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

Pilot Henriksdal 2040. Results from 2021-2022

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In cooperation with

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Preface

This report presents work performed mainly during 2021 but also the first months of 2022, 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, 6 and 7 the reports are in English, see Andersson et al. (2019; 2020; 2021a; 2021b).

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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 2013 and was reconstructed to its current configuration in 2016. In 2017 the MBR pilot was supplemented with a sludge treatment line to study different aspects of sludge digestion.

During 2021, the MBR-pilot was operated at a fixed inflow of 4.1 m^3/h, which is 37 % higher than the design average flow, with externally provided glycerol as well as internally produced VFA as carbon source for post-denitrification. Aluminum (PAX) was used 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 0.07 mg TP/L, respectively, which means that the effluent requirements were met also this year. To achieve this, 8.6 g Fe^{2+}/m^3 and 0.9 g Al^{3+}/m^3 was required. During flux enhancer trials a total of 17.8 g iron (Fe^{2+} + Fe^{3+})/m^3 was added. The glycerol dose was equivalent to 17.3 g COD/m^3 and for internally produced VFA the dose equivalent was 15.5 g COD/m^3. The slightly higher consumption of phosphorus precipitation chemicals compared to 2020, 1.29 mole metal per mole of phosphorus removed, was mainly due to a lower enhanced biological phosphorus removal (EBPR) activity in 2021. In 2021 the phosphorous release rates were low during the spring and showed < 1 g PO_4-P/kg VSS,h in June but recovered in the summer with 5.5 g PO_4-P/kg VSS,h in July after the defoaming agent dosing was stopped. The iron and aluminum content in the activated sludge was 6.2 and 0.7 %, respectively. Average total sludge age during 2021 was 17.2 days and average aerated sludge age was 7 days. Nitrification was always complete with ammonia concentrations below 2 mg/L except week 25.

Test with use of internally produced VFA as carbon source showed that the specific COD consumption was almost the same as for glycerol when comparing the yearly average from 2021 and 2020. Effluent nitrate and total nitrogen removal was similar during the trial with VFA as the rest of the year, when glycerol was used.

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 sodium hypochlorite. The membranes were operated with an average net flux around 21 to 25 L/(m^2·h) but starting from week 25, the flux was increased to 30 L/(m^2·h) which is the design net max flux of the full scale MBR in Henriksdal and was tested in the pilot for 25 weeks. The net TMP varied between 49 and 218 mbar for MT1 and between 51 and 146 mbar for MT2. TMP was reduced after each recovery cleaning (RC) with hypochlorite, but the effect did not last long.

The permeability was generally above 200 L/(m^2·h·bar) throughout 2021-2022 for both membranes. Recovery cleanings were done twice with hypochlorite and once with acids during 2021. During 2022 a final RC, first with hypochlorite then with acids was carried out. The first RC for MT1 resulted in a clear increase in permeability after cleaning. For MT2 the major increase in permeability was the result of a citric acid MC (one week after the hypochlorite RC). The RCs at the end of 2021 and in March 2022 had clear but smaller positive impact on permeability. 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. As a result of the tough operational strategy from week 25 2021, permeability decreased quite quickly after RCs.

MT2 reached a stable level around 250-300 L/(m^2·h·bar) while MT1 decreased even more, to as low as around 200 L/(m^2·h·bar).

Emission of chlorinated compounds in the off-gas ventilation were measured during the final sodium hypochlorite recovery cleaning. The emission process was slower than expected and generally no clear sign of attenuation of emissions was observed during the 21 hours of sampling. Although composite samples of several hours during the night are not providing enough details, it was concluded that the emissions can be harmful during the entire RC process from an exposure perspective. Trichloramine peaked at 36 times the recommended limit, chlorine gas at 73 % of the short-term exposure limit (15 min exposure), and chloroform at 9 % of the occupational exposure limit (8-hour workday average).

To follow up previous measurements of greenhouse gases nitrous oxide (N_2O) and methane (CH_4), a new campaign was performed during several months in 2021. Generally, emissions observed in 2021 were significantly higher than in previous campaigns in the pilot and especially high N_2O-emissions from the membrane-tank could be identified. No clear reason could be identified but the increased incoming load with a maintained effluent quality and a "better" sampling setup may partly be an explanation.

In collaboration with Kemira, tests with a flux enhancer product were performed in 2021. However, no obvious positive or negative change in permeability due to dosing of flux enhancer was possible to identify based on continuously monitored process parameters and commonly observed variations in permeability and effect of membrane cleaning.

As the formation of foam is a common phenomenon in MBR plants, tests with an antifoaming agent were done by dosing in batches and continuously to the biological treatment during the period of heavy foaming (March-June). Even if foaming was not avoided, a good reduction and control of foaming could be achieved. An optimal effect was achieved with continuous dosages of > 10 ppm. However, even though the product has shown to have a positive effect in the MBR-pilot, a permanent use in full-scale may not be economically feasible due to the high consumption.

Test with a reduced RAS flow from the design value of $4\times Q_{in}$ to $2\times Q_{in}$ was done with the aim to reduce energy consumption for pumping. A reduced RAS flow would however imply an increased sludge concentration in the membrane tanks, which may have negative effects on the membrane performance with more clogging and consequently increased aeration for membrane scouring and need for more frequent membrane cleaning. However, no negative effects of the reduced RAS-flow could be seen on the membrane performance.

During 2021, tests with a transition from mesophilic to thermophilic anaerobic digestion, dewatering of digested sludge after mesophilic and thermophilic digestion, and thermophilic digestion at high organic loading rate (OLR) and low hydraulic retention time (HRT) were performed in the sludge pilot. Results show that the transition from mesophilic to thermophilic digestion can be done without any major problems if the load was reduced during the most critical temperatures and that stable operation was achieved after 10-12 days. Evaluating the dewatering of mesophilically and thermophilically digested sludge was more difficult and no clear differences could be observed. However, it was concluded that used methods for determining dewaterability or optimal polymer dose are not reliable. Trials with high organic loading rate at thermophilic digestion showed that the digester performance could be maintained up to an OLR of around 4 kg VS/m^3, d and an HRT of 12 d. When the load is further increased and HRT decreased, the performance in terms of VS reduction and biogas-/methane production decreased although the reactor operation was still stable.

The overall resource consumption in the pilot showed that the consumption of glycerol was the same as for the future Henriksdal design, even though the nitrogen load in the pilot was 21 % higher and the average effluent total nitrogen concentration was 3.9 mg TN/L compared to the design of 6 mg TN/L. Also, the iron/metal consumption was 73 % of the future Henriksdal design, although the phosphorus load to the pilot was about 50 % higher compared to design values and effluent phosphate concentrations were below the target concentration. This is mainly explained by the EBPR activity in the pilot. Also, the consumption of cleaning chemicals was lower than the future Henriksdal design although the inflow to the pilot was 30 % higher than design.

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 FoU-anläggningen Hammarby Sjöstadsverk, som ligger i anslutning till Henriksdals reningsverk. MBR-piloten togs i drift 2013 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 2021 kördes MBR-piloten med ett fast inflöde på 4,1 m^3/h, vilket är 37 % högre än det designade medelflödet, med externt tillförskaffad glycerol och internt producerad VFA-kolkälla för efterdenitrifikation. Aluminium (PAX) användes i stället för trevärt järn (PIX) som komplement till tvåvärt järn ($FeSO_4$) för fosforutfällning. Detta gjordes för att testa driftstrategin för den första MBR-linjen i Henriksdals reningsverk. Medelhalter av kväve och fosfor i utgående vatten var 3,9 mg TN/L respektive 0,07 mg TP/L, vilket innebär att utsläppsvärden uppfylldes även i år. För att uppnå detta krävdes 8,6 g Fe^{2+}/m^3 och 0,9 g Al^{3+}/m^3. Under försök med fluxförhöjare tillsattes totalt 17,8 g järn (Fe^{2+}+ Fe^{3+})/m^3. Glyceroldosen motsvarade 17,3 g COD/m^3, och för användning av internt producerad VFA motsvarande dosen 15,5 g COD/m^3. Den något högre förbrukningen av fosforfällningskemikalier jämfört med 2020, 1,29 mol metall per mol avlägsnad fosfor, berodde främst på en lägre bio-P aktivitet under 2021. År 2021 var fosforsläppshastigheten mycket låg under våren, ex. < 1 g PO_4-P/kg VSS,h i juni men återhämtade sig under sommaren med t.ex. 5,5 g PO_4-P/kg VSS,h i juli, efter att doseringen av skumdämpare stoppades. Järn- och aluminiumhalten i det aktiva slammet var 6,2 respektive 0,7 %. Genomsnittlig total slamålder under 2021 var 17,2 dagar och luftad medelslamålder var 7 dagar. Nitrifikation var alltid komplett med utgående ammoniumkoncentrationer under 2 mg/L, förutom vecka 25.

Tester med användning av internt producerad VFA som kolkälla visade att den specifika COD-förbrukningen var nästan densamma som för glycerol när man jämförde årsgenomsnittet från 2021 och 2020. Utgående nitrat och total kvävereduktion var liknande under försöket med VFA som resten av året, då glycerol användes.

Liksom tidigare år rengjordes membranen i membrantank 1 (MT1) med oxalsyra och membranen i MT2 med citronsyra. Båda membranen rengjordes också med natriumhypoklorit. Membranen kördes med ett genomsnittligt flux på 21 till 25 L/(m^2·h), men med startvecka 25 testades fullskaledesignens maximala nettoflux på 30 L/(m^2·h) i piloten under 25 veckor. Netto-TMP varierade mellan 49 och 218 mbar för MT1 och mellan 51 och 146 mbar för MT2. TMP reducerades efter varje återhämtningsrengöring (RC) med hypoklorit, men effekten varade inte länge.

Permeabiliteten var generellt över 200 L/(m^2·h·bar) under hela 2021–2022 för båda membranen. Återhämtningsrengöringar gjordes två gånger med hypoklorit och en gång med syror under 2021. Under 2022 genomfördes en slutlig RC, först med hypoklorit sedan med syror. Den första RC för MT1 resulterade i en tydlig ökning av permeabiliteten efter rengöring. För MT2 var den största ökningen av permeabiliteten resultatet av en citronsyra-MC (en vecka efter hypoklorit-RC). Följande RC i slutet av 2021 och i mars 2022 hade tydliga men mindre positiva inverkan på permeabiliteten. Före den första RC var permeabiliteten högre för MT1 (rengöras med oxalsyra) jämfört med MT2 (rengöras med citronsyra). Efter de första RC hade båda membranen liknande permeabilitet. Som ett resultat av den tuffa driftstrategin från vecka 25 2021 minskade permeabiliteten ganska snabbt efter RC. MT2 nådde en stabil nivå runt 250-300 L/(m^2·h·bar) medan MT1 sjönk ytterligare till som lägst 200 L/(m^2·h·bar).

Utsläpp av klorerade föreningar mättes under den slutliga återställningsrengöringen med natriumhypoklorit. Utsläppsprocessen var långsammare än förväntat och generellt sett observerades inga tydliga tecken på dämpning av utsläppen under provtagningens 21 timmar. Även om sammansatta prover på flera timmar under natten inte ger tillräckligt med detaljer, drogs slutsatsen att utsläppen kan vara skadliga under hela RC-processen ur ett exponeringsperspektiv. Exempelvis nådde trikloramin sin topp vid 36 gånger den rekommenderade gränsen, klorgas vid 73 % av korttidsexponeringsgränsen (15 min exponering) och kloroform vid 9 % av den yrkesmässiga exponeringsgränsen (genomsnittlig arbetsdag på 8 timmar).

För att följa upp tidigare mätningar av växthusgaserna lustgas (N_2O) och metan (CH_4) genomfördes en ny mätkampanj under flera månader i 2021. Generellt sett var utsläppen som observerades 2021 betydligt högre än i tidigare kampanjer och särskilt höga N_2O-utsläpp från membrantanken kunde observeras. Någon tydlig orsak kunde inte identifieras men den högre inkommande belastningen med bibehållna reningskrav och ett "bättre" provtagningsupplägg kan delvis vara en förklaring.

I samarbete med Kemira genomfördes tester med en fluxförhöjande produkt (flux enhancer). Ingen uppenbar positiv eller negativ förändring i permeabiliteten på grund av dosering av fluxförhöjare kunde dock identifieras utifrån de kontinuerliga processparametrar som övervakades och vanliga variationer i permeabilitet samt effekten av membranrengöring.

Eftersom skumbildning är ett vanligt fenomen i MBR-anläggningar genomfördes tester med ett skumdämpande medel som doserades i pulser och kontinuerligt till den biologiska behandlingen under perioden med kraftig skumbildning (mars-juni). Även om skumning inte upphörde helt så kunde en god minskning och kontroll av skumning uppnås. En optimal effekt konstaterades vid en kontinuerlig dos på > 10 ppm. Men även om produkten har visat sig ha en positiv effekt på skumning i MBR-piloten, framstår inte en permanent användning i fullskala som ekonomiskt genomförbar på grund av den höga förbrukningen.

Tester med reducerat RAS-flöde från $4\times Q_{in}$ enligt design till $2\times Q_{in}$ syftade till att minska energiförbrukningen. Ett minskat RAS-flöde skulle dock innebära en ökad slamkoncentration i membrantankarna, vilket kan ha negativa effekter på membranets prestanda med mer igensättning, vilket i sin tur kan leda till ökad luftning för membranrengöring och behov av tätare membrantvättar. Projektgruppen kunde dock inte observera några negativa effekter av det minskade RAS-flödet på membranets prestanda.

Under 2021 genomfördes tester med övergång från mesofil till termofil rötning, avvattning av rötslam efter mesofil och termofil rötning, samt termofil rötning vid hög organisk belastning (OLR) och låg hydraulisk uppehållstid (HRT) i slampiloten. Resultat visar att övergången från mesofil till termofil rötning kan ske utan större problem om den organiska belastningen minskades lite vid den mest kritiska temperaturen och att stabil drift uppnåddes efter 10-12 dagar. Att utvärdera avvattningen av mesofilt och termofilt rötat slam var svårare och inga tydliga skillnader kunde observeras. En slutsats var dock att de metoder som användes för att bestämma avvattningsbarhet eller optimal polymerdos inte framstår som tillförlitliga. Försök med hög organisk belastning vid termofil rötning visade att rötkammarens prestanda kunde bibehållas upp till en OLR på cirka 4 kg VS/m^3, d och en HRT på 12 d. När belastningen ökades ytterligare och HRT minskade, minskade prestandan vad gäller utrötningsgrad och biogas-/metanproduktion, även om själva reaktordriften fortfarande var stabil.

Den totala resursåtgången i piloten visade att konsumtionen av glycerol var densamma som för den framtida Henriksdalsdesignen, även om kvävebelastningen i piloten var 21 % högre och den genomsnittliga totala kvävekoncentrationen i utgående vatten var med 3,9 mg TN/L lägre än design på 6 mg TN/L. Järn-/metallförbrukningen var också 73 % av den framtida Henriksdalsdesignen, även om fosforbelastningen till piloten var cirka 50 % högre jämfört med designvärden och utgående fosfatkoncentrationerna låg under målkoncentrationen. Detta förklaras främst av EBPR-aktiviteten i pilotprojektet. Dessutom var förbrukningen av rengöringskemikalier lägre än den framtida Henriksdalsdesignen även om inflödet till piloten var 30 % högre än designen.

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	Atmosphères Explosibles
BOD_7	Biochemical Oxygen Demand, 7 days (mg/L)
BR1 to BR6	Biological reactor 1 to 6, sampling points
CAS	Conventional activated sludge process
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 mass of microorganisms (kg BOD_7/kg SS, d)
Flux	Flow rate per unit area (L/(m^2·h) or lmh). 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 solids 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 through which 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)
VFA	Volatile Fatty Acids
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/, retrieved 2023-02-01.

1 Introduction

This report presents the results from year 2021, including the first three months of 2022 (project year 8) of the pilot scale trials with 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 2014-2020 have been presented in previous, separate reports.

In the initial chapters (2-3), the project background and the configuration of the pilot plant is described. An overview of the experimental plan is presented in chapter 0, 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 2021, and thus chapter 2 and 3 are similar to the previous report. However, the flow rates and loading rates to the pilot for 2021 are summarized in section 3.3. Also, the method section is like the description in the previous reports.

2 Background

Within the project Stockholms 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) SVOA's WWTP in Bromma (which is already over loaded with very limited space available for extension) will be decommissioned in 2029 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 tighten 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 preliminary 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, was 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 during high flows. 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 dilute 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, 2010). In USA, China, Japan, South Korea, France, Great Britain and Spain, there are several large MBR plants (50,000-80,000 PE) which

have been in operation for 5-10 years (Judd and Judd Limited, 2017). The largest MBR plant in operation today is 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, retrieved 2022-04-22), is also the largest MBR installation using SUEZ's ZeeWeed ultrafiltration (UF) hollow-fiber membranes, which are the same 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, e.g., *Broad Run* and *King William County* in Virginia, *Ruidoso* in New Mexico and *Cauley Creek* and *Yellow River* in Georgia, that reach very low effluent nutrient 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, odor, 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 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 design mean flow of around 3.2 m³/h. Primary treatment comprise of a fine screen (6 mm openings), pre-aeration, a primary settler, and a fine sieve (2 mm openings). The biological treatment consists of pre- and post-denitrification followed by two parallel membrane tanks. The return activated sludge (RAS) passes through 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.

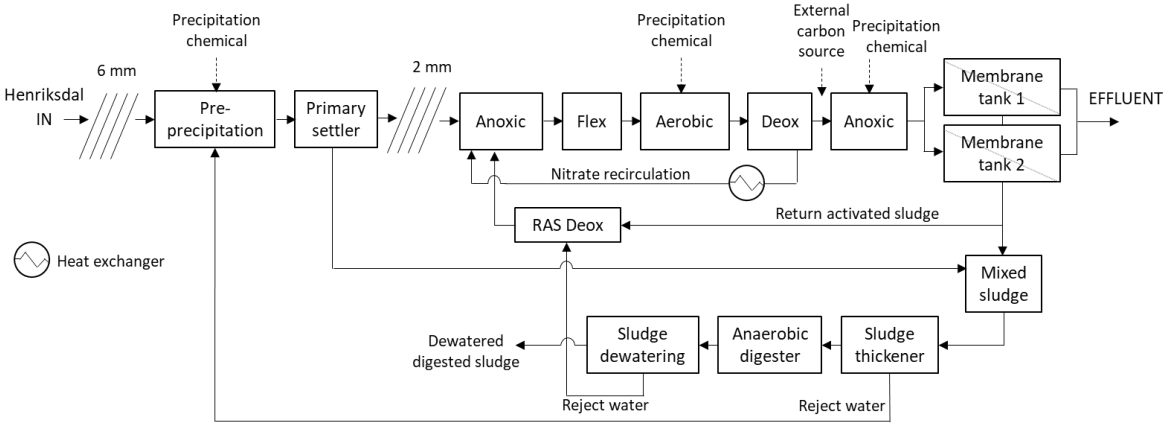


Figure 1. Flow scheme of the pilot WWTP.

The reactor volumes of the pilot plant and the function of each reactor is 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 (m³)	Future H-dal (m³)	Scale factor H-dal/Pilot	Specification
Pre-treatment				
PA (sand trap)	0.7	2 460	-	Pre-aeration. Dosing point 1: Fe^{2+}.
SED	3.3	30 000	9 200	Primary settler. Withdrawal of primary sludge.
Membrane bioreactor (MBR)				
BR1	4.8	33 500	7 000	Anoxic conditions. Stirred. Pre-denitrification.
BR2	4.8	33 500	7 000	Anoxic conditions. Stirred. Pre-denitrification.
BR3	4.8	40 000	8 300	Flex. Stirred/(aerated). Pre-denitrification/(nitrification).
BR4	4.8	31 000	6 500	Aerated. Nitrification. Dosing point 2: Fe^{2+}.
BR5ox	1.5	10 000	6 700	Aerated. Nitrification.
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.
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).

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

*Maximum volume. The actual 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 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 since 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 18.5 °C, which is 2.6 °C higher than the influent wastewater to Henriksdal (15.9 °C). The daily average temperature in the pilot inlet varied between 12.1 °C and 21.9 °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 16.9 °C.

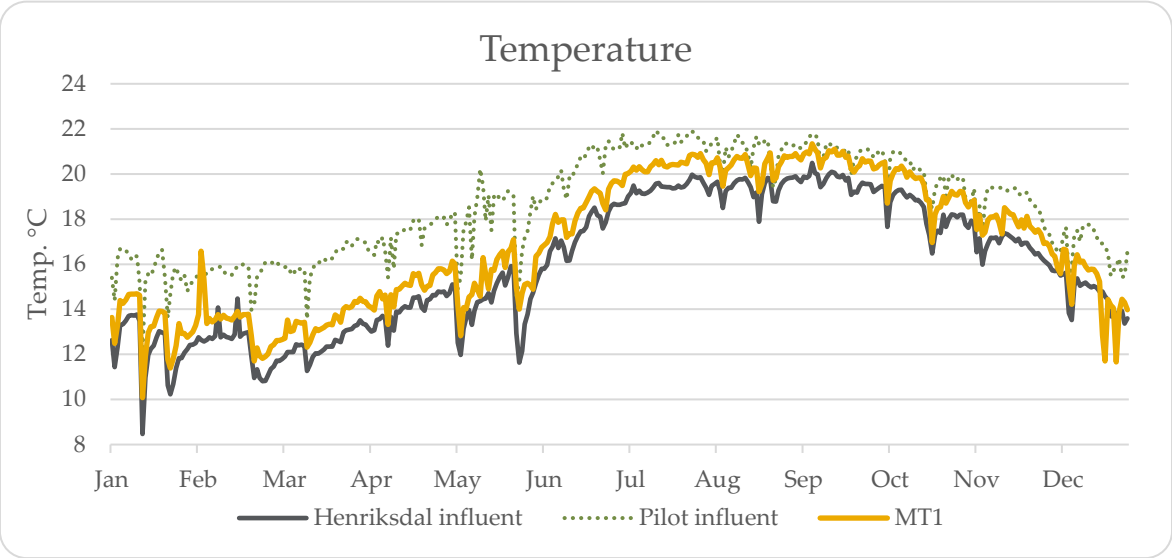


Figure 2. Influent temperatures to the MBR pilot (dotted line) and Henriksdal WWTP (black line) together with temperature in the Membrane tank 1 (MT1) during 2021.

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 ion dosing and a vertical flow primary settler, with a surface area of 1.13 m^2 and a water depth of 4.3 m (scale 1:9 200 compared to the future Henriksdal design), followed by a 2 mm punch hole drum sieve before the biology (Figure 3).

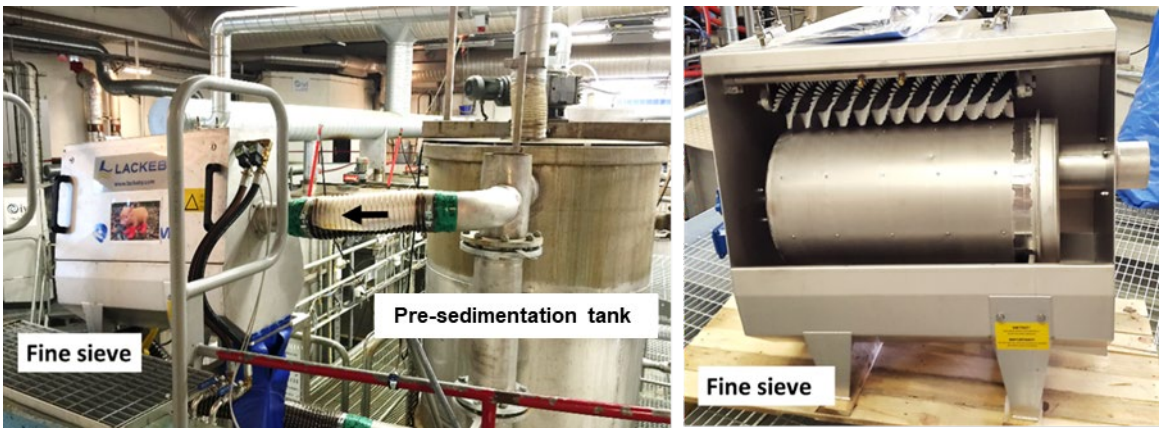


Figure 3. Photo of the fine sieve installation.

3.1.3 Biological treatment

The biological treatment consisted of six identical biological reactor tanks, BR1-6 (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 (design flow 4 × 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 came from 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 membranes 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 × 1.0 m × 0.34 m) consisting of three membrane modules each (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 stretched 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 area 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 in 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.

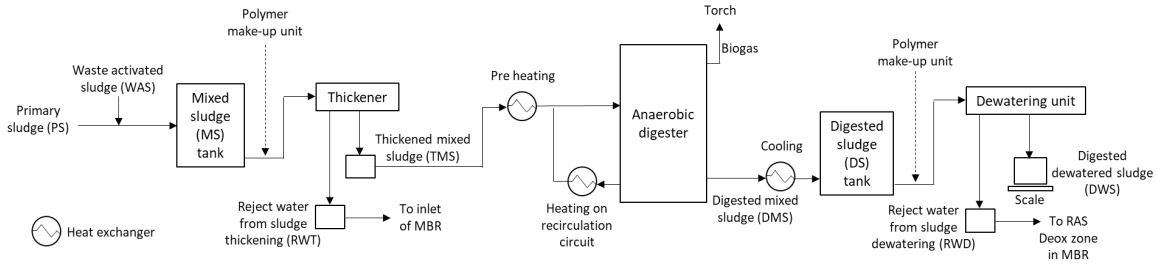


Figure 6. Process set-up for the sludge treatment line.

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.

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 2021 the volume was kept between 3.4 and 5.0 m^3 (instead of 5.7 m^3) to reach different HRT and OLR. 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).

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.

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 (screw press) 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.

Due to the technical design of the dewatering unit, it is challenging to maintain a continuous operation. Therefore, during 2021 the dewatering unit was only operated during two planned dewatering trials.

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

3.3 Flow rate and load

Mean values for flow rates and loading rates of the pilot wastewater and sludge treatment lines during 2021 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 flow rate in 2021 was 4.12 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 (year 2021) and design data (year 2040) for the future Henriksdal WWTP.

Parameter	Unit	Value Pilot 2021	Design Pilot	Design future H-dal	Value Pilot/ Design H-dal[i]
Flow rates					
Average influent flow rate, Q_{in}	m^3/h	4.12	3.16	20 880	0.020 %
Design average flow rate, Q_{dim}	m^3/h		3.32	21 960	0.015 %
Max flow rate	m^3/h	5.5	5.44	36 000	0.015 %
Min flow rate	m^3/h	1.8	1.8	11 600	0.016 %
Nitrate recirculation flow rate	m^3/h	5.1-13.1	3.8-13.3	-	-
Nitrate recirculation flow rate	× Q_{in}	2.8	1.2-4.2[ii]	0-4	-
RAS flow rate	m^3/h	4.1-22.6	3.6-19	-	-
RAS flow rate	× Q_{in}	3.1	1.1-5.9[ii]	4 (3-5)	100 %
Temperatures					
Temperature influent	°C	18.5	-	-	-

Parameter	Unit	Value Pilot 2021	Design Pilot	Design future H-dal	Value Pilot/ Design H-dal[i]
Temperature biology	°C	16.9	-	-	-
Incoming load					
BOD_7 influent	mg/L	220	206[iii]	216	102 %
SS influent	mg/L	276	201[iii]	280	99 %
TN influent	mg/L	41	44[iii]	37	111 %
TP influent	mg/L	5.2	5.7[iii]	4.9	106 %
Primary settler (SED)					
BOD_7 reduction over SED	%	19	46	50[iv]	38 %
SS-reduction over SED	%	25	60	60[iv]	42 %
TN reduction over SED	%	1	10	10[iv]	10 %
TP reduction over SED	%	11	40	40[iv]	28 %
BOD_7 PTW	mg/L	169	112	108	156 %
SS PTW	mg/L	192	80	112	171 %
TN PTW	mg/L	40	40	33	121 %
TP PTW	mg/L	4.6	3.4	3.0	153 %
SS removed over SED	kg SS/d	7.4	13.3[v]	89 300	0.008 %
Primary sludge production	kg SS/d	12.8	17.2[v]	115 000	0.011 %
VS-concentration PS	% of TS	89	77	77	116 %
Biological treatment					
BOD_7-load PTW (at average flowrate)	kg BOD_7/d	16.7	8.6	57 500	0.029 %
Specific WAS-production [vi]	kg SS/kg BOD_7	1.07	1.02	1.02	105 %
WAS production, average	kg SS/d	17.8	8.8	58 600	0.030 %
VSS-concentration WAS	% of TSS	78	64	64	122 %
SS in biological tanks	mg/L	8 000	8 000	8 000	100 %
SS in membrane tanks	mg/L	11 600	10 000	10 000	116 %
Total sludge age	d	17.2	32.0	31.2	55 %
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)	$L/(m^2 \cdot h)$	26.2	17.9	20.9	125 %
Net flux max	$L/(m^2 \cdot h)$	30.4	30.8	30	101 %
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]	$Nm^3/(h \cdot m^2)$	0.136	0.136	0.098	139 %
Specific air demand at Leap-Hi [vii]	$Nm^3/(h \cdot m^2)$	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 mesophilic trial (w.10-17), the thermophilic trial (w. 20-51) and high-OLR trial (w.39-51).

Parameter	Unit	Value pilot w.10-17 (Mesophilic)	Value pilot w.20-37 (Thermophilic)	Value pilot w.39-51 (High OLR)	Design future H-dal
Sludge into digester					
Flow into digester TMS	L/h	16.0	15.8	13.7	118 000
TS-concentration TMS	%	4.7	4.9	5.5	6.0
TS-load TMS	kg TS/d	18.2	17.3	17.3	172 000
VS-load TMS	kg VS/d	86.1	83.4	86.5	124 000
VS-load EOM	kg VS/d	0.0	0.0	0.0	44 000
Digestion					
Digester temperature	°C	37.1	54.7	54.5	55
Rector volume	m^3	5.0	4.3	3.4	38 000
Hydraulic Retention time, HRT	d	13.7	12.0	11.2	13[b]
Specific VS-load (OLR)	kg VS/m^3,d	3.1	3.5	4.4	3.3[b]
Organic degradation rate (ODR)	% of VS_{in}	54.1	51.8	51.8	42[b]
VFA/Alkalinity	mg CH_3COO^- eq/mg $CaCO_3$	0.06	0.07	0.09	-
Out of digester					
Flow Digested mixed sludge, DMS	L/h	17.4	15.5	13.2	123 000
TS-concentration DMS	%	2.0	2.6	2.7	3.9
VS-concentration DMS	% of TS	73.5	70.9	74.2	60
TS-load DMS	kg TS/d	8.6	9.7	8.7	124 000
VS-load DMS	kg VS/d	6.4	6.8	6.4	74 000
Specific biogas production	m^3/kg $VS_{deg.}$	0.53	0.83	0.74	1.0
Specific biogas production	m^3/kg $VS_{in.}$	0.30	0.37	0.40	0.42
Flow biogas	m^3/d	4.7	5.2	5.9	52 000[b]
Methane content biogas	%	52.1	60.2	59.1	65
TS-concentration DDMS	%	23.6[a]	28.1[a]	-	30
SS-concentration RWD	mg/L	544[a]	731[a]	-	900
Polymer consumption dewatering	g/kg TS	20.0[a]	11.1[a]	-	6-10

a) Data from dewatering trials over two weeks. At all other times the dewatering was deactivated.
b) Numbers excluding addition of external organic material to be comparable to data from the pilot.

3.4 Chemicals

During 2021, glycerol and internally produced VFA were used as dosed carbon sources (only one at a time) to the post-denitrification zone. Phosphorus was precipitated using iron(II)sulfate at one dosing point and aluminium(III)chloride in another dosing point for most of the year. During a separate trial, an iron-based flux enhancer product was tested. Another trial was conducted in which a defoaming agent was dosed to the biological treatment. 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

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. Because the tanks are already available, and that glycerol is not flammable, thus not requiring ATEX approved installations and special permits, it is an attractive candidate for post-denitrification. However, the efficiency of glycerol as a carbon source was not known and therefore it has been tested in the pilot since 2019. Glycerol was collected in 25 L containers from Scandinavian Biogas' storage. The measured COD concentration was about 850-1200 g COD/L.

During 2021, glycerol was used as carbon source for most of the year. Starting in November, internally produced carbon source (VFA) was used in the pilot.

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 (2019-2020)
- Ethanol (2020)

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 6.14 Resource consumption.

3.4.2 Internal carbon source

The high future demand of methanol in Henriksdal WWTP, which is designed to be around 9 m^3/d, will be major contributor to the total carbon footprint of the overall process, in addition to the costs and safety risks that its use entails (Metcalf & Eddy, 2014). Replacing this carbon source with an internally produced equivalent will improve the circularity of the process and improve use of internal resources which is of interest in this project.

A recent PhD project at Hammarby Sjöstadsverk focusing on producing high value VFA from co-fermentation of food waste and primary sludge performed denitrification trials that proved fermentation liquid containing VFA to be a better carbon source than methanol and acetic acid. The denitrification rate (mgNOx-N/gVSS·h) was almost three times higher than that of methanol (Owusu-Agyeman et al., 2020). Other studies have also shown the high potential of using VFA as electron donor for denitrification including the ones performed by Liu et al. (2018), where there is a continuous process of alkaline fermentation, dewatering and carbon source utilization with a TN reduction of around 85%. The system showed better economic advantages than a biogas system. Parchami et al. (2020), also found that the continuous production of VFA from sewage sludge and food waste could potentially replace the carbon source based on date of a WWTP in Sweden.

