH were considered good early indicators of a process collapse. The ammonium values on the other hand could possibly be used to foresee process instability.

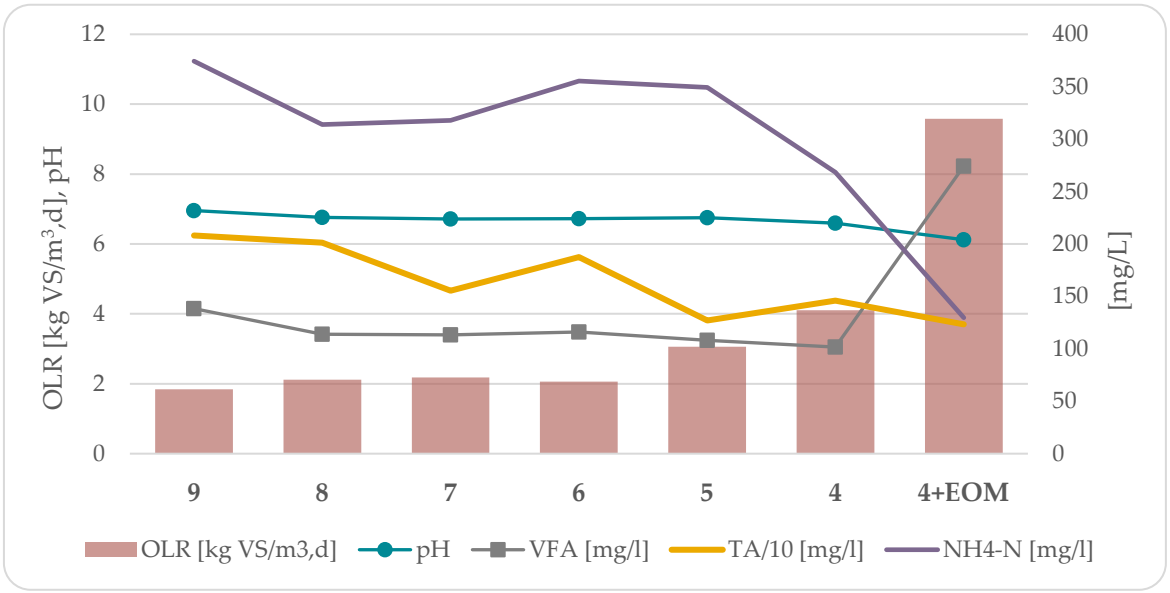


Figure 48. Parameters related to the buffering system in the digester. Note that total alkalinity (TA) values are divided by 10. Averages for each retention time.

Figure 49 shows process parameters related to degradation and biogas production as average values for each retention time. The organic degradation rate (ODR) showed steady values between 49 and 51% of VSin except for the period with 5 days HRT when it was only 41%. When EOM was added, the ODR increased since EOMs are easily available substrates that are almost entirely converted to biogas. The biogas production increased with increasing OLR and the methane content of the biogas was stable at 59-62% until HRT 5 d when it slowly started to decrease to a final value of 54%, which is not that bad. The specific biogas production and the specific methane production decreased slightly with decreasing HRT. One of the parameters that changed the most was H_2S in the biogas, which increased over the trial.

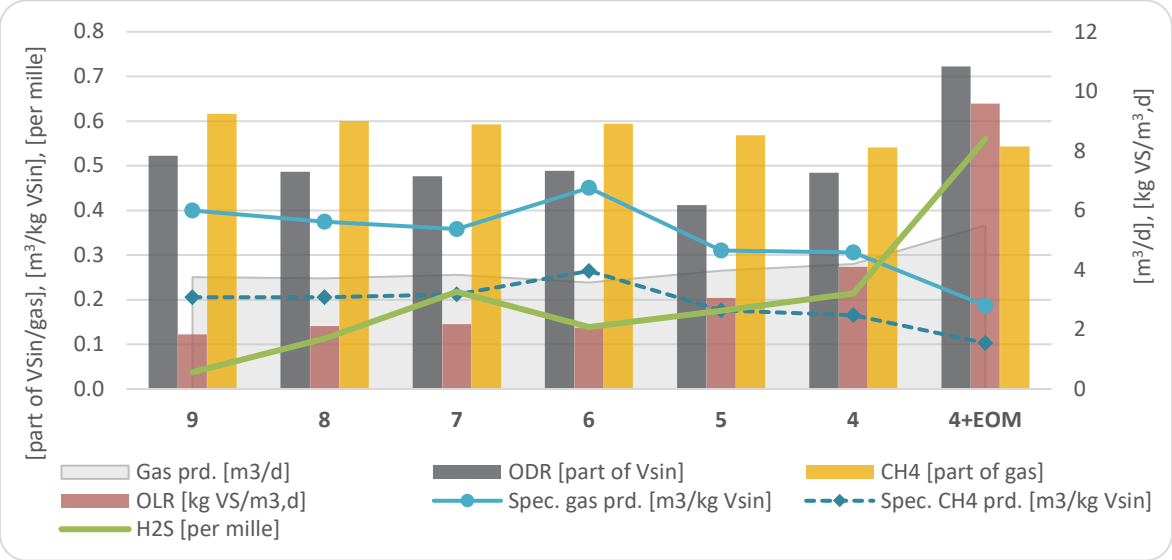


Figure 49. Parameters related to degradation and gas production during the thermophilic trial. Note that H_2S is given in ‰ and not ppm. Averages for each retention time

There was an approximately 200 mm thick layer of foam in the reactor throughout the trial that did not change much over time.

Dewatering

Dewatering of digested sludge was in operation from week 31 (HRT 9 days) to week 10 (HRT 5 days). Data from dewatering is shown in Table 32. Dewatering in the screw press failed when the retention time was lowered from 5 to 4 days. However, no big efforts to optimize the operation were made during the trial since it was not the central focus of the study, and better results could probably have been obtained.

Table 32. Data from dewatering of thermophilic digested sludge. At HRT 4 the dewatering failed.

