

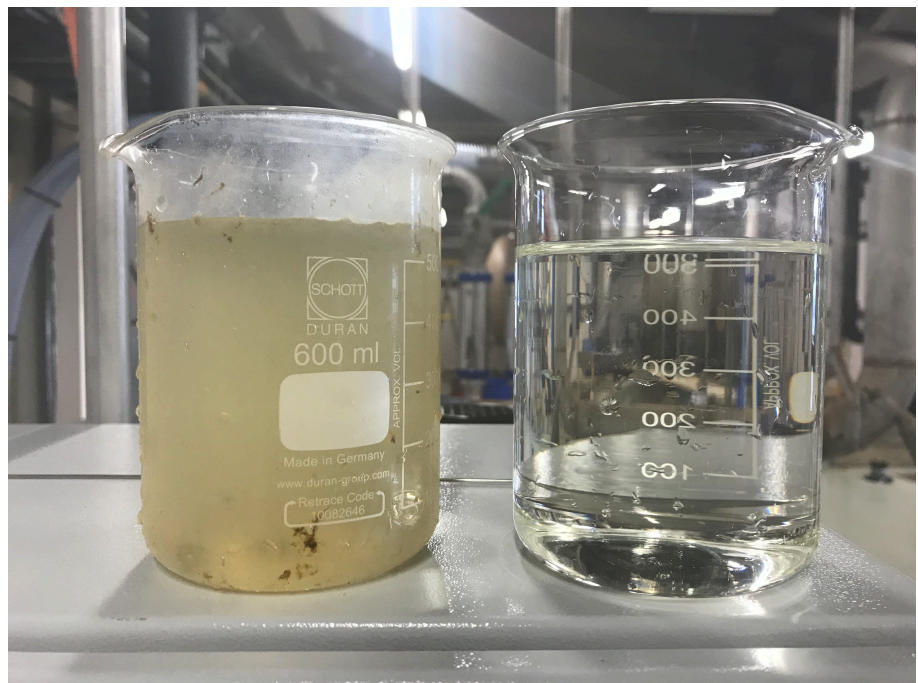
Eline Klaastad

A Comparison of Direct Membrane Filtration and Moving Bed Biofilm Reactor for Municipal Wastewater Treatment in Norway

Master's thesis in Civil and Environmental Engineering

Supervisor: Stein Wold Østerhus

June 2020



Eline Klaastad

A Comparison of Direct Membrane Filtration and Moving Bed Biofilm Reactor for Municipal Wastewater Treatment in Norway

Master's thesis in Civil and Environmental Engineering
Supervisor: Stein Wold Østerhus
June 2020

Norwegian University of Science and Technology
Faculty of Engineering
Department of Civil and Environmental Engineering



Abstract

With an increasing focus on the environment and sustainability, infrastructure like wastewater treatment plants should prepare for future requirements. Future considerations may involve a focus on compactness, low carbon footprint, and energy recovery potential. Two concepts that potentially could offer these qualities are direct membrane filtration (DMF) and moving bed biofilm reactor (MBBR). These concepts will be researched in a case study of a municipal treatment plant in Fredrikstad, South-East Norway.

The feasibility of the emerging DMF concept has been established through a pilot plant study on-site a municipal wastewater treatment plant, consisting of a sandtrap, coagulation and flocculation, microsieving before eventually entering a microfiltration stage. The concept has achieved outstanding results in terms of removal and hence shown that organic matter can be reduced without biological treatment. During operation of the pilot plant, there has been an average of 99% removal of phosphorous, 89% removal of chemical oxygen demand (COD), and 92% of biological oxygen demand (BOD), meaning the Norwegian secondary removal requirements as well as tertiary removal requirements regarding phosphorous were met. Additionally, excellent removal of suspended solids (SS) (99%), total organic carbon (TOC) (83%), and various heavy metals has been obtained.

The pilot plant has also been monitored with regard to operation, where transmembrane pressure (TMP) has been a determining indicator of operational feasibility, as it determines the frequency of chemical cleaning in place (CIP). Different chemical dosages of precipitation metal (PACl) and polymer has been investigated, showing that 9 mg Al/L combined with 1.5 mg polyacrylic polymer/L gave sufficient removal and showed potential for lowering the dosage even further.

A full-scale MBBR-treatment plant and a full-scale DMF treatment plant were designed and compared. This study shows that both processes are compact compared to conventional activated sludge, and energy recovery, i.e., biogas potential has been shown through the use of mass balances. DMF was found to have a higher chemical demand due to chemical cleaning and was also found to likely have a significantly higher energy demand due to a considerably higher requirement for aeration. The thesis, therefore, concludes with MBBR being a more suitable concept for this case study with the current treatment requirements but adds that DMF is a concept that should not be neglected when it comes to future relevance in the treatment of municipal wastewater.

Sammendrag

Med et økende fokus på miljø og bærekraft, bør også infrastruktur som avløpsrensing rustes for fremtiden, med vekt på kompaktet, lavt karbonfotavtrykk og potensiale for energigjenvinning. To konsepter som kan vurderes i denne sammenheng er direkte membranfiltrering (DMF) og biofilmreaktorbasert bevegelig bæremateriale (engelsk: moving bed biofilm reactor, MBBR). Disse to har blitt sett på i en case-studie for et kommunalt avløpsrensaneanlegg i Fredrikstad.

Egnetheten til konseptet DMF har blitt undersøkt gjennom studie av et pilotanlegg som er bygget på et kommunalt renseanlegg. Pilotanlegget består av sandfang, koagulering og flokkulering, mikrosil og til slutt et membrantrinn med mikrofiltreringsmembraner. Konseptet har oppnådd enestående resultater når det gjelder fjerning av organiske stoffer og næringsstoffer, og dermed vist at dette kan oppnås uten bruk av biologi. Gjennom driftsperioden ble det oppnådd gjennomsnittlig 99% fjerning av fosfor, 89% fjerning av kjemisk oksygenforbruk (KOF) og 92% fjerning av biologisk oksygenforbruk (BOF), og de norske sekundærrensekravene samt tertiærrensekravet for fosfor ble dermed oppnådd. I tillegg til dette har det vært effektiv fjerning av suspendert stoff (SS) (99%), total organisk karbon (TOK) (83%), samt god fjerning av tungmetaller.

Drift av prosessen har også blitt undersøkt med særlig fokus på trykkdifferansen over membranene (engelsk: transmembrane pressure, TMP) som er en avgjørende faktor når det kommer til hyppigheten av kjemisk vask. Forskjellige kjemikaliedoseringer av fellingskjemikalie (PACl) og polymer har blitt undersøkt, og det ble vist at 9 mg Al/L kombinert med 1.5 mg polymer/L ga tilfredsstillende fjerning, og viste også potensiale for enda lavere doseringer.

Et fullskala MBBR-anlegg og et fullskala DMF-anlegg har også blitt dimensjonert og sammenlignet. I dette studiet ble det vist at begge disse er kompakte sammenlignet med konvensjonelle aktivslamanlegg, og potensiale for energigjenvinning gjennom biogassproduksjon har blitt undersøkt ved hjelp av massebalanser. DMF har et høyere kjemikalieforbruk enn MBBR på grunn av kjemisk vask, i tillegg til et høyere luftforbruk noe som angivelig også vil føre til et høyere energiforbruk. Konklusjonen i denne studien er derfor at MBBR er et bedre egnet konsept i denne sammenheng, gitt nåværende rensekrav, men at DMF er et konsept som absolutt kan være relevant innenfor avløpsrensing i fremtiden.

List of papers and contribution

In relation to this thesis, an abstract was submitted to the World Water Congress in Copenhagen 2020 (Postponed to 2021). The abstract was accepted for a platform presentation. A full paper will also be submitted.

Paper I: Klaastad, E., Lipnizki, F., Hey, T., and Johannessen, E. Investigating direct membrane filtration (DMF) as a potential treatment concept for municipal wastewater

My contribution to this paper: Experimental work concerning the paper was mainly conducted by me, as well as analysing and interpretation of operational results.

Preface

This thesis concludes my five years at Norwegian University of Science and Technology and marks the end of an era as a student. It has been five wonderful years with a lot of new friends and a lot of new knowledge. The thesis is the final project of the Civil- and Environmental Engineering, specialising in water and wastewater engineering, at the Department of Civil and Environmental Engineering. The project is a collaboration between FREVAR, COWI, and Enwa.

First of all, a great thanks to Erik Johannessen (COWI), and former operations manager of FREVAR, Tor Gunnar Jantsch, without them this project would not be possible. I am very grateful for the opportunity to participate in this project.

I have also received great guidance from my supervisor here at NTNU, Stein W. Østerhus. Thank you for interesting discussion on wastewater treatment, and for providing me with great help on how to write a thesis.

My greatest thanks to Flemming Wessman (Enwa), for sharing his excellent knowledge about MBBR with me, as well as getting me involved with this project. It is highly valued and appreciated.

Thank you Tobias Hey (COWI) for helping me with the understanding of direct membrane filtration and the pilot plant, in addition to your overall support with the thesis.

Thanks to personnel working at FREVAR, especially Raymond, Jørgen, Lisbeth, Merete, and Veronica, for technical help, answering my questions and providing me with data from the pilot plant which has been essential for my analysis. A special thanks to lab technicians who did analyses for me when I could not do so myself. I also want to thank David at Kemira in Fredrikstad, and Emmanuel at Alfa Laval in Denmark.

My friends and family are, and have always been, a great support, not only during the time of writing this thesis. Thank you.

Unless stated otherwise, photos were taken by me, including the cover photo.

Eline Klaastad

Eline Klaastad
Trondheim, June 2020

Table of Contents

List of Figures	xiii
List of Tables.....	xiv
List of Abbreviations.....	xv
1 Introduction	17
1.1 Objectives and motivation	18
1.1.1 Limitations	18
1.1.2 Specialisation project.....	18
1.2 Outline of the thesis	18
2 Theoretical background.....	19
2.1 Sustainability in wastewater treatment	19
2.2 Direct membrane filtration.....	20
2.2.1 Membranes in wastewater treatment	20
2.2.2 Fundamentals of direct membrane filtration.....	21
2.2.3 Chemical precipitation of phosphorous	22
2.2.4 Direct membrane filtration and sustainability	23
2.3 Moving bed biofilm reactor (MBBR).....	23
2.3.1 Fundamentals of MBBR	23
2.3.2 MBBR and phosphorous removal	24
2.3.3 MBBR and sustainability	25
3 Pilot plant study.....	27
3.1 Description of the pilot plant.....	27
3.2 Jar testing	32
3.2.1 Method	32
3.2.2 Results and discussion of jar test	32
3.3 Chemical dosing.....	35
3.4 Internal process behaviour	36
3.4.1 Method	36
3.4.2 Results and discussion of internal process behaviour	36
3.5 Sludge sampling.....	38
3.5.1 Method	38
3.5.2 Results and discussion of sludge sampling.....	38
3.6 Removal efficiency	41
3.6.1 Method	41
3.6.2 Results of removal efficiency	42
3.7 Development of TMP.....	47

3.7.1	Method	47
3.7.2	Results and discussion of the development of TMP.....	48
3.8	Discussion of feasibility of direct membrane filtration	54
3.8.1	Conclusion of feasibility.....	55
4	Design of full-scale treatment process	57
4.1	New treatment requirements	57
4.2	Design criteria	57
4.3	Design of pretreatment	59
4.4	Design of direct membrane filtration.....	61
4.4.1	Coagulation and flocculation	61
4.4.2	Microsieving	62
4.4.3	Microfiltration	63
4.4.4	Theoretical sludge production	64
4.5	Design of MBBR	65
4.5.1	Biological reactors (MBBR)	65
4.5.2	Coagulation and flocculation	67
4.5.3	Sludge separation – dissolved air flotation.....	67
4.5.4	Theoretical sludge production	68
4.6	Mass balances.....	68
4.7	Process flow charts.....	73
4.8	Footprint of the treatment plant.....	74
4.9	Energy consumption	75
4.10	Design summary	75
4.11	Discussion	76
5	Conclusion	81
5.1	Further work.....	81
References	83
Appendices	93

List of Figures

Figure 2.1 BWT X carrier, specific area 650 m ² /m ³ . Courtesy of Biowater Technology....	24
Figure 2.2 Close-up of the BWT-X carrier. Courtesy of Biowater technology.	24
Figure 3.1 Simplified flowchart to illustrate the pilot plant.....	27
Figure 3.2 P&ID of the pilot plant.	28
Figure 3.3 The sandtrap of the pilot plant.	29
Figure 3.4 Coagulation and flocculation reactors of the pilot plant.....	30
Figure 3.5 Interior of microsieve in the pilot plant.....	30
Figure 3.6 Illustration of hollow sheet membrane. Courtesy of Alfa Laval AB.	31
Figure 3.7 SEM photo of hollow sheet membrane. Courtesy of Alfa Laval AB.	31
Figure 3.8 One of the membrane modules in the pilot plant. Courtesy of Tobias Hey.	31
Figure 3.9 Jar test setup.....	32
Figure 3.10 Turbidity with different chemical dosages.	33
Figure 3.11 Orthophosphate concentrations with different chemical dosages.	34
Figure 3.12 Sludge samples from microsieve (left) and membrane tank(right).	39
Figure 3.13 COD removal with initial chemical conditions.....	42
Figure 3.14 Tot P removal with initial chemical conditions.....	43
Figure 3.15 COD removal with the first change of chemical dosage.....	44
Figure 3.16 Tot P removal with the first change of chemical dosage.....	44
Figure 3.17 COD removal with the second change of chemical dosage.....	45
Figure 3.18 Tot P removal with the second chemical dosage.	45
Figure 3.19 TMP development in round 1 of initial operation conditions.	48
Figure 3.20 TMP development in round 2 of initial operation conditions.	48
Figure 3.21 TMP development in round 3 of initial operation conditions.	48
Figure 3.22 TMP development round 4 (operation condition I).....	50
Figure 3.23 TMP development round 5 (operation condition I).....	50
Figure 3.24 TMP development round 6 (Operation condition II).	51
Figure 4.1 Duration curve of Q_{hour} and Q_{maxhour}	58
Figure 4.2 Duration curve of BOD load.	58
Figure 4.3 Duration curve of phosphorous load.....	59
Figure 4.4 DMF mass balance.....	70
Figure 4.5 MBBR mass balance.....	72
Figure 4.6 Flowchart of a full-scale DMF process.....	73
Figure 4.7 Flowchart of a full-scale MBBR-DAF process.....	74

List of Tables

Table 2.1 Membrane types with sizes and rejected materials.....	20
Table 3.1 Results from jar tests.....	34
Table 3.2 Chemical dosages for different operation conditions.....	35
Table 3.3 Phosphorous concentrations internally in the pilot plant.....	36
Table 3.4 COD concentrations internally in the pilot plant.....	37
Table 3.5 SS concentrations from internally in the pilot plant.....	37
Table 3.6 Content of total solids, inorganic, and organic solids in sludge.....	39
Table 3.7 Production of total, inorganic, and organic solids in sludge.....	39
Table 3.8 Content of total solids, inorganic, and organic solids in sludge.....	40
Table 3.9 Production of total, inorganic, and organic solids in sludge.....	41
Table 3.10 Removal rates from accredited lab analyses.....	46
Table 3.11 Removal of heavy metals from accredited lab analyses.....	47
Table 3.12 Gradient of TMP against time.....	49
Table 3.13 Gradient of TMP against time for optimised operation conditions I.....	50
Table 3.14 Gradient of TMP for the second round of optimised conditions.....	51
Table 4.1 Design values for a new treatment plant.....	59
Table 4.2 Capacities of existing units at FREVAR treatment plant.....	60
Table 4.3 Design of flocculation for DMF.....	62
Table 4.4 Design of microsieving for DMF.....	63
Table 4.5 Design of DMF.....	64
Table 4.6 Air demand for DMF.....	64
Table 4.7 Sludge production DMF.....	65
Table 4.8 Design of bioreactors.....	66
Table 4.9 Air demand for MBBR.....	66
Table 4.10 Design of carrier sieves for MBBR.....	67
Table 4.11 Design of flocculation for MBBR.....	67
Table 4.12 Design of flotation for MBBR.....	68
Table 4.13 Sludge production MBBR.....	68
Table 4.14 Footprint of DMF.....	74
Table 4.15 Footprint of MBBR-DAF.....	74
Table 4.16 Summary of the two designs and the current FREVAR plant.....	76

List of Abbreviations

AS	Activated sludge
BMP	Bio methane potential
BOD	Biological oxygen demand
CIP	Cleaning in place
COD	Chemical oxygen demand
DAF	Dissolved air flotation
DMF	Direct membrane filtration
DOC	Dissolved organic carbon
EU	European Union
FREVAR	Fredrikstad vann, avløp og renovasjon (Fredrikstad water, wastewater, and renovation)
HRT	Hydraulic retention time
LCA	Life cycle assessment
MBBR	Moving bed biofilm reactor
MBR	Membrane bioreactor
MF	Microfiltration
MS	Microsieve
N	Nitrogen? (Ikke nødvendig pga grunnstoff?)
OR	Overflow rate
OTR	Oxygen transfer rate
PACl	Polyaluminium chloride
PVDF	Polyvinylidene difluoride
SBOD	Suspended BOD
SEM	Scanning electron microscope
SS	Suspended solids
TBOD	Total BOD
TMP	Transmembrane pressure
TOC	Total organic carbon
TS	Total solids
VS	Volatile solids
VSS	Volatile suspended solids

1 Introduction

The Global Risks Report has been published annually since 2007 by the World Economic Forum. It describes the changes in risks faced by the world in terms of economy, environment, geopolitics, society, and technology. For the first time, in 2020, all the top five risks in terms of likelihood are of environmental concerns, and environmental concerns also make up three of five for the top risks in terms of impact. It appears clear that the environment is a matter to be taken into consideration in all sections of society (World Economic Forum, 2020).

One way of expressing environmental efficiency is by looking at emissions and resource use. However, if sustainability is to be measured, indicators reflecting the economy and society also needs to be included (Balkema et al., 2001). In 2016, there were 2284 wastewater treatment plants in Norway treating the wastewater of at least 50 person equivalents (pe). These receive nearly 800 million m³ of municipal wastewater annually (Borg et al., 2019). With this amount of wastewater to be treated there is an excellent opportunity to shift towards more environmental and sustainable options both in Norway and worldwide.

According to the Norwegian Environmental Agency, 60% of Norwegian wastewater treatment plants have not been changed since 2007, several of which should have been upgraded. The EU is currently evaluating the wastewater directive, which Norway are facing challenges to follow. There is a chance the wastewater directive will be revised, and other directives might also need to be followed, which might result in stricter treatment requirements (The Norwegian Environmental Agency, 2019). This means new treatment concepts might need to be considered, and the resulting improvement of treatment can be combined with environmentally friendly and sustainable solutions.

In 2015, United Nations (UN) developed and published 17 sustainable development goals (SDGs) which are for *peace, prosperity, people, planet, and future*. SDG number 6 is to ensure availability and sustainable management of water and sanitation for all, which is highly relevant when considering wastewater treatment. It can be argued that several of the SDGs are in fact touching upon this topic, e.g. number 9 about building resilient infrastructure, promoting inclusive and sustainable industrialization and fostering innovation. Other SDGs to mention are number 11 about making cities and human settlements inclusive, safe, resilient, and sustainable, and number 14 about life below water in terms of effluent release (United Nations Department of Public Information).

Globally, it is stated to be likely that around 80% of wastewater is left untreated and released to the environment (UN Water, 2017). Further, according to the same World Water Development Report (UN Water, 2017), described as high-income countries had a 30% fraction of untreated water in 2015, while on the opposite end of the scale, 92% of wastewater in low income-countries were left untreated. The UN emphasises that within the EU, there are challenges with ensuring efficient pollutant removal.

Ødegaard (2016) lists the goals of future wastewater treatment plants which includes avoiding negative impact on receiving waters and recovery of resources like energy and

nutrients as well as water reuse. Additionally, the treatment process should be compact due to the increasing urbanisation, as well as being energy self-sufficient with a low carbon footprint. Further, some critical issues to be considered are mentioned. These involve minimisation of energy consumption through the reduction of air required, pumping, and removal of micropollutants.

Two possible wastewater treatment technologies to be considered as potential candidates for future wastewater treatment plants are direct membrane filtration (DMF) and moving bed biofilm reactor (MBBR). While MBBR is a renowned treatment method for wastewater, with more than 1000 treatment plants worldwide as of 2014 (Ødegaard et al., 2014), DMF is far less common, but has shown excellent results of removal in previous pilot studies (Hey et al., 2017).

1.1 Objectives and motivation

The objectives of this study are:

- To investigate the feasibility of DMF by conducting pilot plant experiments.
- To show what a full-scale DMF treatment plant may be like.
- To compare the designed full-scale DMF plant to a designed full-scale MBBR plant.

The motivation behind these is FREVAR (Fredrikstad vann, avløp og renovasjon, Fredrikstad, water, wastewater and renovation, location of wastewater treatment plant) in Fredrikstad Municipality, South-East Norway, needing a new wastewater treatment plant, and Direct membrane filtration (DMF) and moving bed biofilm reactor (MBBR) are concepts to consider.

1.1.1 Limitations

A pilot plant is not a 100% accurate representation of a full-scale treatment plant but can indicate potential outcomes of full-scale application. Another major limitation that during the time of writing this thesis was the COVID-19 virus. This made travel between Trondheim and the location of the pilot plant in Fredrikstad more limited, as the guidelines given by the Norwegian government about travelling was followed. This again resulted in fewer samples than first anticipated. However, the best was made of the situation, and some samples were taken and provided by lab technicians on-site nonetheless.

1.1.2 Specialisation project

Before this master thesis, a specialisation project on the same topic was conducted as a part of the course TVM4510 at NTNU. Parts of the theoretical background section are based on sections from this specialisation project but have been subject to further review since then. This project is not available online.

1.2 Outline of the thesis

Chapter 2 consists of a theoretical background where the focus is on sustainable wastewater treatment, and the two chosen potential concepts for a new treatment plant in Fredrikstad; Direct membrane filtration and moving bed bioreactor. In chapter 3, the DMF pilot plant study at FREVAR is described, including jar testing, operation, and constituent removal. Chapter 4 concerns the full-scale design of the two concepts which are presented, discussed and compared. This chapter also includes mass balances and process flow charts for the two concepts. The conclusion of the thesis can be found in chapter 5.

2 Theoretical background

In this chapter, the theoretical background of sustainability in wastewater treatment, and the two chosen wastewater treatment concepts to consider, direct membrane filtration and moving bed biofilm reactor, will be described.

2.1 Sustainability in wastewater treatment

In this section, sustainability in wastewater treatment is briefly introduced in order to draw conclusions about whether the processes to be evaluated later are sustainable. If technology is to be sustainable, it cannot threaten either the quality or quantity of resources (Balkema et al., 2001). In the study of Balkema et al. (2001) four sustainability criteria were listed: (1) functional criteria i.e. removal efficiency, (2) economic criteria i.e. cost, (3) environmental criteria i.e. resource use and emissions, (4) and social-cultural criteria i.e. institutional requirements and acceptance.

Life cycle assessment (LCA) is one possible strategy to evaluate environmental impact, with the goal of quantification or characterisations of typical material flows of resource processing, manufacturing, using, reusing, recycling, or disposal (Graedel and Allenby, 2015). LCA has been applied for wastewater treatment by several researchers, as reviewed by Friedrich et al. (2007). Several of the studies found that energy use had the most significant contribution on the environmental footprint (Zhang and Wilson, 2000, Emmerson et al., 1995), with approximately 70% of the energy use found to be occurring during treatment plant operation (Zhang and Wilson, 2000). Other opportunities for reduction of the environmental footprint, which is mentioned as a conclusion from LCA is the reuse of material, as well as choosing reusable materials. Additionally, choosing environmentally conscious suppliers and contractors will reduce the environmental footprint (Friedrich et al., 2007).

According to Martinez (2012), a lot of the total energy consumption of a biological wastewater treatment plant comes from aeration. In 2016, the emissions from municipal treatment plants, including wastewater transportation in Norway made up a total of 194,000 tonnes CO₂ equivalents (Statistisk sentralbyrå, 2016). However, indirect emissions will also occur, e.g. from transport and production of chemicals, which can also be accounted for if the system boundaries are expanded to involve such matters (Borg et al., 2019).

In the report *Greenhouse gas emissions, a guidance document for the water sector* published by Norwegian Water BA (2019), a set of examples of measures to make wastewater treatment more climate-friendly is presented. Some of the listed measures include using biogas, resource recovery, and reducing the transport of, e.g. chemicals. There are also suggestions of measures which are less related to the wastewater treatment itself, for example, use of thermal energy and use of Energy Savings Performance Contracts (ESPC) which guarantees energy savings from a supplier (Borg et al., 2019).