SVOA and other WWTP are currently co-founding a PhD student at IVL with supervision by SLU to conduct further research on the fermentation processes for producing a carbon source that could potentially replace part or the entire methanol demand. As a part of the project, a fermentation reactor was continuously producing a carbon source which was used in the post-denitrification zone of the MBR-pilot. Two types of carbon sources containing VFA were tested in the project: one produced from a mix of primary sludge and food waste (43 g COD/L, or 20 g VFA/L) and one from only primary sludge (8 g COD/L, or 6 g VFA/L), no pre-treatment was used. The results were compared to other external carbon sources previously used.

3.4.3 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 (PAX-XL60). Previous years, a combination of iron(II)sulfate and iron(III)chloride (PIX 111) was used.

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 16 and 57 g/L. For the batches used in the experiment, the iron content was determined by density measurement of 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.

3.4.4 Defoaming agent

As the biological treatment and especially the aerated tanks have experienced problems with excessive foaming at times, a defoaming agent (SNF FloFoam H16) was tested in the pilot during 2021. The trial included testing different dosing strategies (batch dosage and continuous dosage) at different locations with varying concentrations. The product was delivered as a solution.

3.4.5 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 2021, 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 50 and 69 g Cl_2/L during the year.

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).

3.4.6 Chemicals for enhanced membrane performance

To test possible improvement of the membrane performance, a flux enhancer product (produced by Kemira) was tested in the pilot during 2021. The flux enhancer consisted of a combination of iron chloride and polymers and was dosed in the biological treatment (BR6). The goal of using a flux enhancer product is to get an increase in flux which results in lower TMP and less need of cleaning (longer cleaning intervals) which in turn results in a net reduction in carbon footprint. The product was delivered as a solution.

3.4.7 Polymers

For sludge thickening Flopam EM 640 HIB (SNF) and Flopam EM 440 L (SNF) was used during 2021.

For dewatering of digested sludge Flopam EM 640 HIB (SNF) was used during 2021 for mesophilic sludge and Flopam EM 145 CT (SNF) for the 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 controlled locally. Implementation of the control has been carried out within the project, which provides great flexibility to adapt and optimize control.

4 Experimental plan year 2021-2022

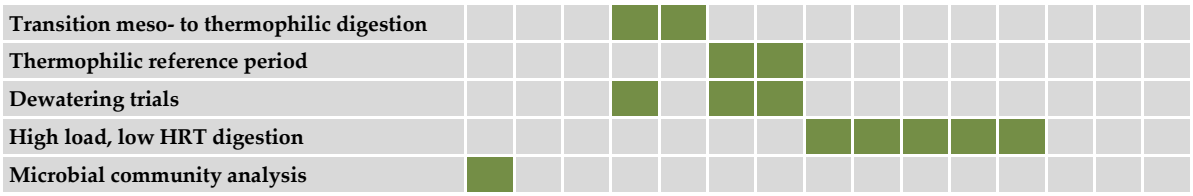
An overview of the experimental plan year 2021-2022 is presented in Table 4 and in more detail in later chapters of the report.

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 that point in time. 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 and 2020, 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 2021 the theme was to push the process towards a maximum performance with high load and high flux while maintaining low effluent concentrations and low resource consumption. In addition, a transition from mesophilic to thermophilic digestion was done.

Table 4. Experimental plan of year 2021.

	2021												2022		
Activity	J	F	M	A	M	J	J	A	S	O	N	D	J	F	M
High fixed inflow															
Reduced RAS-circulation to $2\times Q_{in}$															
High flux															
Evaluation of flux enhancer															
Evaluation of antifoaming agent															
Glycerol as carbon source															
VFA from sludge/food waste as carbon source															
Back pulse instead of relaxing															
Effluent phosphorous target 0.10 mg TP/L															
Al^{3+} with Fe^{2+} for P-removal															
P-release tests															
Leap-Medium															
Oxalic acid and citric acid comparison															
Demand driven hypochlorite MC															
Recovery cleaning															
Measurement of Cl-rich gas during RC															
Greenhouse gas measurement															
Energy modeling & impact assessment															
Sludge thickener in operation															
How low can we go - mesophilic															
Emptying and reseeding the digester															
Mesophilic reference period															



Minimizing the resource consumption, especially for membrane cleaning was continuously in the spotlight this year. The comparison of citric and oxalic acid continues and throughout the period a demand driven approach to the sodium hypochlorite MC was tested (instead of time-based cleanings). Recovery cleaning of the membranes was carried out in May to June 2021, November 2021 and a last time before the project ended in March 2022. Chlorine gas emissions from the process was measured during the last soaking of the membranes in sodium hypochlorite in March 2022.

During most of the year, the production goal for phosphorus and nitrogen was kept low, especially for phosphorus with a target effluent concentration of 0.10 mg TP/L.

As this was the final project year we wanted to carry out as many of the test on the waiting list as possible. Many of them were trials with different chemical additives. Aluminium chloride was dosed in combination with iron sulphate as precipitation chemical. This was tested 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.

Problems with foaming and scum have occurred in the pilot from time to time and was a concern for the full-scale as well. Therefore, an antifoaming agent was tested in pilot.

As mentioned, the performance of the membranes was tested at tougher conditions with higher flux and higher suspended solids concentrations. During this trial, a flux enhancer product was tested and evaluated.

An internally produced VFA (from food waste and primary sludge) was tested as carbon source in post-denitrification in the pilot.

The membranes can be operated with back-pulse instead of relaxing, but this has not been tested in the pilot before. As the programming was in place it was decided to try this operation mode for two months at the end of the project.

For the digestion a transition from mesophilic to thermophilic conditions was repeated a second time followed by a trial with high OLR and low hydraulic retention time.

To answer the question if the sludge dewaterability differs between mesophilic and thermophilic trials some dewatering tests were carried out.

Separate trials during this final year included the evaluation of the resource consumption by use of LCA and some process modelling to quantify energy demand and the potential effects of reducing the RAS flow rate to save energy. A long-term measurement campaign for nitrous oxide emission in the water treatment line was carried out as a complement to previous measurements.

5 Methods

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.

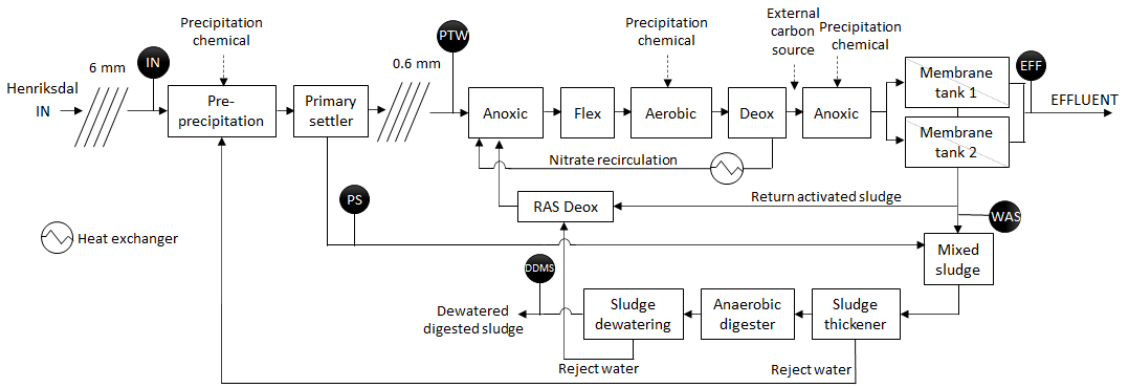


Figure 8. Sampling points in the pilot process marked as black circles (SLUDGE1 and SLUDGE2 sampling points not included in the figure).

Three different sampling strategies 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 4 M sulfuric acid to 100 parts sample volume, except for the samples analysed for TOC which were conserved with 2 M 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 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).

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 standardised cuvette tests. The daily composite samples were analysed according to Table 6. Additional analyses of daily composite samples or grab samples were also done 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 for total solids (TS (%)) and volatile solids (VS (%)) between 2-3 times per week. This includes 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	Flow rate (water)	Measure all recirculation flows
IN	Temperature	Measure the incoming wastewater temperature
IN	Flow rate (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	Flow rate (air)	Measure air flow
BR4	DO	Controlling Dissolved Oxygen
BR4	Flow rate (air)	Measure air flow
BR4	SS	Measure suspended solids concentration in activated sludge
BR5ox	DO	Controlling Dissolved Oxygen
BR5ox	Flow rate (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	Flow rate (water)	Effluent flow of permeate from membrane 1 and 2 (x2)
MT1/MT2	Flow rate (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	Flow rate	Measure flow rate of primary sludge
PS	TS	Measure total solids in primary sludge

WAS	Flow rate	Measure flow rate 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 level in anaerobic digester to detect foam
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, where permeation and relaxation = 16 minutes). The gross values are higher than the net, however the net better represents the average operation. All values for the parameters described below are given as net values in this report.

Flux: Flow rate 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 transport 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, several 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 (2010).

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 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:

$$ODR = \frac{(VS_{MS} - VS_{DMS})}{VS_{MS}} * 100\ \%$$

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 found in literature (Metcalf and Eddy, 2014).

6 Results and discussion

The following sections present and discuss specific results for the different activities and are therefore introduced by the activities time frame according to Table 4.

6.1 Primary treatment

	2021												2022		
Activity	J	F	M	A	M	J	J	A	S	O	N	D	J	F	M
High fixed inflow															

6.1.1 Inlet screen

Different punch hole screens, from 2 mm to 6 mm, have been used for the screening of coarse solids since the start of the project. Since 2017, a 6 mm size has been used to increase the primary sludge production which was required to meet the design loading rate 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 resulted in the sludge being emptied into the inlet channel where the inlet pumps to Hammarby Sjöstadsverk 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 even though the 6 mm screen was used. The main contributing factor was thought 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). In 2021, the inlet concentrations were still affected by the corona pandemic and by technical problems with the inlet pumps resulting in many interruptions in the inflow (> 500 hours in total).

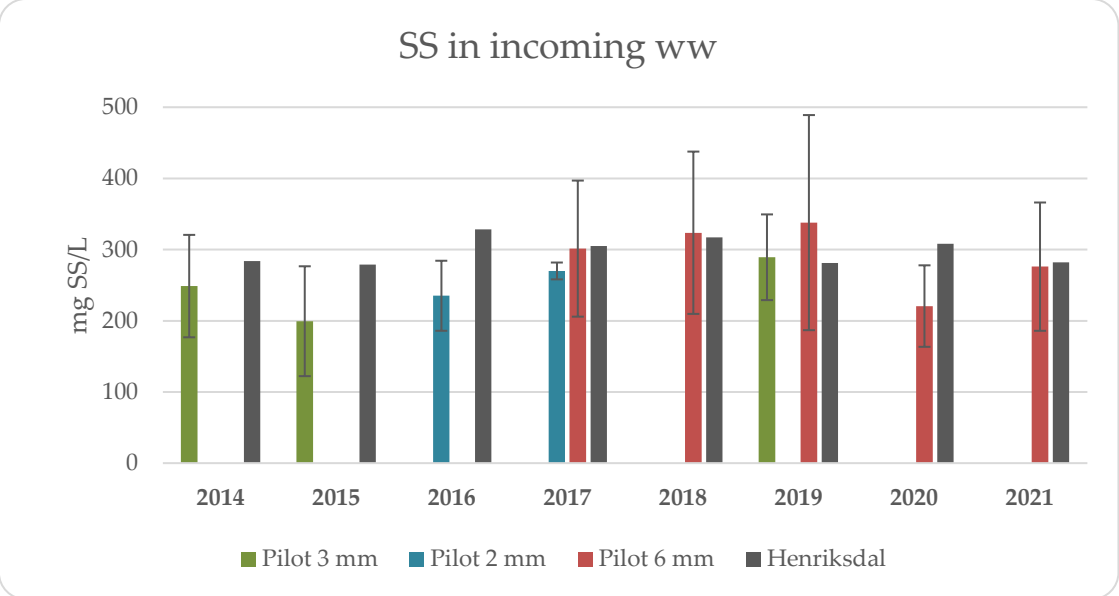


Figure 9. Incoming SS-concentration with standard deviation to the pilot after passing through 2-, 3- or 6-mm punch hole screens and to Henriksdal (3-mm step-screen). Henriksdal data are from the yearly Miljörapport.

6.1.2 Efficiency of primary settler

The primary settling volume in the pilot of 3.3 m^3 and its capacity, is smaller than it should be according to the design value of 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 2021, three additional aspects affected the efficiency of the primary settler. First, the inflow to the pilot was high, 4.1 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. Third, the inlet pumps stopped unusually often this year.

Table 9. Reduction over the primary settler and primary sludge production in the pilot compared to Henriksdal WWTP 2021 and the future design of Henriksdal WWTP.

Parameter	2021	2020	2019	2018	2017	Henriksdal WWTP 2021[a]	Design future Henriksdal
Inflow (m^3/h)	4.12	4.37	3.56	3.50	3.40	12 000	22 000
PE (1 PE = 70g BOD_7/d)	311	292	352	318	314	890 000	1 600 000
Reduction of BOD_7 (%)	19	19	27	25	30	56	50
Reduction of SS (%)	25	32	36	35	37	57	60
Reduction of TN (%)	1	0	4	1	4	2[b]	10
Reduction of TP (%)	11	8	12	10	14	30	40
Reduction of TOC (%)	13	10	20	18	17	35	-
PS-production (kg TS/d)		21.7	18.5	16.1	13.1	60 300[c]	115 000
PS-production (g TS/PE/d)		74	53	51	42	68	72

a) At high flows Enhanced Pre-Precipitation using Al^{3+} is used for half of the primary clarifiers.
b) Sampling point for the Sickla inlet (SIN), does not include 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, for example w3 (Figure 10). No obvious cause could be identified.

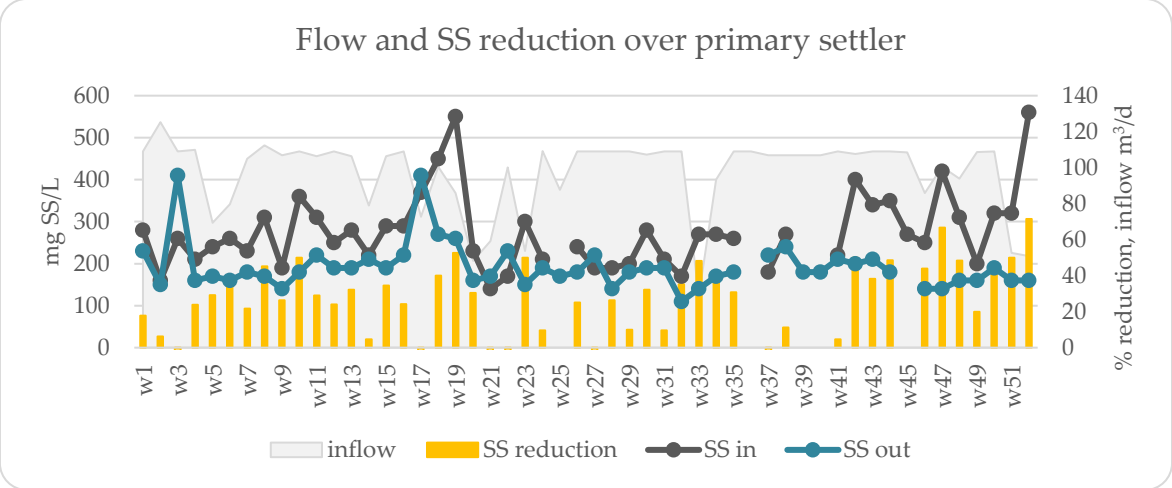


Figure 10. Reduction of SS over the primary settler after a 6 mm screen.

6.1.3 Screen and sieve – effect on trash content

During 2021 the pilot line was operated with a 6 mm punch hole inlet screen (see 6.1.1 Inlet screen) and 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, fibers 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. In 2021, seven measurement of trash content was made. Compared to 2020, the concentration was higher of both larger particles (> 2 mm) and smaller particles (1-2 mm), but still low and in the same range as previous years. According to the membrane supplier, the trash content with a 2 mm sieve should not exceed 2 mg/L. Since 2016 no measurements have exceeded this value.

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. 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 June 2020, 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 indicated 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 screen/sieve configurations. According to the membrane supplier, the trash content with a 2 mm sieve should not exceed 2 mg/L. Screen is located at inlet pump and fine sieve is located before the biological treatment (after primary sedimentation).

Sieves and hole size	Time period	No of analyses	1 mm sieve mg/L	2 mm sieve mg/L
3 mm screen	Dec 2013	2	11.6 ± 5.4	1.0 ± 0.7
2 mm screen	Nov 2016 – Feb 2017	4	6.4 ± 2.4	1.1 ± 0.7
6 mm screen + 0.6 mm fine sieve	Mar 2017 → Dec 2017	8	4.1 ± 3.8	0.6 ± 0.3
6 mm screen + 0.6 mm fine sieve	May 2017 → Dec 2017	6	2.2 ± 1.6	0.6 ± 0.3
6 mm screen + 0.6 mm fine sieve	Jan 2018 → Nov 2018	6	2.0 ± 1.1	0.6 ± 0.5
3 mm screen + 0.6 mm fine sieve	Dec 2018 – Mar 2019	1	2.3	2.0
6 mm screen + 0.6 mm fine sieve	Apr 2019 – Dec 2019	2	5.1 ± 5.4	1.5 ± 2.0
6 mm screen + 0.6 mm fine sieve	May 2020	1	0.4	0.2
6 mm screen + 2 mm fine sieve	July 2020 – Dec 2020	3	1.4 ± 0.5	0.3 ± 0.2
6 mm screen + 2 mm fine sieve	Feb 2021-Dec 2021	7	2.3 ± 0.1	0.5 ± 0.3

6.1.4 Pre-treated wastewater

The quality of the pre-treated wastewater (PTW) is presented in Table 11. The concentrations measured in the pilot were higher than the corresponding concentrations measured in Henriksdal WWTP for SS, BOD_7, TN and TP. This is mainly due to the poor performance of the primary settler and was also observed previous years. The concentrations of SS, BOD_7, TN and TP in pre-treated water were also higher in the pilot than the design values given for the future plant. However, it should be noted that the design values are relatively low since they are 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 2021 and the design data for the future Henriksdal WWTP.

Parameter	Value Pilot 2021	Value Henriksdal 2021	Design future Henriksdal	Value pilot/ Design future H-dal
Pre-treated wastewater (PTW) – into biological treatment				
BOD_7 (mg/L)	169 ± 35	96	108	156 %
SS (mg/L)	192 ± 55	133	113	170 %
TN (mg/L)	40 ± 5	38	33	121 %
TP (mg/L)	4.6 ± 0.6	3.5	3.0	143 %
Fe (mg/L)	7.9 ± 2.0	9.3	12	65 %
Alkalinity (mg/L)	255	245	250	102 %
BOD_7/TN (mg/mg)	4.2	2.5	3.3	127 %

6.2 Nitrogen removal

	2021												2022		
Activity	J	F	M	A	M	J	J	A	S	O	N	D	J	F	M
High fixed inflow															
Reduced RAS-circulation to $2\times Q_{in}$															
Evaluation of antifoaming agent															
Glycerol as carbon source															
VFA from sludge/food waste as carbon source															

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. Two out of 52 weekly composite samples were above the target concentration of 6 mg TN/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 2021.

Parameter	Limit	Mean	Min	Max	No. of weekly samples
TN PTW (mg/L)	-	40	24	52	52
TN EFF (mg/L)	6	3.9	2.8	10.0	52

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 TN/L, as also shown previous years. In the beginning of the year there were problems with nitrogen sensors (mixed problems with both ammonium and nitrate sensors) and the aeration system. In week 4 BR2-BR5 was emptied to replace the bottom diffusors in the aerated tanks. Total nitrogen peaked in week 4-5 because of the low suspended solids concentration after the maintenance work in combinations with faulty sensors making the carbon source dosage too low. A smaller ammonium peak occurred in week 24-25 due to manually limited aeration because of problems with foaming in the aerated tanks.

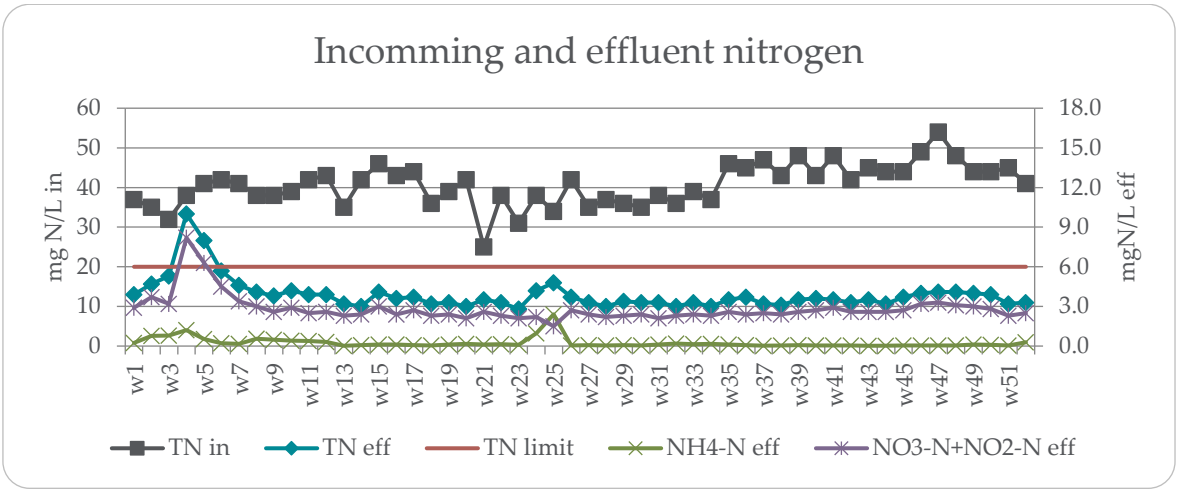


Figure 11. Incoming and effluent nitrogen concentrations from analysis of weekly composite samples during 2020. Limit for effluent total nitrogen was set to 6 mg N/L.

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 2021 (3.6 kg N/d) was the similar as previous years and somewhat higher compared to the full-scale design. Glycerol dosage was used for most of the year (January to mid-November) and the average consumption per day was 1.76 kg COD/d which is almost the same as future Henriksdal design. The yearly average was lower (1.6 kg COD/d) and includes the trial with internally produced VFA as carbon source. Details about the 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 air flow to the pilot was much larger than the estimated for the future Henriksdal WWTP (Table 13). The same has been observed every year in the pilot trials. One reason for the big differences in air flow is that the depth in the basins of the pilot were one fourth of the depth in Henriksdal.

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 2021	Future H-dal design	Value pilot/ scaled future H-dal design[i]
Removed nitrogen (incl. reject water)	kg N/d	3.6	21 000	116 %
Nitrogen removal rate	g N/(kg VSS·d)	16.7	17.6	95 %
Aerated sludge age (incl. MTs)	d	7.0	9.4[ii]	74 %
Aeration, biology (activated sludge)	Nm^3/h	61.2[iii]	68 000	603 %
Specific oxygen demand (SOTR)	kg O_2/d	70.3[iv]	240 000	196 %
Consumption of external carbon	kg COD/d	1.6	12 000	89 %

[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.

6.2.1 Nitrification

Nitrification worked satisfactory most of the year with low effluent concentrations of ammonium except for shorter periods and two peaks in effluent ammonium (w.4-5 and w.24-25). When comparing the aeration of the biology with aeration of the membrane tanks (Figure 12) it can be observed that the air flow to the biology was much higher than the air flow to the membranes for most of the year. The lowest aeration was in week 5 after maintenance on aeration equipment when suspended solids concentration was low. The membranes were operated at either the lower aeration level (Leap-Lo corresponding to 14 m^3/h each, 28 m^3/h in total) or a Leap-Medium level corresponding to 20 m^3/h instead of Leap-Hi (26 m^3/h) from week 9. The average flow rate for membrane aeration was 15.9 m^3/h for MT1 (excluding week 51-52 when MT1 was out of operation) and 17.4 m^3/h for MT2. The air flow to the biology varied as weekly average between 13.6 and 124 m^3/h and was 61 m^3/h as yearly average.

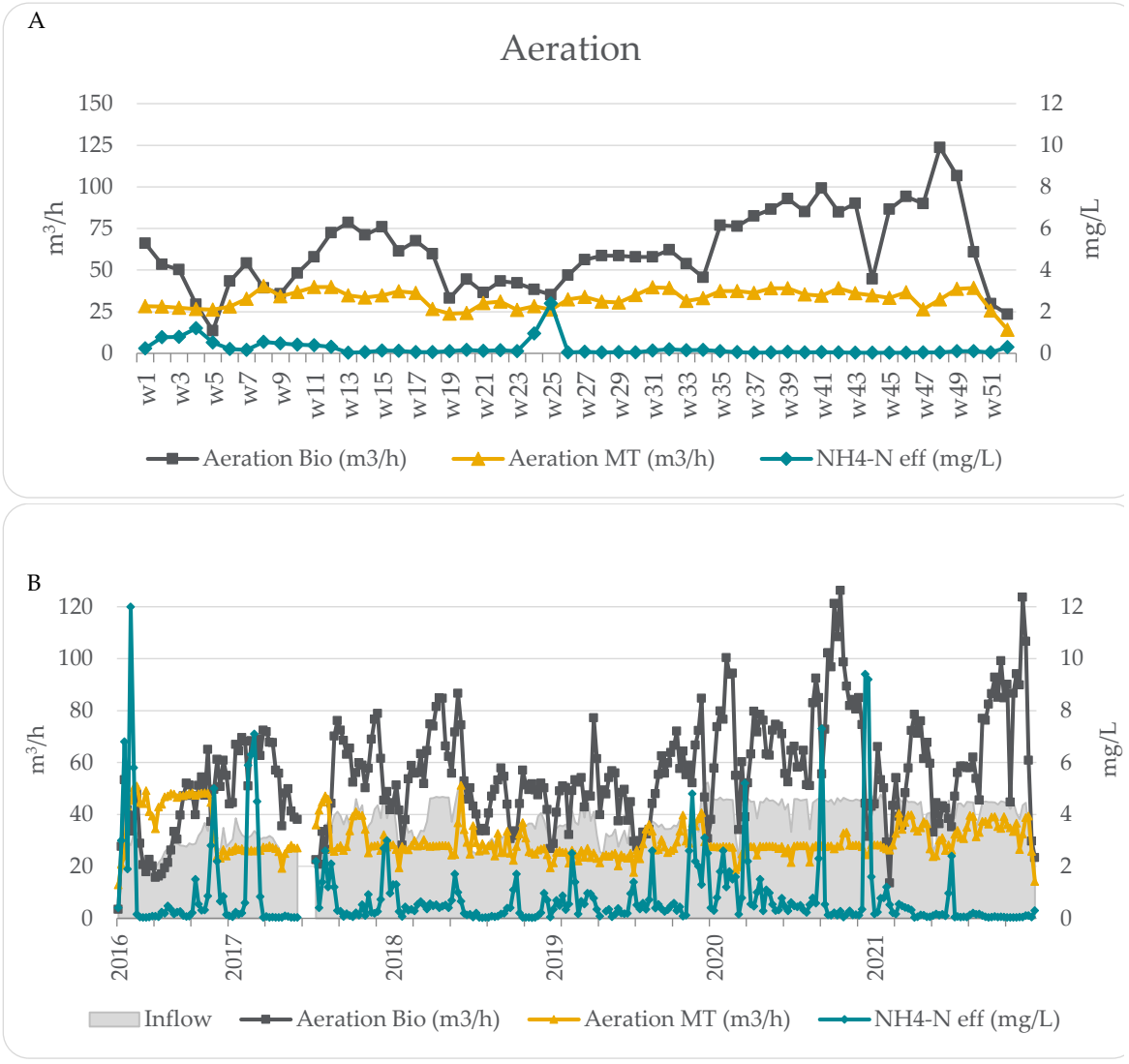


Figure 12. A) Aeration need in biology and membrane tanks (MT) together with effluent NH_4-N during 2021. B) Aeration, influent flow rate, and effluent NH_4-N from 2016.

Table 14 shows the annual average air flows and DO levels in the aerated tanks year 2016-2021 together with the calculated theoretical airflow demand. The airflow year 2020 was exceptionally high compared to previous years and it was suspected that this was partly due to some problems with the aeration system. In the beginning of 2021, the diffusors in BR3-BR5 were replaced with new ones.

Table 14. Air flow and DO in the different zones, year 2016-2021. Calculated air flow 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)	Air flow BR3 (m^3/h)	DO BR3 (mg/L)	Air flow BR4 (m^3/h)	DO BR4 (mg/L)	Air flow BR5 (m^3/h)	DO BR5 (mg/L)	Air flow Tot (m^3/h)	Calc. Air flow Tot (m^3/h)
2021	4.12[f]	27.4	1.35	23.8	2.25	10.0	1.62	61.2	55.7
2020	4.38[f]	23.2	1.42	38.7	2.71	10.5	1.67	72.5	56.8
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 period with incomplete nitrification in week 24-25 was related to manually limited aeration of the bioreactors. This was done as a measure to limit the overflow of sludge due to foaming.

On average, the aeration of the membranes accounted for 34 % of the total aeration, which is in line with results from years 2017 to 2019 although Leap-Hi was replaced with Leap-Medium. One explanation to this is the higher aeration in the biology, another is a more frequent use of the higher aeration mode (Leap-Medium) because of the high flux tests.

As previous years, problems with foaming and unreliable DO sensor readings in the foamy sludge have made aeration control difficult at times. BR3, BR4 and BR5ox was aerated to keep a manually selected DO setpoint between 1 and 4 mg/L throughout the year, with only short periods without aeration of BR3 due to operational disturbances.

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 TSS concentration in the line and the sludge age was not actively controlled and has only been monitored.

Waste activated sludge (WAS) flow rate 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 are presented in Figure 13 together with the WAS flow rate. The peak in sludge age around week 6-8 is the result of lowering the WAS pumping to increase suspended solids concentration after the major loss of sludge in week 4 when BR2-BR5 were emptied for maintenance.

The calculated total sludge age was on average 17.2 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 7.0 days.

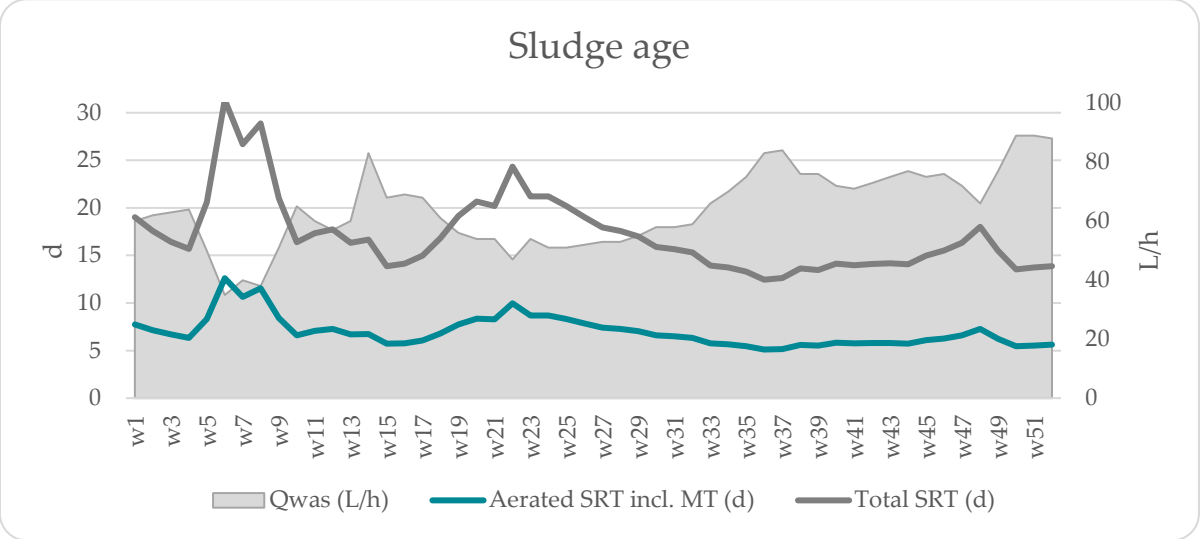


Figure 13. Total and aerated (incl. MTs) sludge age (moving average one month back in time) together with WAS flowrate.

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 Q_{in}, except for a short period February when 1 x Q_{in} was used as a measure to avoid recirculation of oxygen when the process was unstable. 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 3.3 – 7.6 mg/L (average 4.7 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 2 - 4 x Q_{in}. The nitrate concentration in the RAS was 1.1 – 11.6 mg NO_3-N/L with 2.9 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 until mid-November (w. 45) when a new trial started using internally produced VFA from sludge and food waste as carbon source. 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 alternated 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 3 mg NO_3-N/L was used.