HRT [days]	Polymer dose [g/kg TS]	TS [%]	SS in reject water [mg/L]
9	6.8	28.6	863
8	7.4	26.3	723
7	12.2	24.6	412
6	12.6	24.4	956
5	12.7	24.6	2146

Last three weeks

Table 33 shows process data from the last 3 weeks of thermophilic operation when EOM was added to the digester. It can be seen that VFA did not increase to values higher than 100 mg/L and the alkalinity, which decreased inconsistently, didn't drop below 1000 mg/L until the OLR reached around 10 kg VS/m^3,d. VFA

didn´t reach values above 500 mg/L until the last day. Interestingly, the alkalinity increased the last two days. The gas production increased slightly with increasing OLR and the methane content was rather similar until the time of the crash. The specific gas production, on the other hand, decreased day by day indicating decreased activity and/or capacity of the process. Figure 50 shows a high-resolution graph of the changes in pH-value during the last phase of operation (HRT 4 d + EOM).

Table 33. Data from the last three weeks of thermophilic operation at 4 d HRT and addition of EOM according to Table 28. No sampling and analyses were done on weekends and public holidays (10/4).

Date	OLR [kg VS/m³,d]	pH	VFA [mg/L]	ALK [mg/L]	VFA/TA	NH_4-N [mg/L]	Gas prd. [m³/d]	CH_4 [%]	Spec. CH_4 prd. [m³/kg VS_{in}]
06-apr	3.3	6.45	76	1480	0.05	236	4.69	54.3	0.23
07-apr	4.0	6.45	123	2060	0.06	234	4.98	56.3	0.21
08-apr	5.5	6.49	101	2240	0.05	406	5.99	58.6	0.19
09-apr	8.0	6.44	101	1300	0.08	182	5.73	58.3	0.12
10-apr	6.3								
11-apr	6.3								
12-apr	6.3								
13-apr	4.8	6.36	101	1220	0.08	128	5.74	51.9	0.18
14-apr	6.5	6.34					5.06	53.1	0.12
15-apr	9.8	6.22	220	1080	0.20	140	5.92	56.0	0.10
16-apr	8.6	6.32	162	1000	0.16	116	5.59	54.7	0.10
17-apr	11.5	6.25	97	740	0.13	116	5.33	54.8	0.07
18-apr	11.8								
19-apr	11.7								
20-apr	11.9	6.16	209	460	0.45	20	5.32	54.5	0.07
21-apr	13.5	6.09	215	560	0.38		5.54	54.8	0.06
22-apr	14.0	6.10	120	1220	0.10	24	5.58	54.2	0.06
23-apr	14.1	5.30	697	1800	0.39	23	5.73	50.3	0.06
24-apr	15.5	4.88	1083	1200	0.90	44	1.45	46.6	0.03

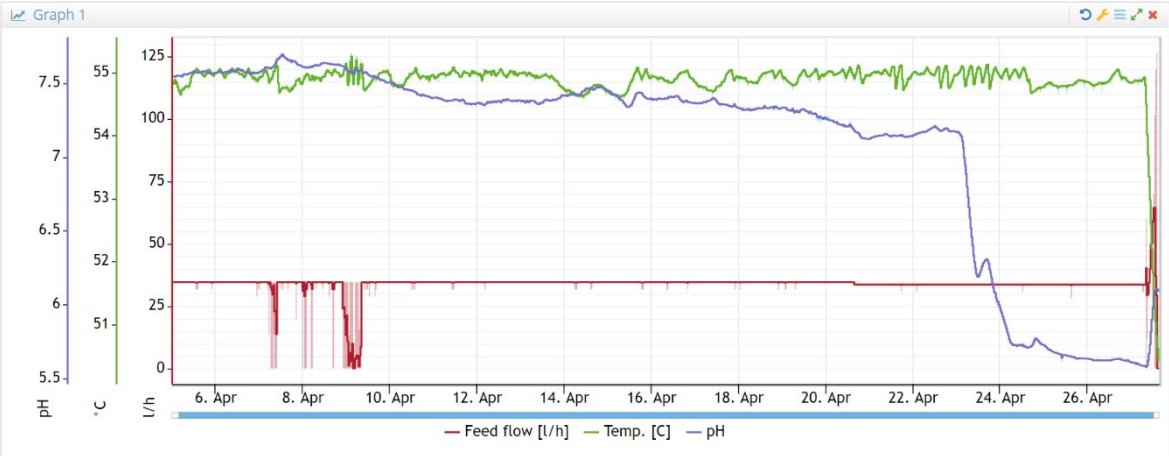


Figure 50. Online data on temperature, feed flow and pH from April 6 to 27 2020. The digester was proclaimed dead on the 24th. Note that the online pH sensor shows a value that is about one unit higher than the actual value (Screenshot from aCurve, resolution 1 min).

Sludge samples were analysed for content of different VFAs by GC as described previously. Samples were taken on the 8th (OLR 5.5 kg VS/m³,d), 11th (OLR 6.3 kg VS/m³,d), and 22nd (OLR 14.0 kg VS/m³,d) of April when the digester was operated at 4 d HRT and with EOM addition. Results are shown in Table 34. The most abundant VFA at both low and high VFA concentration was acetic acid followed by propionic acid.

Table 34. Composition of VFAs in the digester around the time of the thermophilic crash. Value from spectrophotometric cuvette tests as reference.

Date	GC Results (mg/L)						Cuvettes (mg/L)
	Acetic	Propionic	Isobutyric	Butyric	Isovaleric	Total VFA	Acetic eq.
8-apr	48	8	0	0	0	56	100
11-apr	50	5	0	0	0	55	220
22-apr	1047	321	53	7	48	1476	991

Summary of performance

A summary of the results is shown in Table 35. As mentioned before, the digestion process performed stably down to 4 days HTR without being irreversibly acidified. Translated to full-scale, this means that we probably could operate the digesters at considerably lower retention times than expected (though not as low as 4 days), at least if the reduction in retention time is done slowly and the OLR is relatively low. It should be noted that the mixing of the pilot digester is most likely better than in the full-scale digesters due to the relatively high recirculation flow, this could affect the performance since no local shock loadings occur. This information is important when planning for maintenance work or during risk assessment of for example the thickeners efficiency and reliability.

However, the process performance in terms of biogas production (Spec. CH_4 prd., CH_4) and degradation of organic material (ODR, NH_4-N), two factors important for full-scale operation of the digestion process, was unsatisfying below 6 d HRT. Thus, operation at very low retention times should not be a standardized mode of operation. To balance the up to 25% variations in sludge production over the year and keep a bit of a margin, an HRT of less than 8 d is not recommended in the full-scale digesters.

Table 35. Evaluation of process performance.