While reducing the energy use of a treatment plant is of great importance, recovery of energy is also of interest. It is most commonly done either through biogas production, which can be used for electricity, fuel, and heat, or through a heat pump using heat from

the wastewater which can heat the treatment plant itself, or nearby settlement using district heating (Martinez, 2012).

By carrying out a qualitative assessment using numbers given by 26 EU member states, it was found that using bio-methane, especially as biofuel, seems to be able to minimise emissions of greenhouse gases (Tilche and Galatola, 2008). In terms of energy, it was found that biomethane from landfills and wastewater treatment plants in Europe could make up as much as 380 PJ/year, which, according to Tilche and Galatola (2008) can substitute petrol for driving 1.31×10^{11} km annually. In 2010, 33% of the total Norwegian biogas production was from wastewater treatment plants (Bachmann et al., 2015).

2.2 Direct membrane filtration

The term **D**irect **M**embrane **F**iltration (DMF) was coined by van Nieuwenhuijzen (2002), while microfiltration/ultrafiltration membranes were commercialised already in the 1950s (Water Environment Federation, 2006). As described in the recent review by Hube et al. (2020), direct membrane filtration can be operated with various driving forces; pressure-driven, osmosis-driven, thermally-driven, and electrically-driven.

2.2.1 Membranes in wastewater treatment

With the increase in performance and the decrease in price, membranes have been increasing in popularity in recent years (Hube et al., 2020). Their popularity is likely to increase due to its compactness (Van Nieuwenhuijzen et al., 2000) combined with the continuous population increase, especially in urban, densely populated areas (United Nations Department of Economic Affairs (Population division), 2019) with a higher fraction of people connected to the sewer network than in rural areas (Butler and Davies, 2017).

Generally, membranes can be classified by looking at pore size, applied pressure, and molecular weight cut-off. Constituents larger than the pore size of the membrane will not go through but instead gather near the membrane surface described as size exclusion (van Nieuwenhuijzen, 2002). Different membranes and their respective dimensions and materials rejected are subject to interpretation, one of which can be found in Table 2.1 as according to Crittenden et al. (2012).

Table 2.1 Membrane types with sizes and rejected materials.

Membrane type	Nominal pore dimensions [μm]	Materials rejected
Microfiltration (MF)	0.1	Particles, sediment, algae, protozoa, bacteria
Ultrafiltration (UF)	0.01	Small colloids, viruses
Nanofiltration (NF)	0.001	Dissolved organic matter, divalent ions
Reverse osmosis (RO)	Nonporous	Monovalent ions

The biggest challenge found using membrane filtration in wastewater treatment is fouling, which results in flux decline and/or pressure increase over time (Guo et al., 2012). Membrane fouling can be classified as reversible or irreversible. Guo et al. (2012) list six principal fouling mechanisms, which are pore blocking, cake formation, concentration polarisation, organic adsorption, and inorganic precipitation. Pre-

treatment, backflushing, and chemical cleaning are the most common strategies for mitigation of membrane fouling (Tchobanoglous et al., 2014).

Several types of membrane module configurations are available, with tubular, hollow fibre, spiral wound, and flat sheet being the most common ones (Water Environment Federation, 2006). Hybrids and variations of these are also under development (Judd and Judd, 2011). Generally, different module configurations vary in size, dimensions, and packing density, and have varying flow patterns; outside-in or inside out. In wastewater treatment, hollow fibre and flat sheet are frequently used, utilising outside-in flow pattern as it allows for higher fractions of total suspended solids, and also allows backwashing and air scouring (Tchobanoglous et al., 2014).

A variety of membrane materials have been used for membrane production, with the earliest being made of cellulose acetate, a material which has declined in popularity due to operational limitations (Park et al., 2015). Material properties are essential, for example, chemical stability is vital for sufficient membrane cleaning (Warsinger et al., 2018), and fouling problems are more frequent with hydrophobic materials (Meng et al., 2009) like polymers (Judd and Judd, 2011). Membrane materials are often categorised into three types: polymeric, ceramic, and nanocomposites which all have different properties. Ceramic membrane membranes can handle more aggressive cleaning, e.g. use of biocides or higher temperatures, than a polymeric one, while polymeric can be more cost-efficient and have smaller pores (Warsinger et al., 2018). Though ceramic membranes can be described as superior in many required properties, they are less widely used because of their high cost (Judd and Judd, 2011). Nanocomposites are not commercially available yet, but the material is under ongoing research (Warsinger et al., 2018).

2.2.2 Fundamentals of direct membrane filtration

The concept, termed as direct membrane filtration (DMF) in this thesis, as done by several researchers (Hey et al., 2017, Van Nieuwenhuijzen et al., 2000, Ravazzini et al., 2005), has also been described as direct membrane separation (Ahn et al., 2001) and direct sewage microfiltration (Jin et al., 2015). It describes the concept of treating wastewater directly over a membrane e.g. microfiltration or ultrafiltration without the involvement of any biological treatment (Hey et al., 2017).

The most common membrane configuration used in direct membrane filtration is found by Nascimento et al. (2018) to be hollow fibre. For DMF, the most common membrane material was found to be polyvinylidene fluoride (PVDF). In contrast, other polymeric, as well as ceramic membranes, were found to be used to a smaller extent (Nascimento et al., 2018).

As described in section 2.2.1, membrane fouling is the most significant drawback of using membrane filtration in wastewater treatment, and measures to control and mitigate fouling are therefore significant if direct membrane filtration is to be sustainable in long term (Hube et al., 2020). Considering this, several researchers have carried out experiments with different pre-treatment methods prior to direct membrane filtration. Examples of include Gong et al. (2015) investigating coagulation and adsorption, Hey et al. (2017) applying a microsieve and metal salt and/or polymer addition, Im et al. (2019) experimented with ozonation and coagulation, and Zhao et al. (2019) compared different metal salt coagulants. A review of different pre-treatments for direct membrane filtration was conducted by Nascimento et al. (2018).

The use of aeration has also been described in several studies and have been found to be a successful but energy-demanding method of reducing fouling rate (Jin et al., 2015, Mezohegyi et al., 2012). Mezohegyi et al. (2012) found the use of vibration of membranes to be an efficient method. Discontinuous operation, also called relaxation of membranes, has been discussed as another means of physical cleaning, letting foulants diffuse away from the membrane surface when relaxed (Le-Clech et al., 2006). This has also been found to give satisfying results of fouling reduction when combined with aeration (Hong et al., 2002).

Additionally, fouling mitigation in direct membrane filtration using chemical cleaning has been found to be efficient, testing sodium hydroxide (NaOH), sodium hypochlorite (NaOCl), and citric acid. Citric acid and NaOCl proved to be more efficient than NaOH, which led to a more rapid TMP increase (Lateef et al., 2013). Gao et al. (2011) describe oxidising reagents like NaOCl as being effective in removing organic matter, and acids like citric acids as being effective in removing inorganic foulants, which led Lateef et al. (2013) to the conclusion that a combination of these two could efficiently mitigate fouling.

Hey et al. (2017) showed that direct membrane filtration is a process that could be able to meet European wastewater discharge demands in terms of phosphorous and chemical oxygen demand (COD) for small and medium-sized wastewater treatment plants. High removal efficiencies have also been achieved by several other researchers. Bendick et al. (2005) found microfiltration membranes to reduce biological oxygen demand (BOD) and suspended solids below the standards of secondary treatment as well as being able to separate bacterial indicator organisms.

Direct membrane filtration, which is without any biological treatment, and membrane bioreactor, a concept involving utilisation of biology, is not to be confused as both are membrane-based wastewater treatment processes (Ahn and Song, 2000). Direct membrane filtration has been used for treating relatively low strength wastewater, while membrane bioreactors, which rely upon microorganisms, have been more regularly used to treat higher strength wastewater (Ahn et al., 2001).

2.2.3 Chemical precipitation of phosphorous

With membrane filtration being a merely physical process, if phosphorous is to be removed, precipitation to form particles is of interest, as 50-80% of influent phosphorous is soluble orthophosphate, making up the biggest fraction of phosphorous in raw wastewater (Gilberg et al., 2003).

Precipitation using a type of hydrolysing metal salts is a popular method of phosphorous removal in wastewater (Duan and Gregory, 2003). Aluminium and ferric salts are the most widely used precipitants in both drinking water and wastewater treatment in Norway (Ødegaard et al., 2014) but have lately seen a decline in popularity because of environmental concerns (Lee et al., 2014). Aluminium and ferric salts demonstrate comparable phosphorous removal efficiency in both synthetic and actual wastewaters and will have the same type of precipitation reaction when added (Szabó et al., 2008). The precipitation depends on the solubility of the metal, which depends on pH, with iron coagulants having a much broader range of pH than aluminium (Duan and Gregory, 2003).

Addition of metal salts has also been found to enhance COD removal (Jin et al., 2015). This has been agreed upon by several researchers, finding the combination of a metal

salt and polymer would improve control not only of phosphorous but also of SS and COD (Väänänen et al., 2016, Remy et al., 2014).

2.2.4 Direct membrane filtration and sustainability

As described previously, the DMF concept does not involve any biological treatment, which gives it the advantage of simple operation and maintenance compared to, for example, membrane bioreactors (Hube et al., 2020).

Several studies have found direct membrane filtration, specifically with a preceding coagulation step, to be adequate for energy recovery. With the relatively low consumption of energy by the overall process, this can result in energy neutrality (Jin et al., 2016). However, this energy demand will depend on several factors, e.g. required pressure; membrane configuration in terms of both operation and build; and what type of cycle being used, i.e. relaxation; and operation time (Pinnekamp et al., 2018). Nevertheless, in some cases, it is necessary to trade off a more environmentally friendly option. One example of this is the use of cross-flow configuration being more energy demanding than dead-end, but it is frequently chosen as it can handle feedwater with a higher content of constituents (Pinnekamp et al., 2018). Therefore, it might be more sustainable overall to use a crossflow configuration, when looking at all the indicators defining a sustainable treatment process.

Despite membrane aeration being energy-consuming, it has been found that the specific energy consumption of a DMF plant with coagulation and flocculation with metal salt and polymer can be lower than the average small or medium-sized Swedish wastewater treatment plant (Hey et al., 2017). The average electricity demand was found to be 0.4 kWh/m³ for the direct membrane filtration, and 0.75 kWh/m³ for the 105 Swedish treatment plants.

2.3 Moving bed biofilm reactor (MBBR)

Moving bed biofilm reactor is a biofilm process which was introduced and developed by Hallvard Ødegaard and commercialised by Krüger Kaldnes in the late eighties and has in the later years become a success. Per 2014 there were 1000 MBBR treatment plants worldwide (Ødegaard et al., 2014).

2.3.1 Fundamentals of MBBR

In a moving bed biofilm reactor, the biofilm grows on small carriers moving along with water in a reactor, with biofilm being a layer of biomass developing on the surface of a media. Compound diffusion in and out of biofilm is essential, meaning the effective biofilm thickness is of significance. Ideal biofilm is thin due to the depth of full substrate penetration being less than 100 µm (Ødegaard, 2006). The biofilm must be distributed evenly over the carrier surface, which can be ensured by turbulence in the reactor. This turbulence also plays the role of transporting the substrate to the biofilm. When the turbulence, which is caused by either stirring or aeration, is sufficient, the biofilm should be thin and smooth (Ødegaard et al., 1999).

There are several carrier types available with varying size and specific surface area, as well as different material choice, but they are, however in the same density range. With a density of close to the density of water; 1 g/cm³, it is ensured that the carriers will easily follow the turbulent water stream in the bioreactor. This constant movement of the biofilm gives good contact between substrate and bacteria reacting with the substrate,

compared to what of a trickling filter where the biofilm is stationary (McQuarrie and Boltz, 2011). The carriers are retained within the reactor using outlet sieves, which only let biomass and water through (Ødegaard et al., 1999).

An example of what a biofilm carrier can look like is shown in Figure 2.1, with a close-up photo in Figure 2.2. Investigations have been made on the impact of different shapes and sizes of biofilm carriers. It was found that the effective surface area influenced the results to a bigger degree. The shape and size of the carriers will have an impact on the effective area. After establishing the effective area, there seems to be little difference between the different shaped and sized carriers (Ødegaard et al., 2000).



Figure 2.1 BWT X carrier, specific area 650 m²/m³. Courtesy of Biowater Technology.



Figure 2.2 Close-up of the BWT-X carrier. Courtesy of Biowater technology.

The filling fraction of the reactor can be chosen, but the general recommendation is a maximum of 70%, depending on carrier, in order to allow free movement of the carrier suspension. A lower fraction of as much as is needed below this can be used. This is especially advantageous when upgrading existing wastewater treatment plants (Ødegaard, 2006).

An MBBR treatment plant will typically consist of several reactors in either series or parallel, with the reactor volumes being totally mixed and without any unused space (Rusten et al., 1997). Often, pre-settling is used before the biological process in order to avoid clogging of bioreactor sieves (Ødegaard, 2006). Following the bioreactors, a particle separation step is required, due to the biomass in the effluent. The biomass accumulation is independent of settler, meaning MBBR has excellent flexibility in terms of choice of separation unit. Several options have shown success, e.g. sedimentation basins, dissolved air flotation, flocculation with lamella settlers, and membrane filtration (McQuarrie and Boltz, 2011). This flexibility allows for choosing more compact separation processes.

2.3.2 MBBR and phosphorous removal

Biological removal of phosphorous, combined with nitrogen removal, can be achieved using MBBR when operating it in a sequencing batch reactor (Helness and Ødegaard, 2001). However, the most common choice of solution when an MBBR treatment plant is to be used for removal of phosphorous in addition to organic material, is the addition of a chemical as a precipitating agent just after the bioreactor, followed by a flocculation step (Ødegaard et al., 1999). Chemical precipitation of phosphorous was described further section 2.2.3. In Norway, the combination of chemical precipitation and MBBR is the most common combination of biological and chemical treatment (Ødegaard et al., 2014).

2.3.3 MBBR and sustainability

The initial idea behind the moving bed biofilm reactor was to have a treatment process with continuous operation, using a biofilm that could not clog, and have reduced operational cost, i.e. no backwash and low headloss. Additionally, a high specific area was desired, resulting in a low footprint of the treatment process (Rusten et al., 1997). The flexibility to choose the liquid-solid separation process gives the flexibility to make the whole treatment process more compact by choosing, for example, dissolved air flotation (McQuarrie and Boltz, 2011).

Another advantage of MBBR compared to other biological processes like conventional activated sludge (AS) is that in MBBR, there is no sludge recirculation to obtain biomass in the reactor (Ødegaard, 2006). This makes the process very operator friendly, as worries concerning sludge return ratios, proper sludge wasting, sludge volumes, or sludge volume indexes are not of relevance, meaning little supervision is required (Rusten et al., 1997). Drawbacks of MBBR compared to conventional activated sludge involve a generally higher energy demand, and some maintenance matters being more complex, i.e. maintenance of air diffusers which involves removing carriers (Tchobanoglous et al., 2014).

There has recently been reported a case in Norway where carriers have gone through the wastewater treatment plant and out in the recipient (Trondsen, 2015). The carriers have been reported found along the Norwegian south coast as well as the northern coast of Denmark up until 2020 (Løberg Skår et al., 2020). Releasing plastic to the ocean is an obvious drawback; however, according to the operational leader of the wastewater treatment plant where the incident happened, this would only happen due to operational errors, or fatigue, or under-sizing of the sieves holding carriers back. It is not of general concern in an MBBR plant (Geir Breimyr, 2020, personal communication, 20.04).

3 Pilot plant study

In this chapter, the pilot plant set-up will be described, followed by jar-testing experiments. This lays the foundation for the reduction of chemical use. The pilot plant performance in terms of constituent removal and operation in regard to transmembrane pressure is looked into. Additionally, testing of sludge from both the membranes and sieve was conducted to give a foundation for comparison. COD, phosphorous, and SS were all tested within the process to gain a further understanding of how the pilot plant performs between inlet and outlet.

The chapter includes methodology, results, and discussion of those for the respective parts. In the end, the feasibility of the DMF concept based on the pilot plant study is discussed.

3.1 Description of the pilot plant

The pilot plant is located at FREVAR municipal wastewater treatment plant in Fredrikstad, South-East Norway. The pilot plant was built at the beginning of the summer of 2019, based on experimental work by Hey et al. (2017), operating a smaller DMF-pilot plant in Sweden. The DMF-pilot plant in Fredrikstad has been in operation since then, with some time off. The original membrane modules were found not to work as desired in August/September 2019 and were therefore replaced with new ones in October of the same year. The treatment steps of the DMF-pilot plant is shown in the simplified flowchart in Figure 3.1.

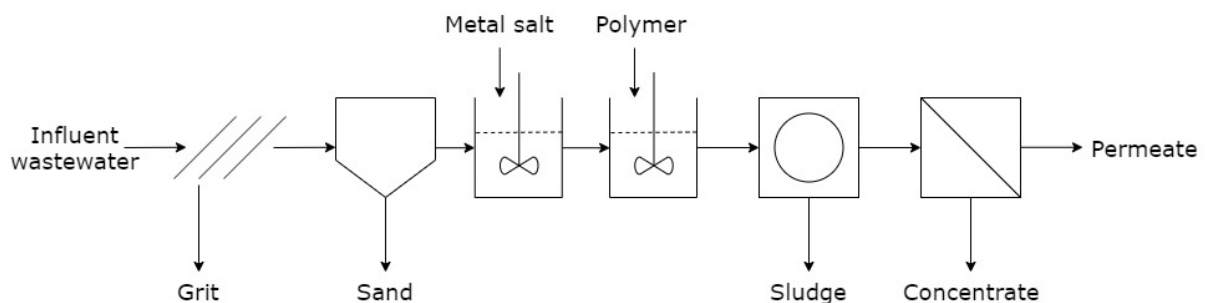


Figure 3.1 Simplified flowchart to illustrate the pilot plant.

The process and instrumental diagram of the DMF-pilot plant is shown in Figure 3.2. The PID shows that the pilot plant included two membrane modules operated in parallel. This was made by Tobias Hey prior to building of the pilot plant, but was later adjusted to resemble how the pilot plant looked like.

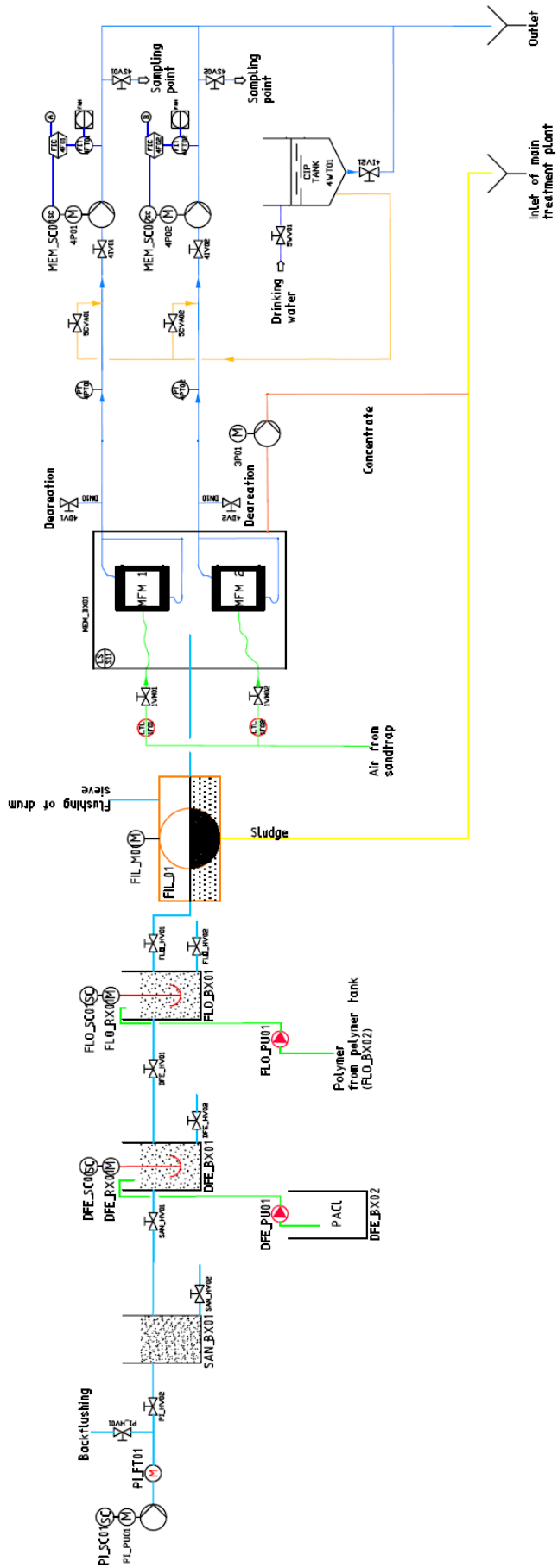


Figure 3.2 P&ID of the pilot plant.

The influent wastewater first goes through coarse screens of the main treatment plant at FREVAR. After the screen, the wastewater was pumped to the pilot plant with a designed flowrate of 4 m³/h, whereas the average flowrate was 3.1 m³/h through this study. The pumped water first enters an aerated sandtrap designed according to the Norwegian guidelines for design of wastewater treatment plants (Ødegaard et al., 2009). The sandtrap has a volume of 1 m³ which results in detention time of more than 10 minutes, as suggested by the Norwegian guidelines, and can be seen in Figure 3.3.



Figure 3.3 The sandtrap of the pilot plant.

In order to remove phosphorous to fulfil the new treatment requirements (County governor of Oslo and Viken, 2019), PAX-18 (Polyaluminium chloride) is added as a precipitating agent. This is followed by the addition of Superfloc N-100, a non-ionic polymer in order to aggregate and strengthen the flocs. The technical data sheet for PAX-18 and Superfloc N-100 can be found in Appendix A and B, respectively. Both chemicals are added directly to the reactors, which both have a volume of 1 m³ and is shown in Figure 3.4. The reactor volumes for metal salt and polymer addition and mixing are designed following the Norwegian guidelines for design of wastewater treatment plants (Ødegaard et al., 2009).



Figure 3.4 Coagulation and flocculation reactors of the pilot plant.

The chemical step is followed by a rotating drum sieve with a polyester filter cloth and pore openings of 100 μm . The designed filtration rate of the microsieve is 5 m/h. According to experiments conducted by Rusten and Lundar (2004), a low filtration rate obtained a filter cake and hence higher removal rate. Furthermore, successful operation using a drum sieve in combination with coagulation and flocculation has been described by Ljunggren et al. (2007) and Remy et al. (2014). The microsieve in the pilot plant can be seen in Figure 3.5.

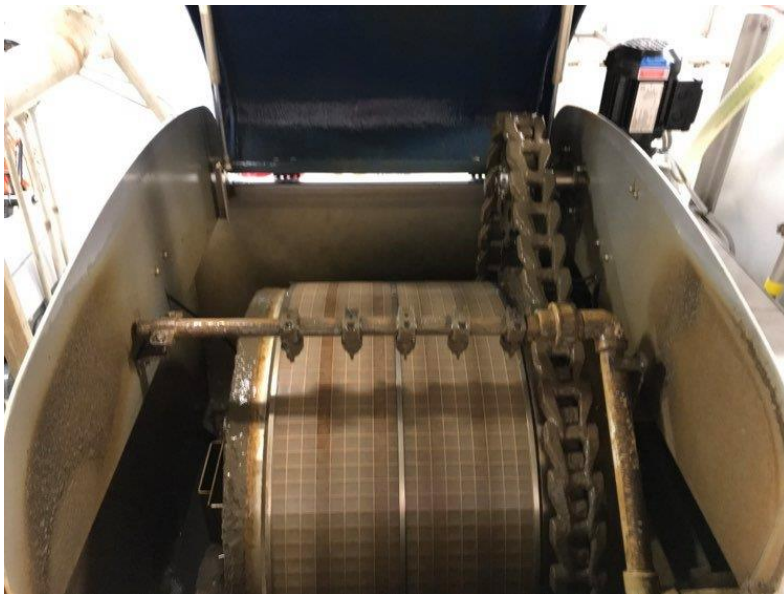


Figure 3.5 Interior of microsieve in the pilot plant.

The final step of the main treatment process consists of two microfiltration membranes with a pore opening of 0.2 μm . These were provided by Alfa Laval AB and are made of polyvinylidene difluoride (PVDF) (Alfa Laval, 2020). The membranes are of hollow sheet configuration, which is a combination of the more commonly used flat sheet and hollow fibre. This relatively novel configuration is illustrated in Figure 3.6 and Figure 3.7.

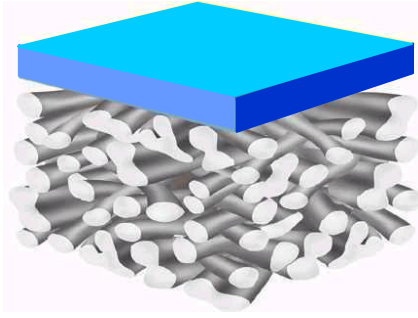


Figure 3.6 Illustration of hollow sheet membrane. Courtesy of Alfa Laval AB.