An overview of the external carbon source addition is given in Figure 14. The effluent nitrate concentration was possible to control and keep below 3 mg NO_3-N/L with both glycerol and VFA. A more detailed evaluation of the denitrification potential of the internally produced VFA is presented in the next section (6.2.3 Comparison of different external carbon sources). It is clear from Figure 14 that the lower nitrate recirculation in week 5-6 resulted in an increased carbon dosage but not lower effluent nitrate, as expected.

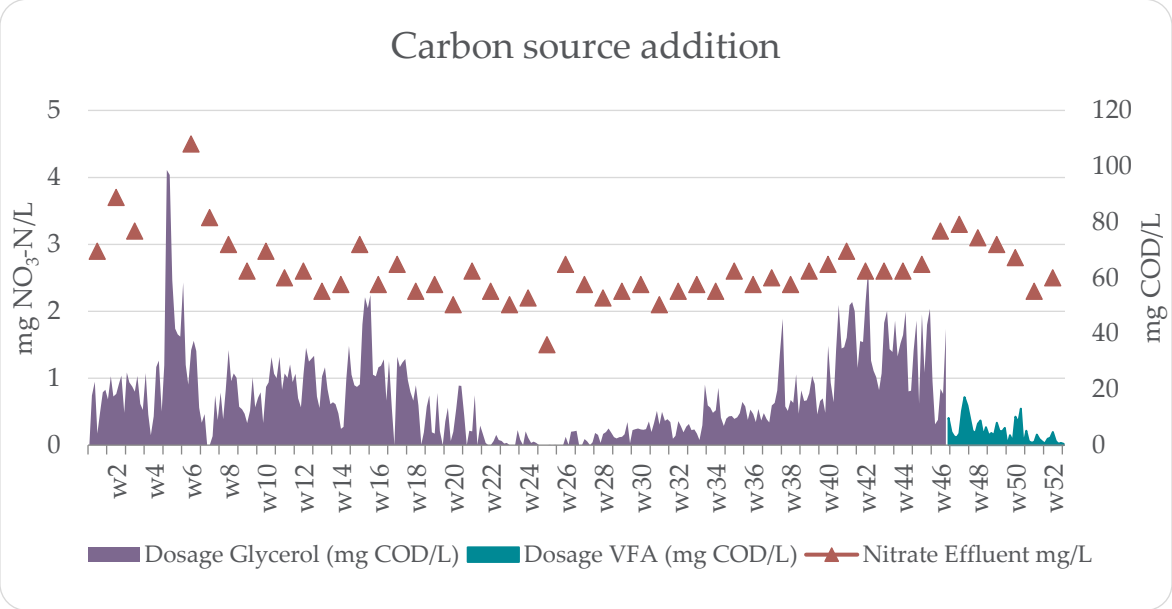


Figure 14. Dosage of glycerol and VFA (mg COD/L incoming wastewater) as daily average values and effluent nitrate analysed in weekly composite samples.

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 15. The peak in week 4-5 was related to low suspended solids concentration after the maintenance work in combinations with faulty sensors making the carbon source dosage too low. The concentrations in BR5deox varied between 3.3 and 7.6 mg NO_3-N/L while the concentration in BR6 varied between 0.9 and 5.1 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 the nitrate concentration in BR5 just prior to the VFA trial were low and only slightly higher than in BR6 and effluent. However, during the VFA trial (week 47 to 52) there is a decrease in nitrate from BR5 to BR6 indicating some effect on denitrification although the dosage was low.

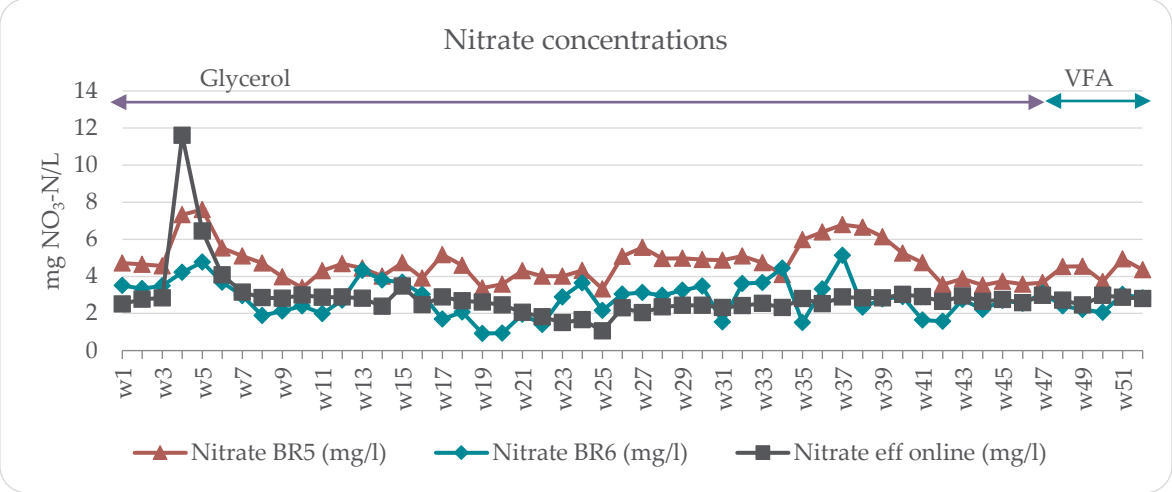


Figure 15. Online nitrate concentrations as weekly average.

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 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 alternative external carbon sources in the pilot since 2020. During the final months of the pilot trials an internally produced VFA was also tested as carbon source dosed in the post-denitrification zone. These results, however, are part of a PhD project and will be published in detail separately.

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 internal VFA from 2021-2022 are shown in Figure 16 and Table 15, previous results using other external carbon sources can be found in Andersson et al. (2021b).

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 16, the specific COD consumption was almost the same for glycerol when comparing the yearly average from 2021 and 2020. The preliminary results for VFA dosage show a slightly lower consumption, but it should be noted that the data for VFA evaluation is only based on a 7 week long trial.

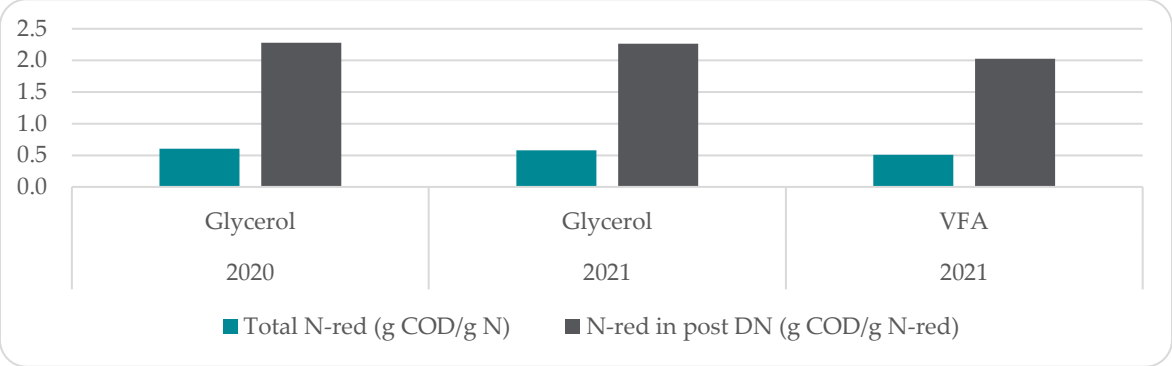


Figure 16. Comparison of COD consumption for glycerol in the pilot during 2020 and 2021 to usage of VFA.

When looking at some of the operating conditions during the trials 2021, both effluent nitrate and total nitrogen removal was very similar during the VFA trial as the rest of the year, when using glycerol (see Table 15).

Table 15. Carbon source dosage and nitrogen removal for trials using glycerol and VFA during 2021.

Year	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)
2021	Glycerol	45	1.7	0.76	3.0	17.1	2.8
2021	VFA	7	1.5	0.73	2.9	15.8	2.8

6.3 Phosphorus removal

	2021												2022		
Activity	J	F	M	A	M	J	J	A	S	O	N	D	J	F	M
Effluent phosphorous target 0.10 mg TP/L	■	■	■	■	■	■	■	■	■	■	■				
Al^{3+} with Fe^{2+} for P-removal		■	■	■	■										
Evaluation of flux enhancer						■	■	■	■	■	■				
P-release tests	■		■	■		■	■	■		■	■	■			

The goal of reaching a stable effluent phosphorus concentration below 0.15 mg P/L as monthly and annual averages was achieved previous years by using a control strategy with dosage of ferrous sulfate 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 sulfate 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. During 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. During 2021, this strategy was kept until June when PAX dosage was stopped as a separate trial using an iron-based flux enhancer product (PIX-XL3164 from Kemira). It started in week 27 and dosed until week 36. When the flux enhancer trial was finished, the pilot was operated using only a 10 mg Fe/L flow-proportional dose of ferrous sulfate in the inlet until the end of the pilot trials in March 2022.

The second dosage point of ferrous sulfate 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 17 below. The yearly average effluent total phosphorus was 0.07 mg TP/L, which is well below the target of 0.10 mg TP/L. It should be noted that the inlet total phosphorus during 2021 (just as 2020) was lower than normal. Total phosphorus in primary treated water has previous years (before 2020) varied between 5.3 and 5.5 mg TP/L as yearly average.

Table 16. Phosphorus concentrations in primary treated water (PTW) and effluent during 2021.

Parameter	Limit	Mean	Min	Max	Nr of weekly samples
TP PTW (mg/L)	-	4.6	2.7	6.1	52
TP EFF (mg/L)	0.20	0.07	0.014	0.23	51

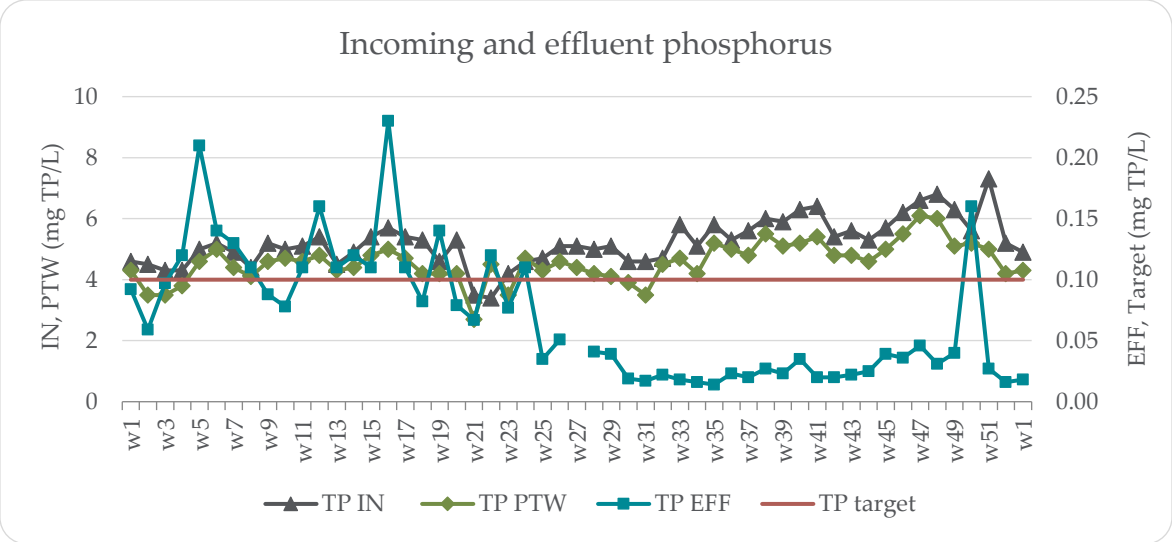


Figure 17. Influent and effluent total phosphorus analysed in weekly composite samples during 2021. Effluent data from week 27 is considered a measurement error and was excluded from calculations.

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 (see section 6.3.1). 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 2021 was 11.0 mg Me/L (10.1 mg Fe/L + 0.9 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 2021	Future H-dal design	Value pilot/scaled Future H-dal design[a]
Phosphorous load influent	kg P/d	0.51	2 594	133 %
Phosphorous load PTW	kg P/d	0.45	1 580	191 %
Phosphorus load reject water	kg P/d	0.002	480	2 %
Total phosphorous load biology	kg P/d	0.45	2 060	147 %
Phosphorus load effluent	kg P/d	0.0069	80	57 %
Phosphorus removed in biology	kg P/d	0.45	1 980	151 %
Iron consumption (Fe^{2+})	kg Fe/d	0.85	10 000	57 %
Iron consumption (Fe^{3+})[c)]	kg Fe/d	0.78	-	-
Iron consumption (Fe^{2+}+ Fe^{3+}) during flux enhancer trial	kg Fe/d	1.75	10 000	117 %
Aluminium consumption	kg Al/d	0.09	0	-
Metal consumption (Fe+Al)	kg Me/d	1.09	10 000	73 %
Metal consumption per removed phosphorus	mole Me/ mole P	1.29	2.80	46 %
Metal consumption (Fe+Al)	kg Me/d	1.75	10 000	117 %
Phosphorus in sludge	% of SS	3.1	5.4[b]	-
Iron in sludge	% of SS	6.2	-	-
Aluminium in sludge	% of SS	0.7	-	
VSS in sludge	% of SS	77	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 the sludge is low.

The total amount of iron and aluminium dosed is presented as daily average values in Figure 18. The base dose of Fe^{2+} was added to the pre-aeration which was controlled flow proportionally to a dose of 8-10 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. When dosage of flux enhancer started, the effluent total phosphorus decreased to even lower values (below 0.05 mg TP/L for most of the rest of the year). Details regarding the flux enhancer trial can be found in section 6.8. The yearly average dosage of aluminium chloride was only 0.9 mg Al/L, and the highest weekly dosage was 4.9 mg Al/L. The maximum weekly average effluent total phosphorus was 0.23 mg TP/L and the yearly average 0.07 mg TP/L.

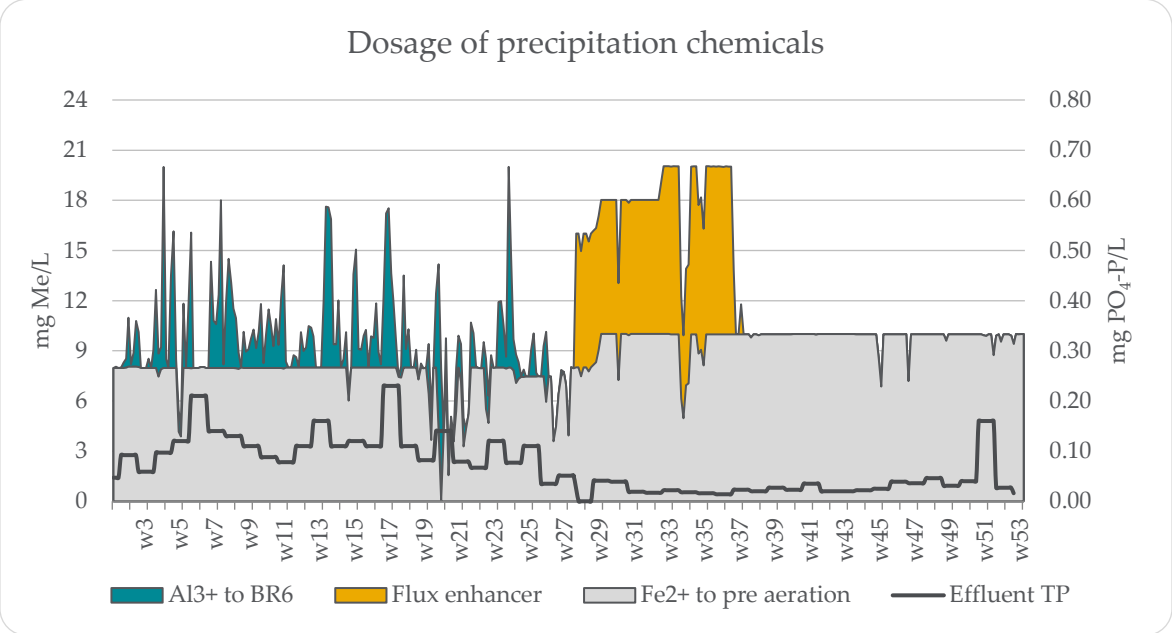


Figure 18. Metal dosage (the sum of Fe^{2+} to pre-aeration, Fe^{3+} in flux enhancer and Al^{3+} to BR6) as daily average and effluent total phosphorus in weekly composite samples during 2021. Note that the iron and aluminium are presented as a sum of added metals (g Me/d).

The iron, aluminium and phosphorus content of the sludge was monitored during 2021. The aluminium content in the sludge was on average 1.2% of TSS when aluminium was dosed, but quickly decreased after dosage stopped (Figure 19). The iron content of the sludge increased rapidly during the flux enhancer trials to a maximum of 12% of TSS at the end of the trial. The phosphorus in the sludge was similar throughout the year but varied a bit more in the first half compared to the second half of the year.

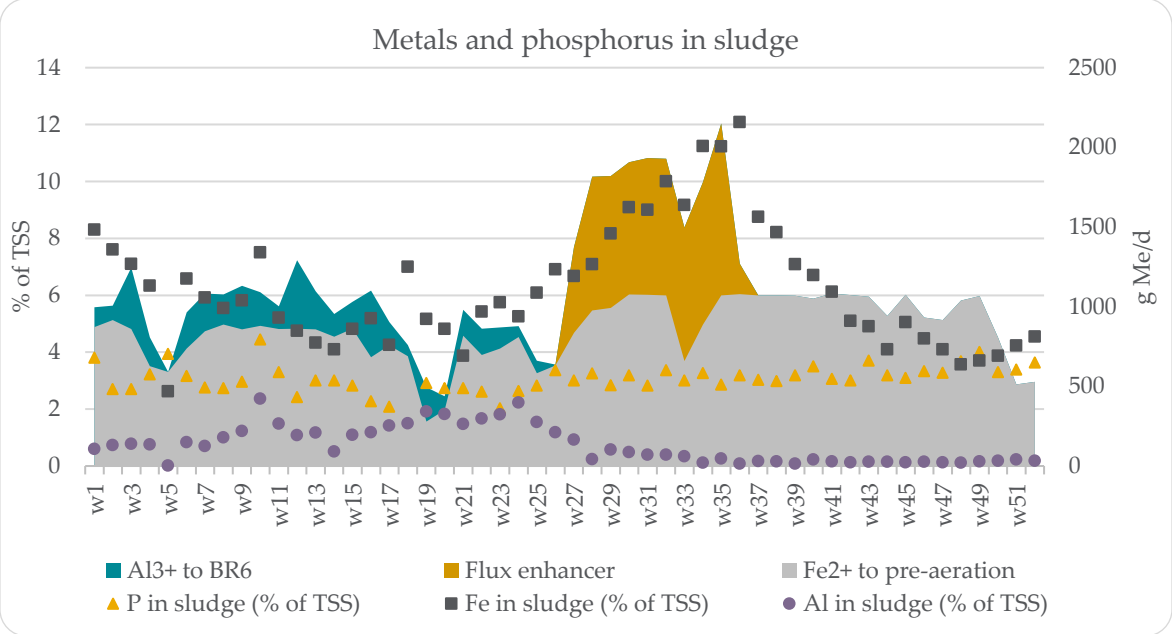


Figure 19. Iron, aluminium and phosphorous in sludge (weekly data). Note that the iron and aluminium are presented as a sum of added metals (g Me/d). The highest total metal dose was during the flux enhancer trials.

6.3.1 Enhanced Biological Phosphorus removal (EBPR)

The enhanced biological phosphorus removal (EBPR) activity, measured as the phosphate release rate of the activated sludge sampled in the aerated zone of the MBR pilot and, as a reference, in one of the CAS lines at Henriksdal WWTP (method described in Tykesson et al., 2005), is shown in Figure 20. On two occasions, 17th of June and 9th of September, additional samples were taken from the MBR line in Henriksdal, which was taken into operation in January 2021, and the results showed that no EBPR could be detected (0.43 g PO_4-P/kg VSS, h and 0.8 g PO_4-P/kg VSS, h respectively).

The P-release rate usually varies over the year. However, unusually low values were obtained from December 2020 until June 2021, when the P-release rates were similar to the negative reference (Henriksdal WWTP) indicating that the EBPR activity completely ceased. One possible reason for this might be the problems related to the RC with hypochlorite that was performed in December 2020, which caused disturbances in the nitrification during approximately 2 weeks, resulting in effluent ammonia peaks as high as 20 mg NH_4-N/L. This might also have affected the polyphosphate accumulating organisms (PAO). Unlike the nitrifiers, which recovered after a couple of weeks, it took months for the PAO to recover and resume the EBPR activity. The reason for the slow recovery is not known, it might because EBPR was lost during winter when the activity normally is lower, or it could be caused by some other operational factor. The only difference in operation during spring 2021 that could have a negative effect on EBPR is the trial with defoaming agent, which was performed from March to June. There is no known component in the defoaming agent that has a negative effect on EBPR activity. But it could be seen that when the dosing of defoaming agent was terminated on the 10th of June a rapid increase in the P-release rate from 0.84 g PO_4-P/kg VSS, h on the 15th of June to 5.45 g PO_4-P/kg VSS, h on the 9th of July was seen. Nevertheless, based on the available data it cannot be concluded if the defoaming agent had a negative effect on the recovery of EBPR or not.

In July 10th, the P-release rate recuperated to, for the MBR pilot, more normal values, although not as high as in 2021. In July 5th, the dosing of flux enhancer also began, which meant a significantly higher Fe dose (from

an average of around < 9 g Me/m³ to 18 g Me/m³) and increased Fe content in the activated sludge, from around 8 % of TSS before the trial to 12% week 37, the beginning of September. This sudden increase in Fe dosing might be the reason of the decreased P-release rate, 2.26 g PO_4-P/kg VSS, h, seen on the 27th of August. The following P-release rate test was performed in mid-October, when the rate was back to the same level as in July, 5,10 g PO_4-P/kg VSS, h, where it stayed throughout the year.

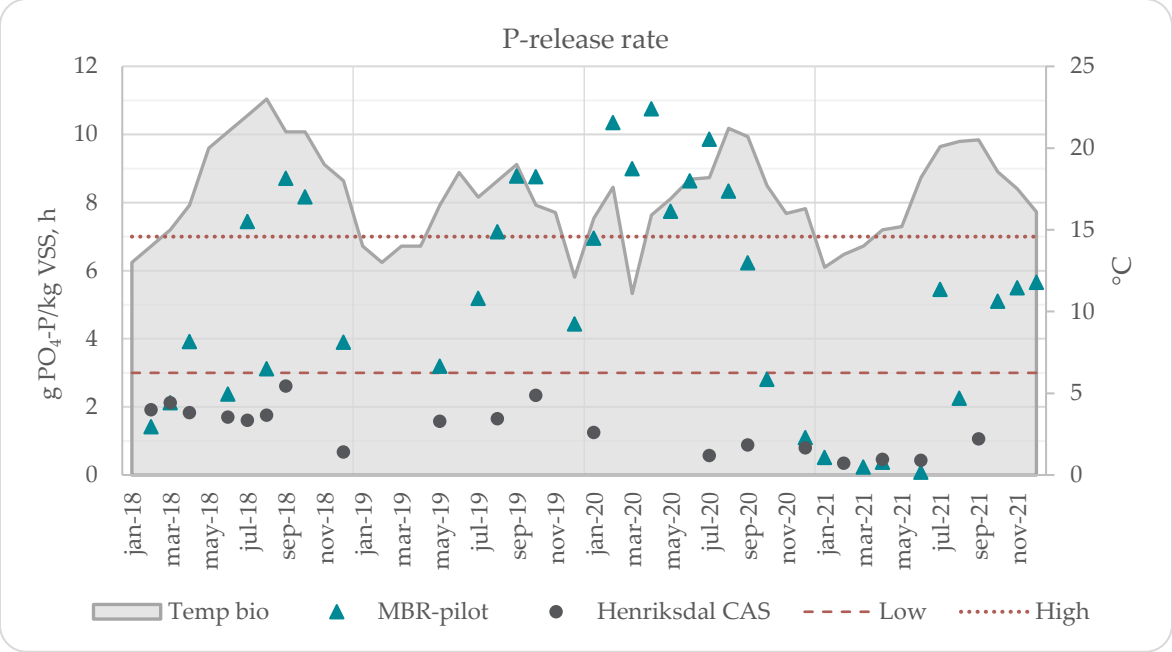


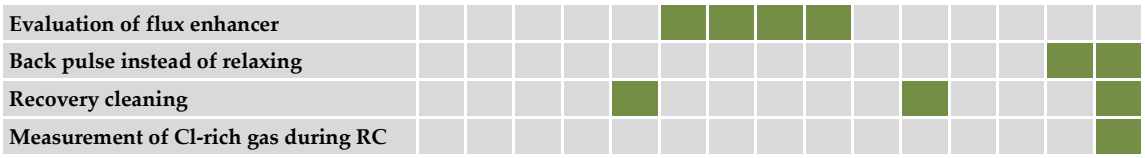
Figure 20. EBPR activity in the MBR-pilot and Henriksdal WWTP (CAS). Dotted lines show reference lines for EBPR-activity (Janssen et al., 2002), below the lower line = poor EBPR, above the higher line = high EBPR. Temp bio is the temperature in the treatment process when the sample was taken.

6.4 BOD reduction

Analysis on BOD_7 from daily composite effluent samples have, since the start-up of the MBR pilot in 2013, until 2016 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 2021 or 2022 and no specific measures have been taken to achieve a higher BOD reduction.

6.5 Membrane performance

	2021												2022		
Activity	**J**	**F**	**M**	**A**	**M**	**J**	**J**	**A**	**S**	**O**	**N**	**D**	**J**	**F**	**M**
Oxalic acid and citric acid comparison															
Demand driven hypochlorite MC															
High fixed inflow															
Reduced RAS-circulation to 2×Q_{in}															
Leap-Medium															
High flux															



During 2021 and 2022 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. The permeate flow control strategy is based on the level in the biological tanks and this means that the membrane flux will be controlled, and the transmembrane pressure (TMP) will be allowed to vary.

During this period 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 not needed to achieve the design flux. However, permeate recirculation was used during several periods to reach higher flux (up to 30 L/(m²·h)).

Table 18 list the main events and experiments regarding the membrane operation during 2021 and 2022.

Table 18. Main events and experiments regarding the membrane operation.

Event	Time period (2021-2022)
Recovery cleaning (RC) with sodium hypochlorite	Week 21 and 47 in 2021 Week 12 2022
Recovery cleaning (RC) with acids (oxalic and citric)	Week 23 2021 Week 13 2022
Demand driven maintenance cleaning (MC) with sodium hypochlorite	Week 13 to 23 2021
High flux	Week 25 to 50 2021
Evaluation of flux enhancer	Week 26 to 36 2021
Backpulse instead of relaxing	Week 7 2022 and continuing 2022
Operation with prolonged permeation cycle (15 min instead of 10 min, capacity increased by 3 %)	Throughout 2021-2022
Operation with Leap-Medium (medium level aeration instead of high)	Week 9 2021 and continuing.
Reduced RAS recirculation	Week 9 to week 42 2021
MT1 taken out of operation due to broken frame-support	Week 50 2021 to week 4 2022

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 21, the permeability was above 200 L/(m²·h·bar) throughout most of the year 2021-2022 for both membranes. The permeability was greatly affected by the different operational and cleaning strategies tested this year. By the end of 2021 there was a breakage in a part holding the membranes in place in MT1 and it was taken out of operation between week 50 2021 and week 4 2022 to repair the membrane fitting.

Recovery cleaning was carried out twice with hypo and once with acids during 2021. During 2022 a final RC, first with hypochlorite then with acids were carried out. The first RC for MT1 in week 21 resulted in a clear increase in permeability after cleaning. For MT2 the major increase in permeability in week 22 was the result of a citric MC (one week after the hypochlorite RC). After this big increase in permeability, the effect of the following RCs with acids in week 23 showed no clear effect on permeability. The following RCs by the end of 2021 and in 2022 had clear but smaller positive impact on permeability.

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. As a result of the tough operational strategy from week 25 2021, permeability quite quickly decreased after RCs to a stable level

around 250-300 L/(m²·h·bar) for MT2 and with a decreasing trend for MT1. The lowest permeability was measured in week 39 with 183 L/(m²·h·bar) for MT1. From week 40 the permeability was similar for the two membrane tanks (excluding the period when MT1 was out of operation).

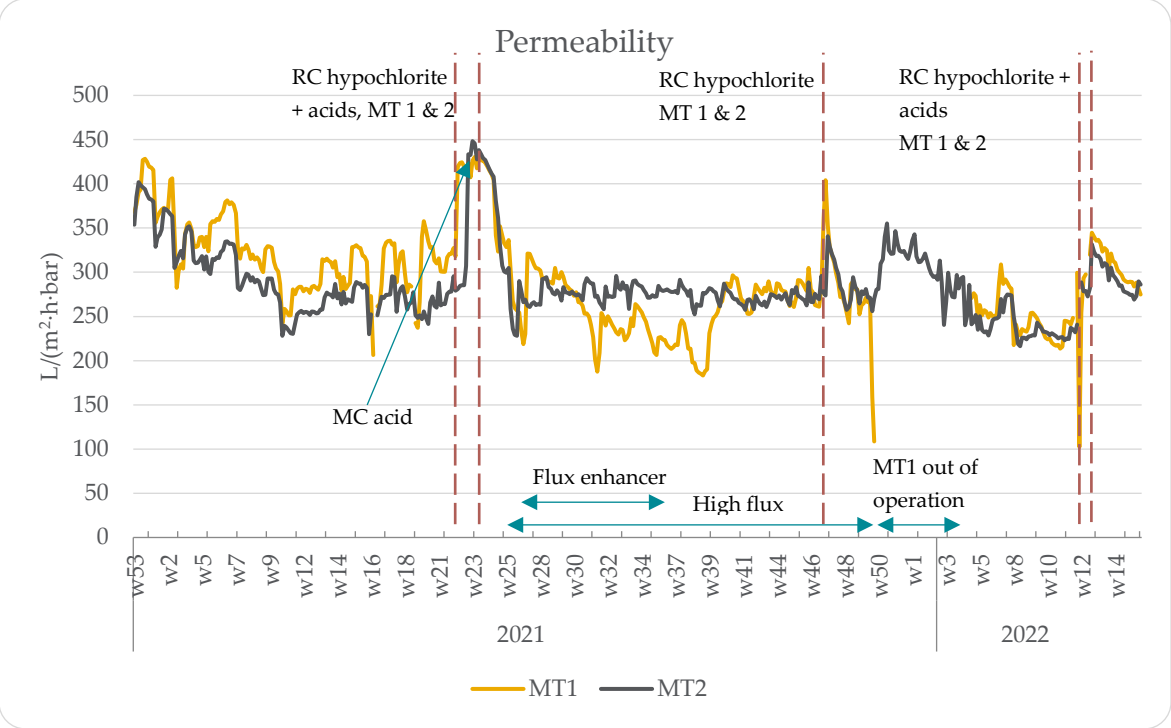


Figure 21. Permeability (temperature compensated) for membrane 1 (MT1) and 2 (MT2) during project year 8-9 (2021-2022). Recovery cleanings (RCs) were carried out with hypochlorite in week 21 and week 47 in 2021 and in week 12 2022. RC with acids were carried out in week 23 2021 and 13 2022 for both membranes. MT1 was cleaned with oxalic acid, MT2 was cleaned with citric acid.

During 2021, the membranes were operated with maximum flux 30 L/(m²·h) for 6 months, in combination with high TSS (11-12 g/L) for about 10 weeks, and with reduced membrane aeration and maintenance cleaning of the membranes. Despite an increase in yearly average flux of 19 % for MT1 and 17 % for MT2 in 2021 compared to 2020, the yearly average permeability was only 13 percent lower for both membranes and only below 200 L/(m²·h·bar) for MT1 (on two occasions excluding the period when it was out of operation). It should be mentioned that during the high flux tests a flux enhancer product was tested, however, evaluation of periods before and after dosing of the flux enhancing product showed no clear effect on permeability.

6.5.2 Flux and TMP

Fluxes for the two membranes are presented in Figure 22. Previous years, the membranes were normally 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). During 2021, the design max net flux was tested in the pilot for 25 weeks (starting week 25 2021). 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 or drops reflects the inflow flow rates.

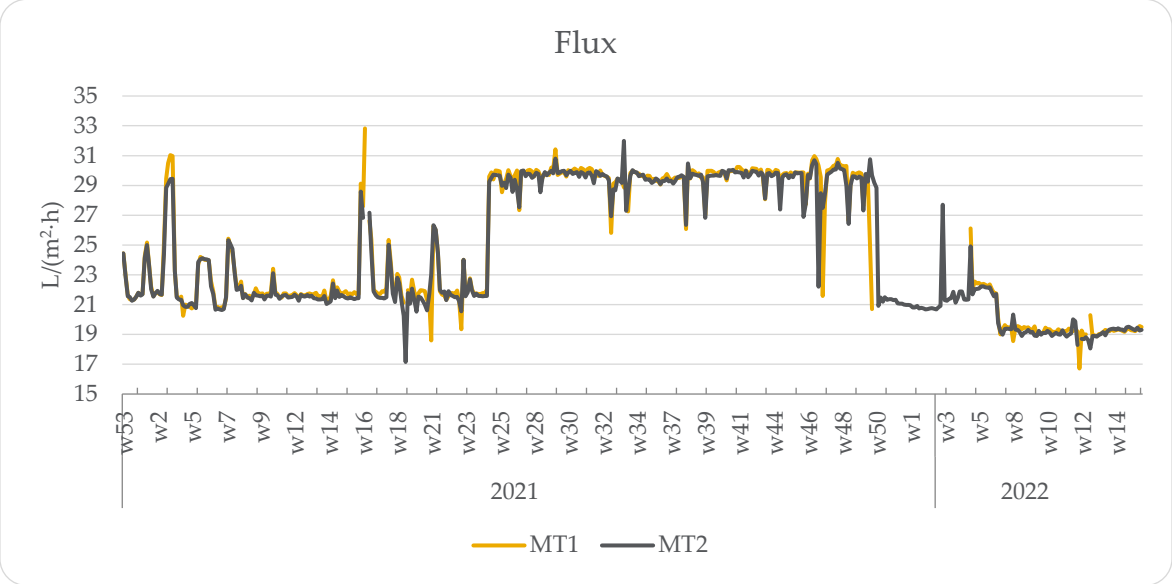


Figure 22. Net flux during 2021 and beginning of 2022.