Parameters	Henriksdal HRT 16 d, mesophilic	Green	Yellow	Red	Termophilic trial					
					HRT9	HRT8	HRT7	HRT6	HRT5	HRT4
Org. degradation rate (% of VS_{in})	50%	>50%	>45%	<45%	52%	49%	48%	49%	41%	48%
Spec. CH_4-prod. (m^3/kg VS_{in})	0.31	>0.25	>0.20	<0.20	0.21	0.21	0.21	0.26	0.18	0.17
pH	7.2	>6.90	>6.70	<6.70	6.96	6.76	6.71	6.73	6.75	6.60
VFA/TA	<0.03	<0.07	<0.3	>0.3	0.08	0.06	0.07	0.12	0.08	0.07
- VFA (mg/L)	<100	<150	<300	>300	138	114	113	116	108	102
- TA (mg/L)	3400	>1500	>1000	<1000	2081	2012	1554	1875	1271	1459
CH_4 (% of biogas)	66	>60%	>55%	<55%	62%	60%	59%	59%	57%	54%
H_2S (mg/L)	-	<50	<100	>100	38	113	218	139	176	214
NH_4^+-N (mg/L)	-	>500	>300	<300	374	314	318	355	349	268
TS after dewatering (%)	28%	>26%	>23%	<23%	28.6	26.3	24.6	24.4	24.6	x

6.8.4 Results mesophilic trial

An overview of the operation of the digester during the mesophilic trial can be seen in Figure 51. As can be seen, the mesophilic digestion process, just like the thermophilic one, was operated rather stably down to 4 days retention time. After 21 days (5.3 x HRT) at 4 days retention time with an OLR of 3.2 kg VS/m^3,d, EOM was added in an attempt to crash the digester (since the HRT could not be lowered further). The OLR was increased over 5 days to 5.3 kg VS/m^3,d on the 15th of January 2021 when the process crashed.

Just like in the thermophilic trial it can be seen in Figure 51 that the actual HRT increased slightly at the end of HRT 10, 9 (a lot) and 5. This was caused by short stops in the inflow to allow the water level in the digester to decrease and thereby decrease the reactor volume. At the end of HRT 9, the inlet pipe got clogged during the lowering of the volume a Friday afternoon and the pipe was not cleared until the following Monday resulting in a high HRT peak. There was also a small failure in the heating system on week 35 with a temperature low at 33°C.

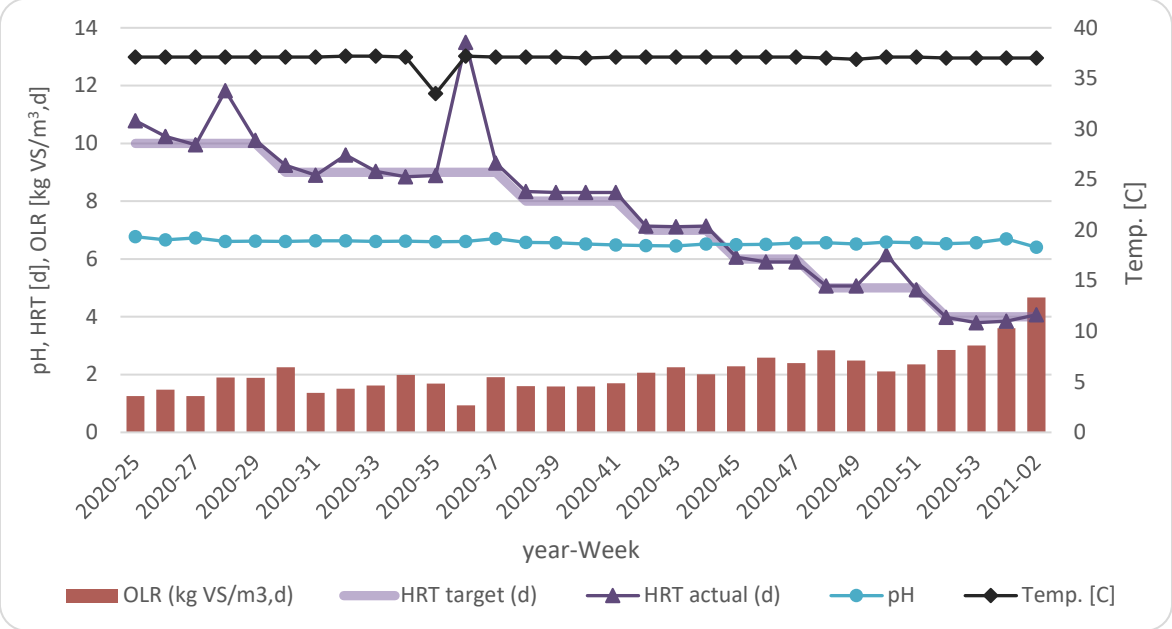


Figure 51. Operation during the mesophilic trial. Weekly averages.

Parameters often used to monitor digestion processes and more specifically the buffering system in the sludge, are shown in Figure 52 as average values per retention time. pH decreased from 6.68 to 6.60 before addition of EOM while VFA was rather constant at 65±10 mg/L until HRT 4 d when it increased to over 100 mg/L. The total alkalinity (TA) went down from 1760 to 650 mg/L from HRT 10 to 5 d and then went up a bit to values around 1000 mg/L. Ammonium decreased almost linearly from 232 to 130 mg/L with exception of a small increase at HRT 7 d. The VFA/TA ratio was below 0.06 until HRT 5 d when it increased to 0.13. The average values for the period when EOM was added shows a steep increase in VFA while alkalinity remained the same. These changes resulted in a VFA/TA ratio of 0.47 and a pH of 6.4. None of the parameters in Figure 52, not even ammonium, provided a clear early warning of a collapse, although some of them clearly indicated irreversible instability at HRT 4 with EOM dosage.