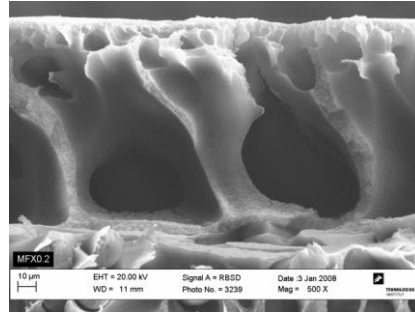


Figure 3.7 SEM photo of hollow sheet membrane. Courtesy of Alfa Laval AB.

The two membrane modules are submerged in a containment vessel below the floor and are connected in parallel. One of the membrane modules is shown in Figure 3.8. They are set to be operated at a constant flux (ca. 2 m³/h) and hence a declining TMP, with operation cycles consisting of 10 minutes of operation and 2 minutes of relaxation. The containment vessel is continuously aerated. The combination of aeration and relaxation has been proved successful for mitigation of fouling by (Hong et al., 2002), while (Mezohegyi et al., 2012) found vibration of membranes more successful than aeration, albeit aeration provided more satisfying results than operation without any kind of fouling mitigation. The membranes are operated at a cross-flow configuration which has shown to be able to reduce membrane fouling to some extent compared to a dead-end configuration (Arndt et al., 2016, Gibert et al., 2016).



Figure 3.8 One of the membrane modules in the pilot plant. Courtesy of Tobias Hey.

The membranes in the pilot plant are cleaned by instructions from Alfa Laval, using citric acid and hydrogen peroxide for cleaning, and sodium hydroxide and hydrogen chloride for necessary pH adjustment.

For a further description of the pilot plant and a more extensive literature related to direct membrane filtration, it is referred to the specialisation project from the autumn semester 2019.

3.2 Jar testing

Jar test apparatus is a usual method for studying mechanisms of coagulation and flocculation processes (Ødegaard et al., 2014). Jar testing was developed in 1921 by Langelier and has been developed further over the years. The purpose of this type of testing is to simulate coagulation and flocculation in the full-scale treatment plant (Crittenden et al., 2012).

3.2.1 Method

A 25-litre grab sample of inlet water was used, which included reject water from the main treatment plant, which was bypassed the pilot plant. Kemira flocculator 2000 was used for the jar tests, with the settings set to 10 seconds of rapid mixing, 10 minutes of slow mixing, and eventually 10 minutes of sedimentation, i.e. no mixing. The chemicals to be tested were added during the phase of rapid mixing. The pH, turbidity, and orthophosphate concentration were measured as soon as the sedimentation phase finished. The jar test setup is shown in Figure 3.9.



Figure 3.9 Jar test setup.

The chemicals used were PAX-18 and Superfloc N-100, as in the pilot plant. The polymer was prepared as a 0.2% stock solution (2 g polymer per L water), as in the pilot plant. First, the PAX dosage was varied, while the polymer dosage was held constant. The different PAX dosages were chosen from the supplier Kemira's experience-based suggestion, which was also the case for the polymer dosage. When seemingly successful dosages were found based on turbidity and orthophosphate, three of these were tested again with two new and different dosages of the polymer.

For the measurement of turbidity, a Hach 2100Qis turbidimeter was used. In order to measure orthophosphate, LCK349 cuvette test was used after filtering the sample through a folded filter. A Hach HQ11D pH meter was used for measurements of the pH.

3.2.2 Results and discussion of jar test

The results from the jar tests are presented in Table 3.1 and visualised in Figure 3.10 and Figure 3.11 showing a reduction of both turbidity and orthophosphate concentration already at the lowest dosage of PAX, at 3.6 mg Al/L. At the lower dosages, both turbidity and orthophosphate concentration decrease almost linearly, with the turbidity curve flattening slightly earlier than the orthophosphate curve. The removal rate of

orthophosphate is assumed to be representative for the phosphorous removal (Gilberg et al., 2003) and reaches 90% already at 7.2 mg Al/L, and keeps increasing until the concentration is below the measuring range, i.e. below 0.15 mg/L, of the cuvettes used. These concentrations are so low, and if the lowest measurable value is used, removal of 96.5% is still obtained.

The range of aluminium concentrations was narrowed down for testing with two new polymer dosages to investigate whether this had a significant impact on the results, and in that case, if less aluminium could work with the addition of more polymer. As shown in the graphs in Figure 3.10 and Figure 3.11, and in Table 3.1, for 5.4 mg Al/L and 7.2 mg Al/L there was a slight improvement in orthophosphate reduction, and the same improvement was also seen for the turbidity at 5.4 mg Al/L. However, at 9 mg Al/L for orthophosphate and 7.2 mg Al/L and 9 mg Al/L for turbidity, the condition was worsened from the lower polymer dosage.

The concentration of orthophosphate was not expected to decrease just by increasing the polymer concentration like it appeared to do. One possible explanation of this can be a slight strengthening of the flocs by the polymer; with the lower dosage, it is possible that some colloidal matter got through the filter. This is more likely to happen if the flocs are weaker, which might be the case with a lower dosage. However, the difference in concentrations and hence removal is not of enormous significance.

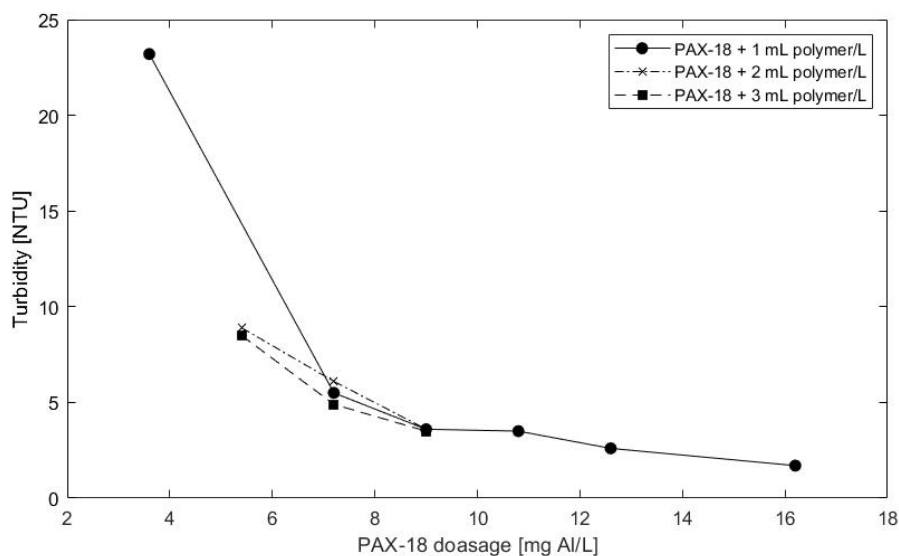


Figure 3.10 Turbidity with different chemical dosages.

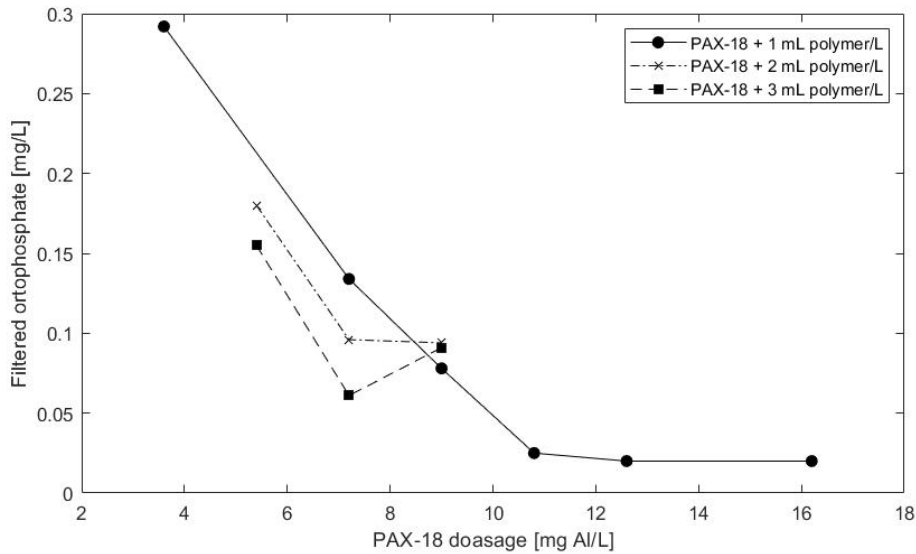


Figure 3.11 Orthophosphate concentrations with different chemical dosages.

Table 3.1 Results from jar tests.

mg PAX-18/L	mg Al/L	mL polymer/L	pH	Turbidity [FNU]	Filtrated orthophosphate [mg PO ₄ -P/L]	Removal
0	0	0	7.45	123	1.47	-
40	3.6	1	7.25	23.2	0.292	80.1 %
80	7.2		7.09	5.5	0.134	90.9 %
100	9.0		7.08	3.6	0.078	94.7 %
120	10.8		6.96	3.5	0.025 ^a	98.3 %
140	12.6		6.93	2.6	0.02 ^a	98.6 %
180	16.2		6.86	1.7	0.02 ^a	98.6 %
60	5.4		2	7.03	8.9	0.18
80	7.2	7.04		6.1	0.096	93.5 %
100	9.0	7.03		3.6	0.094	93.6 %
60	5.4	3	7.01	8.5	0.155	89.5 %
80	7.2		6.98	4.9	0.061	95.9 %
100	9.0		6.96	3.5	0.091	93.8 %

^aUnder measuring range of cuvette (0.05 mg/L).

The strength of the flocs was judged by visual inspection in all the beakers. The dosing of higher polymer concentration did not seem to make a visible difference in the floc strength. As this could not be quantified, there is an apparent uncertainty of the observation. As described in the section above, it may seem that the addition of extra polymer added some floc strength, as less orthophosphate went through the filter. However, the amounts are small, e.g. at 7.2 mg Al/L, 2.6% and 5% increase in removal were measured for 2 mL polymer/L and 3 mL polymer/L respectively, than for 1 mL polymer/L.

Further, it was observed a decrease in turbidity of 80% already at the lowest chemical dosage of 3.6 mg Al/L. The pH also decreased as more PaCl was added, like expected due to the low pH of the substance, as described in Appendix A (PAX-18 Technical

datasheet). The pH value was consequently above pH 6 which is the pH where aluminium salts have the lowest solubility, i.e. precipitates most easily (Duan and Gregory, 2003), but based on the results the orthophosphate seems to be precipitating as it should.

Testing of both turbidity and orthophosphates can give the results relatively quickly, which is one advantage of choosing these two parameters. Turbidity is a commonly used parameter to quantify the physical properties of water, as it implies the reduction of water clarity caused by light scattering by particles (Crittenden et al., 2012).

Orthophosphate was chosen as the testing parameter to indicate phosphorous removal, as 50-80% of the total phosphorous is present as soluble orthophosphate (Gilberg et al., 2003).

There will be multiple inaccuracies and error sources when conducting the jar test. Most noteworthy is human errors, especially in terms of pipetting. This can affect whether the correct concentrations are added or not, particularly when it comes to the polymer which behaves viscously. The risk of pipetting errors also applies when adding samples to the cuvettes. There will also be some uncertainties by using Dr Lange cuvettes.

Based on jar testing, it was decided to reduce the chemical dosing. The high removal efficiency of the pilot plant supports this decision, as the removal requirements for both phosphorous and COD seem to be fulfilled without difficulties. Therefore, the initial dosage of 147 mg PAX-18/L is reduced to 100 mg PAX-18/L, while the polymer is slightly increased to 1 mL polymer/L. This combination is chosen due to being sufficient for meeting the phosphorous removal requirements of 90% more than just marginally in the tests, like 80 mg PAX-18/L did, with a removal of 90.9%.

3.3 Chemical dosing

In the following sections, the analysis result and discussion will be described using the description for the different chemical dosages shown in Table 3.2. The different dosages start with newly chemically cleaned membranes.

Table 3.2 Chemical dosages for different operation conditions.

	PAX-18 [mg Al/L]	Superfloc N-100 [mg polymer/L] ([mL polymer solution/L])	Date from-to
Initial operation conditions	13.2	1.48 (0.74)	23.10.2019 - 29.03.2020
Optimised operation conditions I	9.0	2.00 (1.00)	30.03.2020 - 18.04.2020
Optimised operation conditions II	9.0	1.48 (0.74)	30.04.2020 - 26.05.2020

The initial operation conditions were set by operational personnel mid-October 2019. As the writing of the master thesis started, it was decided to keep these until jar tests had been performed to provide a more educated suggestion to change of dosage.

With the first new process configuration, TMP of -0.5 bar was reached within a week due to fouling. It was decided to try once more, but without CIP, but the membranes were continuously air scoured as usual. Next, based on results from the two previous operation conditions, it was decided to go on using the metal salt dosage from optimised conditions I (Table 3.2), and the polymer dosage from the initial operating conditions.

The inflow is constant, and the chemicals are dosed flow proportionally, meaning the dosing is also constant. The dosages were based on a flow of 3.1 m³/h.

3.4 Internal process behaviour

To understand the performance of the different steps within the process, samples of the wastewater were taken and analysed at several points in the pilot plant. The samples were taken at three internal points – from the sandtrap, from the polymer reactor, and in the filtrate from the microsieve. COD and phosphorous were both tested filtered and unfiltered in the sandtrap and microsieve filtrate, as neither of those constituents is removed between those. Further, as well as the two chemical parameters, SS was also measured later to be able to calculate the mass balance of the system to make a clearer picture of the overall flows.

3.4.1 Method

All samples were grab samples taken directly from their respective reactor. The samples were not taken with a time lag, as the wastewater was expected to have a nearly uniform quality during the relatively short period between the three samples. Sample 1 and 2 were taken on the same day with approximately 4 hours between, while the third sample was taken the following day.

The SS analysis was done following Standard Norge (1983). Weighing of the samples was performed using Sartorius ED224S Extend analytical balance, with a standard deviation of 6x10⁻⁵ g. The analysis of COD and phosphorous were taken using Hach cuvettes LCK114 and LCK314, and LCK348 and LCK 349, respectively, dependent on expected concentration. For the filtered test, the filtrate from filtering the SS was used. Hach HT200 thermostat was used for sample digestion, and Hach DR3900 spectrophotometer was used for the analyses of the cuvettes.

3.4.2 Results and discussion of internal process behaviour

The analyses performed in this section was done during the initial operating conditions (Table 3.2). It was observed how the concentration of phosphorous changes after having passed the chemical precipitation and microsieve step. The total phosphorous concentration has decreased. The fraction of dissolved phosphorous did also significantly decrease from the sandtrap to the microsieve filtrate, which leads to the particulate fraction increasing. The results are shown in Table 3.3.

Table 3.3 Phosphorous concentrations internally in the pilot plant.

	Sandtrap			Microsieve filtrate		
	Total [mg/L]	Filtrated [mg/L] (%)	Particulate [mg/L] (%)	Total [mg/L]	Filtrated [mg/L] (%)	Particulate [mg/L] (%)
1	3.57	1.38 (39)	2.19 (61)	1.93 ^a	0.034 (1.8)	1.896 (98.2)
2	4.34	2.49 (57)	1.85 (43)	2.26 ^a	0.063 (2.8)	2.197 (97.2)
3	4.24	1.45 (34)	2.79 (66)	3.07	0.176 (5.7)	2.894 (94.2)
Avg.	4.02	1.77	2.28	2.42	0.091	2.329

^aOut of the measuring range of the used cuvette, i.e. above 1.5 mg/L.

The natural explanation of the decrease in dissolved phosphorous and increase in particulate phosphorous is the chemical step aiming for precipitation of dissolved phosphorous. By judging from the given results, this seems to be successful in the pilot plant. While there is still a concentration of phosphorous left in the filtrate, the majority

is of particulate origin, which the subsequent microfiltration step is expected to remove efficiently. As the total phosphorous concentration decreased, some of the incoming particulates and some of the precipitate must also have been removed by the drum sieve.

The change of COD from the sandtrap to the filtrate of the drum sieve is shown in Table 3.4. There is some removal of COD over the microsieve, although not very extensive. Just as in the sandtrap, the majority of the COD is present as particulate, which, like for particulate phosphorous, is expected to have efficient removal in the following membrane step.

Table 3.4 COD concentrations internally in the pilot plant.

	Sandtrap			Microsieve filtrate		
	Total [mg/L]	Filtrated [mg/L] (%)	Particulate [mg/L] (%)	Total [mg/L]	Filtrated [mg/L] (%)	Particulate [mg/L] (%)
1	404	79.3 (19.6)	325 (80.4)	351	41.9 (11.9)	309 (88.1)
2	416	123 (29.6)	293 (70.4)	316	110 (34.8)	206 (65.2)
3	553	156 (28.8)	397 (71.8)	462	137 (29.7)	325 (70.3)
Avg.	458	358	338	376	96.3	313

Table 3.5 shows how the concentration of SS changed internally in the pilot plant. As one would expect, there is an increase in the SS concentration from the sandtrap to the influent of the drum sieve due to the addition of chemicals which will produce chemical sludge. The concentration decreases again after going through the microsieve for sample 2 and 3, but not for the first one. One possible explanation for this is the difficulties of grabbing the sample out from the drum sieve. The filtrate can be challenging to reach by hand without touching the internal walls of the sieve. This may result in filter cake residues and other residues entering the sampling bottle, but this was hard to control. This can be what happened for sample 1, causing the SS concentration to increase instead of decrease after going through the microsieve. By averaging the measurements, the overall SS removal over the microsieve was found to be 27%.

Another possible explanation for the increase of SS in sample 1 is the fact that there was no time lag between the sampling. This implies the measurements can have been done correctly, but the incoming and outgoing water of the microsieve was perhaps not of the same quality by the inlet.

Table 3.5 SS concentrations from internally in the pilot plant.

	Sandtrap [mg/L]	Microsieve influent [mg/L]	Microsieve filtrate [mg/L]
1	102	136	167
2	212	254	154
3	212	280	168
Average	175	223	163

There are several factors involved which may have made the samples less representative than desired. The fact that there are no sampling valves makes the grabbing of the samples slightly more complicated. When taking a sample from the sandtrap, the sample has to be taken from the top. The sampling location might be a source of error; however,

the pipe between the sandtrap and the coagulation tank goes from the top of the sandtrap, meaning the error is unlikely to be significant. The same point can be valid for the sample from the polymer reactor. There will also be uncertainties with pipetting, as well as some uncertainty of the analysing methods.

Due to the situation with COVID-19, sampling within the process was decided to be discontinued. It would be beneficial to continue; However, as described in section 3.3, due to the rapid change in TMP, this would not be considered an optimal operation condition. It was therefore of priority to change the chemical dosage instead of getting these analyses done after CIP.

3.5 Sludge sampling

The pilot plant has separated sludge flows from the microsieve and the membrane tank. Both flows are tested for total solids (TS) and volatile solids (VS). This is of interest because high content of VS indicates high methane potential in anaerobic digesters, which has been supported by BMP testing (Paulsrud et al., 2014). Remy et al. (2014) conducted experiments where part of the setup resembled the pilot plant with coagulation, flocculation, and microsieving, and found that the sludge from the microsieve showed a high specific biogas yield compared to mixed sludge from a reference conventional activated sludge plant.

3.5.1 Method

Both the sludge from the microsieve and the membrane tank were tested for total solids (TS) and volatile solids (VS). Both analyses were performed following Norwegian Standard NS 4764 (Standard Norge, 1980). The scale used was Sartorius ED224S Extend analytical balance, which has a standard deviation of 6×10^{-5} g. Samples from both flows were taken 'mid-flow'. In order to ensure a representative result, samples from both sludge flows were tested twice and averaged. Sludge samples from the sieve were taken from the sludge flow right after flushing, while the samples from the membrane tank the sludge pump was turned on for the sake of simplicity.

The sludge flow from the microsieve was measured using a bucket with a volume of 10 L while taking the time. Additionally, the flow and time between the main flows, i.e. between cleanings, were measured. This procedure was done multiple times and then averaged and weighted to find an estimate of the total hourly flow (m^3/h). The sludge flow from the membrane tank is measured continuously with a flow meter.

3.5.2 Results and discussion of sludge sampling

From the results with initial operation conditions, there is a clear difference between sludge from the microsieve and the membranes in the concentration of total solids. There are also differences in the concentration of organic and inorganic solids. This is presented in Table 3.6. The table also shows that sludge from the microsieve has a higher fraction of organics compared to sludge from the membranes. The two analyses were performed on two different days, with approximately 27 hours between. The difference in total solids could also be visually observed, as can be seen in Figure 3.12, with the sludge sample from the sieve containing what looks like much more solids.

Table 3.6 Content of total solids, inorganic, and organic solids in sludge.

		Sludge from sieve		Sludge from membranes		Unit
		Day 1	Day 2	Day 1	Day 2	
Total solids	1	0.34	0.42	0.075	0.10	%
	2	0.36	0.40	0.077	0.12	
	Average	0.35	0.41	0.076	0.11	
Inorganic solids	1	15.65	17.35	55.56	55.88	% of TS
	2	18.22	17.21	62.64	48.62	
	Average	16.94	17.28	59.10	52.25	
Organic solids	1	84.35	82.65	44.44	44.12	% of TS
	2	81.78	82.79	37.36	51.38	
	Average	83.06	82.72	40.90	47.75	



Figure 3.12 Sludge samples from microsieve (left) and membrane tank(right).

Measuring the microsieve sludge flow the way it was done also comes with several uncertainties, including measuring uncertainties of both time and volume. However, it can indicate the mass of total and volatile solids per time. Using these flows combined with the measurements, production of TS, VS, and FS per volume and time was calculated and is presented in Table 3.7. The flows and solid measurements used are averaged over the two days of the samples.

Table 3.7 Production of total, inorganic, and organic solids in sludge.

	From MS	From MF	Unit
Sludge flow	0.075	0.40	m ³ /h
Total solids	3800	940	g/m ³
	285	376	g/h
Inorganic solids	646	517	g/m ³
	48.5	207	g/h
Organic solids	3154	423	g/m ³
	237	169	g/h

Despite the sludge from the microsieve containing a higher fraction of total and organic solids, as shown in Table 3.6, the membrane sludge gives a higher hourly mass of TS, as seen in Table 3.7. The membrane sludge also contains almost the same amount of organics as sludge from the microsieve, when looking at overall production over time, as shown in Table 3.7.

Samples of the sludge were not taken during the operation using the first optimised operation conditions due to lack of time because of the rapid fouling.

When the chemical dosing was changed to optimised operation conditions II (Table 3.2), the sludge was again tested for total and volatile, i.e. organic solids. Observations were similar to the initial operation conditions, although with an increased fraction of total solids in both sludge streams, as can be seen in Table 3.8. It can also be observed in the same table, that while the sludge from the microsieve has a lower fraction of organics than in the previous sample, the membrane sludge has a higher organic fraction than in the previous sample.

Table 3.8 Content of total solids, inorganic, and organic solids in sludge.

		Sludge from sieve	Sludge from membranes	Unit
Total solids	1	0.52	0.21	%
	2	0.57	0.21	
	Average	0.55	0.21	
Inorganic solids	1	24.33	46.85	% of TS
	2	27.53	46.42	
	Average	25.93	46.63	
Organic solids	1	75.67	53.15	% of TS
	2	72.47	53.58	
	Average	74.07	53.37	

More samples over more days would be necessary in order to state if this is the expected percentage of total solids in the sludge streams. It can therefore not be said with certainty that the solid fraction is higher using the chemical dose in this operation condition compared to the previous one. It would be beneficial to do more extensive sampling on this for confirmation; however, the results can still be used in a mass balance to provide insight into how the DMF process works. The results from these measurements are used in the mass balance described in section 4.6.

The sludge flow of the microsieve was not measured in for the operation condition II in this section. Therefore, when calculating the production of solids in this section, the measurements from the initial operation condition was used. This will also cause further uncertainties of the calculations. The flow of sludge from the membranes was read from measurements and averaged over the day.

Using the measurements in this section, significantly higher production of solids can be observed overall, as both the sludge from the microsieve and the membranes had a higher content of TS. This will necessarily also result in higher production of organic solids. These numbers are summarised in Table 3.9.

Table 3.9 Production of total, inorganic, and organic solids in sludge.

	From MS	From MF	Unit
Sludge flow	0.075	0.35	m ³ /h
Total solids	5500	2100	g/m ³
	412.5	735.0	g/h
Inorganic solids	1426	979	g/m ³
	107	342	g/h
Organic solids	4074	1120	g/m ³
	306	392	g/h

Grabbing of the samples as well as performing the actual analysis was not done by the same person in the initial conditions as the optimised conditions II, as operational personnel and lab technicians did it the second time around. This causes some uncertainty as it cannot be ensured the exact same procedure. The sludge samples from the membrane tank were taken by turning on the sludge pump rather than taking a sample when the pump would normally be running. This may have affected the total and organic solids results for the membrane sludge. Another potential source of error and inaccuracies would be the scale. Especially when weighing fluids, the scale experienced issues with stabilising.