The transmembrane pressure (TMP) is presented in Figure 23. Based on daily average data, net TMP varied between 49 and 218 mbar for MT1 (highest value just before it was taken out of operation in week 50) and between 51 and 146 mbar for MT2 during 2021. TMP was reduced after recovery cleaning (RC) with hypochlorite (w. 21 and w. 47 2021 and w. 12 2022) but the effect did not last long. Similar events of decreasing TMP can be observed for MT2 for example in week 15 after an MC with citric acid was carried out and in week 51. From the TMP curve it can be noted that TMP in MT1 increased during the first part of the trial with high flux while TMP in MT2 remained stable, with a small increase towards the end of the high flux trial.

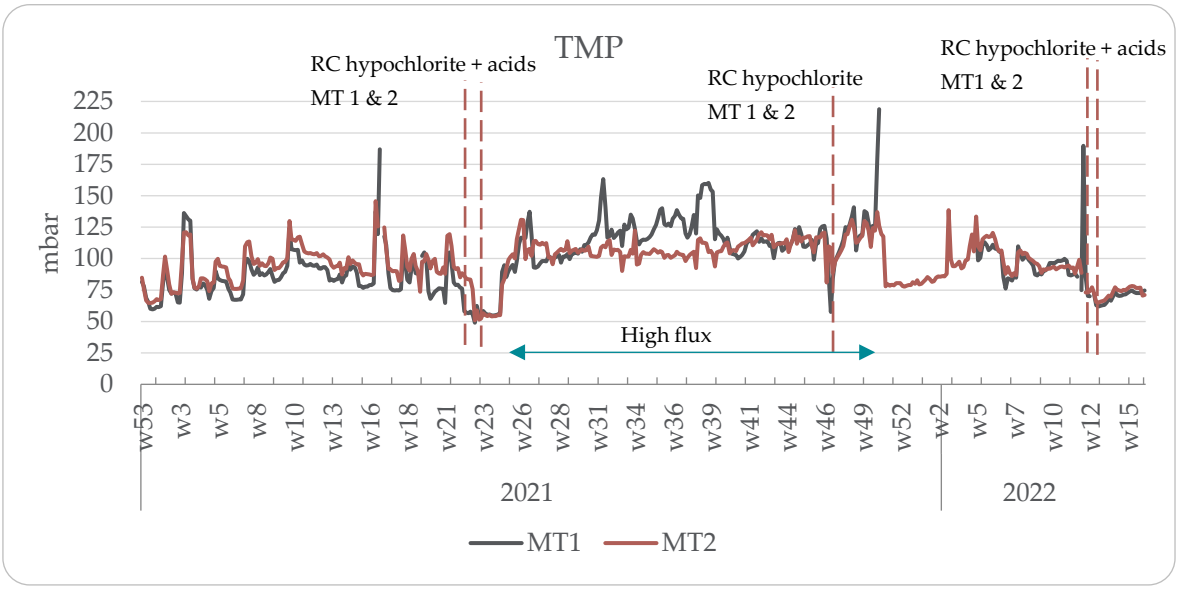


Figure 23. Net TMP during 2021 and beginning of 2022. Recovery cleanings (RCs) were carried out with hypochorite in week 21 and week 47 2021 and in w.12 2022. RCs with acids in week 23 2021 and week 13 2023 for both membranes. MT1 was cleaned with oxalic acid, MT2 was cleaned with citric acid.

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. To keep the treatment line in operation, each membrane was cleaned separately, and the cleanings were scheduled at night. 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 the focus was on reducing the amount of hypo used for MC. In October 2020 a demand-driven strategy for MCs with hypo was introduced, 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. This strategy was used throughout 2021 until the project ended in 2022.

The operational settings have been divided into separate trial periods. An overview of the trials regarding the MCs are presented in Table 19.

Table 19. Overview of trials regarding the maintenance cleaning (MC) of the membranes.

Trial	Start	Description
T1	Sep 2017	Citric vs Oxalic - 7 BP (both MT)
T2	Jun 2018	Recovery Period (short switch between chemicals)
T3	Jul 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 Hypochlorite MC, every 4th hypochlorite for MT1 and every 2nd hypochlorite 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 Hypochlorite MC Citric acid 100 % chemical flow, 7 BP, every 4th Hypochlorite MC
T11	Feb 2020	Oxalic acid 80 % chemical flow, 7 BP, every 4th Hypochlorite MC Citric acid 100 % chemical flow, 9 BP, every 4th Hypochlorite MC

T12	Mar 2020	Hypochlorite MC replaced with water MC
T13	Jun 2020	Hypochlorite 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	Sep 2020	Recovery period
T15	Oct 2020	Fouling-based trigger for hypochlorite MC Acid MCs carried out with half standard interval (once every other week)

To summarize, the maintenance cleaning with sodium hypochlorite has been carried out with reduced backpulses; 2 min initial pumping followed by 6 x 30 seconds at irregular intervals based on the fouling control algorithm. The maintenance cleanings with acid have been carried out with reduced backpulses; 2 min initial pumping followed by 6 x 30 seconds, at regular intervals once every 14 days.

As different operational strategies affecting the membrane permeability were tested during 2021 and 2022, the evaluation of the maintenance cleaning strategy have been divided into several periods described in Table 20.

Table 20. Trials during 2021 and 2022 for which average permeability is evaluated.

Trial	Start	Description
T16	2021-01-01	Fouling-based Hypochlorite MC = Same as T15
T17	2021-03-01	Fouling-based Hypochlorite MC, Leap Med, RAS = 3*Q_{in}
T18	2021-03-29	Fouling-based Hypochlorite MC, Leap Med, RAS = 2.5*Q_{in}
T19	2021-04-09	Fouling-based Hypochlorite MC, Leap Med, RAS = 2.0*Q_{in}
T20	2021-04-22	Fouling-based Hypochlorite MC, Leap Med, RAS=2*Q_{in} + High Flux
T21	2021-04-28	Fouling-based Hypochlorite MC, Leap Med, RAS = 2*Q_{in}
T22	2021-06-21	Fouling-based Hypochlorite MC, Leap Med, RAS = 2*Q_{in}+ High Flux
T23	2021-10-21	Fouling-based Hypochlorite MC, Leap Med, RAS = 4*Q_{in}
T24	2021-12-17	Special period with MT1 out of operation
T25	2022-02-02	Same as T15

The membrane permeability for 2021 and 2022 is presented in Figure 24 as daily average values and as average during separate trial periods with different operational settings regarding flux and return sludge flow rate. The maintenance cleaning strategy was the same throughout 2021-2022 as described for T15 in Table 19. The different settings for each period are presented in Table 20 for the trials during 2021 and 2022.

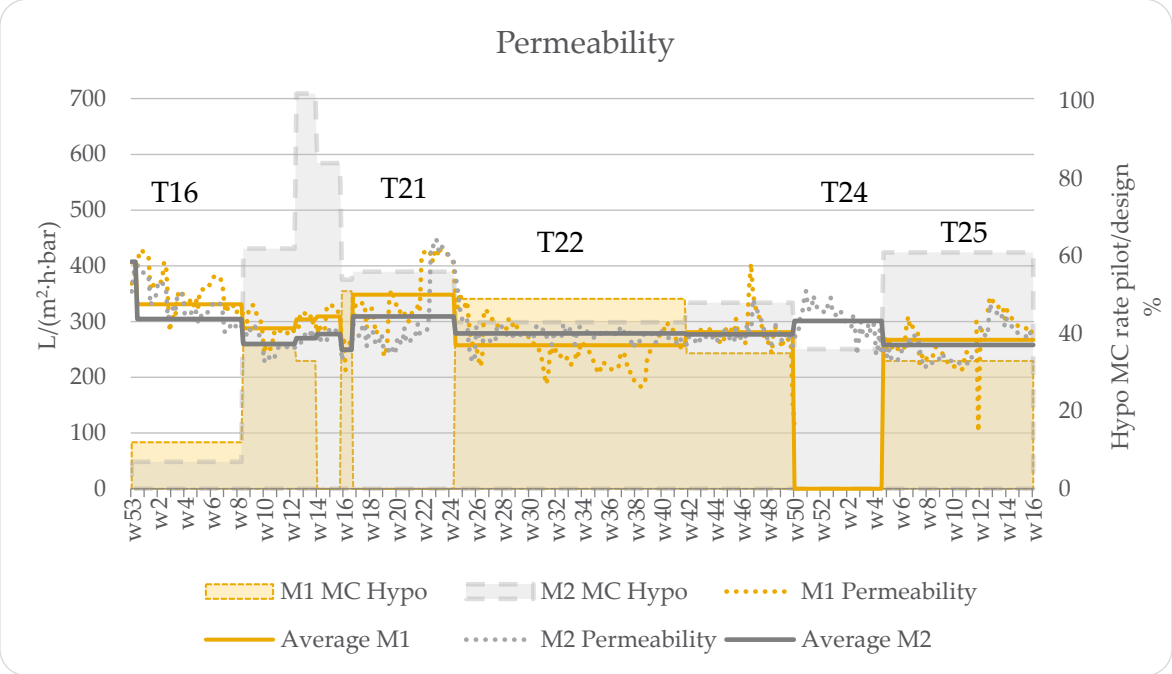


Figure 24. Membrane permeability (M1=MT1, M2=MT2) as daily average and as average during different periods of operational settings during 2021 and 2022 together with the amount of hypochlorite for MC related to design cleaning schedule. MT1 was taken out of operation week 50 2021 to week 4 2022. Trial T16 to T25 are describes in Table 20.

Looking at the trend over the year 2021, it can be noted that the cleaning strategy maintained the permeability at a good level and only below 200 L/(m²·h·bar) for MT1 on two occasions (excluding the period when it was out of operation).

The resulting maintenance cleaning frequency for each period and as total for the full period when the fouling control algorithm was used (T16 to T23) is presented together with average permeability for each period in Table 21. Two interesting results can be highlighted: 1) Although the trials put a lot of pressure on the membranes, cleanings with hypochlorite were less than design in all periods except for MT2 during the 10-day short trial number 18. 2) On average MT1 needed less hypochlorite MC cleanings compared to MT2. During the trials acid cleanings were scheduled at half the frequency of design, and on average using the fouling-based strategy for hypochlorite MC the needed amount of MC with hypochlorite corresponded to 29 % and 38 % of design amount for MT1 and MT2, respectively.

Table 21. Maintenance cleanings and permeability during trial 16 to 25.

Trial	Days	Number of maintenance cleanings				Hypo MC related to design MT1 (%)	Hypo MC related to design MT2 (%)	Permeability MT1 L/(m²·h·bar)	Permeability MT1 L/(m²·h·bar)
		Oxalic MT1	Citric MT1	Hypo MT1	Hypo MT2				
T16	136	11	13	5	3	12	7	335	333
T17	27	2	2	3	5	37	62	288	260
T18	10	2	2	1	3	33	102	304	270
T19	12	1	1	0	3	-	84	309	277
T20	5	0	0	1	1	51	54	250	249
T21	53	3	3	0	9	-	56	348	309
T22	121	15	14	24	21	49	43	257	278
T23	56	9	7	8	11	35	48	281	277
T24	46	0	3	0	5	-	36	-	301

T25	79	4	5	7	13	33	61	267	258
T16-T23	420	43	42	42	56	29	38	297	282

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. RC was performed twice using hypochlorite in 2021 (May and November) but only once using acids (in June). In March 2022 a final set of RCs were carried out first with hypochlorite and then with acid the week after. The schedule for RCs together with amount of chemicals used and conditions at start and end of soaking is presented in Table 22.

Table 22. Settings and operating conditions for recovery cleanings.

Date	Membrane tank	Chemical	Amount	Measurements in tank at the start of soak	Soaking time	Measurements in tank at end of soaking
2021-05-24 to 2021-05-25	MT2	NaOCl (69 g/L)	20.6 L	pH 9.6 Cl_2 675 mg/L	16 h	pH 8.2 Cl_2 340 mg/L
2021-05-25 to 2021-05-26	MT1	NaOCl (68 g/L)	20.0 L	pH 9.5 Cl_2 580* mg/L	19 h	pH 8.3 Cl_2 462 mg/L
2021-06-07 to 2021-06-08	MT2	Citric acid (51 %)	3.6 L	pH 2.8	20.5 h	pH 2.8
2021-06-08 to 2021-06-09	MT1	Oxalic acid (8 %)	22.4 L	pH 2.1	19.5 h	pH 2.0
2021-11-22 to 2021-11-23	MT1	NaOCl (57 g/L)	22.2 L	pH 9.5 Cl_2 812 mg/L	19 h	pH 98.2 Cl_2 455 mg/L
2021-11-24 to 2021-11-25	MT2	NaOCl (69 g/L)	18.0 L	pH 9.3 Cl_2 828 mg/L	20 h	pH 8.0 Cl_2 337 mg/L
2022-03-22 to 2022-03-23	MT1	NaOCl (53 g/L)	25.9 L	pH not measured Cl_2 650 mg/L	22 h	pH not measured Cl_2 395 mg/L
2022-03-24 to 2022-03-25	MT2	NaOCl (53 g/L)	24 L (16 + 8)**	pH not measured Cl_2 510 mg/L	23 h	pH not measured Cl_2 90 mg/L
2022-03-28 to 2022-03-29	MT1	Oxalic acid (8 %)	24.4 L	pH 2.4	23 h	pH 2.1
2022-03-29 to 2022-03-30	MT2	Citric acid (51 %)	3.9 L	pH 3.0	22 h	pH 3.0

*Possibly slightly underestimated as sample was taken 45 min into soaking and after addition of extra permeate (200 L) to reach target level in the tank without mixing with aeration prior to sampling.
** Chemical tubing broke during backpulsing after 16 L of sodium hypochlorite had been pumped. The remaining 8 L was added manually by pouring into the tank while diluting with clean water to the correct final level in the tank.

When comparing the effect of the recovery cleanings on permeability, the RCs resulted in similar final permeability in both MTs both in June and in November 2021 (Figure 25). After the first hypochlorite RCs in May, permeability was not improved much. However, the next MC with oxalic acid one week later had a major impact on restoring membrane permeability for MT1 (see increase in permeability for MT1 from the 2nd to the 3rd of June in Figure 25). After the acid RCs in June 2021 the permeability was around 430 L/(m²·h·bar). After only hypochlorite RC in November 2021 the permeability was lower (around 300 L/(m²·h·bar) and with a decreasing trend. The permeability was back to the same level as prior to the RCs about one week after

cleanings. The general conclusion from the RCs is that the acid cleaning has the largest impact on permeability.

The permeability before and after the final RCs in March 2022 is presented in Figure 26. After RCs the permeability was similar for both membrane tanks, slightly lower for MT2, but above 300 L/(m²·h·bar). This time, there was an effect of both the hypochlorite cleanings and the acid cleanings. During RC with hypochlorite for MT2, MT1 had some operational problems resulting in very low permeability during that day (23rd of March). When both membranes were back in operation permeability increased again and after a couple of days both membranes had similar permeability again.

Figure 25. Permeability before and after recovery cleaning (RC) with sodium hypochlorite (hypo) and acids (oxalic acid and citric acid for MT1 and MT2, respectively) in May/June 2021 (upper) and with hypochlorite in November 2021 (lower).

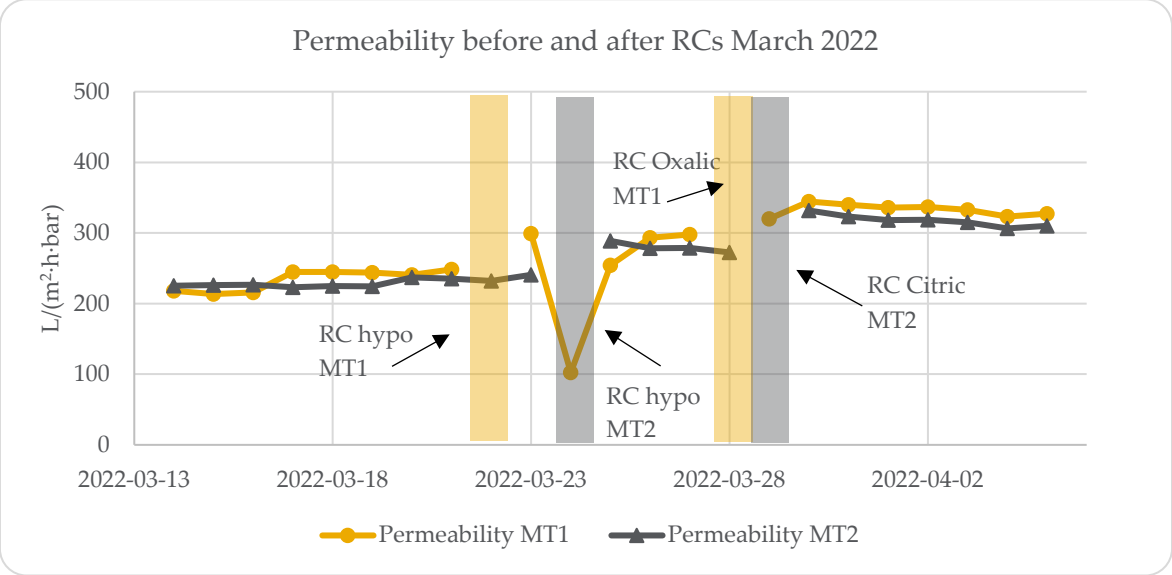


Figure 26. Permeability before and after recovery cleaning (RC) with sodium hypochlorite (hypo) and acids (oxalic acid and citric acid for MT1 and MT2, respectively) in March 2022.

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.

From the 1st of March 2021 (week 9) a new setting, Leap-Medium, with air flow 20 m^3/h was used in the pilot. Leap-Medium replaced Leap-Hi to evaluate if it was possible to save energy without affecting the membrane performance.

Previous years, the membrane aeration was in Leap-Lo mode most of the time. This year, more pressure was put on the membranes with periods of higher flux, higher suspended solids concentrations and reduced maintenance cleaning intervals. The resulting daily average air flow rate to the membrane tanks is presented in Figure 27. The Leap-Medium level was used approximately 38 % of the time for MT1 and approximately 57 % for MT2. The events with lower aeration than the Leap-Lo level (14 m^3/h) corresponds to disturbances in operation, often related to problems with the inlet pumps.

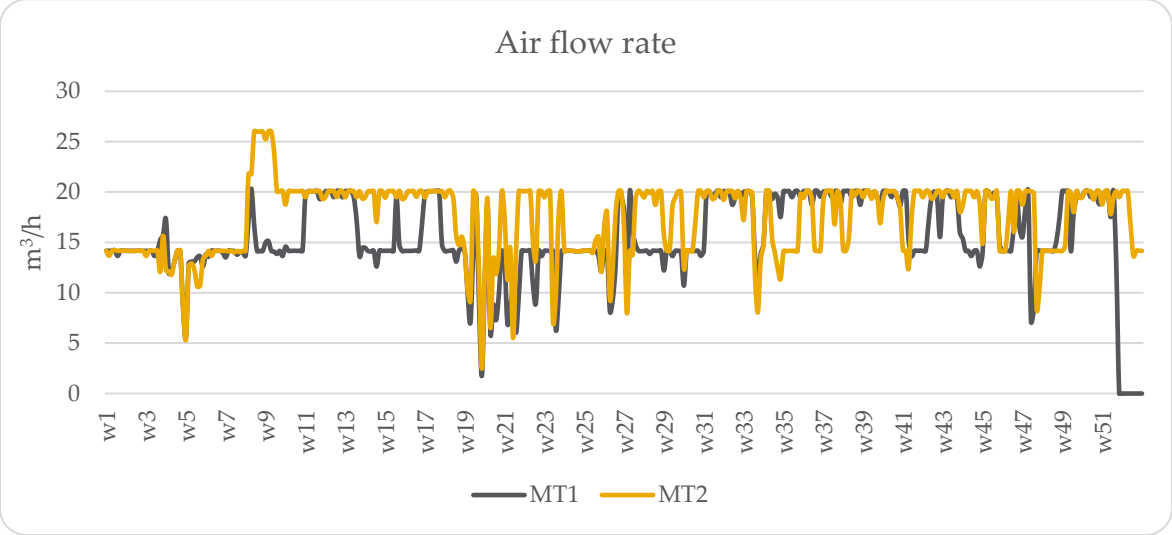


Figure 27. Membrane aeration as daily average during 2021. Leap-Hi corresponds to 26 m³/h, Leap-Medium corresponds to 20 m³/h and Leap-Lo corresponds to 14 m³/h.

6.6 Emission of chlorinated compounds during sodium hypochlorite recovery cleaning

6.6.1 Background

Previous measurements have shown that using sodium hypochlorite (NaOCl) for recovery cleaning (RC) of membranes implies a risk for formation of airborne chlorinated compounds, of which trichloramine is the most prevalent (Andersson et al., 2021b). Health effects ranges from increased problems for people suffering from asthma, to irritated airways, to chronic problems. There is no occupational limit value for trichloramine in Sweden, but the Public Health Agency of Sweden has published a surveillance guideline for environmental supervision in bathhouses, with a suggested reference value of 0.2 mg/m³ (Arbetsmiljöverket, 2018). The World Health Organisation (WHO) has also suggested a reference value of 0.5 mg/m³ for trichloramine (WHO, 2006), and 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/m³ for trichloramine (Wastensson & Eriksson, 2019).

The main objective of the new measurement campaign was to study if the NaOCl RC will result in high levels of chlorinated compounds in the offgas, which can be hazardous from a working environment perspective. This because the membrane tank ventilation will be connected to a technical service tunnel in the full-scale MBR plant. Since chlorine gas is very corrosive and can damage equipment in this tunnel, measurements are also of interest from a technical perspective. Previous testing in March 2020 showed that trichloramine, chlorine gas and chloroform were formed in the pilot during RC with sodium hypochlorite.

6.6.2 Method

Tests were conducted to assess the levels of trichloramine as well as chlorine gas and chloroform emitted during NaOCl RC in the MBR pilot in March 2022. Sampling in the exhaust ventilation channel of the membrane tank (Figure 28) was setup for a period of about 20 hours after initiation of the NaOCl RC. Sampling was started at the same time as addition of sodium hypochlorite to the membrane tank was started.

Filling of the membrane 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 210 m^3/h.

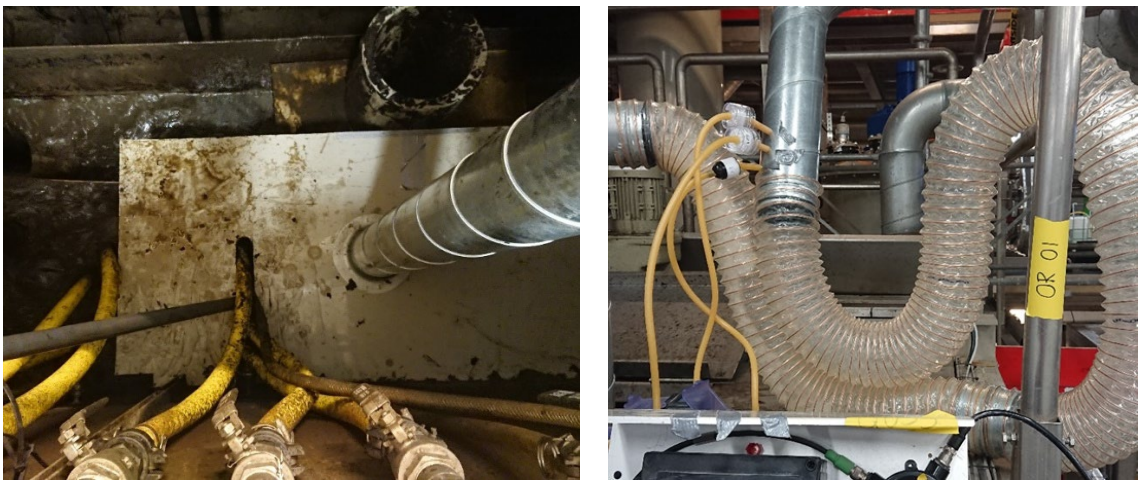


Figure 28. Left: Gas-sampling hood and ventilation channel placed over a membrane tank. Right: Ventilation channel and example of sampling setup for volatile chlorinated compounds.

Trichloramine samples (samplers supplied by Occupational and environmental medicine - University Hospital in Örebro), chlorine samples (SKC 225-9006 samplers) and chloroform samplers (Tenax adsorbent) were collected in series during a 21-hour period after addition of hypochlorite was started (see Table 23). Samples were collected using SKC AirChek TOUCH air sampling pumps. Flow rates were measured using a TSI 4040 flow meter and varied to compensate for different sampling times to make sure that samplers were not oversaturated. Sampling was repeated on two separate occasions, once for each RC of the two sets of membranes in the pilot, two days apart.

Table 23. Sampling schedule.

Analyte	Sampling period	Flow rate (L/min)
Trichloramine	0–1 h	1
	1–2 h	1
	2–3 h	1
	3–4 h	1
	4–5 h	1
	5–6 h	1
	6–7 h	1
	7–8 h	1
	8–9 h	1
	9–10 h	1
	10–21 h	0.5

Analyte	Sampling period	Flow rate (L/min)
Chlorine gas	0–2 h	1
	2–4 h	1
	4–6 h	1
	6–8 h	1
	8–10 h	1
	10–21 h	0.3
Chloroform	0–1.5 h	0.05
	1.5–3 h	0.05
	3–4.5 h	0.05
	4.5–6 h	0.05
	6–7.5 h	0.05
	7.5–9 h	0.05
	9–10.5 h	0.05
	10.5–21 h	0.015

The overall RC-procedure consist of cleaning first with sodium hypochlorite and then with acids a few days later. Section 6.5.3 provides information about RC schedule and amounts of chemicals used and conditions at start and end of the soaking. During the second RC there was an operational problem in the filling phase. A

tube in the sodium hypochlorite line broke and filling had to be stopped after filling about 2/3 of the total volume. The calculated remainder of sodium hypochlorite to be added was poured directly into the tank while filling clean water to the target level. This change in procedure could have had an impact on the results.

6.6.3 Results and discussion

Table 24 provides total amounts and peak concentration of volatile chlorinated compounds emitted for the two sampling occasions during recovery cleaning. In general, the emitted levels were significantly lower the first sampling occasion, with on average 40 % less total emitted analyte compared to the second sampling occasion. The second run showed levels that were closer to what had been observed in previous experiments (Andersson et al., 2021b).

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

Analyte	Sampling occasion	Total emission (mg)	Peak concentration (mg/m³)	Reference level (mg/m³)
Trichloramine	1	3800	0.96	0.2 [1]
	2	12 700	7.3	
Chlorine gas	1	1600	0.67	1.5 [2]
	2	3000	1.1	
Chloroform	1	900	0.28	10 [3]
	2	1400	0.85	

(1) Recommended limit value (Public Health Agency of Sweden).
(2) Short term limit value.
(3) Occupational limit value.

Generally, the results from the two sampling occasions were varying and difficult to assess. Since the pump broke during addition of the sodium hypochlorite (second sampling occasion), about 20 % of the solution was added by hand directly into the membrane tank about 2 hours into RC. As can be seen in Figure 29 (right), a sharp peak in emission of trichloramine was observed short thereafter. This was not observed during the first run Figure 29 (left), where the emission pattern was like what has been observed in previous studies, i.e., with a pronounced dip after the first hour of RC.

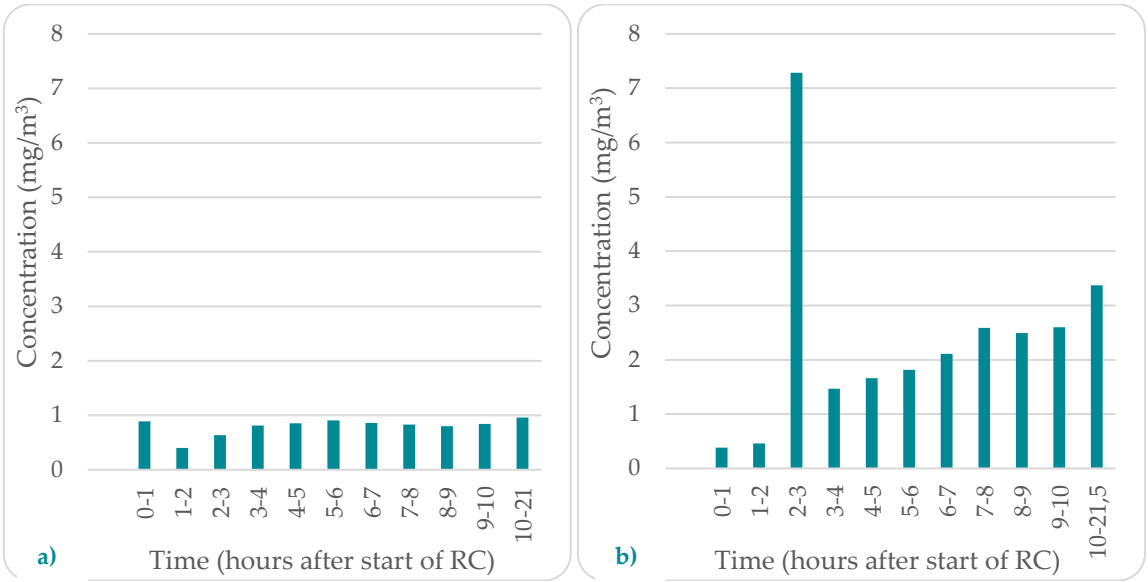


Figure 29. Emission of trichloramine and during a 21-hour period of recovery cleaning with sodium hypochlorite (a – first sampling occasion, b – second occasion).

A clear effect of adding the hypochlorite solution directly into the membrane tank could also be seen for chlorine gas (Figure 30) and chloroform (Figure 31), though not as clear as for the trichloramine.

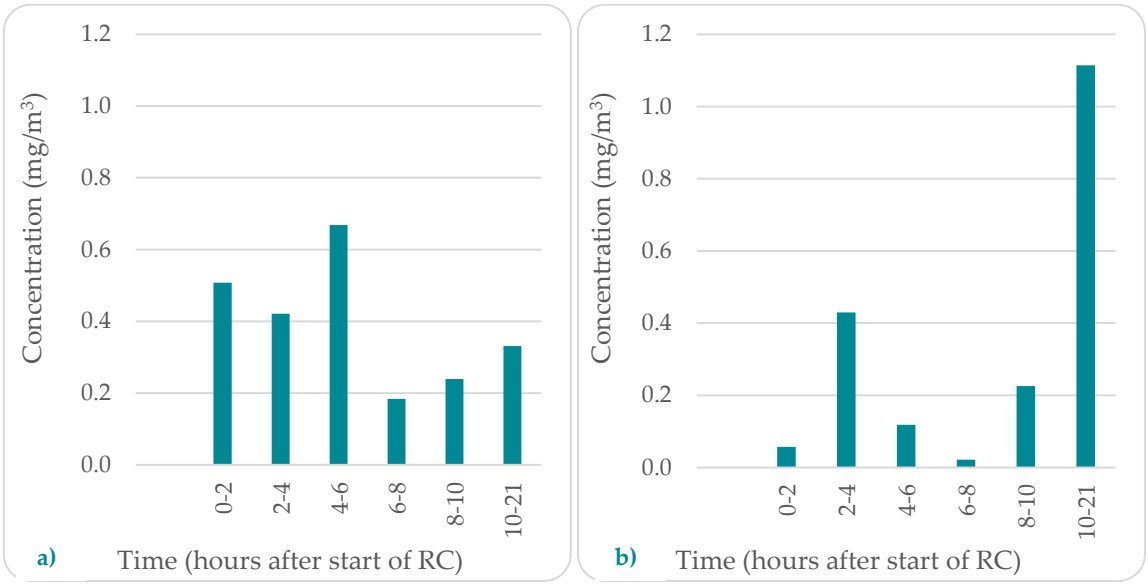


Figure 30. Emission of chlorine gas during a 21-hour period of recovery cleaning with sodium hypochlorite (a – first sampling occasion, b – second occasion).