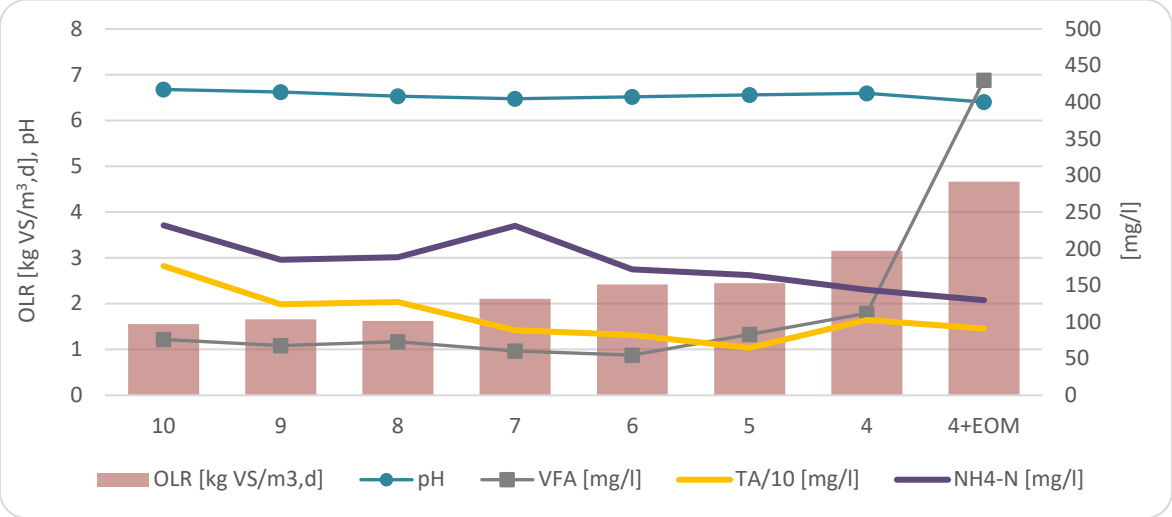


Figure 52. Parameters related to the buffering system in the digester. Note that total alkalinity (TA) values are divided by 10. Averages for each retention time.

Figure 53 shows process parameters related to degradation and biogas production as average values for each retention time. The organic degradation rate (ODR) decreased from 49% to around 30% of VSin. When EOM was added, the ODR increased since EOMs are easily available substrates that are almost entirely converted to biogas. The biogas production increased with increasing OLR until HRT 6 d. After that, it decreased even though the OLR increased. The methane content of the biogas was stable at 58-60% until EOM was added when it decreased slightly to 56%. The specific biogas production and the specific methane production increased initially, from HRT 10 to 8 d where it peaked. Thereafter it was rather similar until EOM was added when it dropped to 0.22 and 0.12 m^3/kg VSin respectively. H_2S in the biogas increased initially but then decreased to low levels again.

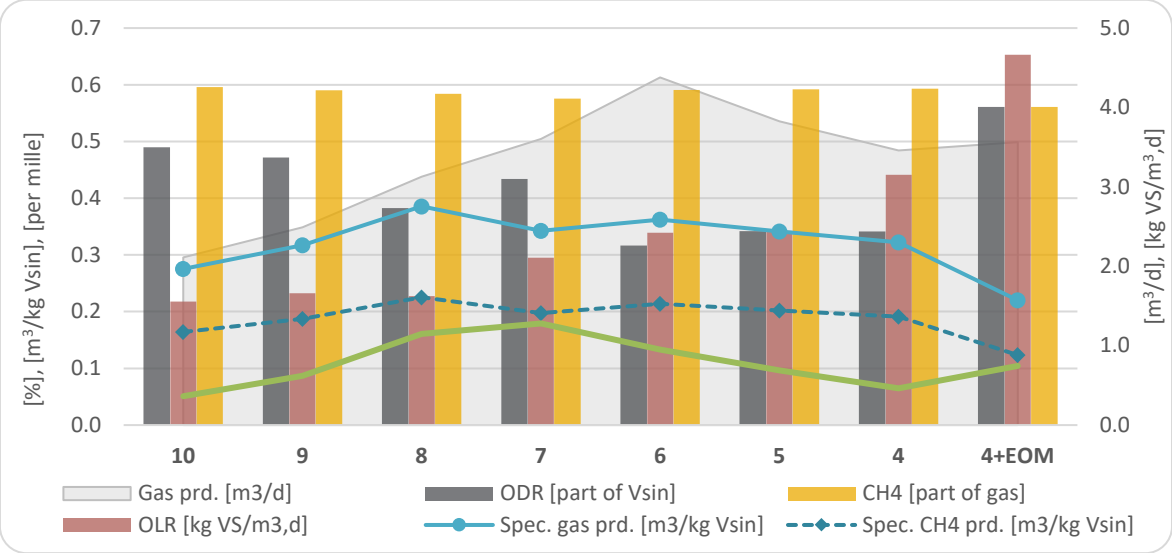


Figure 53. Parameters related to degradation and gas production during the mesophilic trial. Note that H_2S is given in ‰ and not ppm. Averages for each retention time.

The degradation of different components of the sludge at different HRT:s is found in Table 36. The analyses were initiated in September, which means that the first data presented are from 8 days retention time. The

results did not show distinct trends, which could be due to difficulty to accurately analyse sludge (measurement uncertainty 20-30%). The fat analyses gave very varied results in general which could depend on the fact that fat often aggregates making it hard to get a representative sample for the analysis. The results for fat in Table 36 are random making it difficult to draw any conclusions about degradation of fat. Carbohydrates were degraded to slightly higher extent at higher HRTs, which corresponded to the results for VS. For proteins the degradation decreased with decreasing HRT with exception for HRT 7. This correlates with the results from ammonium in the digester presented in Figure 53, showing decreasing concentration with decreasing HRT. The degradation COD was 38 ± 5 % with no correlation to the HRT.

Table 36. The degradation (% of incoming) of sludge components at different retention times. Averages of 3-4 analyses.

HRT	Carbohydrates	Proteins	Fats	VS	COD
8	63%	27%	25%	38%	33%
7	73%	15%	-13%	43%	43%
6	48%	20%	14%	32%	37%
5	47%	18%	57%	35%	41%
4	53%	15%	21%	34%	32%

There was an approximately 280 mm thick layer of foam in the reactor throughout the trial that did not change much over time.

Dewatering

During the period with a retention time of 10 days the dewatering was not in operation. From week 35 (HRT 9 days) to week 46 (HRT 6 days) the digested sludge was dewatered. Data from dewatering is shown in Table 37. Dewatering in the screw press failed after two weeks at HRT 6 days. However, no big efforts to optimize the operation were made during the trial since it was not the central focus of the study, and better results could probably have been obtained. It should also be noted that the polymer used was the one that was tested out for the thermophilic sludge. No test of different polymers was done for the mesophilic sludge. This might affect the results and the comparison with the thermophilic results.

Table 37. Data from dewatering of mesophilic digested sludge. At HRT 5 the dewatering failed.