Hey et al. (2018) reported an expected methane production for the DMF concept of 0.34 Nm³ CH₄/kg VS. Applying this number to the observed production of VS in the pilot plant, the first set of measurements would result in a theoretical methane production of 3.3 Nm³ CH₄/d, and the second set of measurements would result in a theoretical methane production of 5.7 Nm³ CH₄/d. The numbers are for the sludge from the microsieve and the membranes combined. These results are disputable considering the low number of samples taken in this thesis. Although the latter case showed a significantly higher production of organics, it is only based on one measurement. The first case is based on samples over two days, and can by that perhaps be considered more representative, but the two measurements did vary especially in the fraction of total solids. This variation substantiates that more samples should have been taken for both operating conditions in order to comment on the biomethane potential.

3.6 Removal efficiency

Removal efficiency is of interest due to new requirements given by the County Governor of Oslo and Viken (2019), requiring a phosphorous removal of 90%, COD removal of 75% of influent concentration or reduced to maximum 125 mg O₂/L, and BOD₅ removal of 70% or reduced to maximum 25 mg O₂/L. These are requirements found in Pollution regulations (2004). It is therefore used as an indicator of pilot plant performance.

3.6.1 Method

The samples for the influent flow are taken from the inlet of the main treatment plant. The effluent samples are taken from daily composite samples from the permeate from both membrane lines. Sampling was performed according to the Norwegian accredited standard (NA Dok 30) and can be described as accredited composite samples. Cuvettes used are LCK114 and LCK314 for COD, and LCK348 and LCK349 for phosphorous, in combination with Hach HT200 thermostat, and Hach DR3900 spectrophotometer for

cuvette analyses. The removal efficiency of the compounds can be calculated using equation 1.

$$Removal [\%] = \frac{concentration\ in - concentration\ out}{concentration\ in} \times 100 \quad \text{Equation 1}$$

3.6.2 Results of removal efficiency

The removal efficiencies of COD and phosphorous in the pilot plant with the initial operating conditions were well in line within the requirements of removal. The removal efficiencies were consequently above 80% and 90% of COD and total phosphorous respectively. The concentrations of COD in the influent and effluent are shown in Figure 3.13 and had an average removal of 88% during initial testing. All the measurements of effluent COD can also be observed to be below 125 mg/L. Secondly the concentrations of total phosphorous in the influent and effluent are shown in Figure 3.14. The effluent concentration is most frequently close to the limit of detection, and the average removal rate of phosphorous in the period was found to be therefore > 98%. These satisfactory results supported the decision about reducing the chemical dosage.

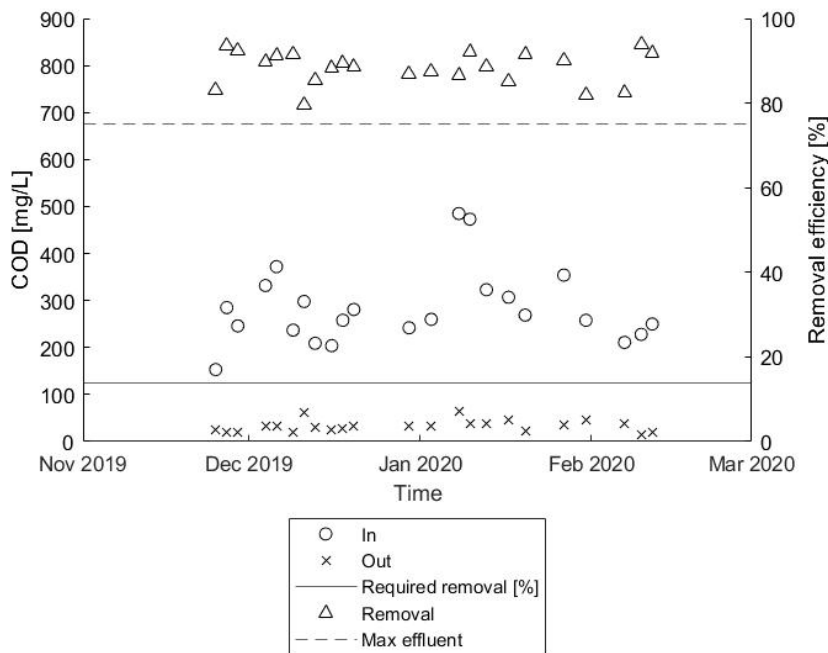


Figure 3.13 COD removal with initial chemical conditions.

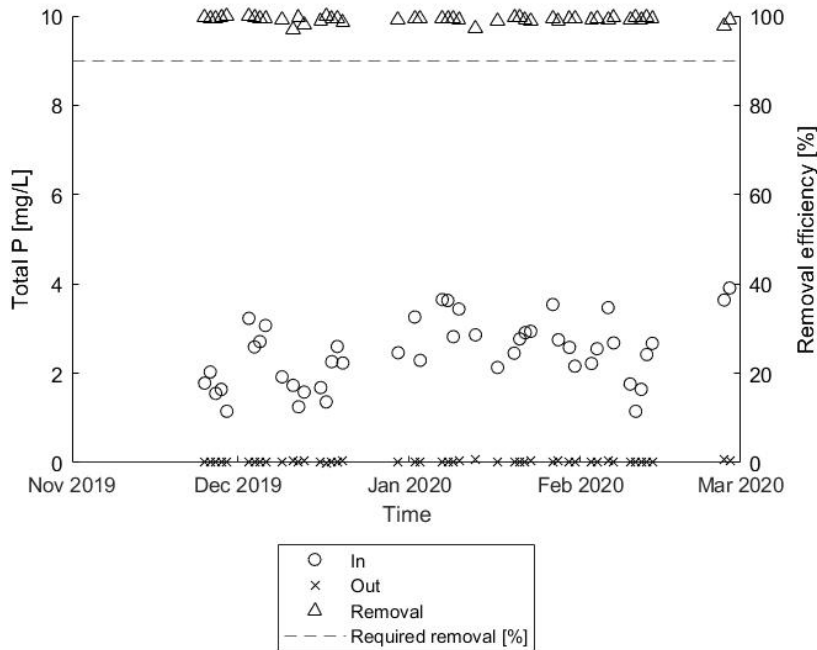


Figure 3.14 Tot P removal with initial chemical conditions.

Despite of that the DMF pilot plant was not operated for long, there were testing of influent and effluent values of COD and P for the optimised operation conditions I (Table 3.2). Regardless of the small amount of data, the results are presented graphically, to provide a similar picture as in the previous section. Influent and effluent values of COD and total phosphorous are shown in Figure 3.15 and Figure 3.16, respectively. The removal of both COD and phosphorus was found to be satisfactory during the short operation time, with average removal efficiency of 87% and 97% respectively, meaning there was no doubt of fulfilling the removal requirements.

With the small number of attained data due to the operation condition running for such a short time, it would be wrong to claim the used chemical dosage always would give satisfactory removal. It would, therefore, be necessary to run the operation condition longer. However, this was not possible due to too rapid development of TMP.

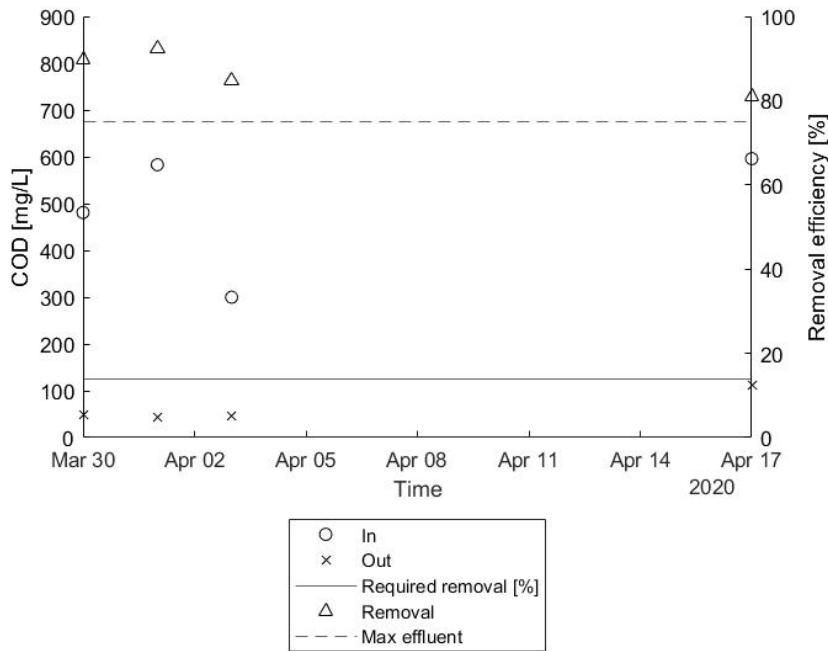


Figure 3.15 COD removal with the first change of chemical dosage.

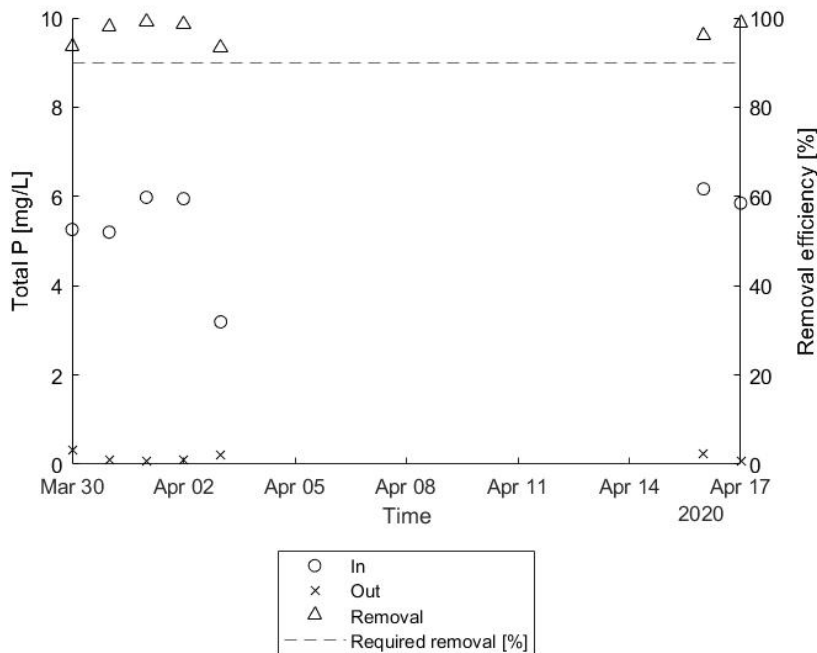


Figure 3.16 Tot P removal with the first change of chemical dosage.

One noteworthy observation is the removal of phosphorous still being adequate with influent concentrations of ca. 6 mg P/L, which is higher influent values than what observed during the previous operation condition described in the preceding section.

Having the same dosage of the precipitating agent PACl, as for the first set of operating conditions, the removal was expected to be similar in terms of phosphorous removal for the optimised operation conditions II (Table 3.2). This turned out to be the case, and yet

again, the removal was beyond satisfaction, as can be seen in Figure 3.17 for COD and in Figure 3.18 for phosphorous.

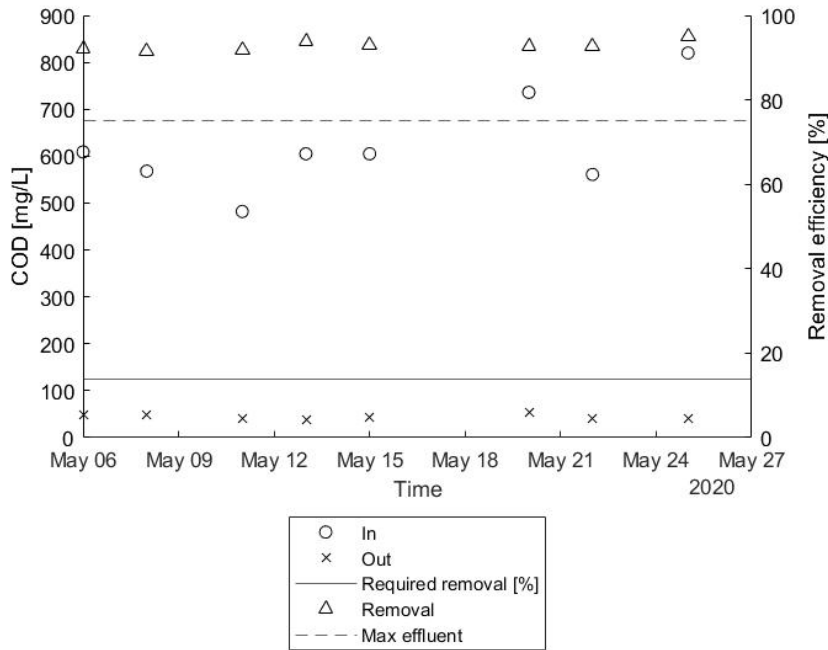


Figure 3.17 COD removal with the second change of chemical dosage.

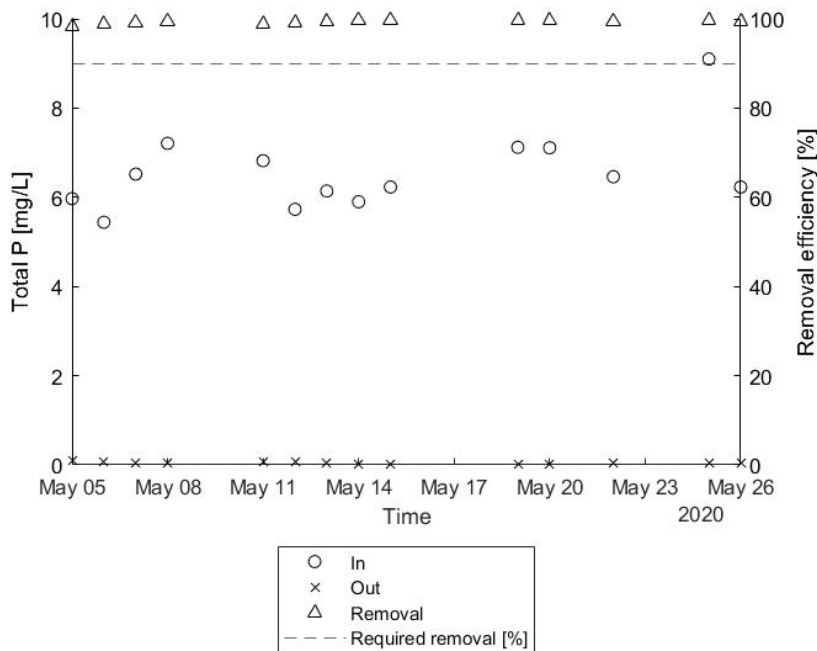


Figure 3.18 Tot P removal with the second chemical dosage.

During the entire period, the treatment requirements of COD and phosphorous were met, with both the coagulant dosages. Perhaps could an even lower concentration of PACl be used, but the time span in which the dosage was tested was not long enough to test a further reduction and see if it would meet the requirements. However, the results show potential for further reduction, considering the removal was well within the requirements for both constituents overall.

Using the chosen dosage of 9 mg Al/L, and considering an influent concentration of phosphorous of 2.9 mg/L (based on Table 4.1), a molar ratio of 3.6(Al:P) was found. This is, for example, lower than what found by (Wang et al., 2005) who found the optimum molar ratio of aluminium to phosphorous of 4.37, while Cai et al. (2020) reported success with a ratio of 2.5 in MBR effluents. The latter supports the potential for further reduction.

The chemical dosages were based on a flow of ca. 3.1 m³/h; however, by taking a closer look at the inflow, it was observed some slight fluctuation around this value. These fluctuations are generally minimal; as can be seen in Appendix C. With this slight instability, there will be some uncertainties about the chemical concentration being precisely as stated at some points. Sometimes the desired concentration of PACl and polymer is exceeded, and sometimes it is not met. This will have an impact on the removal efficiency to some extent. The measured inflow can be found in Appendix C.

It might seem like the incoming phosphorous concentration had seasonal variations, as it can be observed that the influent concentration in was consequently below ca. 4 mg P/L between November and March (Figure 3.14Figure 3.13), while in Figure 3.18 was consequently above ca. 4 mg P/L. The same can be observed for the incoming COD in Figure 3.13 and Figure 3.17, with the incoming COD concentrations in May, generally being higher than those between November and February. This may be coincidental, or it may be the case that there are seasonal variations. Either way, the pilot plant showed sufficient removal of both constituents independent of this variation.

An accredited laboratory, Eurofins in Moss, has also been analysing samples from the pilot plant. These analyses are not as frequent as the COD and phosphorous analyses described in the preceding sections. The results are presented in Table 3.10 and confirm the high removal efficiency shown in the regular day to day analyses performed in the internal lab of the wastewater treatment plant, as well as including some additional constituents.

Table 3.10 Removal rates from accredited lab analyses.

Date	P [%]	TOC [%]	DOC [%]	BOD ₅ [%]	COD [%]	N [%]	SS [%]
10.01.2020	99	81	65	93	91	0	99
08.02.2020	100	82	40	91	88	18	99
27.02.2020	98	79	46	88	86	15	97
02.04.2020	99	86	60	94	93	23	99
14.05.2020	100	89	64	95	94	26	100
Average	99	83	55	92	91	16	99

The pilot process was not designed with nitrogen removal in mind, which is reflected by the low nitrogen removal. As today, there is no knowledge about nitrogen removal requirements in the new treatment plant either.

Weekly composite sample with the removal of heavy metals was only analysed once, at the same time as the last analysis in Table 3.10, and the results can be found in Table 3.11. Comparing these to the average removal of the main wastewater treatment plant, the pilot plant has slightly higher removal for all constituents; the main treatment plant experiences an increase in the concentration of nickel.

Table 3.11 Removal of heavy metals from accredited lab analyses.

Parameter	Inlet [$\mu\text{g/L}$]	Outlet [$\mu\text{g/L}$]	Removal [%]
As	2.40	0.87	64%
Cd	0.16	0.011	93%
Pb	6.0	0.24	96%
Cr	8.3	0.68	92%
Cu	36	1.8	95%
Ni	7.6	4.2	45%
Zn	130	17	87%
Hg	0.0280	0.0025 ^a	91%

^aHalf of the lowest detection value.

3.7 Development of TMP

Transmembrane pressure can be defined as the difference between the average pressure of feed pressure and permeate pressure, effectively being the flux driving force. It is an indication of how much feed pressure is required and can be used in combination with flux to assess fouling (Water Environment Federation, 2006). When operating membranes at constant flux, as in the pilot plant, TMP is the parameter which determines how often cleaning in place (CIP) is needed and is therefore of interest to be able to evaluate performance.

3.7.1 Method

Data collected from the pilot plant was plotted graphically using MATLAB. However, if all the data, which has a frequency of one minute, were to be included, it would cause a very “noisy” picture. Therefore, a script was made to extract data to ensure that one data point is included per operation-relaxation cycle. The script was made using MATLAB, and data points had to be extracted separately from membrane 1 and membrane 2, i.e. they do not always run parallelly and will therefore not start a new membrane cycle at the same time. The script can also be used to look at the flux of the two membranes. The script used can be found in Appendix D.

Additionally, some adjustments were made manually for some cases, when variations were observed in the data. Before making the plot, it was checked that the time between the time points were about 12 minutes, as that would be expected with the operation-relaxation cycle lasts 10+2 minutes. The data sets for the different cycles are starting as the operation is stable, i.e. has the mentioned operation-relaxation cycle.

The TMP data in the data set is based on the measurements and is not standardised for temperature (15°C). This choice was made because the operation is the main focus of this section, and the point where CIP is needed depends on the actual TMP, not the TMP corrected for temperature.

TMP development can be calculated by using equation 2.

$$r_f = \frac{dTMP}{dT} \quad \text{Equation 2}$$

This equation can illustrate that the cake fouling is proportional to the slope of TMP against time (Diaz et al., 2012). The purpose of this concept in the following sections is to illustrate how steep the TMP development of the different cases is, besides, to give an idea of the theoretical frequency of the chemical cleaning in place.

3.7.2 Results and discussion of the development of TMP

There were three cycles of operation with the initial dosage, which are shown in Figure 3.19, Figure 3.20, and Figure 3.21. For all cases, it was observed that the TMP develops differently across the two membranes. Membrane 2 seems to have more operational problems maintaining a stable increase in the vacuum pressure compared to membrane 1, although line 1 was not entirely stable either. It was also observed that the gradient of membrane line 2 seems to be consequently slightly steeper than the gradient of membrane line 1.

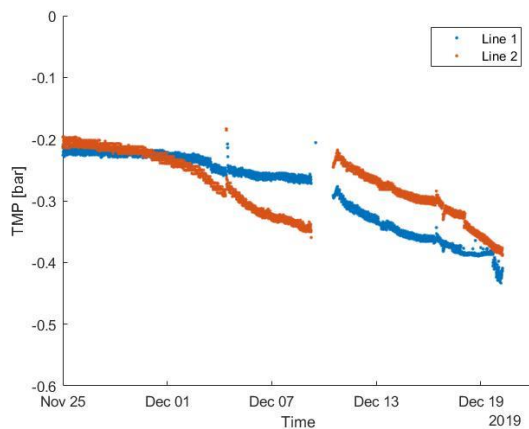


Figure 3.19 TMP development in round 1 of initial operation conditions.

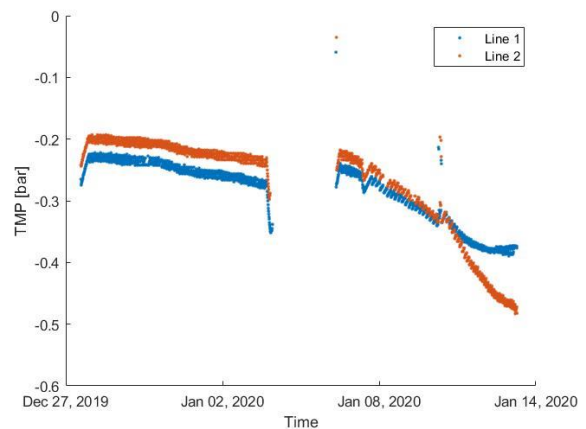


Figure 3.20 TMP development in round 2 of initial operation conditions.

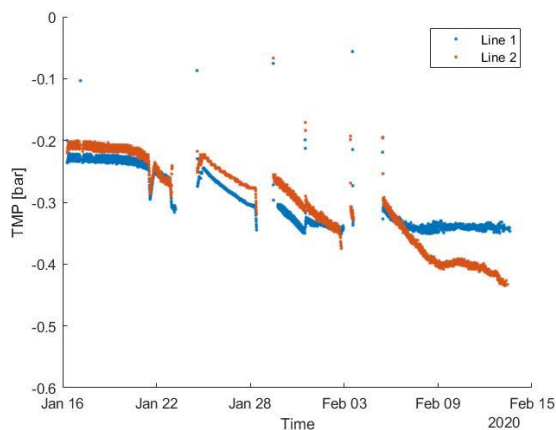


Figure 3.21 TMP development in round 3 of initial operation conditions.

The linear average gradients for the two membrane lines for the three operation rounds are shown in Table 3.12. For all the cycles, membrane line 2 seems to have a significantly larger gradient. Both the lines have varying gradient magnitude for the three cycles. The linear gradients are not a fully accurate representation of the TMP variation as the change of TMP is not of linear nature. However, they provide useful insights into the differences between the two membrane lines to supply the visual representation.

The jump observed in figure Figure 3.19 for membrane line 2 was due to testing physical backflushing, i.e. not CIP. Although the vacuum pressure did decrease, it reached the level of membrane line 1. Between the three figures, there has been performed CIP.

A steeper gradient, as observed for membrane 2 will initiate a more frequent cleaning in place. This could be seen during operation with the initial conditions, where the CIP was triggered by membrane 2 on several occasions, while membrane 1 would still be able to run for some time before reaching -0.5 bar which is the point where CIP is to be performed. This is illustrated in Table 3.12, where in one column, it is calculated how many days it would take to reach -0.5 bar for both of the membranes with the existing data. This is purely theoretical, and just to provide a more relatable number for the gradient and quantifies the variation in the membrane lines as well as the variation over time. These results, and similar results in the following sections, are valid for this study but do not provide a definite answer to reality.

Table 3.12 Gradient of TMP against time.

Line	Start time	End time	Start TMP [bar]	End TMP [bar]	Gradient [bar/d]	Time to CIP [d] ^a
1	25.11.2019 00:45	20.12.2019 06:13	-0.22	-0.41	-0.0075	40
2	25.11.2019 00:45	09.12.2019 06:28	-0.21	-0.36	-0.011	28
2	10.12.2019 13:05	20.12.2019 06:13	-0.24	-0.39	-0.015	20
1	27.12.2019 22:09	13.01.2020 06:35	-0.24	-0.38	-0.0086	35
2	27.12.2019 22:10	13.01.2020 06:40	-0.20	-0.48	-0.017	18
1	16.01.2020 07:28	13.02.2020 14:10	-0.23	-0.35	-0.0040	74
2	16.01.2020 07:37	13.02.2020 10:37	-0.21	-0.43	-0.0078	38

^aIt has been a slight variation on what the TMP returns to. Therefore, a change of 0.3 bars, e.g. from -0.2 to -0.5 bar, was chosen as an example.