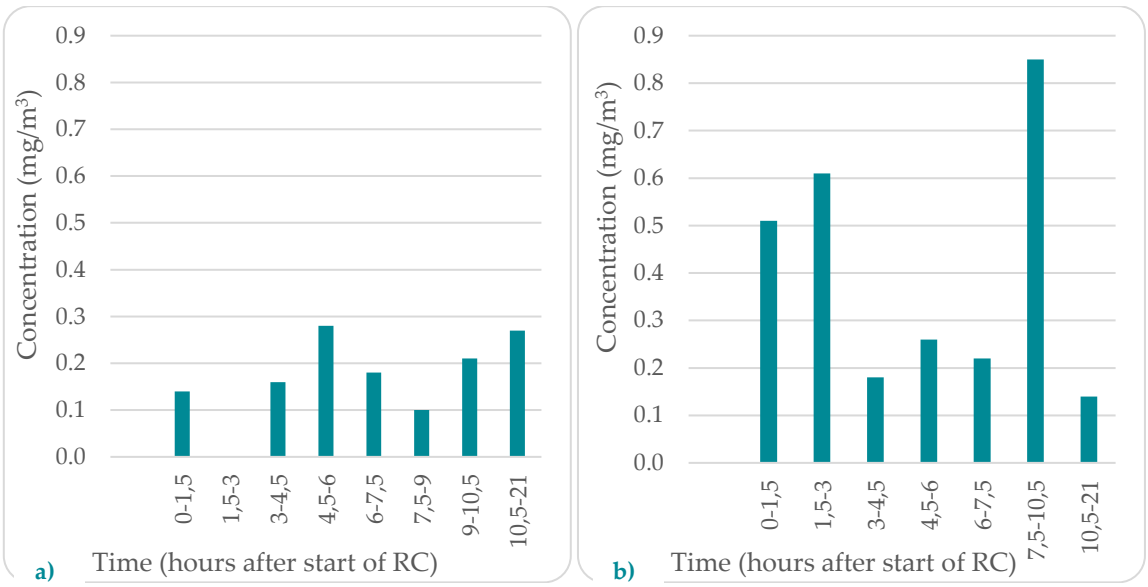


Figure 31. Emission of chloroform during a 21-hour period of recovery cleaning with sodium hypochlorite (a – first sampling occasion, b – second occasion).

6.6.4 Conclusions

The emission process was slower than expected, and, apart from chloroform during the second sampling occasion (Figure 31b), no clear sign of attenuation of emissions was observed during the 21 hours of sampling.

Since the last data point is an average of 10 hours of sampling, it may however be possible that a peak in the emissions occurred during the night. Nevertheless, from an exposure perspective of the results it is safest to assume that the emissions can be harmful during the entire RC process. As a reference, the peak concentrations can be compared to the occupational limit values in Sweden (Arbetsmiljöverket, 2018), or in the case of trichloramine, recommended exposure limit. Trichloramine peaked at 36 times the recommended limit, chlorine gas at 73 % of the short-term exposure limit (15 min exposure) and chloroform at 9 % of the occupational exposure limit (8-hour workday average).

6.7 Emissions of greenhouse gases (GHG)

6.7.1 Background

Direct greenhouse gas emissions from the MBR pilot were previously measured by two shorter campaigns in 2014 and 2018 including measurements of nitrous oxide (N_2O) and methane (CH_4) (Baresel et al., 2022). The two campaigns coved the pilot configuration prior to and after reconfiguration in 2016. It was then found that about 0.004 % and 0.07 % of the total ammonium loads were emitted as N_2O, CH_4 emissions were 0.026 % and 0.12 % of incoming TOC (0.008 % and 0.04 % of incoming COD) in 2014 and 2018. The obtained N_2O-emission values were generally relatively low. The measurements further suggested that a high aeration at the beginning of the treatment line may result in significantly higher emissions of both N_2O and CH_4. A significant change in aeration in the membrane ultrafiltration tank did on the other hand not have the same impact.

To follow up these previous measurements, a new measurement campaign of nitrous oxide (N_2O) and methane (CH_4) was started in July 2021 and continued until November 2021. Previous measurements were carried out in November 2014 and April/May 2018.

6.7.2 Method

Process air from the top of relevant reactors (Figure 32) was collected and analyzed with an on-line instrument (Fresenius, GA2020). As a previous measurement campaigns in 2014 and 2018 showed that the main part of the emissions occurred in the aerated zones even if the formation of N_2O can happen in anoxic zones, only aerated process parts were in focus for the gas measurement campaign. Sampling and analyzing was realized from six sampling points simultaneously by an automatic sample changer. Multiple sampling points could be placed in the same sub-process to catch potential differences depending on sampling location in the reactor.

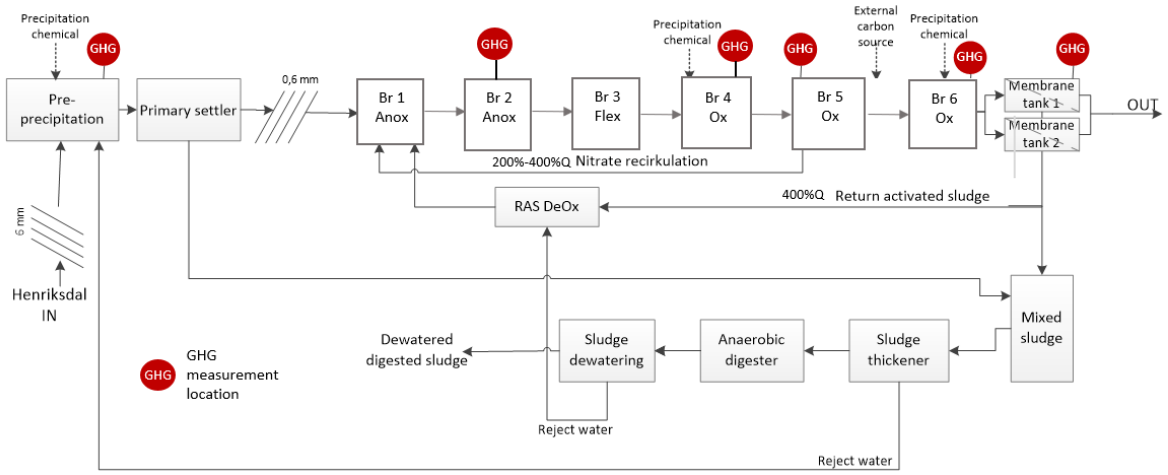


Figure 32. The MBR-pilot including sampling locations.

In difference to measurements in 2014 and 2018, specific gas-sampling hoods were installed in BR3, BR4, BR5 and the MBR tank in 2021. For BR3-BR5, the measurement hood covered half the reactor surface while the other half surface was covered by the reactors original lid, connected with the ventilation system (Figure 33). In this way, external air flows disturbing the measurement were tried to be minimized.

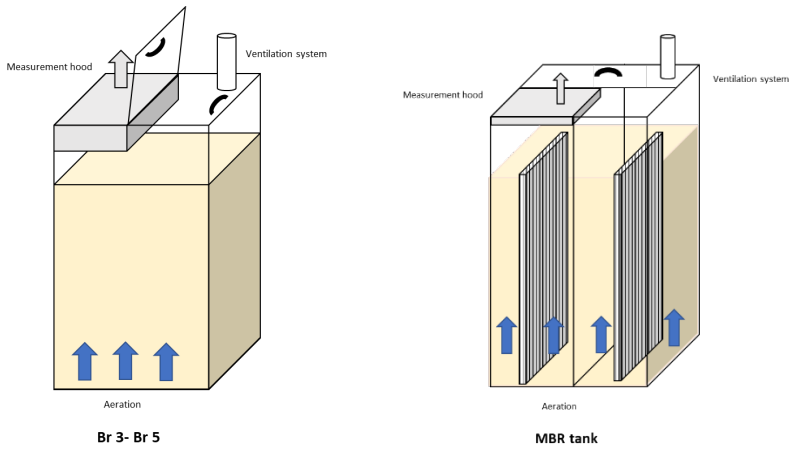


Figure 33. Gas sampling in BR3-BR5 and the MBR-tank.

Regarding the operational differences during the measurement compared to previous campaigns, the higher load (see Table 2) and use of glycerol as external carbon in 2021 can be mentioned. The sludge age in the whole treatment line was 14 days both in 2018 and 2021. However, in 2018 the TSS value was around 9000 mg/L and 11 000 mg/L in 2021, which may affect emissions.

N_2O measurement were carried out both in gas and water phase. The sensor used for measuring the dissolved N_2O in the water phase was a Clark-type microsensor provided by Unisense Environment A/S in Aarhus, Denmark. During the measurement campaign, the N_2O water sensor was placed at different locations in the pilot.

Table 25. Measurement in 2021 for gas phase (G) and water (W).

	July	Aug	Sep	Oct	Nov
Pre-aeration			W	G	

BR1				W	
BR3	G	G	G	G	
BR4	G	G	G	G	
BR5	G	G	G	G	
BR6					W
MT1			G	G	W

6.7.3 Results and discussion

The overall N_2O-emission in 2021 were higher compared to emissions obtained in 2014 and 2018 (Figure 34). Only emissions in the pre-aeration tank gave similar N_2O-emission in 2021 as 2018. The total emission factor (N_2O-N/NH_4-N load) in 2021 was between 0.058 % - 2.345 % (N_2O-N/NH_4-N load) with an average factor of 1.44 %. This in in the same range as reported by earlier published full-scale studies with biological nitrogen removal (0.1 % – 1.9 % of the incoming nitrogen load, cf. Massara et al., 2017). However, compared to average emission factors of 0.004 % in 2014 and 0.07 % in 2018, emissions in 2021 were significantly higher. The main emission source in 2021 was from MT1 with 43.6 g N_2O/d, which was significantly higher than in previous campaigns.

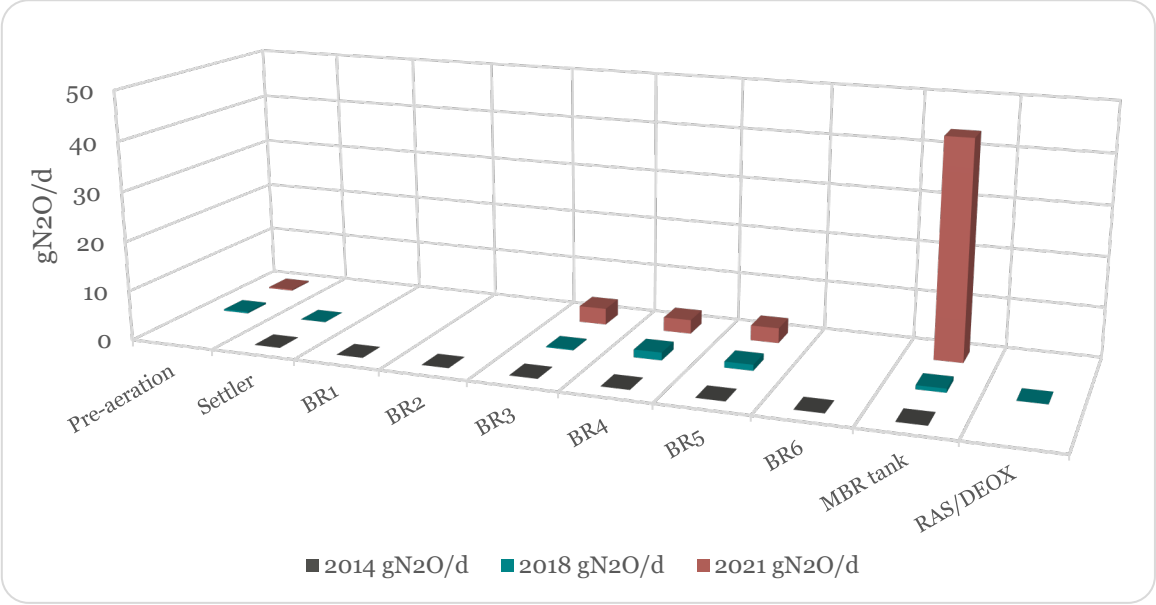


Figure 34. N_2O emission from each reactor from the three campaigns in 2014, 2018 and 2021.

The dissolved N_2O concentrations were generally low but higher in the anaerobic reactors (BR1 and BR6) than in the pre-aeration and the membrane tank. This, as produced N_2O is stripped out in the aerated reactors while it remains in the water phase in the anoxic tanks. The highest N_2O concentration was observed in BR6 with an average value of 0.15 mg N_2O/L.

Methane emissions were also higher in 2021 than during measurements in 2014 and 2018 (Figure 35). The highest emission was observed in the pre-aeration tank, which was the same in 2018. However, the emission from pre-aeration tank was with 46 g CH_4/d almost 5 times higher in 2021 than in 2018 with a value of 10 g CH_4/d. In other tanks, the increase was even more significant.

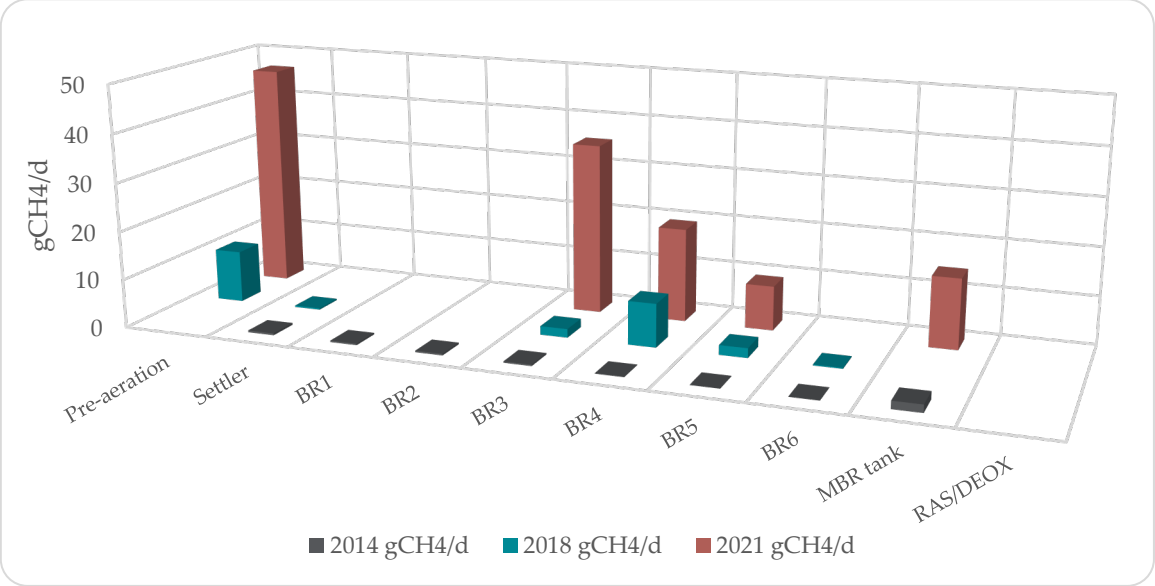


Figure 35. CH_4 emission from each reactor from the three campaigns in 2014, 2018 and 2021.

6.7.4 Conclusions

Emissions of both nitrous oxide (N_2O) and methane (CH_4) observed during the measurement campaign in 2021 were significantly higher than in previous campaigns in the same pilot (Baresel et al., 2022). Considering the higher potency of N_2O as greenhouse gas (298 times more potent than CO_2), especially the high N_2O-emissions from the membrane tank could be identified as crucial for the total greenhouse relevance. Several parameters that could be responsible for these unexpected results were investigated but without being able to identify a clear explanation. The increased load with a maintained effluent quality and a "better" sampling without risk for dilution may be the main explanation but the order of magnitude of the increased emissions leaves room for a more throughout investigation.

6.8 Tests with a flux enhancer product

6.8.1 Background

Kemira has developed a product that is intended to improve the sludge's filtration properties in the MBR process. The product is a mixture of ferric chloride and a polymer and has previously been tested in a shorter experiment with flat sheet membranes where the filtration properties were improved. In our tests with our hollow-fiber membranes, the product was aimed to be tested over a longer period of time.

Based on the short initial tests with the flux enhancer product, an increase in membrane permeability and decrease in TMP was expected. Initial trials at Kemira were performed with constant TMP and thus evaluated against change in flux instead. Secondary effects expected were, at least to some extent, reducing energy requirement for filtration and chemical consumption for membrane cleaning. This, e.g., by increasing the time between necessary MC. Further, sludge properties, including better dewatering ability, were anticipated to improve as sludge-flocks were expected to become more stable.

6.8.2 Method

The product must be dosed continuously to the biological treatment step. The test programs and dosages were proposed by Kemira.

During a 30-day period before the start of the experiment, a recovery cleaning (RC) was performed on the membranes with citric acid/oxalic acid and sodium hypochlorite. On June 21, 2021, when the reference period started, the flux in the membranes was increased from 20 L/(m^2·h) to 30 L/(m^2·h). The high flux was maintained throughout the experiment. On July 5, the flux enhancer was continuously dosed at 8 g Fe/m^3 (equivalent to 63 mL flux enhancer/m^3) in BR6 and on August 9, the dose was increased to 10 g Fe/m^3. On September 8, the dosing of flux enhancer was stopped, and a second reference period started and continued until November 15.

The first 14 days after the dosing stopped (corresponding to one sludge age) were excluded from the evaluation as the concentration of flux enhancer in the sludge decreased continuously during that period. Remaining effects of the product after 14 days were estimated to be negligible. During the second reference period, the process was operated with the same settings except that the return sludge flow (RAS) was changed on October 21 from 2 × Q_{in} to 4 × Q_{in} due to other trials.

The dosing point for flux enhancer replaced the dosing for precipitation chemicals in BR6 (anoxic tank, Figure 1). Dosage of ferrous sulfate (Hepta) before pre-sedimentation was continued as usual with 10 g Fe/m^3 during the experiment. Since flux enhancer consists largely of PIX, the total dose of iron dosed in the biological treatment increased.

Maintenance cleaning (MC) on the membranes with oxalic acid (MT1), citric acid (MT2) and sodium hypochlorite (both MT1 and MT2) was performed continuously throughout the experimental period.

TTF for sludge from the membrane tanks was beside the permeability the main parameter monitored and was analyzed approximately 3 times per week during the dosing period. As TTF is a measure of the sludge's filtration properties, a lower value of TTF implies that it is easier to filter a certain volume of permeate and thus that the sludge has better filterability.

6.8.3 Results and discussions

The permeability decreased by approximately 50 L/(m^2·h)/bar in connection with the increase in flux on June 21 and even thereafter continuously during the first reference period (Figure 36). During the period when the flux enhancer was dosed, the permeability trends observed appeared smoother and generally more stable in MT2 than in MT1. However, no obvious positive or negative change in permeability due to dosing of flux enhancer was possible to identify based on continuously monitored process parameters (see permeability in Figure 36). There are commonly observed variations in permeability and effect of membrane cleaning which made this evaluation difficult. Furthermore, it was not possible to note any effect of the dose increase from 8 g/m^3 to 10 g/m^3.

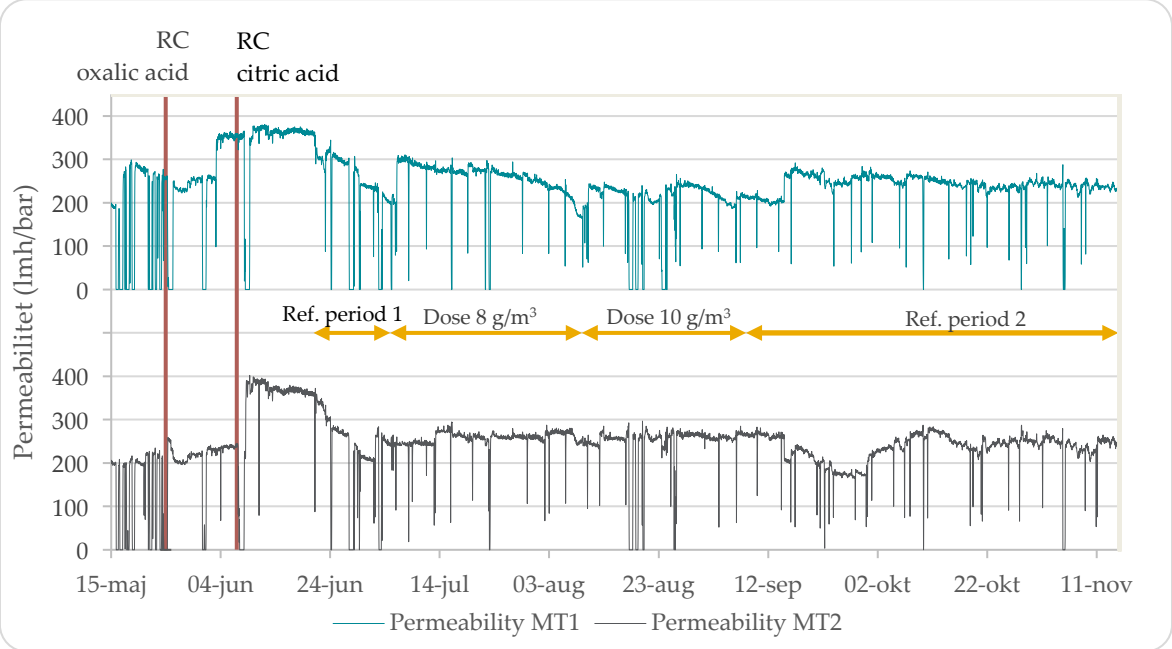


Figure 36. Permeability in MT1 and MT2 during RC, reference periods and dosing with flux enhancer.

During reference period 2, the permeability was smoother and slightly higher than before for MT1 while the permeability for MT2 varied more (Figure 36). This is the opposite compared to when flux enhancer was dosed. As there is only one treatment line, only periods before and after the dosing tests could serve for comparison. In this type of comparison, it is good to remember that it is not known how the permeability in MT1 and MT2 would have been without dosing flux enhancer. In addition, there were differences between the membranes; even and slightly increasing permeability for MT2 while it was more variable and over time decreasing permeability for MT1, which made it difficult to draw any conclusions.

The two air flow modes in the membrane tanks, Leap-Lo (14.0 m^3/h) and Leap-Medium (20.0 m^3/h), are controlled by the TMP and switch thus automatically. An expected improved permeability by dosing flux enhancer should thus imply that Leap-Lo mode would be activated more frequent than without dosing of flux enhancer. However, no clear difference in air flow could be observed between the dosing period and reference periods (Figure 37). The sum of the air flow to the membrane tanks during the reference period before dosing was on average 29.3 m^3/h, during dosing 34.6 m^3/h and 36.6 m^3/h during the reference period after dosing. The initial switch from Leap-Lo to Leap-Medium in MT1 right before the start of the flux enhancer dosing was due to a decreased permeability after the increases of the flux at the beginning of the reference period.

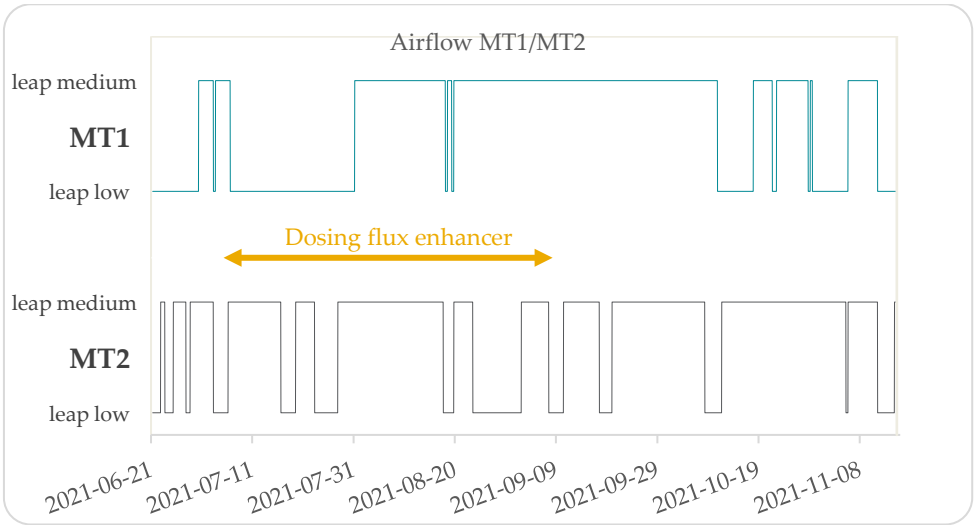


Figure 37. Air flow mode during the test period.

The effect of MC varied and had sometimes a large but sometimes also very small effect on the permeability. Further, no significant difference in the frequency of MC could be observed between the test period with flux enhancer and the reference periods before and after (Figure 38). For MT1, the frequency of MC with hypochlorite increased during the flux enhancer dose, while the frequency decreased slightly towards the end of the trial period for MT2 and then increased again during reference period 2. MC with hypochlorite is performed only when the membranes are aerated with Leap-Medium. This is the reason why no hypochlorite MC was performed in MT1 during the first weeks of dosing. MC with citric acid and oxalic acid takes place after a certain volume of permeate has been produced and thus does not depend on the TMP of the membrane.

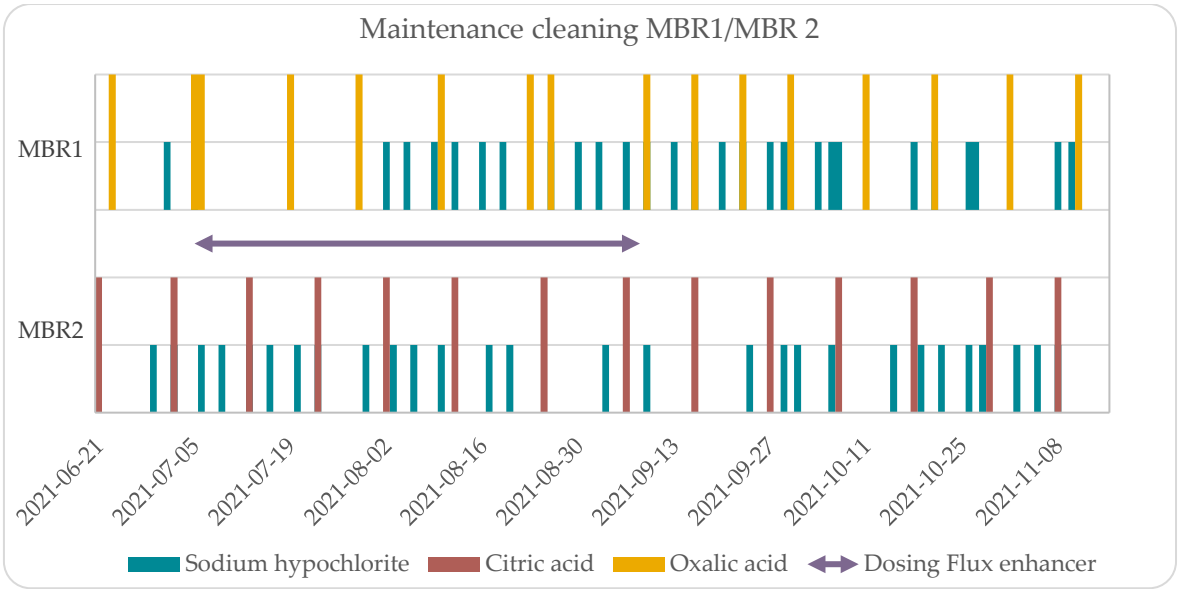


Figure 38. Maintenance cleaning (MC) in the two membrane tanks MT1 and MT3.

The iron content in RAS gradually increased in the beginning of dosing of flux enhancer. The effluent phosphorus decreased as expected as the addition of iron increased with dosing of flux enhancer, however, from already low levels.

When comparing TTF and the mean permeability of the two membrane tanks it can be observed that both follow the same trend. In contrary to expected results, a slight increase of TTF and slight decrease in permeability during the flux enhancer dosing was observed (Figure 39). However, TTF increased even more after dosing stopped and during reference period 2.

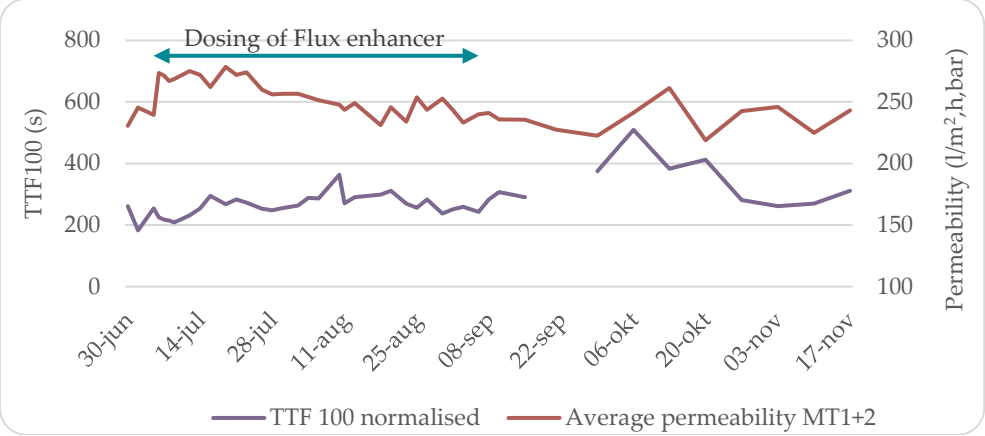


Figure 39. TTF during the period of dosing of flux enhancer.

It was observed that the sludge had a slightly glossy and gelatinous character during the test period with flux enhancer. When filtered through filter paper in SS analysis, the retentate formed a stable gel-like sludge cake. This phenomenon had not been observed before or after dosing with flux enhancer.

When analyzing the sludge volume index (SVI) it happened on several occasions that the sludge's ability to settle was very poor and formed very little separation between sludge and clear phase. At the end of the dosing experiment, weeks 34-37, it happened that the sludge instead floated above the clear phase. It should be added that SVI analyses with poor sedimentation were not consistent. The same sludge sample could settle during an experiment and not sediment if the experiment was repeated.

During the test period with dosing of flux enhancer, an increased dosing of polymer was required to thicken the mixed sludge to achieve the same quality as before with clear flocks and clear supernatant. The polymer in the flux enhancer may have adversely affected the thickener polymer. However, several other factors such as poor admixture of polymer and clogged sieve drum may have caused poorer thickening as well.

6.8.4 Conclusions

Based on evaluated process parameters during the test period with flux enhancer and the two reference periods, as well as regarding various influencing factors such as technical problems and load variations, no clear positive effect of flux enhancer dosing on membrane operation could be observed. Related to the initially expected results, it can be stated that:

- No increase in membrane permeability or decrease in TMP was observed.
- The time between membrane cleaning (MC) did not change more than within normal variations.
- Sludge properties did not improve but rather deteriorated and the sludge dewatering ability deteriorated.
- Effluent phosphate was very low during dosing of flux enhancer due to an increased iron dosing. Phosphorus levels increased gradually after dosing of flux enhancer had ended.

6.9 Tests with a defoaming agent

6.9.1 Background

In MBR plants, the formation of foam is a common phenomenon and, as in conventional active sludge systems, caused by the same agents: bacteria-generated surfactants. However, since MBR systems retain more of the extracellular polymeric substances (EPS) than CAS systems, foaming is largely attributed to excessive amounts of EPS, rather than directly to foam-forming microorganisms. Operational conditions of MBR system that might promote growth of foam-developing bacteria are the retention of fine solids (larger than 0.01-0.1 µm), high TSS concentration, high SRT and low F/M ratio. Therefore, foaming in MBR systems can be considered a standard process condition.

In the pilot, however, enhanced seasonal foaming, mainly in the spring, has also been a recurring problem every year. Foaming mainly occurred in aerated zones (BR3-BR5), in the post-denitrification (BR6) and in the RAS-deox. The buildup of foam often led to flooding of the aerated tanks and loss of SS concentration because of an insufficient freeboard. Foam also complicates online measurements by interfering with submerged sensors. In general, foaming causes the same adverse effects as for CAS including difficulties in estimating solids retention time (SRT) and work environment challenges.

Based on previous experiences with foam in the pilot, the design for the full-scale MBR at Henriksdal WWTP included a surface collection and wasting infrastructure for foam and scum. However, during the startup period of the first full-scale MBR line in the spring of 2021, extensive foaming occurred and the foam separation and wasting system was insufficient to handle the problem.

6.9.2 Method

To investigate other options for foam control in the pilot, a chemical defoaming agent was dosed to the biological treatment during the period of heavy foaming (March-June). The product used was FloFoam H16 from SNF. No specified recommendation for dosage levels were provided by SNF but an approximate dose of 50 ppm was referred to be applied at other similar processes.

Batch dosing

Two trials with batch dosing were carried out with the same dosing point in BR1 (Table 26). Dosage volume was 10 ppm of total sludge volume in line (34 m^3). At each trial, 340 mL defoaming agent was measured in a beaker and poured at once from top of the tank into the sludge.

Table 26. Batch dosing periods.

Dosing point	Start date	Dosage (ppm)
BR1	2021-03-02	10
	2021-03-08	10

Continuous dosing

Two trials with continuous dosing were conducted with different dosing points to investigate the effect of dosage levels and dosing point. The dosing points chosen were BR1 (non-aerated zone), the first tank of the biological line where the incoming and recirculation flows met, and BR3 (aerated zone) the first of the two tanks where the most foaming problems occur.

The first test in BR1 was performed for 43 days (March 11th to April 20th, 2021) and the second test with dosing point BR3 was performed for a duration of 51 days (April 20th to June 10th, 2021). Five levels of dosage were tested as shown in Table 27.

Table 27. Overview of dosage levels during continuous dosing tests.