HRT [days]	Polymer dose [g/kg TS]	TS [%]	SS in reject water [mg/L]
10 – no operation			
9	7.7	23.0	210
8	12.0	21.2	283
7	15.5	20.1	953
6	14.4	18.7	1300

Last week

Table 38 shows process data from the last week of mesophilic operation when EOM was added to the digester. It can be seen that VFA and the VFA/TA ratio increased while total alkalinity (TA) and NH_4-N decreased day by day, indicating that the process was on the verge of giving up already before EOM was added. The gas production initially increased slightly with increasing OLR while the methane content decreased a bit every day. The specific methane production did not change much until the last day when it decreased to 0.09 m^3/kg VSin. Figure 54 shows a high-resolution graph of the changes in pH value during the last phase of operation (HRT 4 d + EOM).

Table 38. Data from the last week of mesophilic operation at 4 d HRT + EOM addition according to Table 29.

Date	OLR [kg VS/m³,d]	pH	VFA [mg/L]	ALK [mg/L]	VFA/TA	NH_4-N [mg/L]	Gas prd. [m³/d]	CH_4 [%]	Spec. CH_4 prd. [m³/kg VS_{in}]
11-jan	3.9	6.67	149	1480	0.10	156	3.67	59.4	0.16
12-jan	4.9	6.66	179	1030	0.17	152	4.04	58.4	0.13
13-jan	4.3	6.51	329	840	0.39	134	4.10	57.1	0.15
14-jan	4.0	6.32	366	870	0.42	128	3.92	55.5	0.15
15-jan	5.3	6.04	847	580	1.46	104	3.34	53.5	0.09

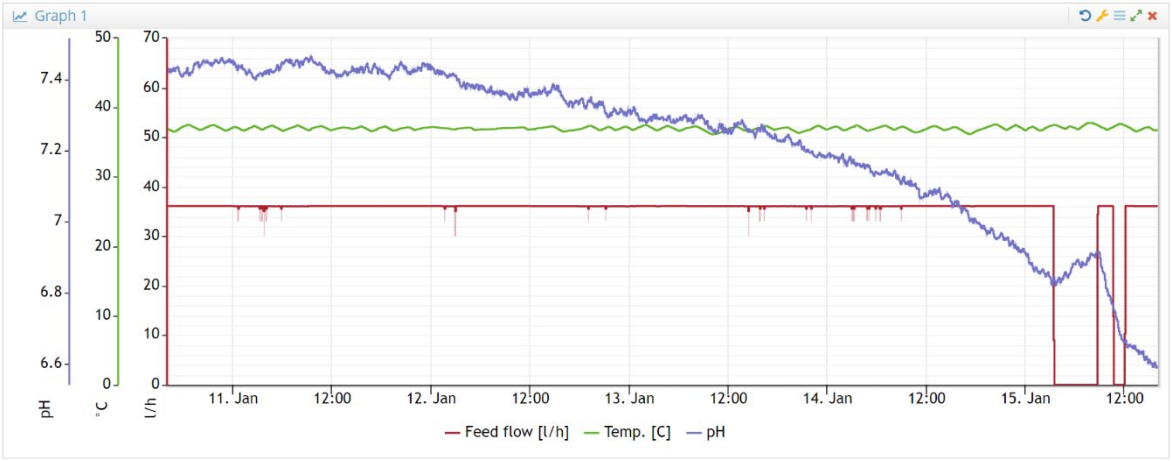


Figure 54. Online data on temperature, feed flow and pH from January 10 to 15 2021. The digester was proclaimed dead on the 15th. Note that the online pH sensor shows a value that is about one unit higher than the actual value (Screenshot from aCurve, resolution 1 min).

In addition to the daily measurements of total-VFA using spectrophotometric methods, GC was used to measure the VFA the day before the crash, January 14 (OLR 4.5 kg VS/m³,d), and the day of the crash January 15 (OLR 5.3 kg VS/m³,d) and 3 days after the crash, on January 18 (OLR not measured). Results can be seen in Table 39. Propionic acid was the most prevailing VFA followed by acetic acid and isobutyric acid.

Table 39. Composition of VFAs in the digester around the time of the mesophilic crash. Value from spectrophotometric cuvette tests as reference.

	GC Results (mg/L)						Cuvettes (mg/L)
Date	Acetic	Propionic	Isobutyric	Butyric	Isovaleric	Total VFA	Acetic eq
14-jan	108	503	17	0	8	636	366
15-jan mor.	100	786	63	0	17	966	711
15-jan aft.	121	950	96	3	18	1188	847
18-jan	92	298	47	7	27	470	393

Summary of performance

A summary of the results is shown in Table 40. As mentioned before, the digestion process performed stably down to 4 days HTR also at mesophilic operation without being irreversibly acidified. However, the process performance in terms of biogas production (Spec. CH_4 prd., CH_4) and degradation of organic material (ODR, NH_4-N), two factors important for full-scale operation of the digestion process, was unsatisfying below 9 d

HRT. With the same reasoning as for the thermophilic results, operation at an HRT of less than 11.5 d is not recommended for the full-scale digester.

Table 40. Evaluation of process performance.

Parameters	Henriksdal HRT 16 d, mesophilic	Green	Yellow	Red	Mesophilic trial					
					HRT9	HRT8	HRT7	HRT6	HRT5	HRT4
Org. degradation rate (% of VS_{in})	50%	>50%	>45%	<45%	47%	38%	43%	32%	34%	34%
Spec. CH_4-prod. (m^3/kg VS_{in})	0.31	>0.25	>0.20	<0.20	0,19	0,23	0,20	0,21	0,20	0,19
pH	7.2	>6.90	>6.70	<6.70	6,62	6,53	6,48	6,52	6,56	6,60
VFA/TA	<0.03	<0.07	<0.3	>0.3	0,06	0,06	0,07	0,07	0,13	0,11
- VFA (mg/L)	<100	<150	<300	>300	68	73	60	55	83	112
- TA (mg/L)	3400	>1500	>1000	<1000	1241	1271	825	890	648	1031
CH_4 (% of biogas)	66	>60%	>55%	<55%	59%	58%	58%	59%	59%	59%
H_2S (mg/L)	-	<50	<100	>100	86,7	160,9	179,2	133,0	96,2	64,9
NH_4^+-N (mg/L)	-	>500	>300	<300	185	189	231	172	164	144
TS after dewatering (%)	28%	>26%	>23%	<23%	23,0	21,2	20,1	18,7	x	x