The gradient for line 2 in the time span corresponding to Figure 3.19 was split in two to see the effect of the backflushing. As already seen in Figure 3.19, the gradient became steeper, and the TMP of line 2 caught up with line 1. After this run, the pilot plant was stopped due to holidays, meaning the pressure did not initiate the CIP.

With the new chemical dosage, operational personnel at the pilot plant quickly reported the membranes seeming to foul more easily than before. This is illustrated in Figure 3.22, where a sudden change in TMP can be observed on membrane line 2 in the late hours of 1st of April, with the gradient before that also having a steady development towards -0.5 bar, with the development appearing much steeper than the TMP development of membrane line 1. As for the first operation condition, membrane line 1 seems to have an overall more stable development of the TMP.

The operation cycle in Figure 3.22 did in reality end on the 2nd of April 2020, but the membranes restarted for some time again the following day, before stopping the pilot plant completely for another two weeks.

In the time between Figure 3.22 and Figure 3.23, there was no chemical cleaning, only air scouring, but they were plotted separately for the sake of clarity. By judging from the figures, a fair fraction of the permeability was restored during this time. However, both membrane lines experience a rapid change right after starting the run again before somewhat stabilising, as seen in Figure 3.23. This change would perhaps not be existent with chemical cleaning between the two operation runs.

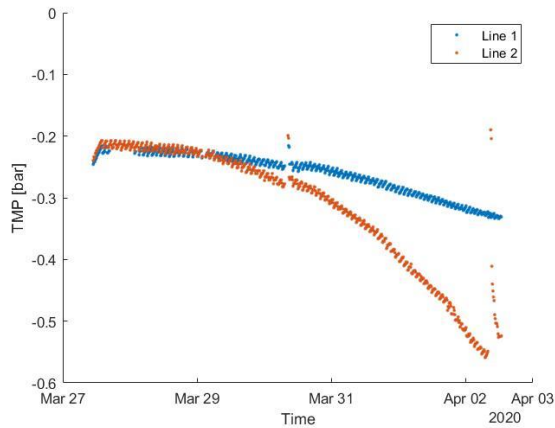


Figure 3.22 TMP development round 4 (operation condition I).

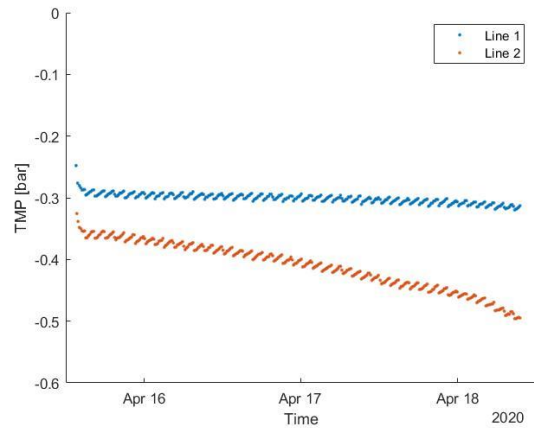


Figure 3.23 TMP development round 5 (operation condition I).

The linear gradient of TMP against time is shown in Table 3.13. Line 2 has, as expected when looking at Figure 3.22 and Figure 3.23, a steeper gradient than membrane line 1. Looking at the transmembrane pressure as well as the gradient, if both membrane lines behaved like line 1, the operation would be better off in terms of the chemical cleaning frequency. However, despite showing a steadier development, just as in the previous section where the TMP development of initial operation conditions was described, neither of the membrane lines have a predictable, e.g. the same gradient.

Table 3.13 Gradient of TMP against time for optimised operation conditions I.

Line	Start time	End time	Start TMP [bar]	End TMP [bar]	Gradient [bar/d]	Time to CIP [d] ^a
1	27.03.2020 13:17	02.04.2020 12:43	-0.22	-0.33	-0.018	16
2	27.03.2020 13:17	02.04.2020 12:45	-0.21	-0.52	-0.052	5.8
1	15.04.2020 15:05	18.04:2020 09.22	-0.30	-0.32	-0.0071	42
2	15.04.2020 15.05	18.04.2020 09.22	-0.37	-0.49	-0.047	6.4

^aIt has been a slight variation on what the TMP returns to. Therefore, a change of 0.3 bars, e.g. from -0.2 to -0.5 bar, was chosen as an example.

The last run of the pilot plant, operational conditions II (Table 3.2), was with the same concentration of PACl but a lower concentration of polymer than the previous one, and can be seen in Figure 3.24. Unlike the other runs, this one was not operated until -0.5 bar before stopping, due to increasing frequency of issues with the inlet pump. Therefore, the development of the TMP in a full run could not be observed for this chemical dosage. Comparing this development to that of the previous runs, it seems like the pilot plant would be able to run for a longer time without any CIP. It can also be noted that the development of the TMP seems more linear, especially in the longest operation interval during the run.

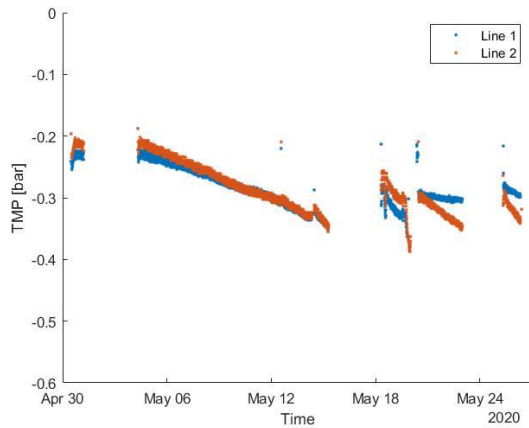


Figure 3.24 TMP development round 6 (Operation condition II).

Estimation of the gradient in this run was based on the entire experiment, like the other experiments described above, and can be seen in Table 3.14. It is observed that the gradient is low compared to the gradient of the other operations, indicating a slow development towards -0.5 bar.

Table 3.14 Gradient of TMP for the second round of optimised conditions.

Line	Start time	End time	Start TMP [bar]	End TMP [bar]	Gradient [bar/d]	Time to CIP [d] ^a
1	04.05.2020 08:28	26.05.2020 07:59	-0.23	-0.30	-0.0029	105
2	04.05.2020 08:18	26.05.2020 09:45	-0.21	-0.32	-0.0049	61

^aIt has been a slight variation on what the TMP returns to. Therefore, a change of 0.3 bars, e.g. from -0.2 to -0.5 bar, was chosen as an example.

For the last run especially, it would be of interest to operate for a longer time, at least until the TMP reached -0.5 bars, considering the other TMP developments seemed to change their shape going towards the end of their respective runs. Additionally, it would be advantageous to observe whether the results were just a one-time result, especially considering that operating the initial conditions for three runs gave three different results.

TMP was found to be varying between the two membrane lines as well as between the different rounds of operation. One rather evident reason for the latter is the fact that the characteristics of the incoming wastewater will vary over time. Therefore, there will also be a variation in the wastewater entering the membrane tank from the microsieve filtrate, and the membranes will experience variable strain from the pollutants, resulting in a variable degree of fouling. Hence a TMP development to maintain the stable flux which will not be the same for the different operation cycles. This was the case for all operating conditions. This will also add to the uncertainty of whether it was the set chemical dosing, or other factors making the TMP develop as it did.

Due to the continuous difference in the performance of the two membrane lines, it was investigated whether it was something wrong the air scouring. It was checked if there was a difference in air scouring of membrane line 2 as compared to the air scouring of membrane line 1, i.e. if it was unevenly distributed however no such matter could be observed.

The membranes were operated at constant flux overall, but the two membranes were not operated at the same flux all the time. Membrane line 2, which is the one having the most rapid increase in the vacuum pressure, is also the line with the highest flux and hence the highest permeate production. The varying flux is likely to be correlated with TMP development. Between 25th of November 2019 and 26th of May 2020, the entire period looked at in this thesis, membrane line 1 had a permeate production of 2505 m³ and membrane line 2 had a permeate production of 2970 m³, meaning membrane line 2 has had an 18.6% higher permeate production than membrane line 1. Looking at specific cases and time ranges, the same pattern is observed, with line 2 having a higher permeate production.

It was investigated whether an equal amount of permeate was produced during each operation round, i.e. if the TMP reached -0.5 bar when a given amount of permeate had gone through the membrane. This turned out not to be the case, and the permeate production for each operation round was differing within a big range. For the cycle in Figure 3.22, 95.1 m³ and 127.5 m³ of permeate was produced by line 1 and 2 respectively, while for the cycle in Figure 3.21, 594.8 m³ was produced by line 1 and 693.9 m³ by line 2.

Looking more into the different operation rounds, a similar pattern was observed, as expected, with membrane line 2 having a higher production in all cases, as well as the highest vacuum pressure at the end of the cycle. It was investigated if the TMP for membrane line 2 was like at the end TMP for membrane line 1, at the permeate production at which membrane line 1 ended. This was not found to be the case for any of the runs.

Although the pilot plant was intended to be operated at constant flux, this turned out not to be the case. The flux was declining over time, but it seemed to be partly restored after chemical cleaning. This is likely to do with the permeate pumps misbehaving. With a constant flux, the TMP would likely develop towards -0.5 bar more rapidly. As suggested, line 2 has a higher permeate production, and hence flux, most of the time. It was observed that the flux is declining over both lines. The total flux of the system was not plotted, as the two membranes do not always have their relaxation at the same time, causing the flux to vary even more. When operating a process with no requirements about permeate production, like a drinking water treatment plant would have, this is not a significant concern, and is not looked further into in this thesis. Operational personnel reports that the reason for this instability is likely to do with the permeate pumps (Jørgen Karlsen, 2020, personal communication, 29.05). Figures showing flux development can be found in Appendix C.

Another reason for the instability could be temperature variation, as the TMP has not been normalised with respect to temperature. The TMP is proportional to the kinematic viscosity of the water, which means the TMP is inversely proportional to temperature. Therefore, increased temperature leads to an increased flux and therefore a decreased TMP, i.e. a lower driving force required to get the water through the membranes (van Nieuwenhuijzen, 2002). During the period of these measurements, the highest and lowest measured temperature in the membrane tank was 15.9°C and 3.0°C respectively, with an average of 9.7°C. The lowest temperature seems to be a one-off deviation, and only lasted for 1 minute, and the second-lowest was 5.2°C which seems to be a more representative value for the lower end. The temperature development over the time of pilot plant operation can be found in Appendix C.

It can be noted that the TMP goes slightly up, i.e. the vacuum pressure goes down when the pilot plant has not been running. As aeration has been going on as usual despite this, the scouring effect did continue, and some physical fouling is likely to have gone off, allowing a lower driving force, i.e. TMP, when the pilot plant starts again. It is not of interest to start and stop the plant like this, but it is useful knowledge that a stop does not have a negative impact on the TMP. It also shows that the air scouring is effective as a fouling mitigation measure.

When calculating the gradients, different choices could be made regarding the time to start and stop. Looking at all cases, a jump in TMP can be observed at the start, although in some cases more significant than others, and for one of the cases (Figure 3.23) the jump was the opposite way. It was chosen to start the gradient calculation after the lines had stabilised again. The TMP graphs were started when a stable flux was reached, based on the aforementioned extraction script in Appendix D, but it may seem like the membranes take some more time to get a stabilised TMP.

Another point, which also makes the gradients more uncertain is the operational breaks due to the issues with the inlet pump, making some gradients lower. This means the time to reach the chosen change in TMP of 0.3 bar will be, perhaps, unrealistically high, like for example case 3, Figure 3.21, which has several operational breaks, making it a longer cycle. A similar observation can be made in Figure 3.24, and Table 3.14, where the time to change 0.3 bar is probably unrealistically long. For this run, by instead looking at the longest interval with constant operation, the gradient looks different, and is perhaps more realistic, with 29 days and 24 days for line 1 and 2, respectively. This resembles the observations from run 1 and 2.

Secondly, the shape of the TMP curves is slightly varying between the two membrane lines in some cases, with membrane line 2 having a more concave decrease. This would perhaps be observed for membrane line 1 as well if the membrane line was operated over a longer time. The shape differences, particularly visible in Figure 3.20 and Figure 3.22, means the gradients of differently shaped TMP developments are calculated. Due to these facts, combined with the previous paragraph, and the fact that the gradient is calculated linearly, it can be concluded that the gradients are not accurate, but still provides a guideline which is their intended use.

During the operation of the pilot plant, there has been repeated trouble with the inlet pump (PI_PU01 in the P&ID chart in Figure 3.2) as it may seem the pump has been overloaded, causing it to stop. This has caused the overall operation to be less stable throughout the research period, which is not ideal but is one of those obstacles one would have to expect when running a pilot plant.

Lastly, the extraction of data turned out to not always be as precise as desired, with the membrane lines ending up with a slightly different number of data points. Despite this, the extraction presents the data more neatly than no extraction at all and makes the graphs easier to interpret than what they otherwise would do. Therefore, for its purpose, it is considered sufficient.

3.8 Discussion of feasibility of direct membrane filtration

Removal of phosphorous and COD in the pilot study has proven to be well within the new requirements, indicating the potential of decreasing the chemical dosing even further. The removal has been good even with a variety concentrations of phosphorous and COD in the influent of the plant. Similarly, removal efficiencies of other constituents like SS, BOD and TOC, tested on a more seldom basis have also shown to be more than adequate.

One issue that has been observed throughout the study when looking at the development of transmembrane pressure is the difference between line 1 and line 2. If the pilot plant were running with only one line, with similar behaviour to membrane line 1, the demand for chemical cleaning would be lower, and it would be possible to keep steady operation for a longer time. The first dosage optimisation would potentially also be more successful, as the operation stop was triggered by the TMP of membrane line 2, with membrane line 1 having more to go on.

The removal rates, as well as the operation, would be advantageous to monitor over a more extended period, to find out the impact of variations throughout a whole year. This would make the seasonal impact clearer. However, the time from the end of November to the end of May provided different temperatures and precipitation both as snow and rain. Based on the existing results, there are seemingly some seasonal variations. As discussed in section 3.7.2, there are several potential explanations of the instability and unpredictability of the TMP of the membranes, like variations in temperature, incoming wastewater characteristics, and varying flux. It does seem unclear how significant the impact of the dosage of the precipitation metal is on the TMP development, compared to the other potential explanations. This causes some uncertainty, and with TMP being an essential parameter for operation, it adds to the reasons that it would be of interest to operate the pilot plant over a longer time. This instability and somewhat unpredictable nature of the pilot plant behaviour is a disadvantage, but it is assumed to be partly due to faulty components like the permeate pumps. A full-scale treatment plant acting this unstable would not be beneficial. However, the transferability from a pilot to a full-scale treatment plant is not guaranteed.

While the pH was measured when conducting jar tests, it could also be advantageous to have a more extensive look into its impact on the removal results and chemical dosage. As the pH was measured to be above 6, which is what is optimum for an aluminium precipitant (Duan and Gregory, 2003), pH correction could be considered. However, it was not observed any issues with phosphorous removal being too inefficient, which is the reason for not taking pH correction further into account in this thesis.

CIP of the two membranes has been performed simultaneously as they are submerged in the same tank. CIP takes approximately one working day. Operational personnel reported that a slight deviation from the original procedure seemed to be more beneficial; letting the membranes be submerged in citric acid for one day, and then hydrogen peroxide for one day (Raymond Jørgensen, 2020, personal communication, 26.05). This made the CIP procedure take three days in practice.

Initially, it was planned to test the biomethane potential (BMP) of the sludge from the pilot plant. However, the existing equipment at FREVAR turned out to be too old and malfunctioning, and the supplier did not perform repairs on the given type anymore. Therefore, with the lack of sufficient equipment, this was not looked into any further. The

sludge sampling did, however, show that most of the sludge from the microsieve was of organic origin. In contrast, the sludge from the membranes had a fraction of organic solids which was slightly lower. It would be beneficial to have more extensive sampling when it comes to testing the sludge in terms of TS and VS, given the low number of samples taken. It is not a certainty that the existing samples are representable for the general condition of the sludge, although it is assumed for the sake of this thesis and the associated calculations.

Operational personnel have reported few problems with the pilot plant overall. There were some issues with one particular pump, which does not have a direct connection to how the treatment process works. The pump has caused the pilot plant to be out of operation occasionally, fatigue being the likely reason. Worn impellers and holes appeared on the housing when the end of the overall operation was approaching, making patching necessary (Raymond Jørgensen, 2020, personal communication, 26.05).

This pilot plant study has achieved results which have been discussed. Therefore, the question remains whether these results give a good representation of reality. A pilot plant will not necessarily behave precisely similar to a full-scale plant, and an optimised pilot plant could not inevitably transfer the given conditions to achieve great results. However, it shows the potential of the concept.

3.8.1 Conclusion of feasibility

The results from the pilot plant study have shown that high removal of COD and phosphorous can be achieved without using a biological process. DMF is not designed with any respect to biology; however, some biological processes can potentially happen in the membrane tank, as it is a submerged an aerated reactor, resembling an MBR in this matter. This was not investigated any more. Furthermore, DMF showed excellent removal efficiency in terms of heavy metals. There are no requirements for this, but it is not unlikely that the requirements will be stricter in the future.

DMF proved sufficient in terms of removal and shown the capability of being operationally feasible for some time in this study. Fouling is frequently described as the main problem using membrane filtration and was also present in this study. However, it has been shown that with some adjustment to the chemical dosage, combined with continuous aeration and cycles consisting of relaxation every 10 minutes, the pilot plant could be operated with satisfaction. It would be beneficial to run the study over a longer time, keeping the same operation conditions to confirm the findings of this study.

In order to compare DMF with other relevant processes, a full-scale design of MBBR-dissolved air flotation (DAF) with chemical precipitation of phosphorous was also designed. This was being considered a viable alternative to achieve similar results, like obtaining sufficient removal of the relevant compounds and being space efficient.

4 Design of full-scale treatment process

It has been decided that MBBR and DMF are the two treatment processes to be considered, and these will be designed for a new full-scale FREVAR plant in this chapter. As a part of the design, theoretical sludge production is also calculated to later give an estimate on biogas production when calculating mass balances for both designs, which are calculated with the purpose of an extended understanding of the processes. The purpose of the design is not to provide a detailed solution of two full-scale treatment plants, but to give an idea of how the processes could look. Options and choices will be discussed along with the design of the processes and their subprocesses.

There are ongoing discussions of including neighbouring municipalities in the new FREVAR treatment plant; however, this will not be taken into consideration in this design. The current treatment plant is a flocculation-sedimentation process, using ferric chloride for precipitation.

4.1 New treatment requirements

Within 31.12.2020 the municipality of Fredrikstad is required to provide the county governor of Oslo and Viken (Former: The county governor of Østfold) a new treatment process solution that can meet secondary treatment requirements and requirements about phosphorous removal (County governor of Oslo and Viken, 2019). The secondary treatment requirements involve as follows: reduction of BOD₅ to either 70% of influent concentration, or to 25 mg O₂/L, and reduction of COD to either 75% of influent concentration, or to 125 mg O₂/L. The phosphorous removal requirements are 90% removal of influent concentration. These requirements are found in Regulations relating to pollution control (Norwegian title: Forskrift om begrensning av forurensning) part 4, wastewater, §14.2 (Pollution regulations, 2004).

The treated wastewater will go out through the current outlet, which is located in the Glomma river, flowing towards the outer Oslofjord, which is defined as a sensitive recipient (Pollution regulations, 2004). As the pollution in the Oslofjord mostly is found to be related to eutrophication because of phosphorus (Ødegaard et al., 2014), implementation of stricter requirements, the pollution in the receiving area can be expected to be reduced in terms of environmental threat (County governor of Oslo and Viken, 2019).

Unless stated otherwise, the Norwegian guidance manual for process design of wastewater (Ødegaard et al., 2009) is the basis for the process design. A new and revised version of these guidelines are expected to be published this year (Skaug, 2020), but was not published at the time of writing this thesis. The new guidelines could provide new insight into some of the design considerations made in the design in this thesis. This makes some of the designs object to change.

4.2 Design criteria

The design criteria for the full-scale treatment process is based on historical data given by FREVAR. The values are found in accordance with Norwegian guidance manual for process design of wastewater treatment (Ødegaard et al., 2009).

The graphs in Figure 4.1 are based on a series of flow measurements from between 01.01.2019 and 31.12.2019. The values were given with a frequency of one hour. In accordance with the Norwegian guidance manual for process design of wastewater treatment, one of the curves represents all hours in the measurement series. The other curve represents the hour with the highest discharge for all the days in the measurement period. Following the guidelines, a design value of Q_{dim} can be found using the median of the $Q_{maxhour}$ curve, i.e. the 50% percentile. The Q_{maxdim} value is found based on how much of the flow to be treated. In this case, a 95% percentile is chosen. These are without the addition of reject water, a decision which was made as the design is mainly for comparative purposes.

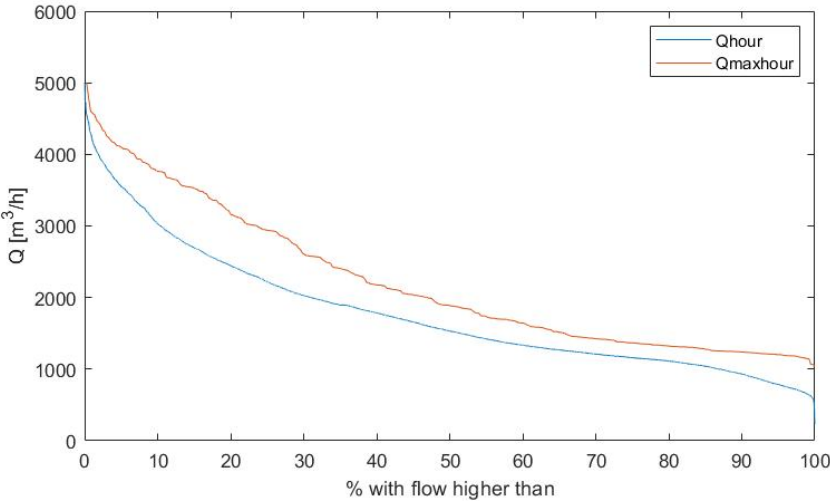


Figure 4.1 Duration curve of Q_{hour} and $Q_{maxhour}$.

The design loads are based on analysis done by accredited laboratories in the period 2017 to 2019. The 85% percentile has been chosen based on being held forward as an example in the Norwegian guidance manual for process design of wastewater treatment. Duration curve for BOD_5 can be found in Figure 4.2, and the duration curve for phosphorous can be found in Figure 4.3.

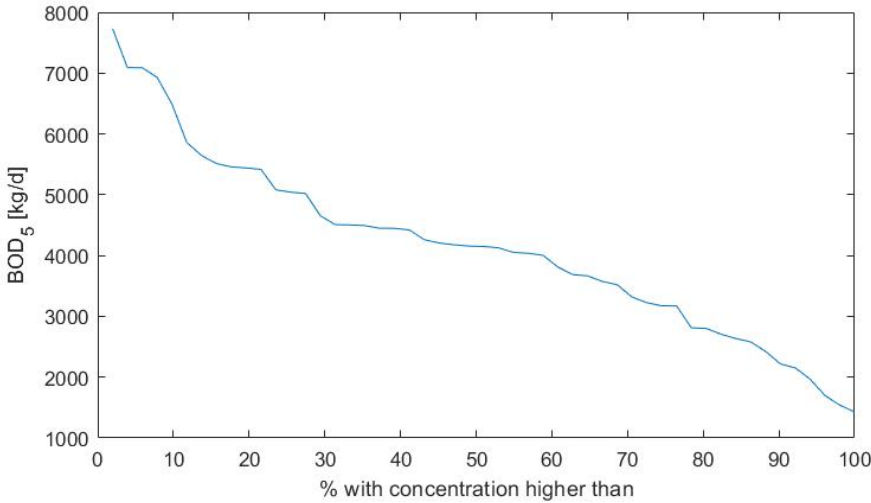


Figure 4.2 Duration curve of BOD load.