Dosing point	Start date	Duration (days)	Dosage (ppm)	Test dosage	Comments
BR1	2021-03-08	7	4.4	Low	Average value.
	2021-03-15	4	24	High	Test with high dose with target 50 ppm. Resulted in 24 ppm. Pumping problems with led to failure to reach high flow. Routines were established for pump calibration.
	2021-03-19	3	0	Pause	
	2021-03-22	7	14.2	Low/ Medium	Average value. Dosage varied between 8.8-15.2 ppm due to operational discrepancies in the pilot.
	2021-03-30	13	10	Low	
	2021-04-13	0	0	Pause	Paused to see effect, but the dosage started after 1 hour due to high foam levels that led to flooding of the pilot.
BR3	2021-04-14	6	20	Medium	

The target for the high dose (Table 27) was 50 ppm based on incoming flow (4.54 m^3/h), but due problems with the dosing pump it was not possible to obtain a higher dosage than 24 ppm. Therefore, the high dose test was canceled after 4 days.

The dosage levels varied occasionally during the second and third test period with dosing point BR3 due to other operational issues in the pilot. These problems, however, was remedied after a few hours to one day, apart from the weekends where there were not staff present.

Further, microscopic investigations of the foam and sludge were performed but without any useful information as a result.

6.9.3 Results

Batch dosing

No effect on the foam was observed in the batch trials. However, the oxygen content dropped momentarily (approximately 10 min) after dosing in the aerated zone but recovered after approximately 30 min. In addition, the SS content of BR4 and RAS increased rapidly shortly after dosing and then dropped to a level lower than before dosing (Figure 40). During the following days, the sludge content recovered slowly towards the initial condition. Most likely the changes in SS was because the sensor was affected by the product. For DO, it is unclear whether the fast drop was due to an actual change in oxygen (which could be the case if the product contained available COD) or because the sensors were affected negatively.

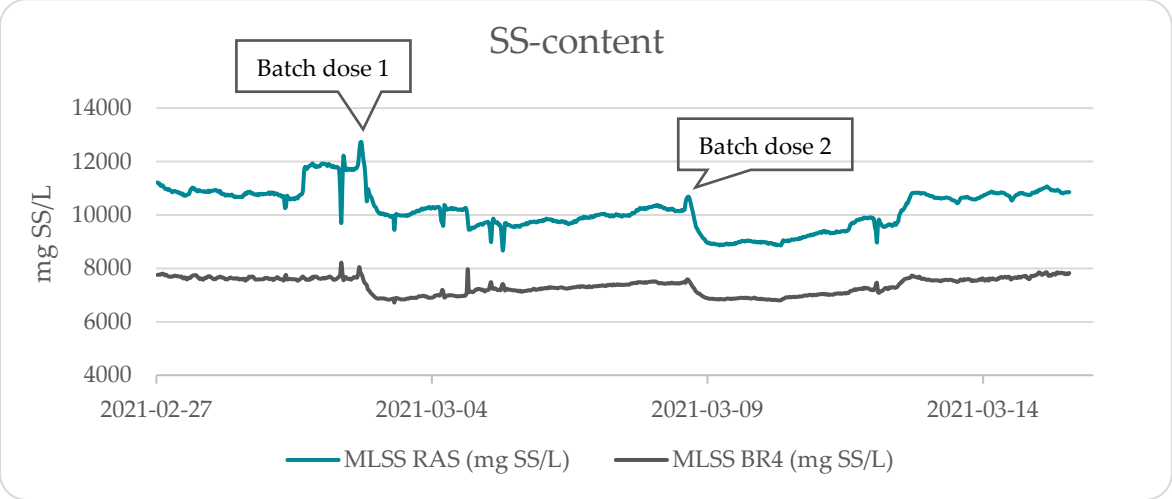


Figure 40. SS content in BR4 and RAS during batch dosing.

Continuous dosing

BR1: Very low dose (4.4 ppm)

The test started initially with very low dosage. No visible mitigating effect on the foam level and no negative effect on the process performance was observed from the dosage, and hence the dosage was changed to high dose.

BR1: High dose (24ppm)

A significant reduction in foaming by approximately more than half the intensity was achieved already at the dose of 24 ppm, despite the recommendation that had been 50 ppm. It was therefore decided to lower the dose to see if it was possible to reduce the consumption and maintain the effect. Before the trial with lower dosage, the dosage was stopped completely for 3 days to allow the effect of the previous dose to subside. During the 3 days without dosing the foam levels increased to the same levels as before the dosing started.

BR1: Low dose (10 ppm)

A test with dosage of 10 ppm in BR1 started on March 22, but due to operational issues in the pilot, the dosage varied between 8.8 to 15.2 ppm during the first week of this trial, and therefore no evaluation is included in this report between March 22nd to 30th 2022. A stable dosage of 10 ppm started on 30th of March 2021. The amount of foam in terms of foam thickness in the tanks was reduced already after one hour and problems with foam flooding over the tank's freeboard stopped. While the dosage was stable, the observed levels of foam were low in all tanks. A test to stop the dosage of defoaming agent resulted in a quick return of foaming problems. To avoid flooding, the dosage had therefore to be turned on again already after 1 hour without dosing. This indicates that the defoaming agent has no lasting effect once dosing has ended.

BR1: Medium dose (20 ppm)

To investigate whether it was possible to remove the foam completely, the dose was increased to 20 ppm on the 14th of April 2021. Initially, the foam level dropped in the tanks, especially in BR3 where it was possible to spot the water surface in the vicinity of the stirrer. However, the foam returned slightly after one day, and the overall effect was only slightly better compared to 10 ppm dosage (Figure 40).

Figure 41. Foam in BR4 at no dosage of defoaming agent (left), a dose of 10 ppm (middle), and a dose of 20 ppm (right).

BR3: Medium dose (20 ppm)
To investigate the effect of the defoaming agent when dosed in different parts of the process, the dosing point was moved from BR1 (anoxic, foam-free zone) to BR3 (aerated zone with foaming), i.e., the beginning of the foaming zone of the line. The dosage tested was 20 ppm, as it had the best observed effect in the test with dosing point BR1.

Compared to the dosage in BR1, the result was very similar except for a slight increase in foaming in BR4. However, some increase in foam levels in BR3 and BR4 was already observed the days before, and thus it was difficult to define whether the change of dosing point had a positive or negative effect.
During June, foaming usually decreases in the pilot due to higher and more stable water temperatures. Therefore, the dosage was stopped on June 10th, 2021, and the test with defoaming agent was completed.

BR3: Observation after complete dosage stop
The foam returned gradually in the pilot, but more slowly than in previous trials with stopped dosage. After 5 days, however, the foam level reached the edge of the freeboard of BR4 which resulted in foam flooding. Thereafter, sporadic flooding occurred in BR4 and BR5.

The foam also changed character and became a "dense mousse" compared to the foam during dosing of defoaming agent, when it was more labile, with larger bubbles (Figure 41).

Figure 42. BR3 before planned dosing stop 10th of June (left), BR3 during the stop (middle) and BR4 during the stop (right), both on the 24th of June 2021.

The dosage stop did not result in continuous foaming problem, and the foaming ceased circa 30 days after dosage stop. This could however have been different if the dosage stop had occurred earlier in spring, as the foaming generally ceases during July in the pilot and the full-scale plant.

6.9.4 Conclusions

The overall effect of the defoaming agent is assessed as useful to minimize foam during periods of problematic foaming. Optimal effect in the pilot was achieved with continuous dosages of > 10 ppm when the foam was subdued to the level without causing any flooding and uncontrolled loss of sludge from the line. This dose could be adjusted as the foaming ability varies in different circumstances such as temperatures, and how much foaming is to be reduced. However, further investigation is required to determine the control parameters.

Further investigation is also required to study the change in sludge properties upon dosing, which might be an aspect to be considered in foam and sludge handling. All dosage changes (start, dosage concentration and stop) resulted in an immediate response, mainly within 1-2 hours. Using defoaming agent changed the density of the foam, and the bubbles became larger and unstable. The oxygen levels in the process may also be affected when dosing defoaming agent, as observed in the first batch trials. However, it was not possible within the trials to determine whether this was due to disturbances caused by the defoaming agent on DO sensors or whether the defoaming agent lowers the actual oxygen level in the process.

Even though the product has been shown to have a positive effect in the MBR pilot, a permanent use in full-scale may not be economically feasible due to the high consumption. With average inflow of 6 m^3/s at the future Henriksdal WWTP, even the lowest dose applied in the trials with acceptable result (10 ppm) would require about 5 tons of antifoaming agent per day, which amounts to ca 5.5 MSEK per month with the prices at the time of trial (prior to COVID-19).

6.10 Test with reduced RAS flow

6.10.1 Background

The return activated sludge (RAS) flow is designed to be $4 \times Q_{in}$. The reason for this is that the membrane supplier does not recommend sludge concentrations higher than 10 000 mg TSS/L in the membrane tanks at the same time as a high sludge concertation, 8 000 mg TSS/L, is desired in the biological treatment tanks. The RAS flow must be pumped back to the inlet of the biology. In the first MBR line at Henriksdal WWTP the

lifting height for the RAS pumps is around 2 m, more than the required hydraulic height loss, which is a result of the configuration of the old, activated sludge tanks, the cavern height, and other constructional parameters. The large flow and lifting height entail a large energy consumption.

One way to cut the energy consumption in half is to reduce the RAS-flow from 4 × Q_{in} to 2 × Q_{in}. Theoretically this would result in a sludge concentration of 12 000 mg TSS/L in the membrane tanks, if the sludge concentration in the biological tanks is to be maintained at 8 000 mg TSS/L. According to Judd (2018) the membrane capacity, in terms of the average and maximum flux, depends on the TSS-concentration, the air scouring and the crossflow filtration effect from the RAS-flow. Operation at high sludge concentrations in the membrane tank normally requires that the fluxes shall be adjusted (reduced). Energy wise lower fluxes will lead to a higher specific energy consumption per m^3 of permeate and reduced hydraulic capacity.

6.10.2 Methods

The target value of the RAS-flow was changed in three steps:

1. Week 9-12: 3 × Q_{in}
2. Week 13-20: 2.5 × Q_{in}
3. Week 21-41: 2 × Q_{in}

Due to many interruptions in the inflow during the trial, the target values were not always met. From week 42 onward the target value for RAS was returned to its normal value of 4×Q_{in}.

6.10.3 Results and discussion

The weekly average inflows and RAS flows during 2021-2022, as well as the ratio between the RAS flow and the inflow, is shown in Figure 43. As can be seen, the target ratio or a value close to the target ratio was reached most of the time. Figure 44 shows the weekly average concentration of TSS in the biological reactors and the RAS as well as the ratio between the TSS in the RAS and the biology. At a RAS flow of 4×Q_{in} the TSS ratio should theoretically be 1.25 and when the RAS flow was lowered to 2×Q_{in} it should increase to 1.50. Both the TSS concentrations (primarily in the RAS) and the ratio between TSS in the RAS and the biology varied a lot throughout the year. One explanation for this is the problems with sludge bulking and overflow that occurred during spring and the beginning of summer (until week 28 approximately) which made it difficult to maintain the sludge concentration in the biology at 8 000 mg/L. The sludge bulking also made the measurement of the sludge concentration in the RAS tank more difficult. The RAS tank has a very high headspace and can accommodate a large volume of bulking and/or scum that obstructs sampling and measurements by sensors. Between week 24 and 32 the TSS ratio was above 1.5 and the TSS concentration in the RAS was at or above the theoretical value of 12 000 mg/L, which makes it a good period for evaluation.

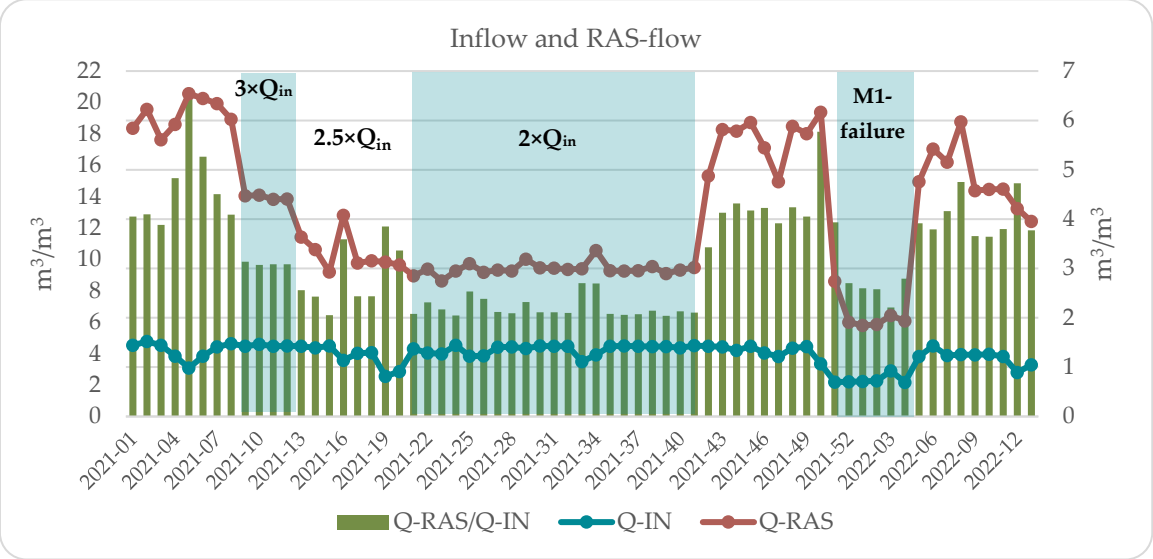


Figure 43. Inflow and RAS flow to the pilot during 2021-2022. Weekly averages.

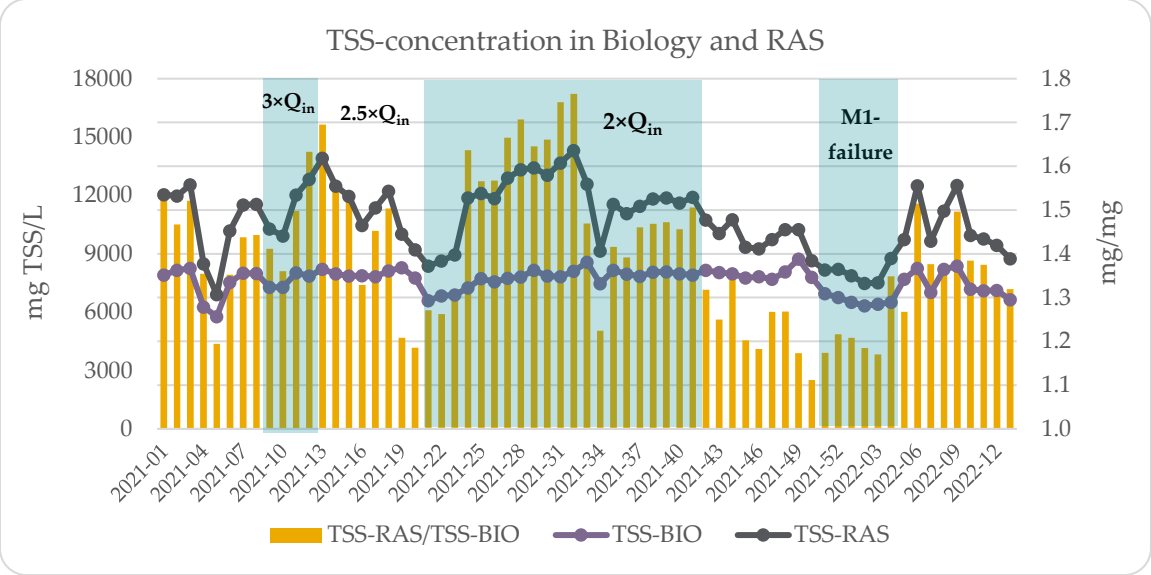


Figure 44. TSS concentration in biology and RAS during 2021-2022. Weekly averages.

To assess if the reduced RAS flow and increased TSS concentration in the membrane tanks had any visible effect on the membrane performance, the TMP, permeability and the membrane aeration was evaluated.

Figure 45 shows the average and the weekly maximum TMP during 2021-2022 together with the time points for recovery cleaning, RC, and the period of the trial with high flux. The TMP increased to around 120 mbar during the period with a RAS flow of $2\times Q_{in}$, although an RC was performed in the beginning of the period. This value can be compared to the annual average TMP year 2020 which was 80 mbar. However, the increased TMP continued after week 41 when the RAS flow was adjusted back to $4\times Q_{in}$. Unfortunately, the trial with reduced RAS flow coincided with another trial with high flux. A high flux result in a higher TMP which is a more likely reason for the TMP increase. All trials related to membranes that are likely to affect the evaluation of the trial with reduced RAS flow are specified in Table 18.

Figure 46 shows the weekly average permeability and the weekly average airflow to the membrane tanks. Average weekly air flows at or below 14 Nm^3/h indicate that the aeration had been in Leap-Lo mode all week.

Air flows of 20 Nm³/h indicate that the aeration had been in Leap-Medium mode all week. Air flows in between indicate that the air flow oscillated between Leap-Lo and Leap-Medium.

From the graphs (Figure 45 and Figure 46) it is difficult to see any effects on the membrane performance that can be direcly and clearly derived from the changes of the RAS flow.

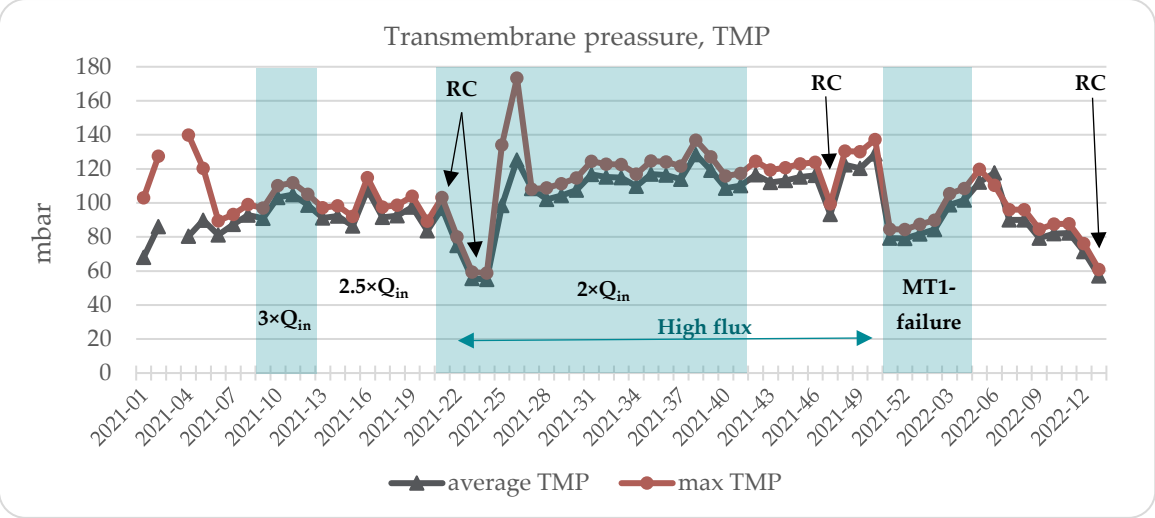


Figure 45. TMP shown as weekly average value (average of MT1 & MT2 when in operation) and maximum value per week.

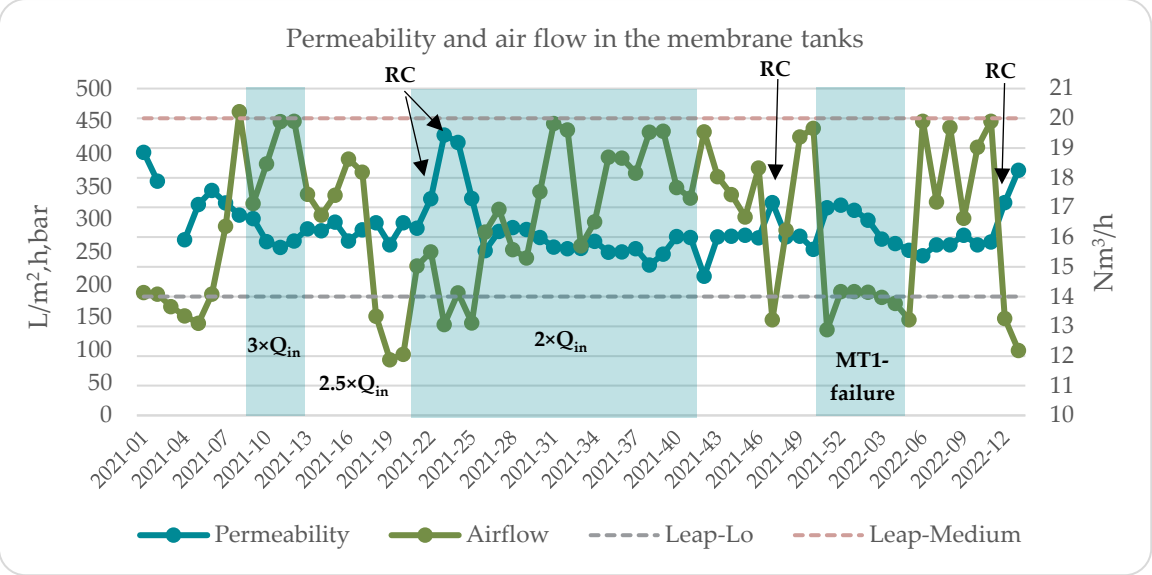


Figure 46. Permeability and air flow per membrane tank shown as weekly average value. Permeability and air flow are average of MT1 & MT2 (when in operation). Dotted lines show air flows for Leap-Lo and Leap-Medium.

6.10.4 Conclusion

No direct effects of the reduced RAS flow could be seen on the membrane performance. However, it is likely that any potential outturn was drowned by effects of other ongoing trials that affected the membranes to a larger extent.

6.11 Back-pulse tests

The operation of both membranes was changed to use back-pulse instead of relax mode between February 21 to March 21, 2023, after the required programming was implemented and tested. The evaluation period also included a reference period of one month where the membranes were operated with relax.

The evaluation of the permeability and TMP indicated no significant differences between the reference period with relax and the back-pulse period. However, due to normal variations in the incoming wastewater and related process conditions, a direct comparison of the two periods is challenging. In addition, the evaluation period was quite short due to the necessary shutdown of the pilot. The effect of back-pulse may therefore be evaluated more thoroughly in future tests.

6.12 Sludge production and sludge properties

In Table 28, data from the pilot related to sludge production and sludge quality is compared to data from the Henriksdal WWTP (annual average 2021) and design data for the future Henriksdal WWTP according to the SFA project design. 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, 2021a, 2021b). The WAS production was slightly higher in 2021, 17.8 kg SS/d compared to 15.8 kg SS/d in 2020 and 13.3 kg SS/d in 2019. The higher WAS production could be explained by the high inflow and reduced reduction in primary sedimentation. The primary sludge production, however, was much lower in 2021 (12.8 kg TS/d) compared to 2020 (21 kg TS/d) and 2019 (18.5 kg TS/d). One possible explanation of the poor separation over the primary clarifier and the low primary sludge production could be the higher inflow, giving too high surface loads for the clarifier, together with the many flow interruptions. In addition, some sludge flow could not be accounted for, i.e. formed sludge cakes on top of the primary clarifier had to be manually removed frequently, and foaming implies several sludge flooding events in 2021.

Table 28. Sludge data from the pilot year 2021 compared to data from the Henriksdal WWTP 2021 and design data for the future Henriksdal WWTP.

Parameter	Pilot data 2021	Henriksdal WWTP data 2021	Design future Henriksdal
WAS production (kg SS/d)	17.8	24 600	59 000
Part of total sludge production (%)	58 %	29 %	34 %
VSS in WAS (% of SS)	75	68	63
Fe in WAS (% of SS)	6.2 %	11	-
Al in WAS (% of SS)	0.7 %	-	-
PS-production (kg TS/d)	12.8	61 500	117 000
Part of total sludge production (%)	42 %	71 %	66 %
VS in PS (% of TS)	89	78	80
Total sludge production (kg TS/d)	30.6	86 100	176 000
Total sludge age, SRTtot (d)	17.2	21.8 (MBR line)	28
Aerated sludge age, SRTox (d)*	7.0**	6.5 (MBR line)	7***
SVI, jan-jun (mL/g)	194	131	-
SVI, jul-dec (mL/g)	338	119	-

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

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

***including membrane tanks, yearly average

A summary of sludge properties analysed in this project that might affect the membrane performance are shown in Figure 47 and Figure 48 below, together with normalized permeability – representing the membrane performance. In Figure 47, other trials that might affect the sludge properties and/or the permeability are also visualised in the graph. The Fe dose was increased significantly during the flux enhancer trial leading to a slow increase of metal in the sludge. From Figure 48 it was seen that the Time To Filter (TTF) was lower during the warm summer/autumn period. The colloidal TOC (cTOC) followed the seasonal changes like TTF. Sludge volume index (SVI), however, did not follow the same pattern, confirming what we have seen previous years that SVI is not related to filterability. Week 34, a huge increase in the SVI was seen coinciding with the increased dose of Flux Enhancer. It is, however, not proven that the two events are related. TTF and cTOC does not seem to be affected by neither addition of defoaming agent nor flux enhancer. None of the parameters showed any clear correlation ($R^2 < 0.1$) with the permeability which probably is due to a generally high permeability and the assumption that the RCs and MCs have a greater effect on permeability than the sludge properties do.

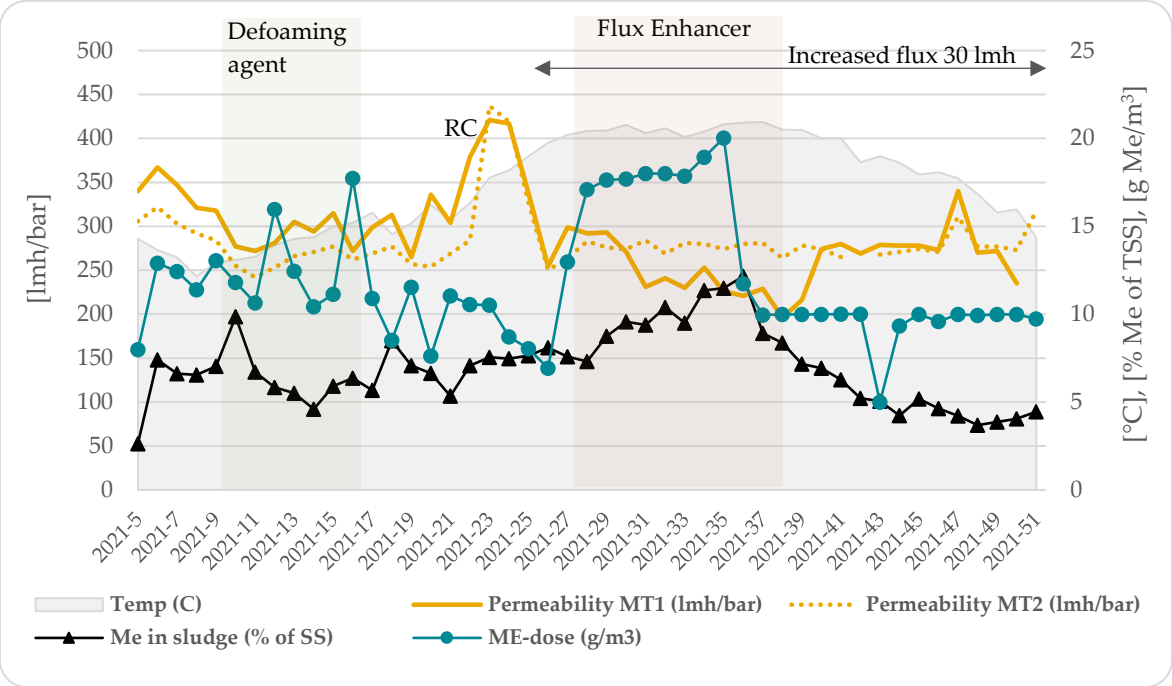


Figure 47. Normalised permeability, temperature, Metal (Me) dose and Metal (Me) content in sludge over the year (weekly averages).

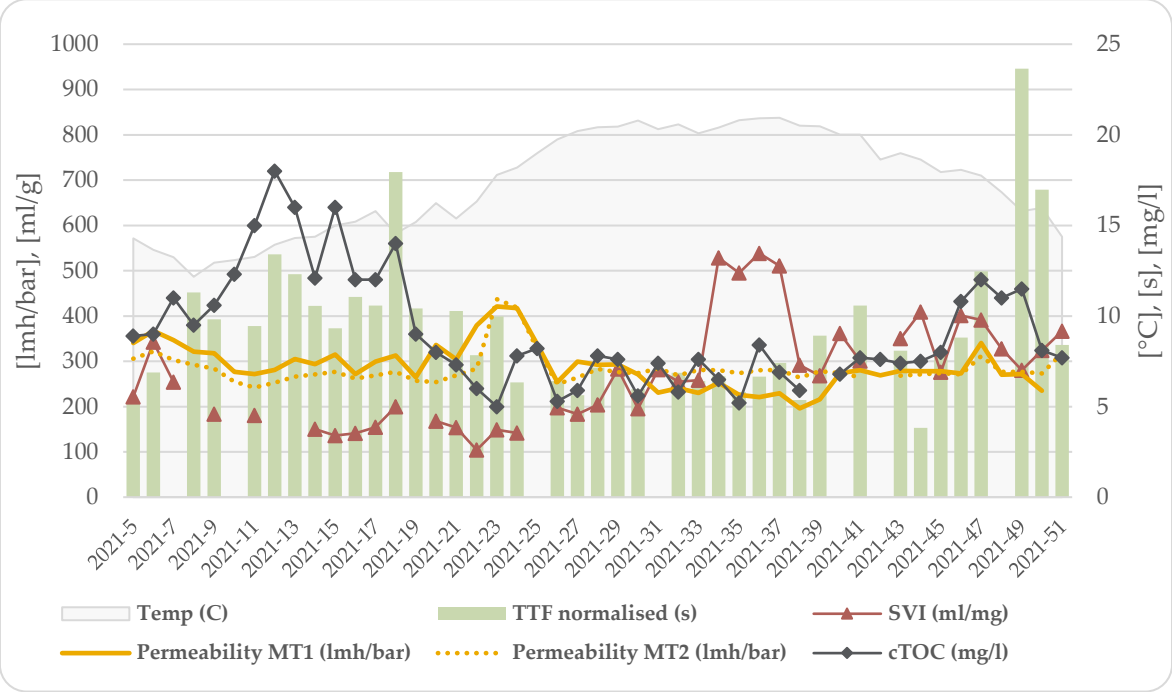


Figure 48. Normalized permeability, temperature, TTF-100 (normalized to a TSS concentration of 10 000 mg/L), SVI and cTOC over the year (weekly averages).

In Table 29 the TSS concentration in waste activated sludge (WAS) as well as the content of iron, aluminium, phosphorus and VSS are listed as annual 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. During 2021 the trial with flux enhancer resulted in 9 weeks of overdosing of iron (compared to what is required for phosphorus removal) which, together with the fact that EBPR was lower than in 2020, gave an increased Me/P-ratio.

Table 29. WAS composition (annual average) in the pilot over 8 years of operation (Me = metal (Fe + Al), n = number of samples). Values for 2022 are from January to end of March.

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)
2022	8 115	4.9	0.2	3.2	80	0.9
n	*13*	*13*	*2*	*13*	*12*	*13*
2021	10 154	6.2	0.7	3.1	78	1.4
n	*52*	*52*	*52*	*51*	*51*	*51*
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*

6.13 Sludge pilot

	2021												2022		
Activity	**J**	**F**	**M**	**A**	**M**	**J**	**J**	**A**	**S**	**O**	**N**	**D**	**J**	**F**	**M**
Sludge thickener in operation															
How low can we go - mesophilic															
Emptying and reseeding the digester															
Mesophilic reference period															
Transition meso- to thermophilic digestion															
Thermophilic reference period															
Dewatering trials															
Optimization of thickener															
High load, low HRT digestion															
Microbial community analysis															
Closing of sludge pilot															

During 2021, three main trials were performed in the sludge pilot:

1. a transition from mesophilic to thermophilic anaerobic digestion
2. dewatering trials with digested sludge after mesophilic and thermophilic anaerobic digestion
3. thermophilic anaerobic digestion at high organic loading rate (OLR) and low hydraulic retention time (HRT)

During January the "how low can we go" trial under mesophilic conditions was completed, this study was presented in the project report for 2020 (Andersson et. Al, 2021b) and will not be discussed here. After the trial, the digester was emptied, cleaned and reinoculated with sludge (week 4) pumped from one of the full-scale digesters at Henriksdal WWTP (RK5, Mesophilic, HRT 18 d, OLR 2,5 kg VS/m^3, d). The thickener was taken into operation four weeks after the inoculation (week 8) and was kept in operation throughout the year.

In March (week 9) a mesophilic reference period was initiated with the goal to operate the digester at 37 °C, HRT 12-14 d and on OLR of at least 3 kg VS/m^3, d. When stable operation was achieved, dewatering trials including operation of the pilot dewatering unit, filtration tests and capillary suction time (CST) tests, were performed.

During the first two weeks of May (week 18-19), the digestion was transitioned gradually from mesophilic to thermophilic operation in order to study the process stability, gas production and digested sludge quality during transition at high OLR. The results were compared to the previous transition trial which was performed in 2019.

The transition was followed by a thermophilic reference period (week 20-30) with the aim to operate at 55 °C and similar HRT and OLR as during the mesophilic reference period. When stable conditions were reached, dewatering trials were performed in the same way as during mesophilic conditions.