6.8.5 Comparison

Figure 55 below shows comparative graphs from the thermophilic and mesophilic trials related to degradation of substrate and production and quality of biogas. Values from the thermophilic trial are generally higher than in the mesophilic trial except for the methane concentration that was quite similar in the two trials. The higher biogas production could be due to the slightly higher OLR in the thermophilic trial (see Figure 44). The higher degradation rate (ODR) during thermophilic digestion was also confirmed by the higher ammonium concentrations (Figure 56) in the reactor, derived from degradation of proteins. Figure 56 shows a comparison of VFA, TA, VFA/TA, NH_4-N and pH during the thermophilic and mesophilic trials. Both VFA and alkalinity was significantly higher during thermophilic digestion compared to mesophilic although the difference decreased with decreasing HRT. VFA/TA did not differ much but was a bit higher at thermophilic digestion initially, but from 5 d HRT the ratio was higher during mesophilic digestion. Ammonium was around twice as high in thermophilic digestion except for the last phase (4 d HRT + EOM) when both temperatures gave equal values. pH was below 7 in both trials but was about 0.3 units higher during thermophilic digestion until HRT reached 4 days.

Overall, thermophilic digestion showed better process values at low HRTs than mesophilic digestion.

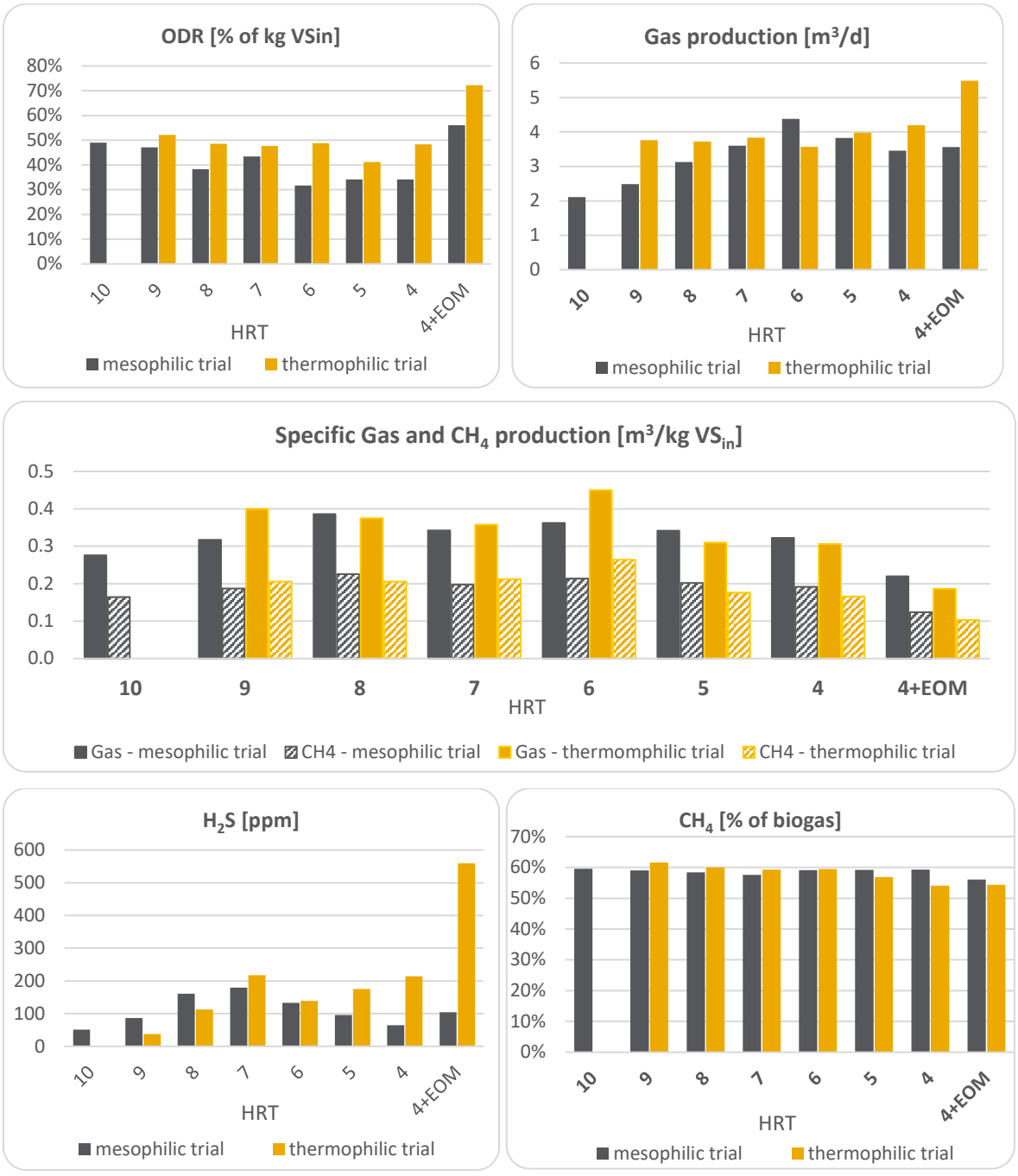


Figure 55. Graphs on degradation and biogas comparing the thermophilic and mesophilic trials.