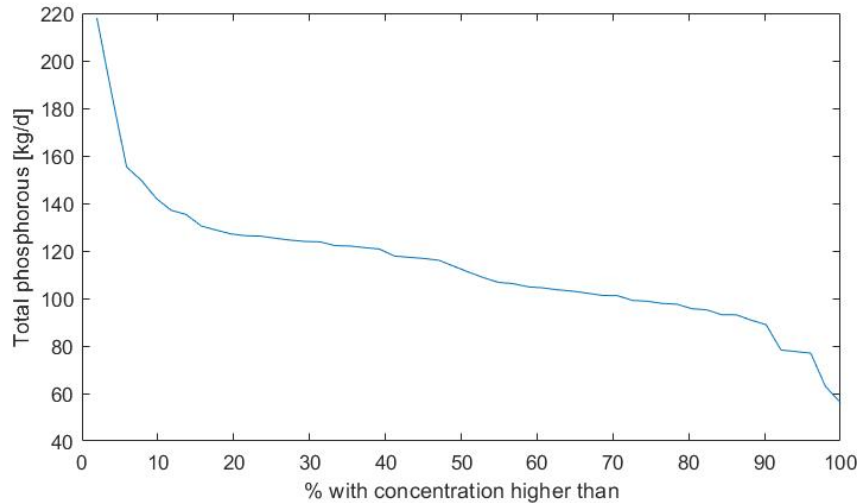


Figure 4.3 Duration curve of phosphorous load.

The findings from the duration curves are found in Table 4.1, together with the chosen design temperature, the max temperature based on measurements from 2019, and the calculated person equivalents. The data has been provided by personnel at FREVAR.

Table 4.1 Design values for a new treatment plant

Parameter	Magnitude	Unit
Q_{dim}	1884	m^3/h
Q_{maxdim}	3552	m^3/h
Q_{avg}	1779	m^3/h
BOD ₅	5538	kg/d
P	131	kg/d
Design temperature	8	°C
Max temperature	19	°C
PE	92300	-

Q_{dim} and Q_{maxdim} from the design of the current plant back in 1989 was found to be 1800 m^3/h and 3600 m^3/h , respectively. With the increase in population, it would be natural to expect an increase in the flow pattern. One possible explanation of the higher Q_{maxdim} 30 years ago can be the fact that only one year of data was used to find the current design values. It may appear that the old values were based on estimates and not measurements considering Q_{maxdim} being exactly twice the size of Q_{dim} . However, according to Ødegaard et al. (2009), the time frame used for the new calculation is sufficient. A likely reason for the change being so small is the decreasing fraction of combined sewer systems in Norway (Lien, 2013). An increased fraction of separate sewers will lead to less stormwater entering the treatment plant, and the inflow will by that naturally decrease. The new design values will be used in this thesis.

4.3 Design of pretreatment

The MBBR and DMF process will have the same pre-treatment with sand- and grease trap. In Norway, aerated sandtraps are the most common ones (Ødegaard et al., 2009). The existing pre-treatment and sedimentation basins are presented in Table 4.2, together with their performance with the new Q_{dim} and Q_{maxdim} . This shows that there is a

potential for reuse when looking at design guidelines. This would not only cut down on capital expenses, but also reduce the environmental footprint of demolition, building, and use of new materials, as described in section 2.1.

Table 4.2 Capacities of existing units at FREVAR treatment plant

Process step	Number of units	Capacity (Per unit)	HRT at Q_{dim} [min]	HRT at Q_{maxdim} [min]	O.R. at Q_{dim} [m/h]	O.R. at Q_{maxdim} [m/h]
Sand- and grease trap	4	765 m ³ (191.25 m ³)	24	13	-	-
Screens	2	4680 m ³ /h (2340 m ³ /h)	-	-	-	-
Flocculation chambers	10x2	650 m ³ (32.5 m ³)	21	18	-	-
Sedimentation basins	5	1590 m ² (318 m ²)	177	94	1.2	2.2

O.R. = overflow rate

The current treatment plant has as mentioned above sedimentation basins. These have the potential of being retrofitted as a pre-treatment step for an MBBR-DAF process, meeting the requirements in terms of overflow rate at both Q_{dim} and Q_{maxdim} , which should not be higher than 2.4 m/h and 4.8 m/h respectively (Ødegaard et al., 2009). This will not only reduce the investment cost of the new treatment process but can also reduce the load on the following treatment step. For an MBBR-DAF step, this decrease in pollution load would reduce the needed biofilm area and hence the volume, footprint, and air demand, as well as avoiding clogging of bioreactor sieves (Ødegaard et al., 2004).

For a DMF step, the volume demand with pre settling would be the same due to it being designed by flow and not load. However, it could have benefits in terms of less fouling due to fewer particles entering, which could reduce fouling and hence increase the time between each CIP (Nascimento et al., 2018). It is beyond the scope of this thesis to test the impact this would have and has therefore not been quantified. Pre-settled wastewater has been investigated as influent for DMF successfully by several researchers (Bendick et al., 2005, Diaz et al., 2012). In case of implementing a settling step, additional separation through microsieves would likely be excess. This would give the process a different structure than the pilot plant, with pre-treatment followed by chemical precipitation, before the settling and microfiltration as the final separation step.

Another option for reuse of the sedimentation basins in a DMF process is to use them as equalisation tanks to retain peak flows. This is particularly of interest for a process operated at a constant flow, like DMF. Determination of volume requirement of an equalisation basin can be done graphically by plotting cumulative inflow versus time of day, as described by Tchobanoglous et al. (2014). By using this method, an equalisation volume of 1600 m³ was found to be necessary by using the inflow data from 2019. The diagram used is shown in Appendix E. However, this basin should be of larger volume in practice than theoretically, an additional volume will vary between 10-20% (Tchobanoglous et al., 2014), resulting in a needed volume between 1760 and 1920 m³. By using this value, the volume of the existing sedimentation basins can be sufficient as a flow equalisation volume.

A third option for using the sedimentation as a part of the full-scale DMF process is to use it as a form of stormwater treatment. This would imply to run the DMF only when the incoming flow is below or at 1884 m³/h and to open the existing sedimentation basins and combine them with flocculators when the incoming flow is above 1884 m³/h. When the inflow is bigger than 1884 m³/h this is expected to be more diluted as one of the reasons for the increased flow is precipitation. Based on given data from FREVAR, the average removal efficiency in 2019 of phosphorous and COD was 82% and 66% respectively. By combining these facts, it can be concluded with this being a satisfactory solution.

4.4 Design of direct membrane filtration

In this design, the choice was made to use the existing sedimentation basins as stormwater treatment, rather than pre-settling or equalisation. The design resembles the pilot plant the most, and with the incoming wastewater composition and characteristics, the pilot plant has been stable in terms of removal efficiency. In this study, the combination of microsieving and membrane filtration has been confirmed to be sufficient for this case, which the combination of settling and membrane filtration has not.

4.4.1 Coagulation and flocculation

It has been observed that chemical addition directly to mixed reactors is less efficient than addition to pipe mixing units, due to short-circuiting streams often present in the mixed reactors. However, in wastewater treatment processes, using pipe mixing units may result in clogging, and chemical addition in an overflow channel is more common in Norwegian treatment plants. In most cases, a narrowing pipe will also be a valid option (Ødegaard et al., 2014). Based on these considerations, an arrangement with the wastewater going through a chamber with rapid mixers is chosen for dosing of PACl.

Generally, when designing the flocculation step, it is necessary to consider the number of chambers. In order to achieve plug flow, it is necessary with at least two chambers in series (Ødegaard et al., 2009), with more chambers making the conditions more plug flow-like. Furthermore, a higher number of chambers will also lead to a lower overall footprint of the process step due to the required hydraulic retention time being shorter. However, increasing the number of chambers to more than four will have an insignificant effect. Despite these advantages of four chambers compared to two, both the capital and operational expenses will be lowered using two chambers, as it will require less electricity with fewer mixers, which will also be a smaller investment. This point will also be viable for design of flocculators for MBBR in section 4.5.2.

Based on this, as well as the fact that the pilot plant gave satisfactory results using two flocculation reactors, two reactors were chosen. The volume is designed based on the hydraulic retention time. The design can be seen in Table 4.3.

Table 4.3 Design of flocculation for DMF.

	Magnitude	Unit
Total retention time at Q_{dim}	15	min
Number of lines	4	-
Number of chambers per line	2	-
Needed volume	471	m ³
Length of chamber	3	m
Width of chamber	5	m
Depth of chamber	4	m
Total volume	480	m ³

The chemical dosage was based on results from the pilot plant, where the chosen dosage was the one which gave the best operational results, and they also met the removal requirements with good margins. This could potentially be prone to reduction, but this was not tested, and the dosage used with success was therefore chosen; 100 g PAX-18/m³ combined with 1.5 g polymer/m³.

4.4.2 Microsieving

According to Ljunggren et al. (2007), using a disc filter can give similar removal rates as using a drum filter with a prior chemical addition. A footprint of 1 m² corresponds to approximately 1.5 m² drum filter area, but as much as 7 m² disc filter area (Ljunggren et al., 2007). From email correspondence with Nordic Water (André Østgård, 2020, personal communication, 19.05), it was informed that their largest freestanding drum filter has a filtration area of 10.4 m² and a footprint of approximately 5.2 m². This corresponds to 2 and 10.2 m² filter area/m² footprint. This is an example, and it was informed that this varies between the different filters.

In this case, rotating drum filters were chosen for the microsieving process, similar to what of the pilot plant. Choosing filter type depends on the incoming wastewater characteristics in terms of concentrations and volume, and the filter types have differences in operation (André Østgård, 2020, Nordic Water, personal email communication, 19.05). The aforementioned drum sieve with a filtration area of 10.4 m² was chosen.

When designing the microsieving step, the filtration rate determines the sieve area. In the full-scale design, the filtration rate will be higher than in the pilot plant. The pilot plant was designed for a filtration rate of 5 m/h but was running at approximately 4 m/h for most of the time. In terms of economy, this does require more filters, which leads to not just a higher capital expense, but also a higher space and electricity demand. In the full-scale design, a filtration rate of 20 m/h was chosen, based on the study by (Ljunggren et al., 2007), where SS was reduced by 90%, phosphorous by 70%, and COD and BOD by 60% at filtration rates between 10 and 30 m/h, where the design in this thesis will use the middle value between these filtration rates.

The chosen design for the microsieving step of the DMF process is presented in Table 4.4.

Table 4.4 Design of microsieving for DMF.

	Magnitude	Unit
Filtration rate	20	m ³ /m ² h (m/h)
Needed filtration area	94.2	m ²
Area of one drum filter	10.4	m ²
Number of filters	10	-
Chosen filtration area	104	m ²
Filtration rate at Q _{dim}	18.1	m ³ /m ² h (m/h)

4.4.3 Microfiltration

To match the pilot plant, the same type of membranes were used for the design; however, a module with a bigger area was chosen in order to reduce the number of membrane modules needed. A module used by Alfa Laval for large scale projects, MFM320 which is made up from MFM080 and MFM240 stacked on top of each other (Emmanuel Joncquez, 2020, Alfa Laval, personal email communication, 07.05), was chosen. This module has a total area of 515 m² according to the datasheet of MFM080 and MFM240 (Alfa Laval, 2020). Alternatively, it could be beneficial to consider other membrane types, for example, receptive for a higher flux in order to reduce the footprint and number of modules needed. Another membrane type will most likely need to be designed differently concerning, i.e. air demand.

Further, constant flux was assumed for the design of the membrane filtration step due to the pilot plant also being based on this assumption. The flux determines the needed membrane area. Although intended to be operated at constant flux, the pilot plant turned out to have a flux which was decreasing over time. This observation would perhaps imply that the full-scale design should be based on a varying flux or a lower flux. However, it is assumed that the inconsistency in the pilot plant flux is due to permeate pump issues or controlling of permeate pumps, which would be expected to be in order in a full-scale treatment plant. Therefore, the design is based on a constant flux after all.

According to Alfa Laval (Emmanuel Joncquez, 2020, personal email communication, 07.05), a suggested arrangement of the membrane modules is submerged tanks of 60 (6x10) modules in each tank.

The operation will occasionally be disturbed by CIP, though unlike the pilot plant, an entire full-scale plant cannot be taken out of operation. Due to the nature of the CIP procedure, which is applied for one tank of membranes, the safety of one extra tank, i.e. 60 extra membranes are added to the design. This would also allow for the flux to have a slight decrease, though less than observed in the pilot plant. Even higher safety could have been chosen. However with the chosen design leading flows above Q_{dim} to flocculation and sedimentation, it was chosen to have one extra tank. It could even be considered not to have the extra safety when having the sedimentation basins as a backup option. However, there could also appear other reasons for a membrane module being out of operation than just CIP.

The chosen design of the membrane step based on these decisions are presented in Table 4.5.

Table 4.5 Design of DMF

	Magnitude	Unit
Design flux	10.2	LMH
Needed area	185400	m ²
Increased area (Safety)	216300	m ²
Area per membrane module	515	m ²
Chosen number of membranes	420	-

Initially, a flux of 10 LMH was chosen. This resulted in need of 366 modules, and it was looked into how much the flux had to be increased in order only to need 360 modules, i.e. 6 tanks, and it was found that a flux increase of 0.2 LMH would result in 359, rounded to 360 modules. This was considered sufficient, as the chosen flux was already in the lower end of what can be found on the datasheet of the modules where between 10 and 30 LMH is set as typical net flux range. Furthermore, if all tanks and modules are in operation, which is assumed as it is unlikely with CIP or operational issues daily based on the pilot plant study, the resulting flux at Q_{dim} would be 8.7 LMH.

Recommended aeration rate for the membranes can be found in Judd and Judd (2011). The recommended aeration rate decreases as the membrane area of the module increases, and the chosen aeration rate for this design was based on the recommendation for a module of 462 m², and the median value is chosen as the design air demand and can be found in Table 4.6.

Table 4.6 Air demand for DMF

	Magnitude	Unit
Specific air demand	0.27	Nm ³ /m ² h
Total air demand	58401	Nm ³ /h

4.4.4 Theoretical sludge production

This process will produce chemical and mechanical sludge. There will be two sludge outtakes, one from the drum sieves and one from the membranes.

The mechanical sludge production can be found using equation 3.

$$SP = SS_{in} - SS_{out} \quad \text{Equation 3}$$

SS_{in} is the incoming concentration of suspended solids, and the SS_{out} is the outgoing concentration of suspended solids. SS_{in} is based on the PE from Table 4.1, SS_{out} is based on the removal results from Table 3.10.

Moreover, chemical sludge production can be found using equation 4.

$$SP = K_{chem} \times D \quad \text{Equation 4}$$

K_{chem} is the amount of SS produced per chemical added and is set to 6 g SS/g Al added, and D is the chemical dosage.

Table 4.7 shows the calculated sludge production for a DMF full-scale design.

Table 4.7 Sludge production DMF.

	kg/d
Mechanical sludge production	6396
Chemical sludge production	2442
Total	8838

4.5 Design of MBBR

In cases where the Norwegian guidance manual for process design of wastewater treatment plants (Ødegaard et al., 2009) was not adequate, the MBBR design is based on personal conversations with Flemming Wessman in April and May 2020, where he kindly shared of his knowledge and experience with the topic. As described, the existing sedimentation basins will be used as pre-settling for the MBBR in this design.

4.5.1 Biological reactors (MBBR)

A generic filling media with a specific area of 500 m²/m³ and a maximum filling degree of 67% was chosen. However, by choosing a lower filling degree, a very flexible solution is achieved in terms of future treatment needs, with an increase in the biological load being likely in the years to come. Therefore, a lower filling degree was chosen for the design, meaning if the treatment plant were to be expanded for future needs, using the maximum filling degree, this would allow for an incoming biological load of 7200 kg BOD/day. The incoming BOD to the reactor is based on the assumption of 20% removal in the pre-treatment (Ødegaard et al., 2009).

The arrangement of the reactors was decided based on hydraulic conditions, i.e. having a length:width ratio close to 1, having two reactors in series in order to avoid short-circuiting and meeting the recommendations about a hydraulic retention time of at least 30 minutes for Q_{maxdim} .

A deep reactor is an advantage in order to distribute the air more sufficiently; however, if the reactors get too deep, it is necessary with extra equipment in order to provide enough air. The organic loading rate was corrected for temperature, as the rate given by the guidance manuals is at a temperature of 15°C. The temperature correction was done using a correction factor found using equation 5 and accounts for the rate being lowered at temperatures below 10°C. The design temperature was presented in Table 4.1.

$$F = 1.06^{T-10} \quad \text{Equation 5.}$$

The needed biofilm area is chosen based on the organic loading rate and the specific area of the chosen carrier. The chosen design of the bioreactors in the MBBR process are presented in Table 4.8.

Table 4.8 Design of bioreactors.

	Magnitude	Unit
Organic loading rate	11.5	g BOD/m ² d
Organic loading rate, corr.	10.2	g BOD/m ² d
Incoming BOD	4430	kg BOD/d
Needed biofilm area	432869	m ²
Specific area of carriers	500	m ² / m ³
Volume of biomedica	866	m ³
Filling degree	41	%
Total volume of biological step	2100	m ³
Number of lines	3	-
Number of reactors in series	2	-
Water depth	5	m
Width per reactor	8	m
Length per reactor	9	m
HRT at Q _{maxdim}	35	min

The air demand of the bioreactor was found using the oxygen transfer rate in water, which was corrected using correction factor α which accounts for the water in the reactor not being clean water, combined with the oxygen demand of 1.0 kg O₂/kg BOD_{5, in}. This can be found in the Norwegian guideline for design of wastewater treatment processes. An α value of 0.7 was used. To find the air demand of the bioreactor, equation 6 was used.

$$\text{Design air demand} = \frac{\text{Oxygen demand}}{\text{OTR} \times \alpha} \times \frac{\text{Saturation concentration}}{\text{Saturation concentration} - \text{desired oxygen concentration}} \quad \text{Equation 6}$$

The saturation concentration is the saturation concentration of oxygen in water at the highest temperature (19 °C) and can be found in Tchobanoglous et al. (2014), and the desired oxygen concentration is set to 3 mg/L. The remaining parameters, as well as the resulting air demand, is presented in Table 4.9.

Table 4.9 Air demand for MBBR

	Magnitude	Unit
Oxygen transfer rate (OTR)	11.5	g O ₂ /Nm ³ m
OTR $\times\alpha$	8.05	g O ₂ /Nm ³ m
Oxygen demand	1	kg O ₂ /kg BOD in
Design air demand	6994	Nm ³ /h

The sieves to keep the carriers in their reactors were assumed to have a sieving rate of 50 m/h, based on experience. Similarly, the diameter and length, and hence the area of the sieves are experience-based. For this design, cylindrical bar sieves were chosen, which can be mounted horizontally or vertically. This has been the most common option in later years. There are also other possible options available, like vertical mounted rectangular mesh sieves (Ødegaard, 2006). The chosen design is shown in Table 4.10.

Table 4.10 Design of carrier sieves for MBBR

	Magnitude	Unit
Filtration rate	50	m ³ /m ² h (m/h)
Necessary sieve area	23.7	m ²
Area per sieve	3.97	m ²
Number of sieves per reactor	6	-

4.5.2 Coagulation and flocculation

Succeeding all the three bioreactor lines, the wastewater was chosen to be collected back to one flow, in order not to have an unnecessary amount of dosage points for the precipitation chemical. The wastewater goes through a rapid mixing chamber on its way to the flocculators. The polymer is added between the flocculation step and the flotation tank, meaning there will be one dosage point per line of flocculation/flotation. The dosage of chemicals is set to the same as for the DMF, based on the pilot plant results.

In the chosen design, there are flocculation reactors before each of the dissolved air flotation lines. As described in section 4.4.1, at least two flocculation reactors in series should be chosen. Since there are several lines, having more than two would require a higher number of mixers and hence have higher electricity demand and investment costs. The chosen design is shown in Table 4.11.

Table 4.11 Design of flocculation for MBBR

	Magnitude	Unit
Hydraulic retention time	15	min
Number of lines	4	-
Number of chambers per line	2	-
Needed volume	471	m ³
Length of chamber	3	m
Width of chamber	7	m
Depth of chamber	3	m
Total volume	504	m ³
Result retention time at Q_{dim}	16	min

4.5.3 Sludge separation – dissolved air flotation

As described in section 2.3.1, MBBR has excellent flexibility in the choice of particle separation. Dissolved air flotation has been chosen as the sludge separation for the biological step, due to its compactness. Operational matters have to be taken into consideration when choosing the dimensions of the flotation tanks, as well as the physical layout. By choosing a wider tank than 7-8 meters, there would be necessary with multiple sludge scrapers per reactor as opposed to one. The design for the dissolved flotation step is presented in Table 4.12.

Table 4.12 Design of flotation for MBBR

	Magnitude	Unit
Area using Q_{dim}	376.8	m ²
Area using Q_{maxdim}	355.2	m ²
Number of lines	4	-
Water depth	2.5	m
Length of tank	7	m
Width of tank	14	m

In accordance with (Ødegaard et al., 2009), the dispersion system should have a pressure between 4 and 6 bars, and a degree of air saturation of between 80 and 90%. The dispersion water volume should be 15% at Q_{dim} 7.5% at Q_{maxdim} and should be controlled by the flow after the bioreactor, meaning a flow meter is necessary.

4.5.4 Theoretical sludge production

The MBBR-DAF solution will produce mechanical sludge from the sedimentation, chemical sludge from the precipitation, and biological sludge from the biological step. There will be a sludge outtake from the sedimentation step where the mechanical sludge is collected, and from the dissolved air flotation where the biological and chemical sludge is collected.

The chemical and mechanical sludge production is found using equation 3 and 4 in section 4.4.4. The biological sludge production is found using equation 7.

$$SSP = 0.6 \times \left(\frac{TSS_{in}}{TBOD_{in}} + 1 \right) - \frac{0.72 \times F}{B_{A,BOD} + 1.33 \times F} \quad \text{Equation 7}$$

TSS_{in} and $TBOD_{in}$ are the total incoming suspended solids and BOD in g/m³ respectively to the MBBR reactor, $B_{A,BOD}$ is the organic load in g TBOD₅/m²d, and F is a temperature correction factor which is found using equation 8.

$$F = 1.072^{T-15} \quad \text{Equation 8}$$

The calculation of the biological sludge is a variation of the method presented for finding sludge production from an activated sludge plant (ATV-DVWK, 2000). The sludge production for the MBBR-DAF is presented in Table 4.13.

Table 4.13 Sludge production MBBR

	kg/d
Mechanical sludge production	5815
Chemical sludge production	2451
Biological sludge production	3997
Total	12243

4.6 Mass balances

Mass balances can be seen as a quantitative description of a treatment process, based on the law of conservation of mass (Crittenden et al., 2012). The mass balances in this thesis were made using Microsoft Excel. To avoid issues with circle referencing, iterative calculation needed enabling. In order to make the mass balances, some assumptions had to be made regarding the processes.

The incoming BOD (Total BOD, TBOD) and flows were based on Table 4.1, while SS was found using the number of PE combined with the assumption of a production of 70 g/PE d (Ødegaard et al., 2009). The suspended BOD (SBOD) is found based on two assumptions; that VSS makes up 75% of SS, and that 1 mg SS equals 0.56 g BOD (Tchobanoglous et al., 2014).

In the thickening and dewatering of sludge, a solid capture of 95% was assumed, leading to 5% of the solids being in the reject water flow. In sludge handling, biogas production is also involved, where the assumptions are based on guideline values (Ødegaard et al., 2009). The sludge handling is an example of what sludge handling might look like and is not definite.

The daily biogas production was found using equation 9.

$$\text{Biogas production} = (SS_{in} - SS_{out}) \times Y_{biogas} \quad \text{Equation 9}$$

Biogas production is the daily methane production, measured in Nm³ biogas/day, SS_{in} is the incoming suspended solids to the digester, SS_{out} is the SS going out of the digester, and Y_{biogas} is the biogas yield, assuming 0.9 Nm³ biogas/kg VSS reduced. The content and reduction of volatile suspended solids (VSS) is implemented in the SS_{out} which is found using equation 10.

$$SS_{out} = SS_{in} - SS_{in} \times VSS \times VSS_{red} \quad \text{Equation 10}$$

VSS is the assumed fraction of SS being volatile and VSS_{red} is the assumed reduction of VSS in the digester. All these parameters are in accordance with Ødegaard et al. (2009), where it is also stated that the produced biogas contains 60-70% methane and 30-40% carbon dioxide. In the mass balance, methane content of 65% was assumed.

The mass balances were used to estimate biogas production for both the processes, as well as providing insight in necessary sludge storage volume. The mass balances can also be used for scenario simulation, like increased load and flow, showing what kind of impact this could theoretically have on the processes. The reject flows and loads are also demonstrated in the mass balances.

In the DMF mass balance, some measurements from the pilot plant were used instead of making assumptions. This includes the TS out of both the microsieving step and the microfiltration step, as described in section 3.5.238. The removal of SS and BOD over the microsieving step is also partly based on measurements in section 3.4.2, and partly assumptions. The overall removal of SS is based on numbers from accredited sampling in section 3.6.20. It was assumed that equal concentrations of SS and BOD are entering the sedimentation step and the microsieving step when the inflow exceeds Q_{dim}. If the flow is at exactly Q_{dim}, the amount of wastewater entering the flocculation-sedimentation will be equal to the reject water flow. This mass balance allows for the simulation of different inflows, and their impact on the overall removal, considering the existing flocculation-sedimentation will not be as efficient as the newly designed DMF. The equations used in the DMF mass balance are shown in Appendix F.