In August (week 33), the function of the thickener was optimized to prepare for the subsequent trial with increasing OLR and decreasing HRT. The digester volume was decreased, to give an appropriate HRT with

the low inflow (due to thick sludge). The trial was carried out for 4 months. During the trial, the polymer dose to the thickener was optimized continuously since the trials with flux enhancer (described in chapter 6.8 above) affected the sludge quality and the function of the thickener.

In January 2022, the sludge pilot was stopped, and no further trials were performed.

6.13.1 Feed characteristics

As shown in Figure 6, the mixed sludge (MS) tank receives Primary Sludge (PS) and Waste Activated Sludge (WAS) from the MBR pilot-line. The mixed sludge was thickened and pumped into the digester. Before the trial the thickener had been slightly modified and the drum had been cleaned thoroughly to regain a good function. Throughout the year the thickener drum was regularly cleaned manually to maintain a good function.

Like previous years, the proportion of PS to WAS, in terms of TS (kg TS/d), was adjusted to 60/40 by discharging a fraction of the WAS so the mixed sludge would be more like the mixed sludge at Henriksdal WWTP. The total and volatile solids (TS and VS, respectively) in the feed to the digester varied during the year which can be seen in Figure 49 together with the organic loading rate (OLR).

The TS concentration in the feed was higher than previous years when the thickener was bypassed. During the first part of the trial, when the thickener had just been meticulously cleaned, and after optimization of the thickener in week 33, the TS was above 5 % when the thickener was operated and reached an average of 5.5 % after optimization. The dips in TS week 13 and 19 was due to problems with the thickeners that had to be temporarily bypassed. The high TS achieved in the pilot is still lower than the design values for the future sludge system in Henriksdal where the digester feed is assumed to reach a TS of 6-7 %. It should be noted that in the full-scale plant other methods for thickening will be used (centrifuges for WAS and gravity belt thickener for PS). From week 23 until week 33 the TS dropped although the thickener was in operation. This coincide with the staff's summer holiday when the maintenance usually is decreased. At this time, the VS in the sludge also dropped slightly from somewhere between 82-86 % to around 80 % of TS.

Low VS usually depend on sedimentation and accumulation of sand and other inorganic particles in the mixed sludge tank, which has been a reoccurring problem over the years. However, the mixed tank was regularly drained of sand by opening the bottom vent for a short time, discharging the heavy inorganic fractions, and making sure the accumulation of sediment did not affect the sludge quality. The VS concentration also depend on the dosing of metal (Fe and Al) for chemical phosphorus removal (see 6.3 above). Week 27-35, a trial with Flux enhancer was performed (see chapter 6.8 above). The main component of the flux enhancer is iron (Fe), which affect the VS-fraction of the WAS. From week 25 to 34, a decline in the VS concentration of WAS was measured, from around 75 % to around 70 % (VS of TS). The VS of primary sludge (PS) always varies more than VS in WAS but no trends of increasing or decreasing concentrations could be seen over the year.

The organic loading rate varied during the mesophilic reference period (week 5-17), partly because of the variations in TS and partly because the flow to the digester was increased stepwise after inoculation week 4. During transition from mesophilic to thermophilic digestion the organic loading rate was adjusted according to the VFA concentration in the digester. During the thermophilic reference period (week 20-30) both the TS, VS and OLR was lower than previously this year which depends on the relatively lower TS concentration, which was discussed above.

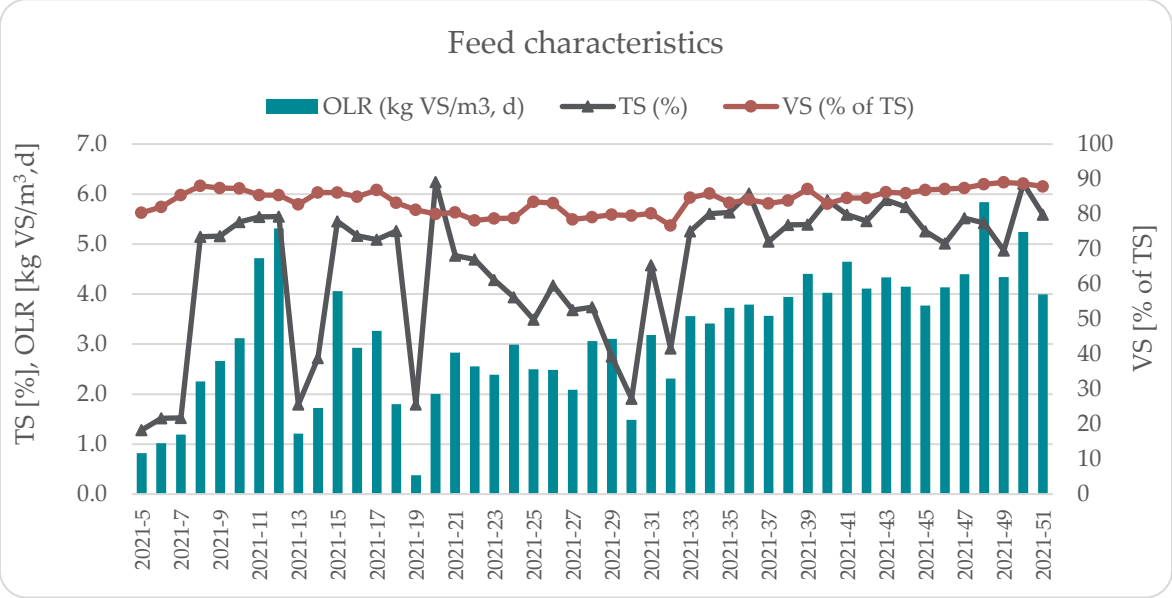


Figure 49. Feed characteristics in terms of TS and VS in thickened mixed sludge and total organic loading rate (OLR) during 2021.

6.13.2 Transition from meso- to thermophilic digestion

To maintain a stable operation of the anaerobic digestion process with thicker sludge (TS), higher organic loading rate (OLR) and lower hydraulic retention time (HRT) at Henriksdal WWTP in the future, thermophilic digestion at 55 °C is planned to be implemented instead of mesophilic at 37 °C which is currently in use. However, it is known that the transition of an anaerobic digester (AD) from mesophilic to thermophilic temperatures could pose problems to the microbial process and operation as well as causing odor problems, since the methanogenic community shifts in the range from 47 to 50 degrees (Schnürer & Jarvis, 2017) see Figure 50, causing accumulation of volatile fatty acids (VFAs).

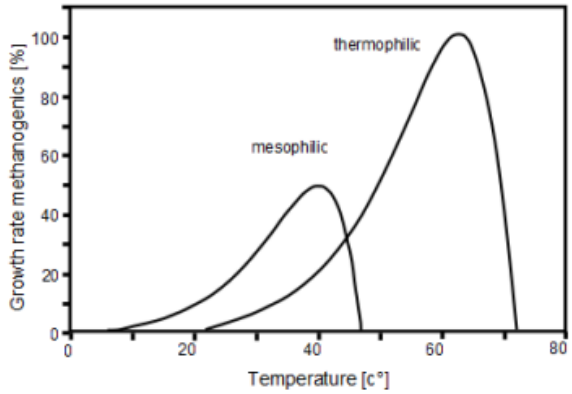


Figure 50. Temperature optimum for mesophilic and thermophilic methanogens.

A laboratory study performed at Ryaverket (Bittlingmayer, 2017) comparing different transition strategies, suggested that as fast increase in temperature as possible is the best way. Also Olsson and colleagues (2016) tested a fast transition (24 hours) in a lab environment (HRT 14 d, OLR 2.4 kg VS/m³,d) which resulted in stable operation after one retention time. At a lab-scale study performed by VEAS in Oslo, the transition was done by a one-step temperature increase over 4 hours (Haualand, 2020). After the change in temperature the

feeding was stopped for 15 days. Shortly after the change high VFA concentrations were obtained (> 3000 mg HAc/L, alkalinity/VFA = 0.7), a foul odor emitted, and biogas production and methane concentration dropped. Stable operation at target OLR was achieved after 33 days. At Käppala WWTP in Stockholm, a pilot-scale transition study was made where the increase from 37 to 55 °C was done linearly over 7 days (Lundwall, 2021). Shortly after 55 °C was reached the feeding was stopped due to high VFA concentrations (4 500 mg HAc/L). At the same time CH_4 content in the biogas dropped to below 30% and the H_2S concentration went up to 80-100 ppm. Stable operating conditions were reached after 50-60 days.

The one-step transition is theoretically better since it selects for the true thermophiles instead of the thermotolerant microbes that would be favored with a slow temperature increase. This is believed to be the reason why this strategy often requires shorter stabilization time (De la Rubia et. al, 2013). This strategy often requires a reduced OLR or a complete stop in feeding due to high VFA concentrations (> 500 mg/L). A slow transition usually allows for a maintained OLR but takes longer time to reach stable operating conditions.

The heating system at the future Henriksdal WWTP will limit the speed with which the sludge can be heated. The maximum heating capacity in the digesters at Henriksdal will be around +2 °C per day in average giving a minimum time for transition of 9 days. Since the difference between the primary heat source and the digester is larger in the beginning of the transition, a bigger increase of around 3.5 °C will be possible the first day and then decrease with time. Already in 2019, a trial with transition from mesophilic to thermophilic digestion was carried out in the pilot (Andersson et. al, 2020) showing that a relatively stable transition is possible although a VFA concentration peak, just below 2500 mg/L, was reached in the reactor at 47 °C even though the feeding was reduced and eventually stopped for a period. This resulted in a drop in methane concentration, specific biogas production and alkalinity. Stable thermophilic operation was achieved around 40 days after 55 °C was reached. During the 2019 transition trial, the thickener and the gas flow meter were not completely reliable. To confirm the results, it was decided to repeat the trial. This time with more reliable equipment in operation.

Experimental plan

The trial consisted of three phases:

1. Mesophilic reference period week 8-17 with the goal to maintain HRT 12-14 d and OLR above 3 kg VS/m^3, d.
2. Transition period week 18-19 (2-11 May) with temperature increase of approximately 1-3.5 °C per day, except weekends when no increase was made. During the transition period TS and VS in and out of the reactor; VFA, alkalinity and NH_4-N in the reactor; and CH_4, H_2S and CO_2 in the biogas was measured daily, except weekends. The feed to the reactor was supposed to continue like previously unless VFA-values > 500 mg/L was measured when the feeding was planned to be decreased. The digester was given one week after the transition to stabilize before the next reference period.
3. Thermophilic reference period week 21-29 with the goal to maintain HRT 12-14 and OLR above 3 kg VS/m^3, d. During the first week of the thermophilic reference period, analyses were made daily in the same way as during the transition. When stable operation was achieved sampling and analysis went back to normal (see chapter 5.1).

During the trial, samples in and out of the digester were analyzed for COD and on two occasions, once during the mesophilic reference period (28th of April) and once during the thermophilic reference period (17th of June) samples were sent in triplicate to an external laboratory for analyses of proteins, fats, and carbohydrates using the same analysis methods and specific methane yields as described in Sellin (2021). The reactor volume was kept at 5 m^3 throughout the trial.

Results

The actual temperature and the temperature increase are shown in Figure 51 and the inflow to the reactor together with data from the online sensors is shown in Figure 52.

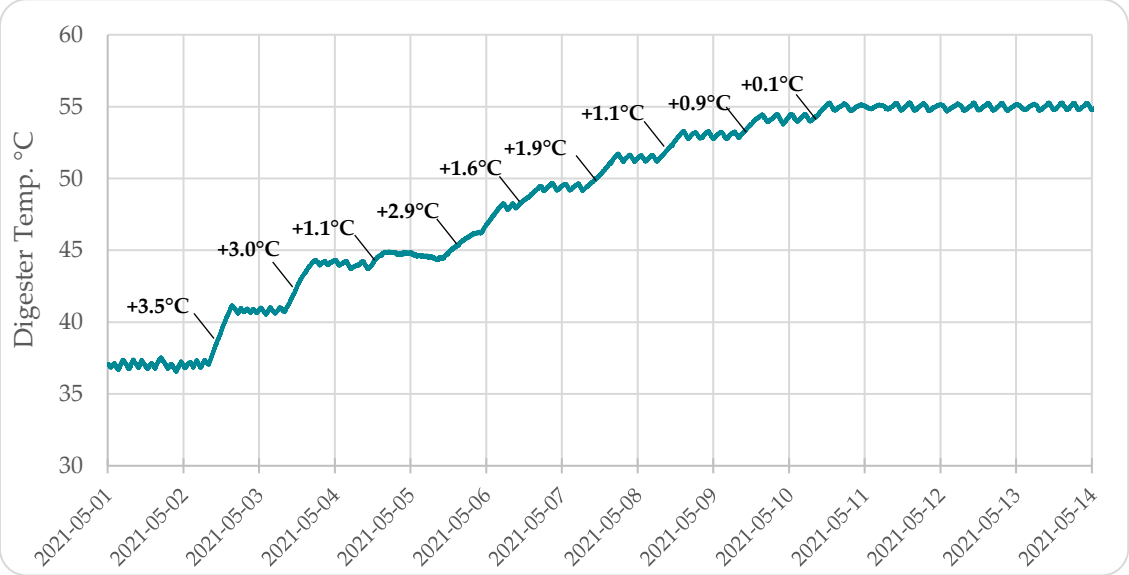


Figure 51. Temperature and temperature increase in the anaerobic digester during transition from meso- to thermophilic digestion (6 min values).

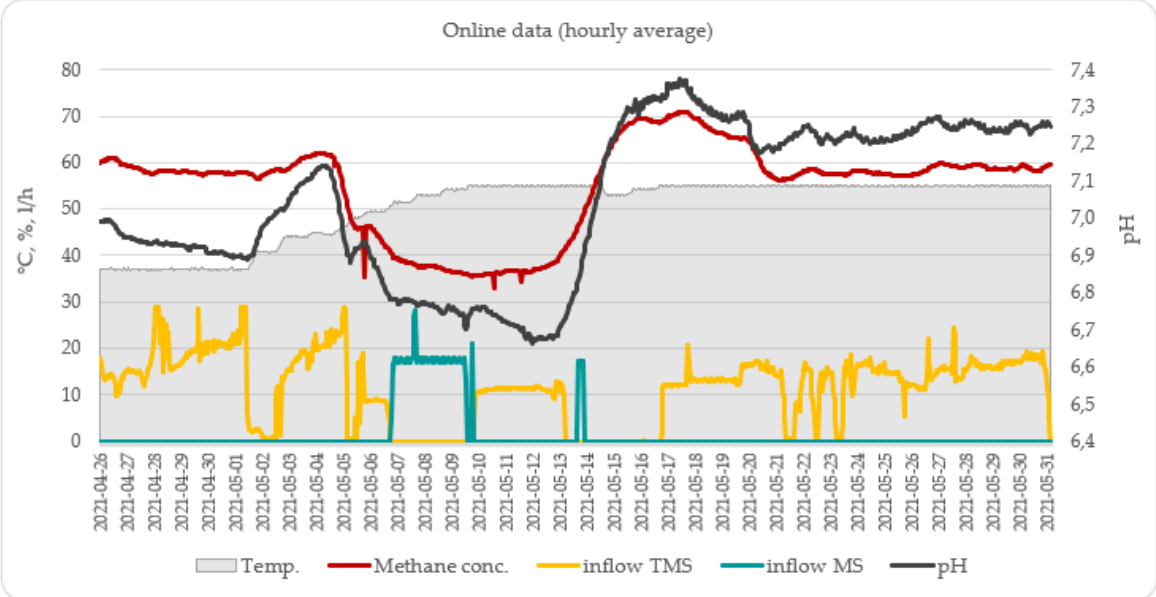


Figure 52. Online data from the digester: temperature, pH, methane concentration in biogas and inflow of thickened mixed sludge (TMS) and mixed sludge (MS). Hourly data from 2022-04-27 to 2022-05-31.

From Figure 52 it can be seen that the results for pH and methane content in the biogas show similar patterns with a slight increase, to pH 7.1 and 62% CH_4, when the temperature started to increase followed by a drop that started when the temperature reached 45 °C and continued to decreases for three days after 55 °C was reached when a fast recovery was seen (minimum values 6.7 pH and 35% CH_4) pH stabilized at around 7.2 and methane content of gas stabilized around 58% eleven days after 55 °C was reached.

Figure 53 shows lab-analysis results of VFA, alkalinity and ammonium in the digester during the trial. When the temperature increased from 46.2 to 49.1 °C VFA started to increase and kept rising to a maximum value above 2700 mg/L. Alkalinity started to decrease just before VFA was peaking and dropped to a lowest value of 1120 mg/L. Around 10 days after 55 °C was reached the VFA and alkalinity were almost stabilized on VFA

values around 300 mg/L and alkalinity > 3000 mg/L. Ammonium increased during the transition from around 500 mg/L at mesophilic operation to around 1000 mg/L at thermophilic operation, but no peaks or drops were seen during the transition phase. The increased ammonium concentration at thermophilic operation indicates a higher degradation of protein at higher temperatures.

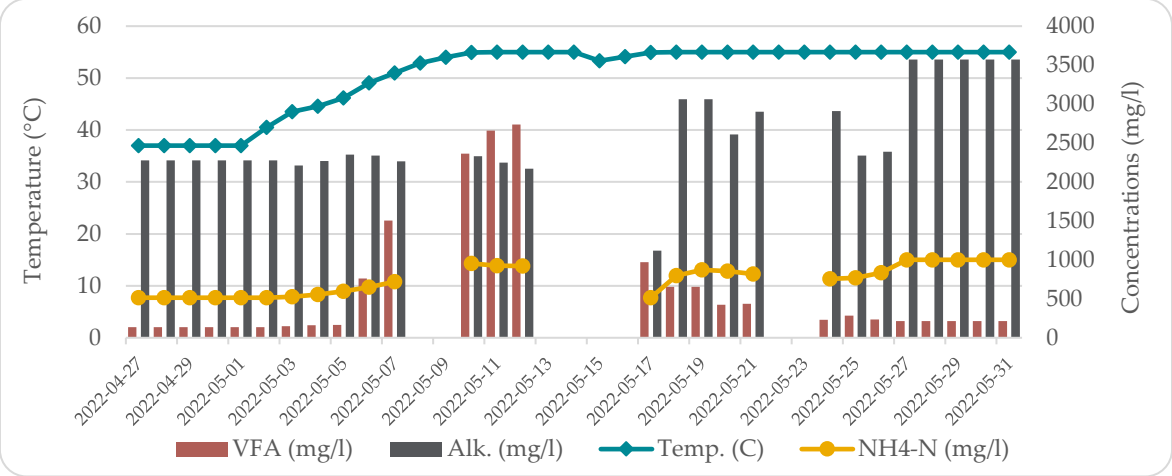


Figure 53. Analysis data related to the digesters buffer system during the transition period.

The hypothesis that more protein was degraded during thermophilic than mesophilic conditions was confirmed by the results from analysis of proteins, fats and carbohydrates which showed 50 % degradation of protein during thermophilic digestion (3.1 kg degraded/d) and 24 % during mesophilic (0.8 kg degraded/d). This difference corresponds well to the difference in NH_4-N concentration in the reactors during thermophilic and mesophilic digestion. All results from the analyses of protein, fats, and carbohydrates together with VS, COD and theoretical biogas production are shown in Table 30. As can be seen the values of COD and VS are not always corresponding to each other and the composition of thickened mixed sludge looks different in the two samples (22 % and 38 % of VS was protein on the 28th of April and 17th of June respectively). These discrepancies affect the evaluation of the results and the reliability of the conclusions. More samples would be needed to get a better statistical basis for evaluation.

Table 30. Results from analysis of proteins, fats, and carbohydrates in thickened mixed sludge (TMS) and digested sludge (DS). Dates of sampling: mesophilic reference period 2021-04-28, thermophilic reference period 2021-06-17. Average values from triplicates. The measured biogas production the 28th of April was 3.8 Nm^3/d and the 17th of June 6.4 Nm^3/d (average weekly).

Parameter	Unit	COD	VS	Protein	Fat	Carboh.	Sum of P-F-C	COD/VS
Mesophilic reference period								
TMS	kg/d	15.9	15.3	3.4	1.0	10.8	15.2	1.0
DS	kg/d	9.2	5.6	2.6	0.4	2.4	5.4	1.6
Degradation	%	42	63	24	60	78	64	
Biogas prod.	Nm^3/d	8.7	2.3	0.4	0.6	2.8	3.7	
Thermophilic reference period								
TMS	kg/d	33.8	16.5	6.3	1.1	8.8	16.3	2.1
DS	kg/d	16.4	8.9	3.2	1.0	4.0	8.3	1.8
Degradation	%	51	46	49	7	55	49	
Biogas prod.	Nm^3/d	6.1	6.8	1.4	0.1	1.6	3.1	

The organic degradation rate (ODR), however, stabilize on a lower value, 43 % of VS_{in}, at thermophilic operation compared to almost 56 % at mesophilic operation (Figure 54), which is contradictory to the protein degradation trend. The ODR decreased significantly during the transition, starting at the same temperature step as the other parameters (from 46.2 to 49.1 °C). Around 11 days after the temperature reached 55 °C, the ODR started to recover and two days later ODR values like those under stable mesophilic operation was achieved. The biogas production followed a similar pattern as the ODR, which was expected.

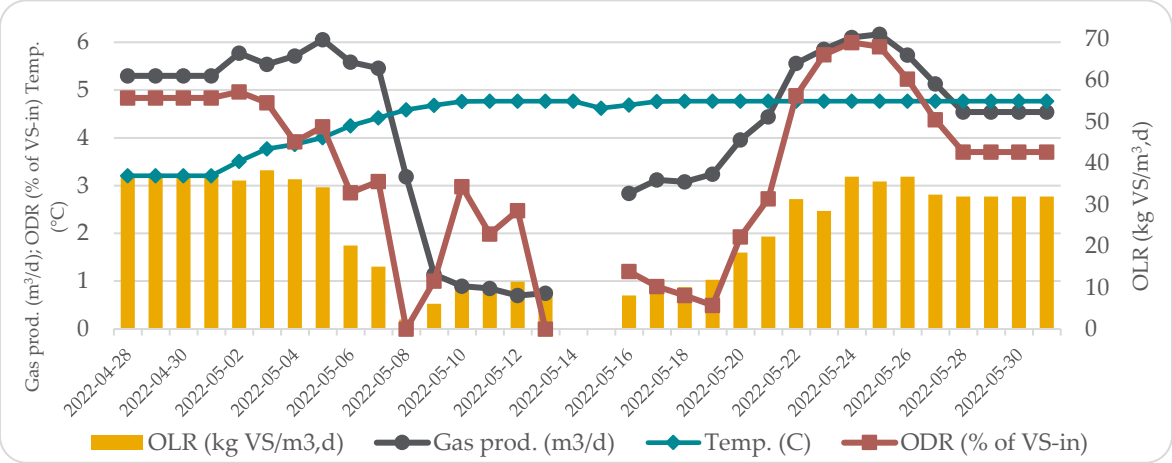


Figure 54. Operational data from the digester during transition; organic loading rate (OLR), biogas production and organic degradation rate (ORD).

Figure 55 shows how the H_2S concentration in the biogas increased at the same time as the temperature was increased and how it stabilized, on a higher level, as soon as 55 °C was reached.

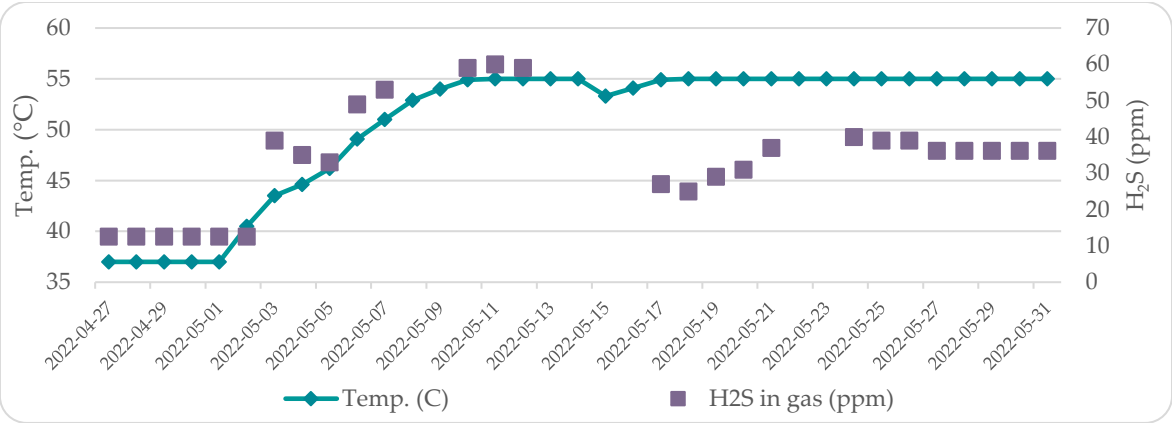


Figure 55. Temperature, VFA/Alk. And H_2S in the biogas.

A summary of all results show that the transition can be done without any major process interruptions, although the feeding had to be decreased or stopped for short time periods, and that stable operation was achieved around 10-12 days after the target temperature of 55 °C was reached.

A comparison between the reference periods can be seen in Table 31. Most of the parameters point at a slightly better performance with higher specific biogas and methane production and higher degradation of proteins during the thermophilic reference period, however, the organic degradation rate was significantly lower during thermophilic digestion, resulting in a higher specific biogas production per degraded unit of VS. The results presented in Table 31 are somehow contradictory and it is not clear that the thermophilic digestion is superior to the mesophilic digestion under the prevailing operational parameters.

Table 31. Summary of operational parameters during the mesophilic and thermophilic reference period.

Parameter	Unit	Mesophilic reference period	Thermophilic reference period
Retention time, HRT	d	13.2	12.9
Organic loading rate, OLR	kg VS/m^3,d	3.2	2.8
Specific biogas production (based on VS in)	Nm^3/kg VS_{in}, d	0.31	0.35
Specific biogas production (based on VS degraded)	Nm^3/kg VS_{deg}, d	0.56	0.81
Methane in biogas	%	57.6	59.0
H_2S in biogas	ppm	13	36
ODR	%	56	43
Degradation of protein	%	24	49
VFA in digester	mg/L	138	215
Alkalinity in digester	mg/L	2300	3600
NH_4-N in digester	mg/L	500	1000

6.13.3 Dewatering trials

One of the goals with the sludge pilot was to determine if there is a difference in dewaterability between sludge digested at mesophilic and thermophilic conditions. According to literature thermophilic sludge is supposedly more difficult to dewater than mesophilic sludge (Gebreeyessus and Jenicek, 2016) although results are often inconclusive.

To study the dewaterability of sludge digested at different temperatures, dewatering trials were performed during the mesophilic and thermophilic reference periods respectively, with stable digestion at similar HRT and OLR. The tests comprised four trials:

1. Optimizing polymer type and dose
2. Filtration trials
3. Capillary suction time (CST) trial
4. Pilot trials – using the optimized polymer at the optimized dose, digested sludge was dewatered in the pilot scale equipment described in chapter 3.2.3. The settings of the screw press were similar in both trials.

Optimizing polymer type and dose

Flocculation test with different types of polymers were performed in laboratory scale together with the polymer supplier, SNF Nordic, with the aim to find an optimized polymer type and dose for the mesophilic and thermophilic sludge respectively.

For the sludge digested under mesophilic conditions FLOPAM 640 HIB was selected and the optimal dose was determined to be 8.2 g dry weight of polymer/kg TS.

For the sludge digested under thermophilic conditions FLOPAM EM 145 CT was selected and the optimal dose was determined to be 9.2 g dry weight of polymer/kg TS.

These polymers were used in the other dewatering trials.

Filtration trials

The aim of the filtration trials was to test and further optimize the polymer dose and to get a first result on dewaterability of the sludge. Digested sludge (mesophilic and thermophilic) was mixed (by inverting a tube six times) with different doses of polymer and filtered through a cylinder with fine mesh sieve bottom (see Figure 56). For mesophilic sludge 0-35 g polymer/kg TS were tested and for thermophilic 0-40 g polymer/kg TS. The volume of the filtrate was monitored over time by weighing using a lab scale (see Figure 56). A larger filtrate volume and faster filtration indicated good dewaterability. The filtrate volume was recalculated to % of total volume (sludge sample 200 ml + polymer solution 0-139 ml).

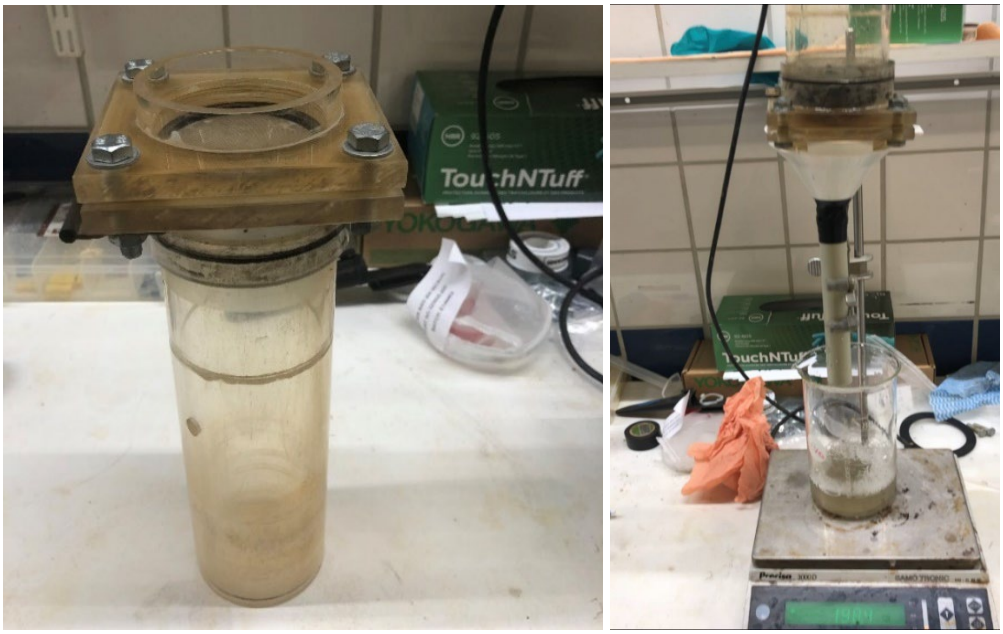


Figure 56. The filtration trial laboratory setup. To the left the cylinder with mesh sieve bottom (upside down in the picture) and the setup with the filter and the lab scale.

The results from the filtration trial are displayed in Figure 57. For the mesophilic sludge the best water-sludge separation was achieved at a dose of 20 g/kg TS. Higher doses gave equal separation percentage but since the sludge sample gets more diluted with a higher polymer dose the actual dewatering function is not better. The filtration time started to decrease from a high level (60 s) at a dose of 15 g polymer/kg TS and continued to decrease until the dose 30 g/kg TS. The results from the trial shows that a polymer dose over 15 g/kg TS is required to obtain good dewatering.

For thermophilic trials the filtrate as percentage of total volume increased rapidly between the doses 15 and 20 g/kg TS but then continued to increase almost linearly to the highest tested dose of 40 g/kg TS. The filtration time was longest (35 s) at a polymer dose of 20 g/kg TS and about the same (around 23 s) for all higher doses.

In general, the mesophilic sludge gave higher filtrate volumes and, at polymer doses over 20 g/kg TS, lower filtration times than thermophilic sludge. This might indicate that mesophilic sludge is easier to dewater than thermophilic.

Both the mesophilic and the thermophilic sludge gave their worst results (smallest filtrate volume and long filtration time) at a polymer dose around 10 g/kg TS which is close to the dose suggested by the supplier. The results from this trial indicate that very high polymer doses are required for good dewaterability.

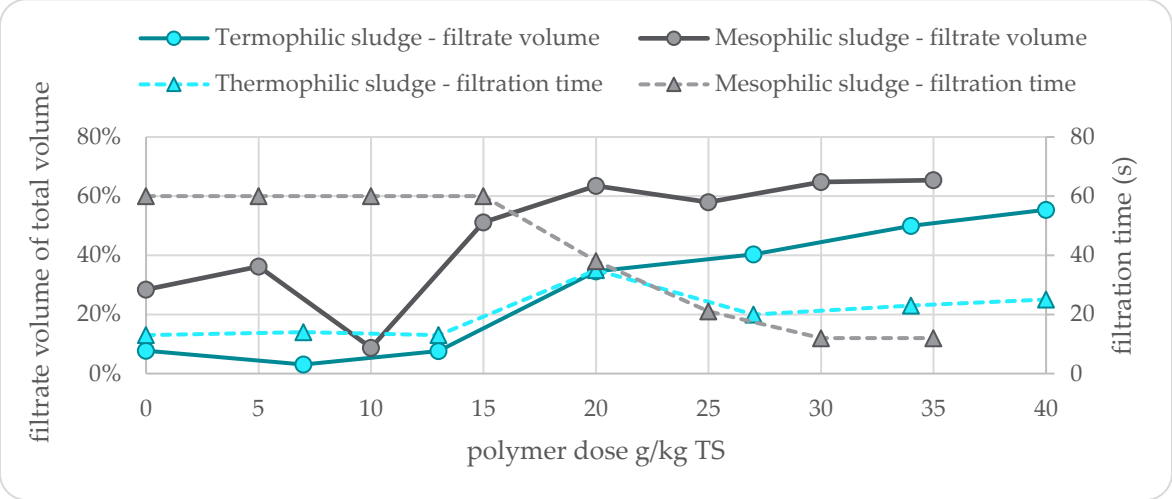


Figure 57. Results from the filtration trial.