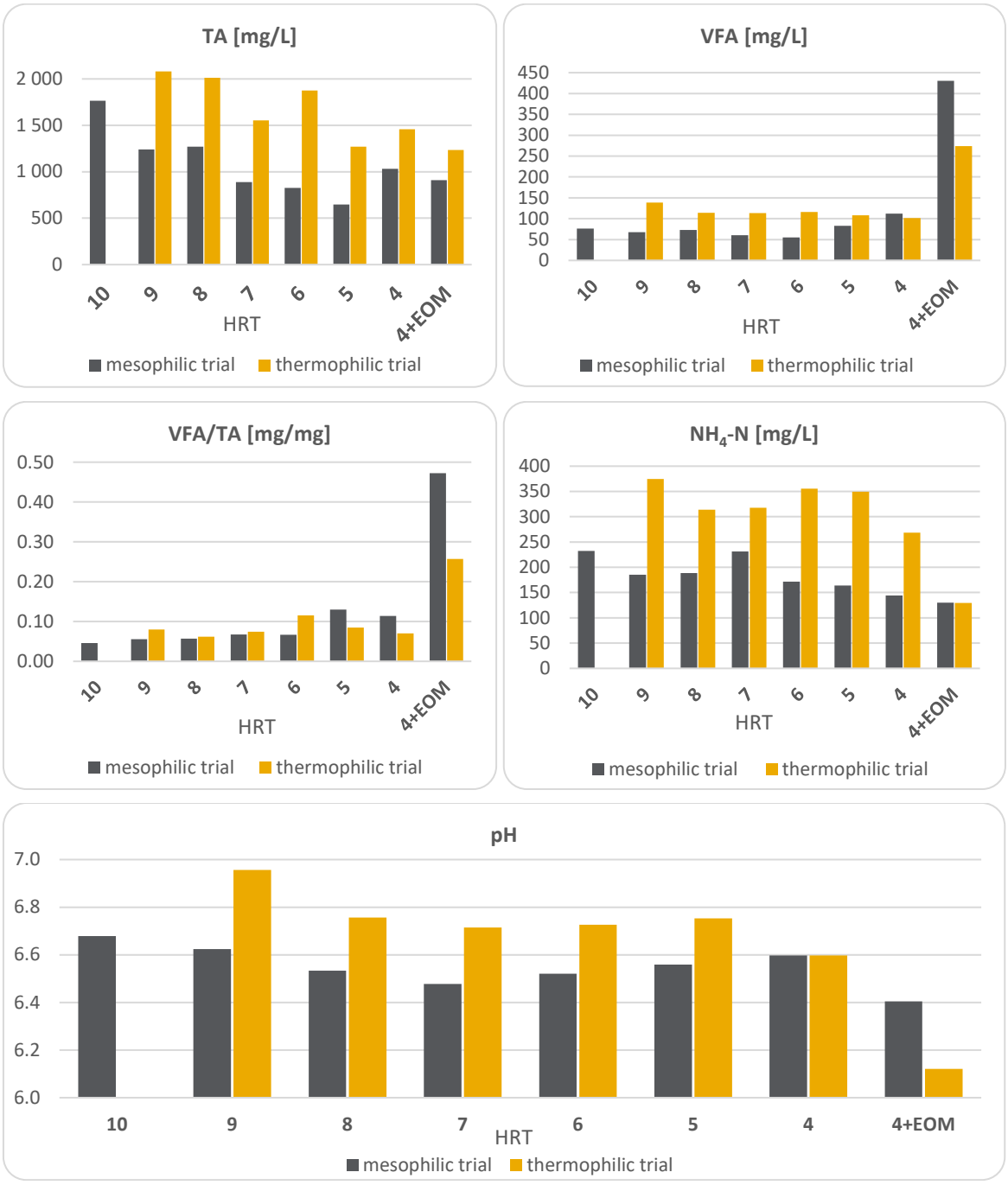


Figure 56. Graphs on parameters related to pH and the sludge buffer system, comparing the thermophilic and mesophilic trials.

6.8.6 Method comparison for VFA and CH_4 production

Jessica Sellin (2021) compared different methods for analysing VFA in the digesters and for estimating the methane production as part of her Master thesis. The results show that the spectrophotometric cuvette tests are best at low concentrations, <100 mg/L, while both HPLC, titration and GC give more reliable results at high concentrations, >200 mg/L (Figure 57). Cuvette analysis of pure acetate and propionate at concentrations between 80 and 1600 mg/L showed a 100% accuracy for acetate but only 66% for propionate. Since the VFA in the digester was low during normal operation and acetate was the dominating VFA, even at low HRTs, it was concluded that the cuvette tests are best suited for monitoring of the digestion process.

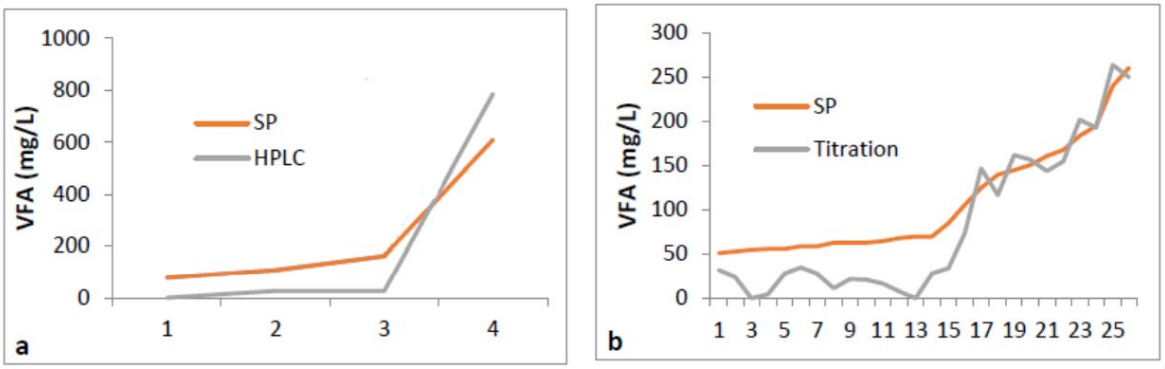


Figure 57. Comparison of VFA analysis methods. SP = spectrophotometric cuvette test. From Sellin (2021).

During a short period of the trials, there were problems with the gas flow meter. Therefore, Jessica also investigated how well a theoretically calculated biogas and methane production based on either VFA, COD or protein/fat/carbohydrate degradation corresponds to the measured values (Sellin 2021). Results, shown in Figure 58, show that calculations based on protein/fat/carbohydrate degradation gave the most accurate results with a 98% (±9%) accuracy. Calculations based on VS underestimated the methane production with around 13% while COD overestimated the production with 22% as an average.

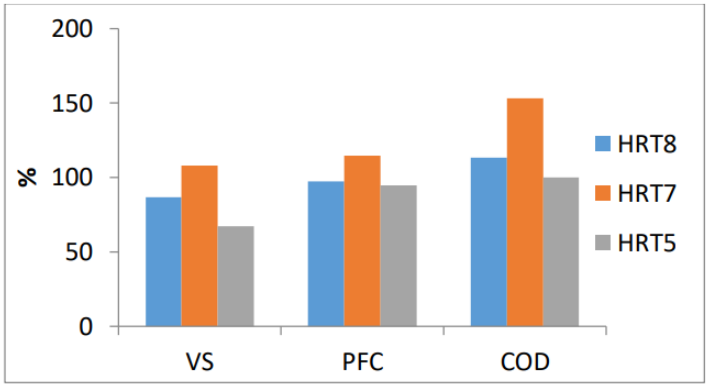


Figure 58. Theoretical methane production for mesophilic trial at HRT 8, 7 and 5 based on degradation of VS, PFC (protein, fat and carbohydrate) and COD as percentage of measured methane production. From Sellin (2021).