An example of the mass balance for DMF is shown in Figure 4.4. In the case shown in the figure, Q_{dim} is used as the inflow value, the load of BOD and SS is also according to design values. An example showing what the full-scale DMF might look like with Q_{maxdim} instead of Q_{dim} can be found in Appendix G, where it can be seen that with a similar TBOD, but higher flow, removal of 80% is still obtained.

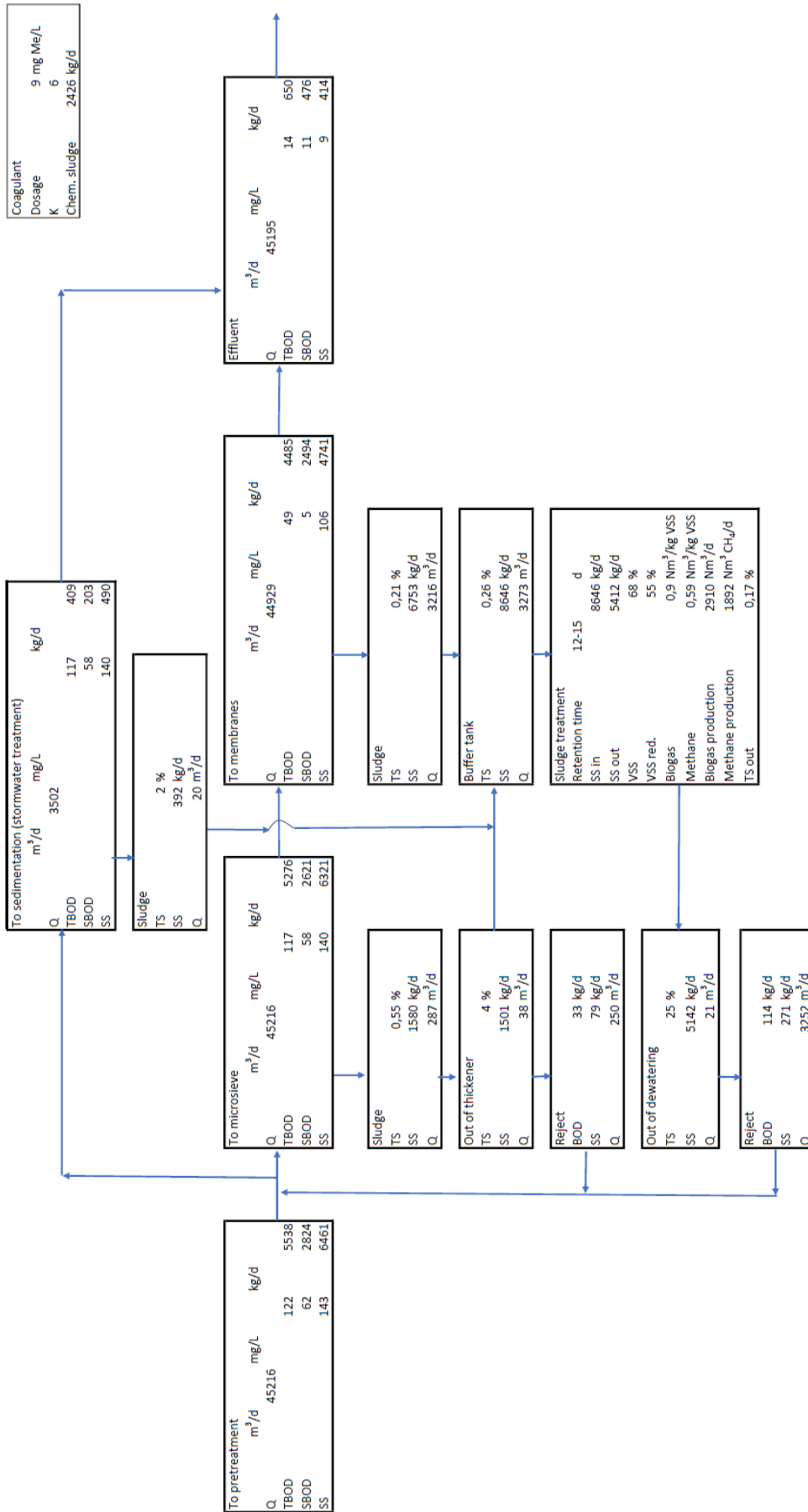


Figure 4.4 DMF mass balance.

In the MBBR mass balance, it was assumed that the pre-settling would remove 50% of the incoming SS and 20% of the incoming BOD, and in the overall process, removal of 90% is assumed for both SS and BOD (Ødegaard et al., 2009). The TS in the two sludge outtakes was based on typical values found in Tchobanoglous et al. (2014). The equations used in the MBBR mass balance are shown in Appendix H. An example using the mass balance with the design values for both flow and pollutants is shown in Figure 4.5.

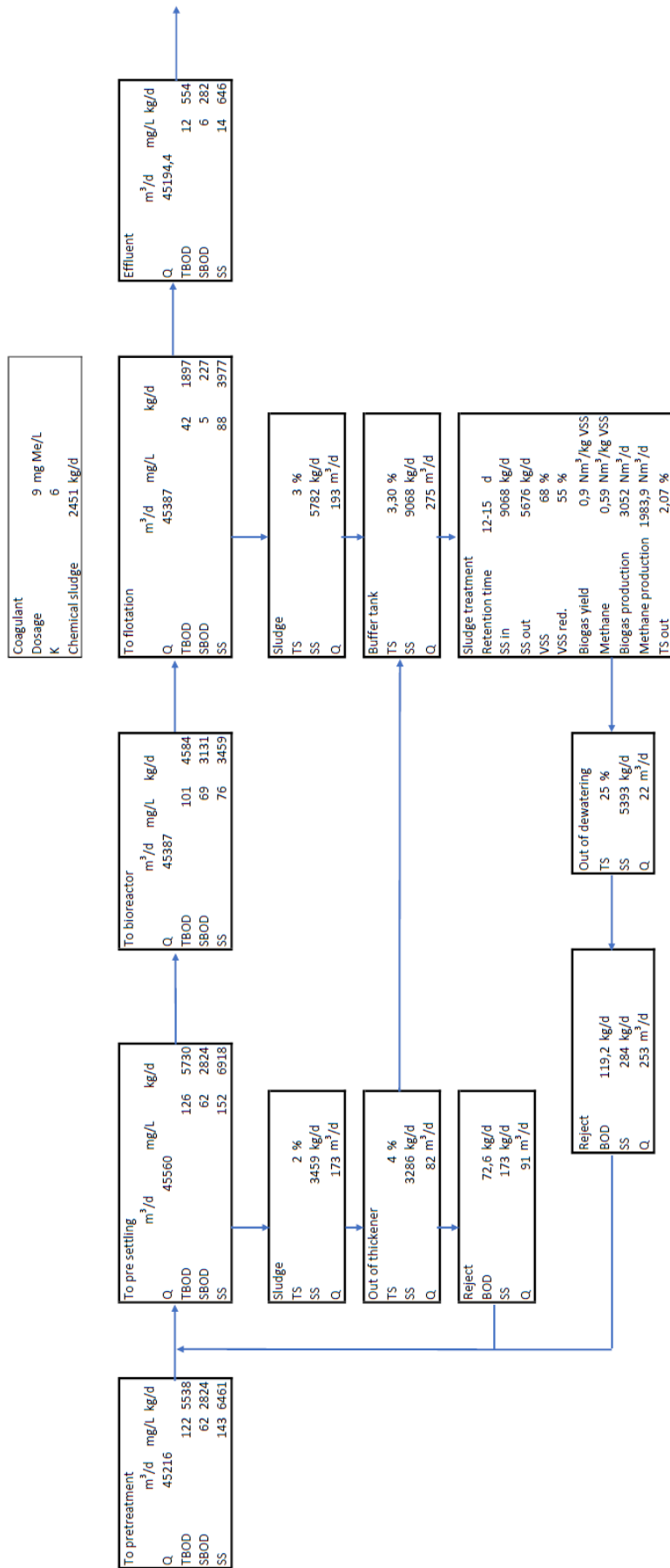


Figure 4.5 MBBR mass balance.

4.7 Process flow charts

Simplified process flow charts were made using AutoCAD. Their purpose is to show the general layout and flows of the process. Hence all valves, sensors, and the like are not included, neither are the dosing arrangement of chemicals, or sludge treatment. All of these components are beyond the scope of the design, which is process design, and not a detailed design. The aim of the flow charts is mainly to show unit processes and reactors, and how the flows are arranged in this design.

The process flow chart for a potential design of the DMF concept is shown in Figure 4.6. For the microsieving step, all the drum sieves will have their own sludge flows and filtrate flows, but for the sake of clarity, they are all collected into one in the flowchart. Similarly, the membrane filtration is divided into seven tanks containing 60 membranes, as described earlier, which are all collected into one sludge flow and one permeate flow in the drawing. However, these would, in reality, be separated. Lastly, the existing flocculation and sedimentation steps used for stormwater treatment is not included in this drawing for simplicity. FT represents a flow transmitter, while PT represents a pressure transmitter.

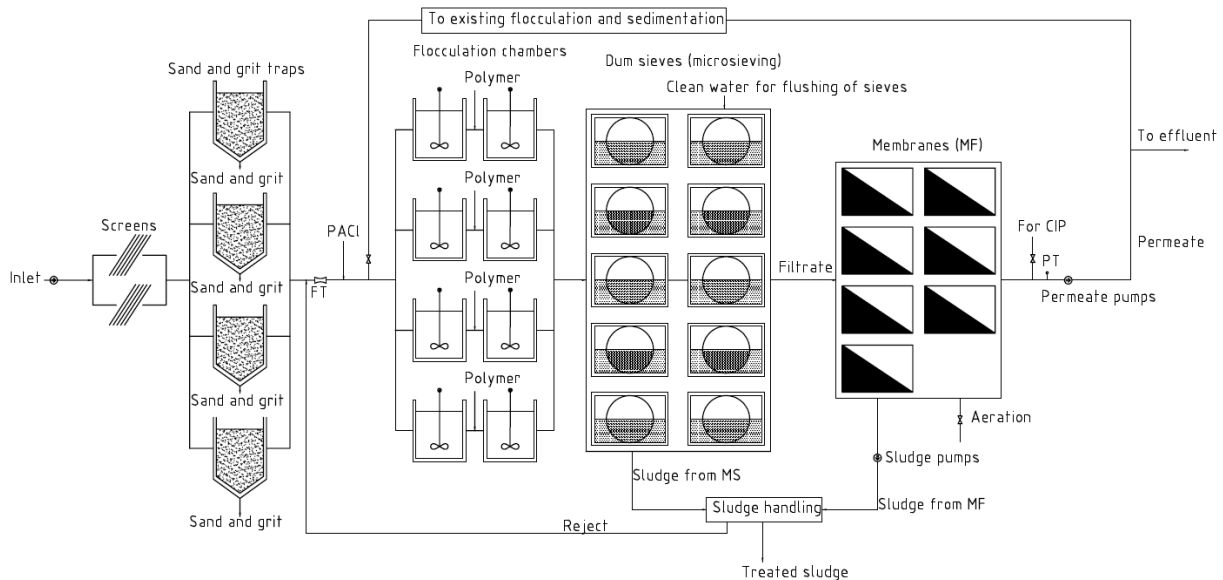


Figure 4.6 Flowchart of a full-scale DMF process.

The process flowchart for the designed MBBR-DAF process is shown in Figure 4.7. Production of aerated dispersion water, as well as the dosing arrangements, is not described in detail in the drawing, as for the DMF process. As in the DMF flowchart, the FT represents a flow transmitter.

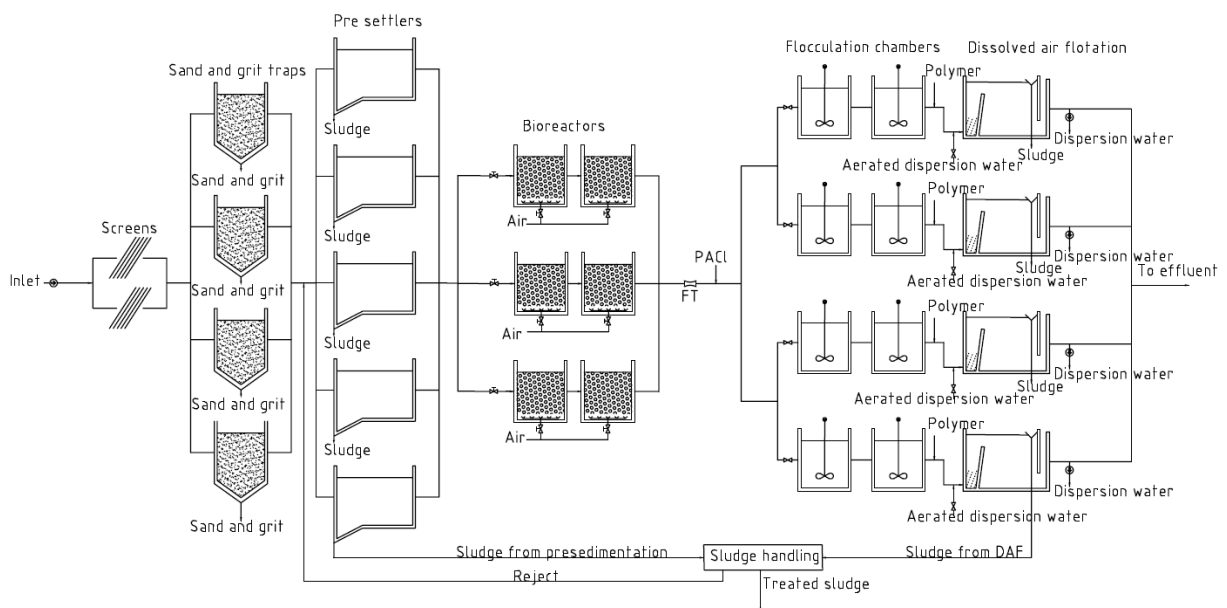


Figure 4.7 Flowchart of a full-scale MBBR-DAF process.

4.8 Footprint of the treatment plant

The footprints of the treatment processes were only based on the designed process steps, not piping and other necessary equipment. Furthermore, neither the existing pre-treatment, i.e. inlet screens, sand- and grease trap nor the sedimentation basins used for pre-sedimentation and stormwater treatment for MBBR and DMF respectively, are accounted for in the area estimation, as they would be the same for both processes. The sludge handling, e.g. thickening, dewatering, and digestion, are also not included. The estimated area of DMF and MBBR can be found in Table 4.14 and Table 4.15, respectively.

Table 4.14 Footprint of DMF.

Step	Area [m ²]
Coagulation and flocculation	128
Microsieving	52
Microfiltration	879
Total	1059

Table 4.15 Footprint of MBBR-DAF.

Step	Area [m ²]
Bioreactors	432
Coagulation and flocculation	168
Flotation	392
Total	992

Using the PE estimate based on the incoming BOD load, the specific area demand can be found for the two processes. A specific area demand of 0.0115 m²/PE was found for DMF, and 0.0107 m²/PE was found for MBBR.

The specific area demand for DMF was found by Hey et al. (2018) to be 0.024 m²/PE, but this included pre-treatment as well as equalisation and CIP for the MF. When the sludge handling was included, the specific area was found to be 0.046 m²/PE, while if just the inclusion of the same steps as in the calculation of this thesis, the specific area demand would be 0.0093 m²/PE. Both the DMF and the MBBR-DAF in this study were found to have a smaller specific area than activated sludge plants, comparing to five AS plants in Sweden found to have specific area demands between 0.11 m²/PE and 0.23 m²/PE (Hey, 2016).

4.9 Energy consumption

In this thesis, energy consumption is not calculated, but experiences from previous studies are presented to provide insight into how energy consumption would be. Looking to the literature, Lateef et al. (2013) report an estimate of energy consumption of a full-scale DMF to be <0.5 kWh per m³ treated wastewater. A similar estimate was made by Hey et al. (2018) who found an energy consumption of 0.41 kWh per m³ treated wastewater for DMF, including energy consumption for mixing and microsieving. In comparison, the new Norwegian guidance manual for process design of wastewater treatment plant reports typical energy consumption for MBBR and DAF for BOD removal can be assumed to be typically 0.08-0.10 kWh per m³ treated wastewater and 0.04 kWh per m³ treated wastewater, respectively. This adds up to the total of 0.12-0.14 kWh per m³ treated wastewater for MBBR-DAF (Johannessen et al., 2020).

There will be other electricity consuming components in a treatment plant. These components will be approximately the same for the two processes, and neither are the most dominant regarding energy consumption. The same is assumed for pretreatment, while handling of sludge is not taken into consideration. For example, energy use for precipitation is small and mostly due to the pumping of chemicals and mixing. The estimation of energy use of mixing will be assumed the same for the two processes in this thesis; typical energy consumption for flocculation chamber 15-20 W/m³ reactor volume or 10-15 W/m³ reactor volume for reactors with and without bottom scraper respectively. Equally, the sedimentation, as pre-settling for the MBBR and stormwater treatment for the DMF, will have the same specific energy consumption for the two, typically 0.005 kWh/m³ treated water (Johannessen et al., 2020). It can be expected that the sedimentation basins in the DMF will be used less frequently and will, therefore, give a lower energy consumption. This is a gravity-based process that consumes little energy nonetheless compared to the other components.

4.10 Design summary

In this section, the design of the two treatment concepts is summarised, in terms of design, chemical demands, energy demands, and space demands. The purpose of this section is to provide insight to operational demands, which gives a quick overview of how they compare, and how they compare with the current treatment plant at FREVAR. A summary is shown in Table 4.16

Table 4.16 Summary of the two designs and the current FREVAR plant.

	DMF	MBBR-DAF	Current FREVAR
Chemical demand			
Precipitation	4522 kg PAX/d (3.3 m ³ /d)	4522 kg PAX/d (3.3 m ³ /d)	19194 kg PIX/d
Flocculant	66.5 kg polymer/d (33.2 m ³ solution/d)	66.5 kg polymer/d (33.2 m ³ solution/d)	-
Fouling removal (CIP)	7.5 kg H ₂ O ₂ +35 kg citric acid/CIP of 1 module	-	-
pH adjustment (CIP)	2 kg HCl+1 kg NaOH/CIP of 1 module	-	-
Footprint	0.0115 m ² /PE	0.0107 m ² /PE	0.0191 m ² /PE
Sludge production	8838 kg/d	12243 kg/d	12223 kg/d
Air demand	58401 Nm ³ /h	6994 Nm ³ /h	-
Energy demanding components	- Permeate pumping - Microsieves	Dispersion pump	-
Energy consumption	0.4-0.5 kWh/m ³ treated water	0.12-0.14 kWh/m ³ treated water	0.05 kWh/m ³ treated water

For the chemical precipitation step, the estimate of the consumption of chemicals is based on the last operation condition of the pilot plant. Daily consumption is based on Q_{dim} .

The chemical demand for cleaning of membranes has been provided by the membrane supplier, Alfa Laval, based on a full-scale design consisting of a total of 720 membrane modules and an average flow of 2700 m³/h (Emmanuel Joncquez, 2020, personal email correspondence, 11.05.).

The estimates of the energy consumptions are based on literature and have not been calculated. Nor do they include every component of the processes, as described in the previous section. For the current treatment plant, only the sedimentation is included.

For the current treatment plant, the estimated use of precipitation chemical is based on the new Q_{dim} and the average use (g/m³ from 2019). Polymer is being used at the treatment plant today, but in the sludge handling, and is therefore excluded. The sludge production is an estimate based on equation 3 and 4, based on current removal rate of SS and the incoming SS based on PE. The footprint only includes the flocculation and sedimentation.

4.11 Discussion

Neither of the designs in this thesis took the reject water flow or reject pollution loads into account, as one would do in an actual design (Ødegaard et al., 2009). This would have some impact on the result in both cases, but it was decided not to include it. The purpose of the designs was not to be complete but to provide a foundation for comparing the novel DMF process to more renowned processes, like the MBBR-DAF process. The reject water flows and loads were shown in the respective mass balances, but are mainly

theoretical, and not necessarily complete. This is another reason for exclusion from the overall design.

The chemical demand of the DMF could perhaps be lowered, especially the precipitation chemical due to the high removal of phosphorous. The polymer could also perhaps be reduced. It is not a certainty that this would work efficiently, but there was not enough time to confirm or disprove this statement. Reducing the dosages would result in a lower chemical use overall for DMF, and the same dosages would also have been used in the design for the MBBR. It was shown that the DMF process would demand more chemicals overall than the MBBR, due to the necessary chemical cleaning. Both processes need chemicals for phosphorous precipitation and flocculation. In this design, the same chemical concentrations (mg Al/L and mg polymer/L) are assumed for MBBR as for DMF, which is perhaps conservative, and the dosage for MBBR would, in reality, be lower. The exact dosage for the MBBR is not clear. The reason for expecting a lower dose for the MBBR is some of the phosphorous which the chemical precipitation is aimed at is particulate and hence removed in pre settling. Some will also be assimilated, i.e. taken up by removed cell mass in the biological step (Gilberg et al., 2003). Either way, the overall chemical consumption will be lower for the MBBR.

Comparing the chemical demand to the current usage at FREVAR, the dosage of precipitation chemical is significantly higher than necessary. In 2019, 492 g ferric chloride (Kemira PIX-118A05)/m³ (51.7 g Fe³⁺/m³) was added for precipitation on average, resulting in average daily use of 19194 kg ferric chloride, comparing to an estimated daily use of 4522 kg PAX for the new designs. This is a different chemical from what has been used in the pilot study and designs. It has another amount of active metal, but the chemical use is still overall much higher. To put this into further perspective, a CIP using the amount of chemicals suggested in Table 4.16, for all 420 modules of the DMF design, would require a lower amount of chemicals than one current daily dose of ferric chloride at FREVAR, although obviously with different chemicals.

Furthermore, by looking at the molar ratio for the current dosage, it is of significantly higher magnitude, at a Fe:P ratio of 9.89, comparing to the Al:P ratio 3.57 used in the pilot plant. Although these are different chemicals, lower optimum Fe:P ratios have been reported in the literature (Cai et al., 2020, Wang et al., 2005). This observation implies that perhaps it would be of the interest of the current treatment plant to change their chemical as well as dosage even before starting the building of the new treatment plant, as it would be advantageous in terms of both finance and environment.

Both processes demand air for different purposes. MBBR requires supplied air for the bioreactors, while the DMF has an air demand in order to mitigate fouling of the membranes using air scouring. The latter was found to have almost 10x the magnitude of the first. This could perhaps be lowered, as the assumed air demand for the DMF was based on a membrane module with lower area and the air demand increasing inversely proportionally to the module area. Although it could perhaps be lowered, it is highly unlikely that it could be lowered to the levels of the MBBR. This leads to the conclusion of DMF having a higher energy consumption than MBBR-DAF. The energy demands of the two processes were not calculated, but can be evaluated from literature, which demonstrated a higher energy demand of DMF than of MBBR-DAF. Comparing to a conventional activated sludge, the MBBR-DAF will have an energy consumption of the approximately same magnitude, when looking at the new Norwegian guidelines for wastewater treatment (Johannessen et al., 2020).

MBBR and DMF are according to the results from this thesis, both space-efficient compared to a conventional activated sludge treatment plant. It was found that for this exact design, the full-scale MBBR alternative takes up slightly less space than a full-scale DMF. The coagulation/flocculation step of both processes are designed under the same conditions but have different areas due to layout arrangement. When judging the treatment plant footprint, it would conceivably be reasonable to use the same area for both processes.

The two processes were also designed differently; in the DMF design, assumptions about filtration rate and flux had to be made. The flux was chosen at the lower end of the given capacity of the chosen membrane modules. Being less conservative could perhaps still be sufficient, and by increasing the flux, the necessary membrane area and hence the footprint of the membrane step would decrease. The chosen membrane modules were chosen from convenience, using the same ones as in the pilot plant. Membranes with higher specific area exist and could be considered. The same can be said about the microsieving step, although the assumption made for designing this was less conservative in the first place. Furthermore, more space-efficient microsieves exist, for example, by looking at disc sieves, rather than drum sieves.

Also the bioreactor step of the MBBR-DAF could be made more space-efficient if a carrier with a higher specific biofilm area, meaning a lower reactor volume would be necessary, also resulting in a smaller footprint. The impact of this could have been looked further into but was not prioritised, and a non-specified carrier with the same specific area as the original K1 carrier was chosen.

Additionally, looking into the overall footprint, not just the designed process steps, the inclusion of sludge handling would change the total picture. It was found in the mass balance, that due to the more diluted sludge coming from DMF compared to MBBR, DMF would demand a higher storage volume. This would add to the overall footprint.