Capillary suction time (CST) trials

CST was used to study the dewaterability of mesophilic and thermophilic sludge with and without polymer addition. The principle of the method is that dewatering is achieved by the suction applied to the sludge by the capillary action of an absorbent filter paper. A part of the absorbent paper area is exposed to the sludge, while the remaining area is free for absorbing the filtrate. The rate at which the paper becomes wetted with filtrate is a measure correlated to sludge dewaterability. It is measured by the time necessary for the filtrate to cover the space between two probes which detect the advancement of the liquid front on the paper.

Digested sludge (100 ml, TS 1.9-2.1%) was mixed with polymer solution (0.25%) at different doses in a beaker. Three different mixing times (10 s, 40 s and 100 s) were used. The sludge mixtures were then sieved through a 1 mm mesh filter and the filtrate volume was measured, see Figure 58, except for the samples without polymer addition which were not pre-sieved. The partially dewatered sludge was then used in the CST equipment (Triton. 18 mm reservoir), see Figure 58. Each sample was run in triplicate.

Figure 58. Laboratory setup for CST. To the left, the pre filtering and to the right the CST-equipment.

Results from the CST tests without polymer addition are shown in Figure 59. Mesophilic sludge gave significantly shorter CSTs than thermophilic sludge which might indicate better dewatering ability. A short mixing time seemed to give slightly shorter CST for both mesophilic and thermophilic sludge, likely due to rupture of flocs during stirring without flocculant.

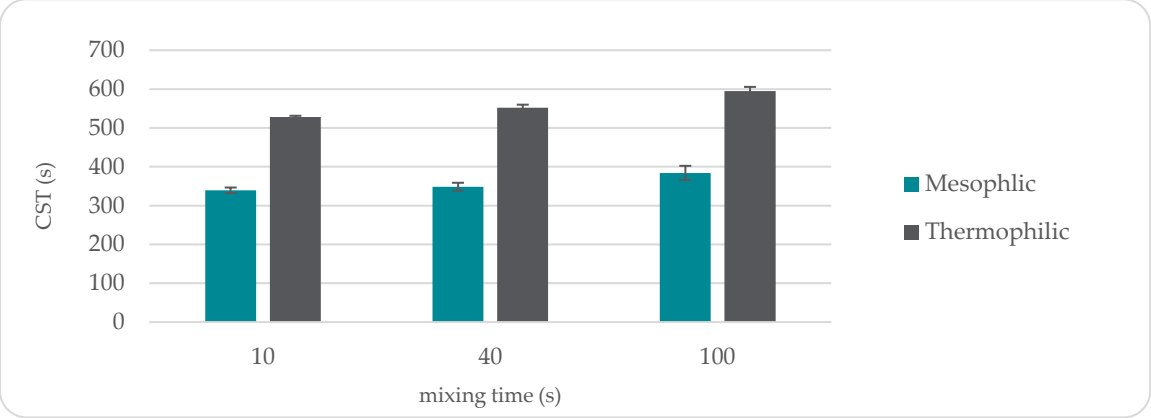


Figure 59. Results from CST test with digested sludge without polymer addition. Average of triplicates with standard deviation.

Looking at the results from CST tests with polymer addition, significantly longer CSTs were seen during tests with thermophilic sludge compared to the mesophilic, another indication that mesophilic sludge has a better dewatering ability than thermophilic. The graph below, Figure 60, shows the CSTs as average values for each polymer dose regardless of mixing time. Values deviating more than 20 % from the mean value for each dose were removed from the data set. From the tested polymer doses 13 g/kg TS gave best results for mesophilic sludge and 14.5 for the thermophilic. These values are lower than in the filtration test but higher than what was determined in the flocculation tests performed by the supplier.

At the polymer dose with the best result the volume of filtrate, as percentage of total volume (sample + polymer) was slightly higher for the thermophilic sludge, 66 %, compared to the mesophilic, 59 % (Figure 61). In general, the filtrate volumes were slightly higher for thermophilic sludge when comparing similar doses, which, oppositely to the CST results, indicate better dewatering properties for thermophilic sludge.

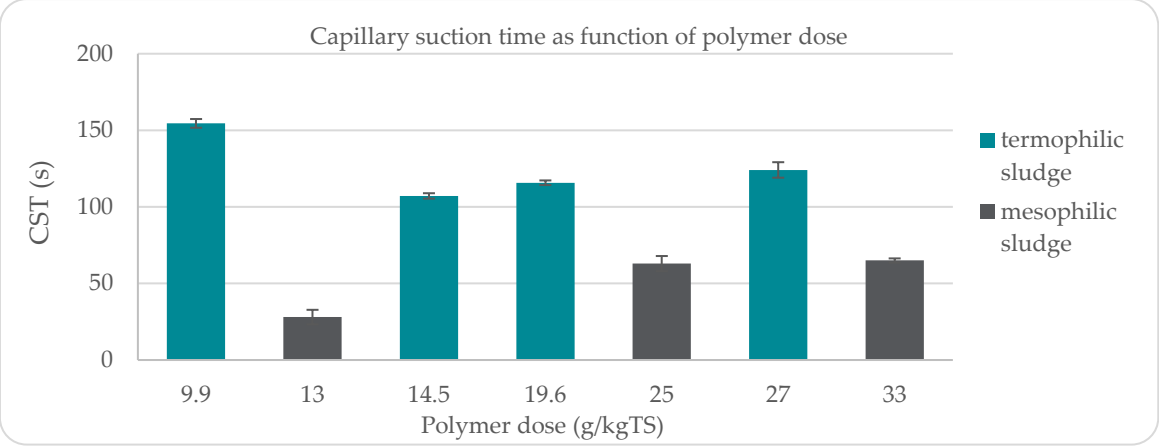


Figure 60. Capillary suction time (CST) for mesophilic and thermophilic sludge at different polymer doses.

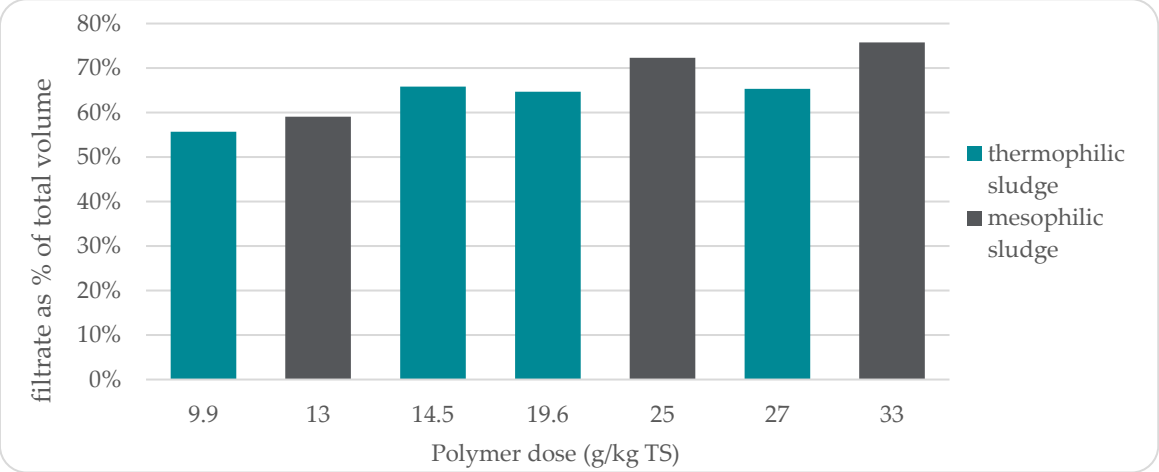


Figure 61. Filtrate volume from pre-sieving before CST-test at different polymer doses.

For both mesophilic and thermophilic sludge, a longer mixing time gave a larger volume of filtrate in the initial sieving step which might indicate that a longer mixing time gives better dewatering properties. However, for thermophilic sludge a shorter mixing time in general gave a shorter CST which indicate better dewatering properties. For mesophilic sludge the optimal mixing time varied between the polymer doses and no clear trend could be seen. These results are contradictory, and no conclusion can be made regarding the effect of the mixing time.

The CST trial, considering both the sieve pretreatment and CST test, gave inconclusive results regarding which sludge has better dewatering abilities. If only the CST test is considered the mesophilic sludge gave best results and if only the sieving is considered, the thermophilic sludge gave best results.

Pilot trials

Dewatering trials in pilot scale were performed in the sludge pilot using the screw press dewatering system described previously (chapter 3.2.3). The plan was to run the dewatering equipment continuously for five weekdays (Monday to Friday), during the last week of the mesophilic (26 - 30 April) and thermophilic (26 June – 7 July) reference periods and measure TS of the dewatered sludge, TS of the polymer and TSS in the reject water twice a day. Apart from the type of polymer and polymer dose, which was optimized for each type of sludge (mesophilic and thermophilic), all settings on the dewatering equipment were the same.

Due to problems with the equipment the operation was limited to daytime and downtime occurred regularly. The dewatering unit was operated for a total of 39 hours during the mesophilic phase and for 48 hours during thermophilic phase. The results presented below are from those hours of operation.

The polymer dose in the mesophilic trial was miscalculated which resulted in a dose that was almost three times larger than the dose that was recommended by the supplier based on the flocculation tests.

The doses used in the trials were:

- For mesophilic sludge, FLOPAM 640 HIB dose 23.4 g/kg TS (0.20% TS)
- For thermophilic sludge, FLOPAM 145 CT dose 11.3 g/kg TS (0.2-0.3% TS)

For both trials the setting of the dewatering unit were an inflow of 11 l/h, pressure after the screw of 1.0 bar and HRT 91 min in the polymer mixing tank.

In Table 32, the average results from the two trials are summarized. For the mesophilic sludge the TS concentration of digested sludge spanned between 23-26 % and the TSS in the reject water was in the range of 210-1600 mg/L. For the thermophilic sludge the TS results spanned between 26-30 % and TSS in the reject water varied between 140-1700 mg/L.

Table 32. Average results from the dewatering trials in pilot scale.

Trial	No. of analyses	TS digested sludge (%)	VS digester sludge (% of TS)	TS dewatered sludge (%)	TSS in reject water (mg/L)
Mesophilic	5	2.2		24	544
Thermophilic	10	2.3		28	730

According to the results in Table 32 the thermophilic sludge was easier to dewater to high TS than mesophilic sludge. However, the dose in the mesophilic trials was higher than intended which might have a negative effect on the results which complicates the evaluation of the results. Thermophilic digestion is proposed to result in more fine particulate matter which could be the reason for the higher TSS in the reject water of the thermophilic sludge. Again, the relatively lower value in the mesophilic sludge could be due to the higher polymer dose. Both trials resulted in relatively high values of TSS in reject water compared to for example the full-scale applications in Bromma and Henriksdal WWTPs. This is most likely due to the different machine installations, the pilot uses a screw-press which is known to use more polymer and give reject water with higher TSS than conventional dewatering centrifuges, which is the type of machine installed at the full-scale plants.

Summary and conclusions from dewatering trials

It is difficult to draw any conclusion about differences in dewatering abilities of mesophilically and thermophilically digested sludge based on these results. In Table 33 the results are summarized in a simplified way, and it shows that the different methods gave very different results.

One parameter, which was not studied in these trials was if different polymers require different methods of injection and mixing. If that is the case it could, to some extent, explain the divergent results.

The only conclusion that can be made is that the laboratory-based methods used in this study are not reliable methods for determining the dewaterability of sludge or to estimate an optimal polymer dose. The difficulty to find a good correlation between lab-scale and full-scale dewatering efficiency of sludge has also been stated in previous research (Lanko et. Al, 2021). It can also be concluded that trials in pilot- or full-scale should be performed over a longer period to produce reliable results.

Table 33. Summary of results from the dewatering trials

Trial	Mesophilic sludge	Thermophilic sludge
Filtration trial	+	
CST trial:		
• **Pre-filtration**		+
• **CST**	+	
Pilot trial		+

6.13.4 Thermophilic digestion at high organic load

After the thermophilic reference period, a trial with high organic loading rate was initiated. This had not been possible to do before since the thickener was not functioning well the previous years.

The thickener was optimized during week 33 with the aim to achieve > 5 % TS in the feed to the reactor. The maximum OLR that could be achieved was limited by the TS of the thickened sludge and the minimum reactor volume. The experimental plan is shown in Table 34.

Table 34. The experimental plan for the trial with thermophilic digestion at high OLR.

Weeks	Time from start (weeks)	OLR (kg VS/m^3, d)	HRT (h)	Digester volume (m^3)

33-37	1-5	3,5	14	5
38-46	6-14	4,0	12	3.4
47-50	15-18	5,0	<10	3.4

The actuals values of the organic loading rate (OLR) and retention time (HRT) varied around the target value according to Figure 62, due to difficulties in achieving a constant TS out from the thickener, occasional stops of pumps due to clogging or electrical failures, etc.

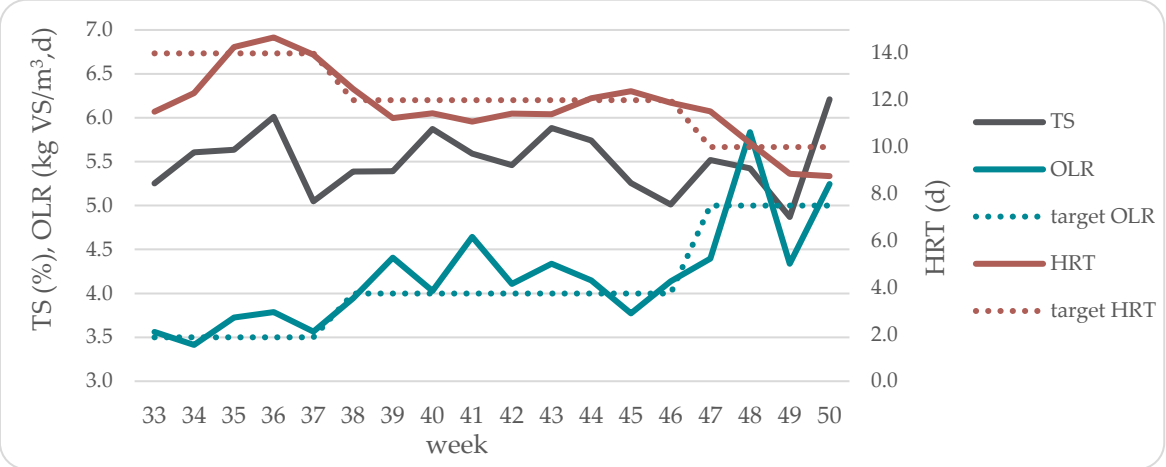


Figure 62. Data on ingoing TS, organic loading rate (OLR) and hydraulic retention time (HRT)

The digester performance was stable throughout the trial. Results are compiled in Figure 63. The organic degradation rate was always above 50% but showed a small decrease with increased OLR. The biogas production increased with increasing OLR, which would be expected. The methane concentration in the gas, however, decreased slightly with increased OLR. The specific biogas and methane production (based on VS in the inflow to the digester) increased when OLR was increased from 3,5 to 4,0 kg VS/m^3, d but decreased when the load was further increased to 5 kg VS/m^3, d.

From these results it was concluded that the digester performance is maintained at similar levels up to an OLR of around 4 kg VS/m^3, d and an HRT of 12 d. When the load is further increased and HRT decreased, the performance in terms of degradation and gas-/methane production decreases although the reactor operation (pH, VFA, alkalinity, NH_4) is still stable.

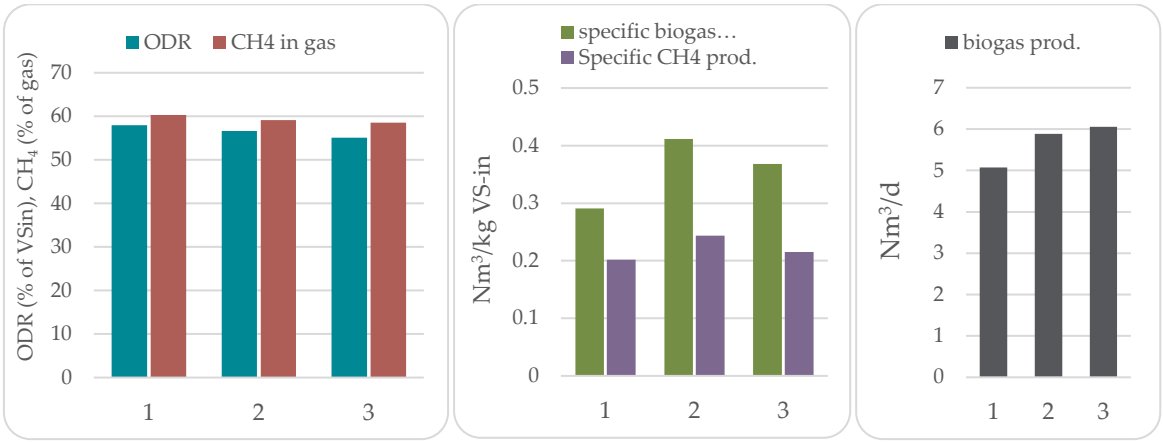


Figure 63. Results from the trial with thermophilic operation at high OLR. The three settings, 1-3, on the x-axis are described in Table 34.

6.14 Microbial community analysis during crash test

A trial, "How low can we go[2]", to test the stability and performance of the digester at decreasing retention time (HRT) and increasing organic loading rate (OLR) was performed at thermophilic conditions from week 28, 2019 to week 17, 2020 and under mesophilic conditions from week 25, 2020 to week 2, 2021. The digester was operated for > 3 HRTs at each retention time, starting at HRT 9 d and OLR 1.9 kg VS/m^3, d for the thermophilic trial and HRT 10 d and OLR 1.5 kg VS/m^3, d for the mesophilic trial. Decreasing the retention time by 1 d at the time until HRT 4 d and an OLR of 3.4-4.1 kg VS/m^3, d was reached (the lowest possible HRT for the sludge pilot). Since the digestion process did not crash at this point, external organic material, EOM (glycerol and kitchen oil), was added to stepwise increase the OLR until the pH dived and the process crashed. Table 35 shows the general setup for the trials. The results from this trial are presented in last year´s report (Andersson et al., 2021).

Throughout the trial, samples were taken regularly for subsequent analysis of the microbial community. The aim of this was to see what happens with the key bacteria and archaea when the retention time decreases and when the process finally collapses. The hypothesis was that the diversity of microorganisms, especially archaea, would decrease along the experiment and that that would be one of the reasons for the increased process instability finally leading to the crash.

Table 35. The experimental plan for the "how low can we go" trial.

	HRT [d]	Date [year-week]	Duration [xHRT]	OLR [kg VS/m^3, d]	pH
Thermophilic	9	2019-28 to 2019-33	4.2	1.91	7.0
	8	2019-34 to 2019-41	6.2	2.09	6.8
	7	2019-42 to 2019-49	8.0	2.18	6.7
	6	2019-50 to 2020-6	10.5	2.89	6.8
	5	2020-7 to 2020-10	5.6	3.06	6.8
	4	2020-11 to 2020-14	7.0	4.11	6.6
	4 + EOM	2020-15 to 2020-17	5.0	9.98	6.5→4.9
Meso	10	2020-25 to 2020-29	3.5	1.56	6.7
	9	2020-30 to 2020-37	6.2	1.66	6.6
	8	2020-38 to 2020-41	3.5	1.63	6.5

[2] trial soundtrack: https://open.spotify.com/track/2qMJ81UauLoSxzl4h5SQzS?si=d50c27c3422d4c79

	7	2020-42 to 2020-44	3.0	2.11	6.5
	6	2020-45 to 2020-47	3.5	2.42	6.5
	5	2020-48 to 2020-51	5.6	2.45	6.6
	4	2020-52[3] to 2020-01	5.3	3.21	6.6
	4 + EOM	2021-02	1.8	4.66	6.6→6.0

Method:

During the thermophilic trial, grab samples of sludge was taken from the digester once per week starting week 43 (HRT 7 d), resulting in a total of 28 samples. During the mesophilic trial, grab samples of sludge was taken from the digester on three occasions during the last 1-3 weeks of each retention time, resulting in 24 samples (8 "triplicates"). The samples were immediately frozen. At the end of the last trial tree samples were sent to Swedish University of Agricultural Sciences (SLU) in Uppsala where DNA was extracted, bacterial and archaeal gene fragments were amplified by polymerase chain reaction (PCR) with primers targeting 16SrDNA specific for bacteria and archaea, and finally the amplified gene fragments were sequenced, and the results ran against a gene database. Because of the pandemic prevailing at the time, the DNA-analysis could not be completed until the end of 2021.

Results and discussion:

The evaluation of the results is still ongoing with support by the Swedish University of Agricultural Sciences (SLU) and findings are planned to be published as an article in a scientific journal during 2023. A general summary of the results is given here.

The results from the thermophilic and mesophilic trials were compiled separately for the two domains bacteria and archaea and are presented on different taxonomy levels from phylum to species (Figure 64).

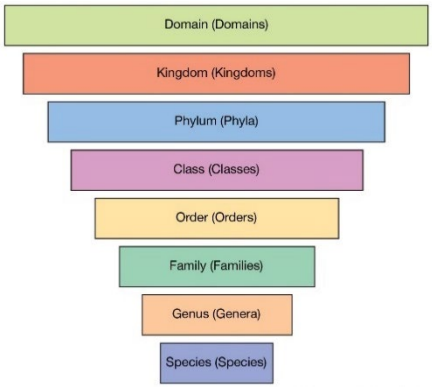


Figure 64. The general taxonomy of microorganisms

In Figure 65 the community composition of archaea in the digester is shown on a Class level over time. The results when visualised on other taxonomy levels show similar trends. The conclusion that can be drawn in this stage of result assessment are:

- The diversity of archaea was always smaller during thermophilic digestion compared to mesophilic digestion.
- The community composition and diversity of archaea changed with changing retention time at mesophilic digestion. The diversity of phylums/classes/orders/families/genus decreased with

[3] Year 2020 had 53 weeks.

decreasing retention time and community shifts could be seen where the most dominant phylums/classes/orders/families/genus changed.

- During the thermophilic trial, the archaeal community composition and diversity remained rather similar over time.
- During thermophilic digestion, the most dominating class was *Methanobacteria*, making up more than 75 % of the archaeal community, followed by *Methanosarcinia*.
- During the mesophilic trial, *Methanosarcinia* was the dominating Class of archaea throughout the experiment. *Methanomicrobia* and *Thermococci* were relatively abundant at the beginning of the trial but decreased with decreasing retention time until around 5 d HRT when they almost disappeared. Around the same time, *Nanoarchaeia* became an increasingly important part of the archaeal community. However, when OLR was increased by addition of EOM at 4 d HRT, *Nanoarchaeia* disappeared completely.

A more in-depth analysis of the archaeal community structure over time will be done, as well as an assessment of the results from the bacterial DNA analysis will be performed and published in a scientific journal further ahead.

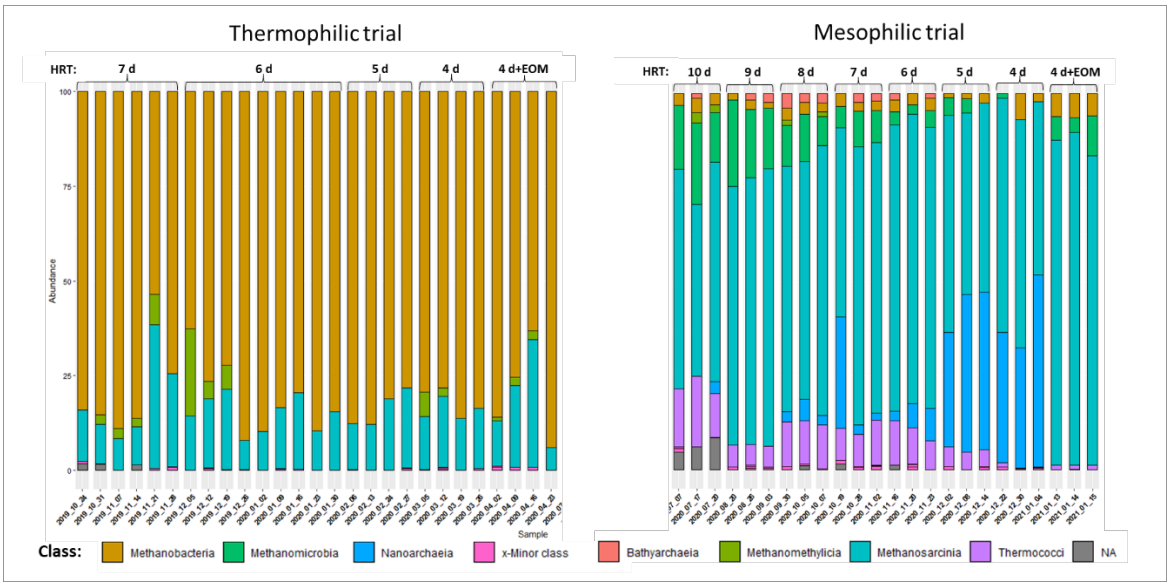


Figure 65. Composition of archaea during the mesophilic and thermophilic trials shown on Class-level.

6.15 Resource consumption

Resource consumption in the pilot for 2021 is summarized in Table 36. A comparison with the future Henriksdal design was made 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 from process calculations based on design manuals and experiences from other WWTPs.

This year, two different external carbon sources were tested. Glycerol, which has been tested previous years, was used in the pilot for about 45 weeks in 2021. A shorter 7-week trial with internally produced VFA was conducted for the remainder of the year (and continued in 2022). No comparison of the two carbon sources, glycerol and VFA is presented in this section as the results from the VFA trials will be further evaluated within an ongoing PhD project.

The consumption of glycerol, expressed as COD, was the same as for the future Henriksdal design, even though the nitrogen load on the biology in the pilot was 21 % higher and the average effluent total nitrogen concentration was 3.9 mg/L compared to the design target of 6 mg TN/L.

The daily iron/metal consumption was 73 % of the future Henriksdal design, although the phosphorus load to the pilot was about 50 % higher compared to design values. This is reflected in the ratio between mole metal dosed and mole phosphorus reduced, which was lower in the pilot compared to the design. The effluent phosphate concentration was below the target concentration and the measured values were lower than those set in the design calculations. The low metal consumption is mainly explained by the EBPR activity in the pilot.

Also, the consumption of cleaning chemicals was lower than the future Henriksdal design, even though the inflow to the pilot was 30 % higher than design. In addition, permeate was recycled to achieve a higher flux in the pilot. This implies that the permeate pumping was about 70 % higher than the design inflow. In total, the amount of chemicals used for RC was around the same as in the design for sodium hypochlorite since RC was performed twice with hypochlorite in 2021. The consumption of acids was about 50 % of the full-scale design as only one RC with each acid was performed in 2021. Chemical consumption for MC, however, was only 56 % of pilot design values for oxalic acid, 58 % for citric acid and 27 % 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 based on the pilot results.

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 176 % higher 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. It could be noted that by design of the pilot, the membrane aeration is already 150 % of the full-scale at Leap-Lo. In addition, the yearly average flux was 25 % higher than design and the suspended solids concentration in the membrane tank was 16 % higher than design.

Table 36. Chemicals consumption during 2021.

Resource	Unit	Value pilot 2021	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, 321 days)	kg COD/d	1.76	-	-
	g COD/g N*	0.54	-	-
External carbon source (COD)	kg COD/d	1.76	12 000	98 %
Iron (PTW)	kg Fe/d	0.85	10 000	57 %
	mole Fe/mole P		2.8	-
Aluminium (BR6)	kg Al/d	0.09	-	-
	mole Al/mole P	-	-	-
Flux enhancer (Fe^{3+})	kg Fe/d	0.15#		
	mole Fe/mole P			
Metals (Iron + Aluminium)	kg Me/d	1.09	10 000	73 %
	mole Me/mole P	1.29	2.8	46 %
Citric Acid (51 %)	L/d	0.077**	1 100	47 %
Sodium hypochlorite (12 %)	L/d	0.24***	1 600	79 %
Oxalic acid (8 %)	L/d	0.40**	5 000	54 %
Aeration, biology	m^3/d	1 469	1 600 000	615 %
Aeration, MT	m^3/d	787	3 000 000	176 %

* N removed in total, from inlet to effluent.
# Actual average consumption during 2021. During the dosing period the consumption was 0.78 kg Fe/d.
** Number of MCs with each acid, multiplied with time settings and number of back-pulses using settings of flowrate of chemical. Measured consumption for one RC preformed with each acid.
*** Number of MCs with hypochlorite was multiplied with time settings and number of back-pulses using design flowrate of chemical. Measured consumption for four RC preformed in total (2 per MT).

6.16 Initial energy modeling and environmental impact assessment

The consumption of chemicals and energy for pumping, aeration and membrane cleaning as presented e.g., for year 2021 in Table 36, represent the main costs and environmental impact for the MBR process during operation (investment and membranes excluded). As such, the savings achieved in the pilot compared to the full-scale design, especially chemicals for membrane cleaning, provides potential for significant cost savings and a reduction of the environmental impact from MBR process operation. But also, for energy consumption, the adapted aeration strategy if corrected for full-scale conditions, would imply savings in cost and environmental impacts.

The project has therefore initiated an evaluation of the savings in costs and environmental impacts based on the project findings for use of chemicals and energy for aeration. This evaluation includes not only results from 2021 but also previous years and will provide a general indication of potential savings. Further, results from the implementation of some of the project findings in the first full-scale treatment line at Henriksdal WWTP and their effect on costs savings will be included in the assessment.

The LCA modelling for the chemicals included is based in the software GaBi, version 9.2.1 (Thinkstep, 2020a). LCA data are taken from databases Thinkstep (2020b) and Ecoinvent (Wernet et al. 2016). To evaluate savings for the energy consumption in full-scale based on pilot findings, developed simulation models for the MBR process will be utilized to investigate actual operation impacts. As these activities are still not finalized, findings will be published at a later stage.

As the cleaning of the membranes with chemicals also is the main driving factor for membrane ageing requiring the replacement, e.g., after 10 years in the case for Henriksdal WWTP, a prolonged usage of the membranes could also be achieved. Every extra year of membrane operation would imply huge costs savings. In addition, even if the total environmental impact of production of the used membrane would not be changed, a prolonged usage would reduce the environmental impact per treated unit water. How much the membrane lifetime may be affected by the adapted cleaning strategies of the pilot project is yet too early to know. Therefore, the project partners have agreed on a continuation of the project for a long-term evaluation of the membrane operation.

7 Related publications

Several conferences and seminars had limited knowledge transfer activities also in 2021 due to COVID-19. In addition, COVID related restrictions were also in 2022 limiting the general cooperation with different partners at the R&D facility which affected the publications produced from the project. The project team, however, was trying it's best and attended both the World Water Congress digitally and the Nordic Wastewater Conference in 2021.

Related to the project, one scientific paper was published:

- Baresel, C., Jingjing, Y., Niclas, B., Kåre, T., Linda, K., Klara, W. 2022. Direct GHG emissions from a pilot scale MBR-process treating municipal wastewater. Advances in Climate Change Research 13, 138–145. https://doi.org/10.1016/j.accre.2021.09.006

The project had seven contributions at international conferences held in 2021:

- Andersson, S.L., Westling, K., Andersson, S., Karlsson, J., Narongin, M., Baresel, C. 2021. Stockholm's Future Wastewater Treatment – long term pilot trials with an MBR process. NORDIWA - Nordic Wastewater Conference, 28 September - 1 October 2021.
- Andersson, S., Roberts, R., Fridh, B., Narongin, M., Muñoz, A.C., Karlsson, J., Andersson, S.L., Westling, K. 2021. Where did the phosphorus go? NORDIWA - Nordic Wastewater Conference, 28 September - 1 October 2021.
- Andersson, S., Sellin, J., Karlsson, J., Carlsson, A., Muñoz, A.C., Narongin, M., Andersson, S.L., Westling, K., Schnürer, A. 2021. How low can we go? – mesophilic and thermophilic digestion of WWT sludge at short retention times. NORDIWA - Nordic Wastewater Conference, 28 September - 1 October 2021.
- Andersson, S.L., Muñoz, A.C., Westling, K., Andersson, S., Sellin, J., Karlsson, J., Narongin, M., Baresel, C. 2021. How low can you go – Resource efficient membrane cleaning in municipal membrane bioreactor pilot. NORDIWA - Nordic Wastewater Conference, 28 September - 1 October 2021.
- Andersson, S., Karlsson, J., Carlsson, A., Carranza Muñoz, A., Narongin, M., Andersson, S.L., Westling, K. 2021. Retention time crash test for mesophilic and thermophilic digestion of WWT sludge. IWA Digital World Water Congress, 24 May – 4 June 2021.
- Westling, K., Närhi, K., Andersson, S., Baresel, C. Narongin, M., Wahlberg, W. 2021. Micropollutants and Microplastics in a Membrane BioReactor (MBR). NORDIWA - Nordic Wastewater Conference, 28 September - 1 October 2021.
- Yang, J., Bornold, N., Baresel, C., Westling, K. 2021. Nitrous oxide emissions from MBR-processes and their impact in the total environmental performance of the process. IWA Digital World Water Congress, 24 May – 4 June 2021.

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