6.9 Resource consumption

Resource consumption in the pilot for 2020 is summarized in Table 41. A comparison with the future Henriksdal design was made where possible (where design values are available). Pilot values contain uncertainties due to problems with pumps, air in the tubes, difficulties in manually reading levels and degradation of some chemicals, to mention a few sources for errors. Also, the design values for the future Henriksdal WWTP are uncertain since they are theoretical values derived process calculations based on design manuals and experiences from other WWTPs.

This year, two different external carbon sources were tested. Glycerol, which have been tested previous years was used in the pilot for about 42 weeks in 2020. A shorter 10 weeks trial with ethanol was conducted before switching back to glycerol as external carbon source for the remainder of the year. A pause in dosage of external carbon source week 41 to 48 has been excluded from resource calculations. Comparing the two carbon sources, ethanol and glycerol were similar in consumption per day and per kg N removed. The consumption of external carbon source, expressed as COD, was the same as for the future Henriksdal design, even though the nitrogen load on the biology was 18% higher and set-point for nitrate in the effluent was >25% lower in the pilot.

The daily iron/metal consumption was about half compared to the future Henriksdal design, although the phosphorus load to the pilot was about 50% higher compared to design values. This is reflected in the mole Fe to mole P consumption, which was lower in the pilot compared to the design. The effluent phosphate concentration was below the target concentration and the measured values were about half of those set in the design calculations. The low metal consumption is explained by the EBPR activity in the pilot. This year, both consumption of precipitation chemicals and effluent phosphate were the lowest measured in the history of the pilot.

The consumption of cleaning chemicals was considerably lower than the pilot design values and the future Henriksdal design. The effort to reduce the cleaning chemicals is described in chapter 6.5.3. In total, the amount of chemicals used for RC was around the same as in the design, since RC was performed twice in 2020. Chemical consumption for MC, however, was only 22% of pilot design values for oxalic acid, 31% for citric acid and 24% for hypochlorite. Design values are based on the design inflow of 3.2 m^3/h while the inflow in 2020 was 4.4 m^3/h. Since MC frequency is based on the volume of water passing the membranes this 38% increase in inflow should also render a 38% increase in chemical consumption. Based on the actual inflow, the chemical consumption for MC was reduced by 84% for oxalic acid, 78% for citric acid and 82% for hypochlorite. The results indicate that costs and environmental impact can be significantly reduced in the full-scale plant by reducing the consumption of cleaning chemicals.

The air demand for the biological treatment cannot be compared in a reasonable way since the configuration of the reactors are completely different with a water depth of 12 m in the full scale, compared to 3.2 m in the pilot. The airflow to the membranes was 145% larger than the design value for future Henriksdal, which could mainly be explained by the fact that the design values given in the table are based on design inflow and membrane aeration at LEAP-lo.

Table 41. Chemical consumption during 2020.

Resource	Unit	Value pilot	Future H-dal design	Value pilot/ scaled future H-dal design
External carbon source (Methanol, not used in pilot)	kg COD/d		12 000	-
	g COD/g N		-	-
External carbon source (Glycerol, 253 days)	kg COD/d	1.75	-	-
	g COD/g N*	0.60	-	-
External carbon source (Ethanol, 68 days)	kg COD/d	1.80	-	-
	g COD/g N*	0.55	-	-
External carbon source (COD)	kg COD/d	1.80	12 000	100%
Iron (FV)	kg Fe/d	0.7	10 000	47%
	mole Fe/mole P		2.8	-
Aluminium (BR6)	kg Al/d		-	-
	mole Al/mole P	0.05	-	-
Metals (Iron + Aluminium)	kg Me/d	0.75	10 000	50%
	mole Me/mole P	0.95	2.8	
Citric Acid (51%)	L/d	0.066**	1 100	40%
Sodium hypochlorite (12%)	L/d	0.150***	1 600	63%
Oxalic acid (8%)	L/d	0.27**	5 000	36%
Aeration biology	m^3/d	1 734	1 600 000	726%
Aeration MT	m^3/d	648	3 000 000	145%

* N removed in total, from inlet to effluent.

** Number of MCs with each acid, multiplied with time settings and number of backpulses using settings of flowrate of chemical. Measured consumption for two RC preformed with each acid.

*** Number of MCs with hypo was multiplied with time settings and number of backpulses using design flowrate of chemical. Measured consumption for four RC preformed in total (2 per MT).

7 Related publications

Several activities such as world water congress were canceled in 2020 due to CIVID-19. Further, the general cooperation with different partners at the R&D-facility had to be restricted. As such, fewer publications were generated from the project during 2020.

The project had one contribution at a digital international conference held in 2020:

- Karlsson J., Carranza Muñoz A. (2020) How low can you go – retention time reduction until microbial inhibition in a thermophilic digester treating wastewater sludge. IWA Latin American Meetings on Anaerobic Digestion, 22 October – 12 November 2020.

The project hosted three master students and following master theses were published:

- Fridh, B. (2020) Mapping of spontaneous biological phosphorus removal in MBR-process. KTH Royal Institute of Technology. TRITA-CBH-GRU; 2020:116
- Roberts, R.A. (2020) Mapping spontaneous biological phosphorus removal observed in a membrane bioreactor process without the anaerobic condition. KTH Royal Institute of Technology, TRITA-ABE-MBT 20734.
- Sellin, J. (2021) Comparison of mesophilic and thermophilic anaerobic sludge digestion at Hammarby Sjöstadsverk MBR pilot plant for wastewater treatment. Master thesis in Environmental Science 2021:09, SLU Swedish University of Agricultural Sciences, Uppsala.

A related project, financially supported by Svenskt Vatten Utveckling, investigated what happens to different types of micro pollutants in a WWTP with membrane technology was published early 2021:

- Närhi, K., Westling, K., Andersson, S., Baresel, C. & Wahlberg, C. (2021) *Micro pollutants in wastewater treatment plant with MBR-process. Comparison with conventional treatment plant and calculation of impact on recipient*. SVU-report 2021-2.

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