Both processes considered in this thesis, MBBR and DMF, are flexible solutions in terms of expansion. While the design of MBBR was made with expansion in mind, choosing a lower filling degree than maximum, DMF is a modular solution, meaning it can easily be expanded by adding more membrane modules and microsieves. Unlike with an MBBR which has not been completely filled in regard to filling degree of carriers, the expansion of DMF would need an enlarged area as well as investment in additional membrane modules.

Looking at the mass balances and the biogas estimates, a slightly higher daily biogas production can be expected from the MBBR than the DMF. These are both only estimates and will have different types of sludge, and it will therefore not represent full correctness to use the same values for estimation. For the DMF pilot plant, another estimate was also made in terms of methane production, where methane production of 5.7 Nm^3 was found for the last sludge analysis. Converting this to production per volume of incoming water, and upscaling it to the size of the full-scale DMF, a result of $3464 \text{ Nm}^3 \text{ CH}_4/\text{d}$ was obtained. Comparing to the methane production from the mass balance, where it is assumed 65% of the biogas is methane, $1892 \text{ Nm}^3 \text{ CH}_4/\text{d}$ is obtained. The difference in these results is high, underlining that there is a general uncertainty in the calculations, not only for the DMF process but also for the MBBR.

Another uncertainty in the designs, including the mass balances, is sludge production, as the amount of TS was assumed based on literature for MBBR and on a minimal number

of samples for the DMF. This also supports the decision of excluding reject water from the designs, as this is dependent on the TS fraction. However, if this were investigated closer, the mass balances would be useful tools for comparing the two concepts on these terms.

According to literature, MBBR-DAF has a low operational demand as described in section 2.3.3, while as described in section 3.8, the pilot plant was mainly maintained by personnel for CIP, as well as pump issues which are not directly related to the DMF concept. These statements cannot be quantified directly and are therefore hard to compare, but as a general assumption, it can be said that both processes are operator friendly.

As stated by researchers introduced in section 2, the reuse of material can reduce the environmental footprint for treatment plants. The design showed that reuse of already existing components, like sandtrap and sedimentation basins, is highly possible, and the latter was found to have several possible applications. Retrofitting the sedimentation basins does not only save the cost and energy of demolishing, but also saves space and energy for the MBBR-DAF by reducing the incoming organic load. Similarly, it can save space and energy for the DMF working primarily as a stormwater treatment instead of using membrane for this purpose.

Both designs are subject to change when designing in accordance with the new and revised guidance manuals for wastewater treatment processes. However, the design principles are the same, and the same conclusion is likely to be drawn for designs using said guidance manual as well. The new guidelines were published at the time of the deadline of this thesis. Therefore, it has not been included except for when looking at energy consumption.

While MBBR-DAF might seem more suitable for the new treatment plant to be built at FREVAR right now, stricter requirements might come up in the future, making membrane technology highly relevant, especially in terms of removal of small particles like microplastics and bacteria-free wastewater, e.g. bathing water qualitative. Although it is positive for the environment of the recipient, it is generally not of concern to remove such constituents and to obtain higher removal than required if it requires extra expenses, like this thesis has found that DMF would. However, in the future, this might be the case.

If choosing an MBBR-DMF process, retrofitting with DMF, or membranes could be an option as a last polishing step, if the requirements were to become stricter. Amongst others, Ødegaard et al. (2012) obtained success with a similar setup, testing various scenarios with coagulation. It was concluded with the combination MBBR-DAF-UF having lower fouling, higher possible flux, and lower consumption of both backwashing water and chemicals compared to having the UF step right after bioreactors. A future expansion like this one is possible if an MBBR-DAF process is chosen.

5 Conclusion

From the pilot plant study in this thesis, DMF turned out to be a sufficient treatment solution, operationally but especially with regards to removal. The pilot plant showed outstanding efficiency in removal of phosphorous, organic matter like COD and BOD, and SS. Although there are no requirements in terms of heavy metal removal, the removal of those can be considered as excellent too. This showed that it is indeed possible to achieve sufficient removal rates without the use of biology. The chemical dosage could perhaps be lowered even more, but this could not be attempted due to lack of time. However, the results with the tested precipitation metal dosage do already imply that even before building a new treatment plant, FREVAR can potentially reduce their chemical consumption.

The two processes considered are different in many ways; how they are designed, as well as the nature of the removal, with DMF being physiochemical, and MBBR being biological and chemical. Compared to a conventional activated sludge process, this study has found that both processes will take up small amounts of space, and based on literature, both concepts have a lower energy demand.

In this study, MBBR-DAF requires both fewer and less chemicals than DMF, as well as significantly less air and will hence likely have a much lower energy demand. Therefore, for this case study, the MBBR is likely to be a better choice of process, but DMF should not be excluded from the consideration. If the treatment requirements were stricter, for example in terms of removal of heavy metals and microplastics, DMF would likely be the preferred option based on literature and the pilot plant study, meaning it might, therefore, be even more relevant in the future.

5.1 Further work

99.9 % removal of microplastics using an MBR treatment process and 97% removal using dissolved air flotation have been confirmed by Talvitie et al. (2017). It could be of interest to investigate the microplastics and other micropollutants removal of DMF, as well as a full MBBR-DAF process.

It was concluded that due to the great air demand compared to MBBR-DAF, DMF would have a higher energy demand. For DMF to be an even more relevant treatment process for a municipal wastewater treatment plant, it would be beneficial to explore whether it could be done more efficiently in terms of energy use.

References

- Ahn, K.-H. & Song, K.-G. 2000. Application of microfiltration with a novel fouling control method for reuse of wastewater from a large-scale resort complex. *Desalination*, 129, 207-216.
- Ahn, K.-H., Song, K.-G., Yeom, I.-T. & Park, K.-Y. 2001. Performance comparison of direct membrane separation and membrane bioreactor for domestic wastewater treatment and water reuse. *Water Science and Technology: Water Supply*, 1, 315-323.
- Alfa Laval. 2020. *Alfa Laval MBR membranes* [Online]. Available: https://www.alfalaval.com/globalassets/documents/products/separation/membranes/mbr-membranes/mbr-membranes_200000414-2-en-gb.pdf?_ga=2.233094814.1878842416.1591621442-2014291205.1588369150 [Accessed 08.06 2020].
- Arndt, F., Ehlen, F., Schütz, S., Anlauf, H. & Nirschl, H. 2016. Influence of operating parameters and membrane materials on fouling of ceramic hollow fibre membranes. *Separation and Purification Technology*, 171, 289-296.
- Atv-DvWK 2000. Dimensioning of Single-Stage Activated Sludge Plants. ATV-DVWK A 131E Standard. *German ATV-DVWK Rules and standards*. Hennef, Germany: German Association for Water, Wastewater and Waste.
- Bachmann, N., La Cour Jansen, J., Bochmann, G. & Montpart, N. 2015. *Sustainable biogas production in municipal wastewater treatment plants*, IEA Bioenergy Massongex, Switzerland.
- Balkema, A. J., Preisig, H. A., Otterpohl, R., Lambert, A. J. D. & Weijers, S. R. 2001. Developing a model based decision support tool for the identification of sustainable treatment options for domestic wastewater. *Water Science and Technology*.
- Bendick, J. A., Miller, C. J., Kindle, B. J., Shan, H., Vidic, R. D. & Neufeld, R. D. 2005. Pilot Scale Demonstration of Cross-Flow Ceramic Membrane Microfiltration for Treatment of Combined and Sanitary Sewer Overflows. *Journal of environmental engineering*, 131, 1532-1539.
- Borg, A., Dejardins, A. K. & Larsen, H. N. 2019. 251-2019 Greenhouse gas emissions, a guidance document for the water sector *Norsk Vann Rapport*. Hamar, Norway: Norwegian Water BA.
- Butler, D. & Davies, J. 2017. *Urban Drainage, Third Edition*.
- Cai, Q., Li, H.-Q. & Yang, P. Evaluation of four different coagulants used for chemical dephosphorization of membrane bioreactor effluent. E3S Web of Conferences, 2020. EDP Sciences, 01009.
- County Governor of Oslo and Viken 2019. Permission to activity by pollution requirements for Fredrikstad municipality for operation of municipal wastewater treatment plant, p. 58-79 (In Norwegian). <http://www.frevar.no/globalassets/pdf/styresaker/2019-09-17-sakspapirer-styret.pdf>: FREVAR kf.
- Crittenden, J. C., Trussell, R. R., Hand, D. W., Howe, K. J. & Tchobanoglous, G. 2012. *MWH's Water Treatment: Principles and Design, Third Edition: Principles and Design*, Hoboken, NJ, USA, John Wiley & Sons, Inc.
- Diaz, S. D., Peña, L. V., Cabrera, E. G., Soto, M. M., Cabezas, L. M. V. & Sánchez, L. R. B. 2012. Effect of previous coagulation in direct ultrafiltration of primary settled municipal wastewater. *Desalination*, 304, 41-48.
- Duan, J. & Gregory, J. 2003. Coagulation by hydrolysing metal salts. *Advances in colloid and interface science*, 100, 475-502.
- Emmerson, R., Morse, G., Lester, J. & Edge, D. 1995. The life-cycle analysis of small-scale sewage-treatment processes. *Water and Environment Journal*, 9, 317-325.

- Friedrich, E., Pillay, S. & Buckley, C. A. 2007. The use of LCA in the water industry and the case for an environmental performance indicator. *Water SA*, 33, 443-451.
- Gao, W., Liang, H., Ma, J., Han, M., Chen, Z.-L., Han, Z.-S. & Li, G.-B. 2011. Membrane fouling control in ultrafiltration technology for drinking water production: A review. *Desalination*, 272, 1-8.
- Gibert, O., Lefèvre, B., Ferrer, O., Prats, G., Bernat, X. & Paraira, M. 2016. Composition and reversibility of fouling on low-pressure membranes in the filtration of coagulated water: insights into organic fractions behaviour. *Desalination and Water Treatment*, 57, 26313-26326.
- Gilberg, L., Hansen, B., Karlsson, I., Nordström Enkel, A. & Pålsson, A. 2003. *About water treatment*, Helsingfors, Finland, Kemira Kemwater.
- Gong, H., Jin, Z., Wang, X. & Wang, K. 2015. Membrane fouling controlled by coagulation/adsorption during direct sewage membrane filtration (DSMF) for organic matter concentration. *Journal of Environmental Sciences (China)*, 32, 1-7.
- Graedel, T. E. & Allenby, B. R. 2015. *Industrial ecology and sustainable engineering*, Uttar Pradesh, Pearson.
- Guo, W., Ngo, H.-H. & Li, J. 2012. A mini-review on membrane fouling. *Bioresource Technology*, 122, 27-34.
- Helness, H. & Ødegaard, H. 2001. Biological phosphorus and nitrogen removal in a sequencing batch moving bed biofilm reactor. *Water Science and Technology*, 43, 233-240.
- Hey, T. 2016. *Municipal wastewater treatment by microsieving, microfiltration and forward osmosis*. Doctoral Thesis, Lund University.
- Hey, T., Bajraktari, N., Davidsson, Å., Vogel, J., Madsen, H. T., Hélix-Nielsen, C., Jansen, J. L. C. & Jönsson, K. 2018. Evaluation of direct membrane filtration and direct forward osmosis as concepts for compact and energy-positive municipal wastewater treatment. *Environmental technology*, 39, 264-276.
- Hey, T., Väänänen, J., Heinen, N., La Cour Jansen, J. & Jönsson, K. 2017. Potential of combining mechanical and physicochemical municipal wastewater pre-treatment with direct membrane filtration. *Environmental Technology (United Kingdom)*, 38, 108-115.
- Hong, S., Bae, T., Tak, T., Hong, S. & Randall, A. 2002. Fouling control in activated sludge submerged hollow fiber membrane bioreactors. *Desalination*, 143, 219-228.
- Hube, S., Eskafi, M., Hrafnkelsdóttir, K. F., Bjarnadóttir, B., Bjarnadóttir, M. Á., Axelsdóttir, S. & Wu, B. 2020. Direct membrane filtration for wastewater treatment and resource recovery: A review. *Science of The Total Environment*, 710, 136375.
- Im, D., Nakada, N., Kato, Y., Aoki, M. & Tanaka, H. 2019. Pretreatment of ceramic membrane microfiltration in wastewater reuse: A comparison between ozonation and coagulation. *Journal of Environmental Management*, 251.
- Jin, Z., Gong, H., Temmink, H., Nie, H., Wu, J., Zuo, J. & Wang, K. 2016. Efficient sewage pre-concentration with combined coagulation microfiltration for organic matter recovery. *Chemical Engineering Journal*, 292, 130-138.
- Jin, Z., Gong, H. & Wang, K. 2015. Application of hybrid coagulation microfiltration with air backflushing to direct sewage concentration for organic matter recovery. *Journal of hazardous materials*, 283, 824-831.
- Johannessen, E., Rusten, B., Ødegaard, H., Bjørn, E. & Paulsrud, B. 2020. 256-2020 Guidance manual for process design of wastewater treatment plants (In Norwegian). *Norsk Vann Rapport*. Hamar, Norway: Norsk Vann BA.
- Judd, S. & Judd, C. 2011. *The MBR book : principles and applications of membrane bioreactors for water and wastewater treatment*, Amsterdam, Elsevier.
- Lateef, S. K., Soh, B. Z. & Kimura, K. 2013. Direct membrane filtration of municipal wastewater with chemically enhanced backwash for recovery of organic matter. *Bioresource Technology*, 150, 149-155.
- Le-Clech, P., Chen, V. & Fane, T. a. G. 2006. Fouling in membrane bioreactors used in wastewater treatment. *Journal of Membrane Science*, 284, 17-53.

- Lee, C. S., Robinson, J. & Chong, M. F. 2014. A review on application of flocculants in wastewater treatment. *Process safety and environmental protection*, 92, 489-508.
- Lien, O. 2013. *VA-tjenestene blir stadig bedre (Norwegian)* [Online]. Norsk Vann. Available: <https://norsk vann.no/component/content/article/10-nyheter/645-va-tjenestene-blir-stadig-bedre> [Accessed 15.04. 2020].
- Ljunggren, M., Alm, L., Nilsson, B., Persson, E., Strube, R., Svensson, L., La Cour Jansen, J., Hahn, H., Hoffmann, E. & Ødegaard, H. 2007. Microscreening in chemically enhanced primary treatment. *Chemical Water and Wastewater Treatment IX*, 49-58.
- Løberg Skår, K., Gelius, J. & Egeland, G. I. 2020. Fem år etter uhellet dukker disse plastbitene fortsatt opp. *NRK Sørlandet*, 20.02.2020.
- Martinez, R. 2012. 194-2012 Optimal energy design of wastewater plants (In Norwegian). *Norsk Vann Rapport* Hamar, Norway: Norwegian Water BA.
- McQuarrie, J. P. & Boltz, J. P. 2011. Moving Bed Biofilm Reactor Technology: Process Applications, Design, and Performance. *Water Environment Research*, 83, 560-575.
- Meng, F., Chae, S.-R., Drews, A., Kraume, M., Shin, H.-S. & Yang, F. 2009. Recent advances in membrane bioreactors (MBRs): Membrane fouling and membrane material. *Water Research*, 43, 1489-1512.
- Mezohegyi, G., Bilad, M. R. & Vankelecom, I. F. 2012. Direct sewage up-concentration by submerged aerated and vibrated membranes. *Bioresource technology*, 118, 1-7.
- Nascimento, T. A., Fdz-Polanco, F. & Peña, M. 2018. Membrane-Based Technologies for the Up-Concentration of Municipal Wastewater: A Review of Pretreatment Intensification. *Separation & Purification Reviews*, 1-19.
- Park, H.-D., Chang, I.-S. & Lee, K.-J. 2015. *Principles of membrane bioreactors for wastewater treatment*, CRC Press.
- Paulsrud, B., Rusten, B. & Aas, B. 2014. Increasing the sludge energy potential of wastewater treatment plants by introducing fine mesh sieves for primary treatment. *Water Science and Technology*, 69, 560-565.
- Pinnekamp, J., Schröder, M., Bolle, F.-W., Gramlich, E., Gredigk-Hoffmann, S., Koenen, S., Loderhose, M., Miethig, S., Ooms, K. & Riße, H. 2018. Energie in Abwasseranlagen. *Handbuch NRW*. Nordrhein-Westfalen, Germany: Ministerium für Umwelt, Landwirtschaft, Natur- und Verbraucherschutz des Landes Nordrhein-Westfalen.
- Pollution Regulations. 2004. *Regulations relating to pollution control (Norwegian title: Forskrift om begrenning av forurensning)* [Online]. <https://lovdata.no/dokument/SF/forskrift/2004-06-01-931>. [Accessed 29.11 2019].
- Ravazzini, A. M., Van Nieuwenhuijzen, A. F. & Van Der Graaf, J. H. M. J. 2005. Direct ultrafiltration of municipal wastewater: comparison between filtration of raw sewage and primary clarifier effluent. *Desalination*, 178, 51-62.
- Remy, C., Boulestreau, M. & Lesjean, B. 2014. Proof of concept for a new energy-positive wastewater treatment scheme. *Water Science and Technology*, 70, 1709-1716.
- Rusten, B., Kolkinn, O. & Odegaard, H. 1997. Moving bed biofilm reactors and chemical precipitation for high efficiency treatment of wastewater from small communities. *Water Science and Technology*, 35, 71-79.
- Rusten, B. & Lundar, A. 2004. PRIMÆRRENS prosjekt 10: Utnyttelse av kjemisk forbehandling på finsilannleg. Case Namsos. (In Norwegian). Aquateam - Norsk vannteknologisk senter A/S.
- Skaug, H. 2020. Avløp - regulering, behandling, dimensjonering og planer. *Vannspeilet*, 01.03.2020.
- Standard Norge. 1980. *Water analysis, Total residue, and total fixed residue in water, sludge and sediments* [Online]. Available: <https://www.standard.no/no/Nettbutikk/produktkatalogen/Produktpresentasjon/?ProductID=134402> [Accessed 24.02. 2020].
- Standard Norge. 1983. *Water analysis, Determination of suspended solids in wastewater and their residue on ignition* [Online]. [Accessed 09.03 2020].

- Statistisk Sentralbyrå. 2016 *Klimagasser, etter kilde og komponent (F) 2009-2016* [Online]. Available: <https://www.ssb.no/statbank/table/08940/> [Accessed 02.05. 2020].
- Szabó, A., Takács, I., Murthy, S., Daigger, G. T., Licskó, I. & Smith, S. 2008. Significance of Design and Operational Variables in Chemical Phosphorus Removal. *Water Environment Research*, 80, 407-416.
- Talvitie, J., Mikola, A., Koistinen, A. & Setälä, O. 2017. Solutions to microplastic pollution – Removal of microplastics from wastewater effluent with advanced wastewater treatment technologies. *Water Research*, 123, 401-407.
- Tchobanoglous, G., Abu-Orf, M., Bowden, G. & Pfrang, W. 2014. *Wastewater engineering : treatment and resource recovery*, New York, McGraw-Hill Education.
- The Norwegian Environmental Agency. 2019. *EUs avløpsdirektiv: Implementering, oppfyllelse og revidering? (In Norwegian)* [Online]. Available: <https://www.norskvann.no/index.php/kompetanse/arrangementer/dokumentasjon-fra-arrangementer?folder=fagtreffV19%252F2AAvlop> [Accessed 10.03. 2020].
- Tilche, A. & Galatola, M. 2008. The potential of bio-methane as bio-fuel/bio-energy for reducing greenhouse gas emissions: a qualitative assessment for Europe in a life cycle perspective. *Water Science and Technology*, 57, 1683-1692.
- Trondsen, R. 2015. Massivt opprydningsarbeid etter utslipp av plastbiter (In Norwegian). *Fædrelandsvennen*.
- Un Water 2017. 2017 UN World Water Development Report, Wastewater: The Untapped Resources. Facts and Figure. *The United Nations World Water Development Report*.
- United Nations Department of Economic Affairs (Population Division). 2019. *World Population Prospects 2019: Highlights* [Online]. https://population.un.org/wpp/Publications/Files/WPP2019_Highlights.pdf?fbclid=IwAR3wCFy9DLdpJE9xws8EjzTmeEPADS65SmZCeOPRr5PibSgsaU34jxv3YVg. [Accessed 13.12 2019].
- United Nations Department of Public Information. *Sustainable development goals* [Online]. <https://sustainabledevelopment.un.org/sdgs>: United Nations (UN). [Accessed 09.06.2020].
- Van Nieuwenhuijzen, A., Evenblij, H. & Van Der Graaf, J. 2000. Direct wastewater membrane filtration for advanced particle removal from raw wastewater. *Chemical Water and Wastewater Treatment VI*. Springer.
- Van Nieuwenhuijzen, A. F. 2002. *Scenario Studies into Advanced Particle Removal in the Physical-Chemical Pre-treatment of Wastewater*. PhD thesis, Delft University.
- Väänänen, J., Cimbritz, M. & La Cour Jansen, J. 2016. Microsieving in primary treatment: Effect of chemical dosing. *Water Science and Technology*, 74, 438-447.
- Wang, Y., Han, T., Xu, Z., Bao, G. & Zhu, T. 2005. Optimization of phosphorus removal from secondary effluent using simplex method in Tianjin, China. *Journal of Hazardous Materials*, 121, 183-186.
- Warsinger, D. M., Chakraborty, S., Tow, E. W., Plumlee, M. H., Bellona, C., Loutatidou, S., Karimi, L., Mikelonis, A. M., Achilli, A., Ghassemi, A., Padhye, L. P., Snyder, S. A., Curcio, S., Vecitis, C. D., Arafat, H. A. & Lienhard, J. H. 2018. A review of polymeric membranes and processes for potable water reuse. *Progress in Polymer Science*, 81, 209-237.
- Water Environment Federation 2006. *Membrane systems for wastewater treatment*, New York, WEF Press/McGraw-Hill.
- World Economic Forum. 2020. *The Global Risks Report 2020* [Online]. Available: http://www3.weforum.org/docs/WEF_Global_Risk_Report_2020.pdf [Accessed 10.03. 2020].
- Zhang, Z. & Wilson, F. 2000. Life-Cycle Assessment of a Sewage-Treatment Plant in South-East Asia. *Water and Environment Journal*, 14, 51-56.
- Zhao, Y. X., Li, P., Li, R. H. & Li, X. Y. 2019. Direct filtration for the treatment of the coagulated domestic sewage using flat-sheet ceramic membranes. *Chemosphere*, 223, 383-390.

- Ødegaard, H. 2006. Innovations in wastewater treatment: –the moving bed biofilm process. *Water Science and Technology*, 53, 17-33.
- Ødegaard, H. 2016. A road-map for energy-neutral wastewater treatment plants of the future based on compact technologies (including MBBR). *Frontiers of Environmental Science & Engineering*, 10.
- Ødegaard, H., Gisvold, B. & Strickland, J. 2000. The influence of carrier size and shape in the moving bed biofilm process. *Water Science and Technology*, 41, 383-391.
- Ødegaard, H., Mende, U., Skjerping, E. O., Simonsen, S., Strube, R. & Bundgaard, E. 2012. Compact tertiary treatment based on the combination of MBBR and contained hollow fibre UF-membranes. *Desalination and Water Treatment*, 42, 80-86.
- Ødegaard, H., Norheim, B. & Norsk Vann, B. A. 2014. *Vann- og avløpsteknikk*, Hamar, Norsk Vann.
- Ødegaard, H., Rusten, B. & Siljudalen, J. 1999. The development of the moving bed biofilm process- from idea to commercial product. *European water management*, 2, 36-43.
- Ødegaard, H., Rusten, B., Storhaug, R. & Paulsrud, B. 2009. 168-2009 Guidance manual for process design of wastewater treatment plants (In Norwegian). *Norsk Vann Rapport*. Hamar, Norway: Norsk Vann BA.
- Ødegaard, H., Rusten, B. & Wessman, F. State of the art in Europe of the moving bed biofilm reactor (MBBR) process. Proceedings of the 77th Annual Water Environment Federation Technical Exposition and Conference, New Orleans, Louisiana, Oct, 2004. 2-6.

Paper I

Investigating direct membrane filtration (DMF) as a potential treatment concept for municipal wastewater

Klaastad, E.*, Lipnizki, F.**, Hey, T.*** and Johannessen, E.***

*Department of Civil and Environmental Engineering, Norwegian University of Science and Technology, Trondheim, Norway, elinekl@stud.ntnu.no

Department of Chemical Engineering, Lund University, Sweden, frank.lipnizki@chemeng.lth.se *
COWI AS, Water Division, Department of Industry and Process, Norway.

Abstract: The FREVAR municipal wastewater treatment plant is required to improve its current treatment and the aim of this study is to investigate if direct membrane filtration (DMF) using microfiltration membranes can be a potential solution for the future. Norwegian requirement for plants in this category is typically to remove 70 % BOD₅ and 75 % COD, as well as 90 % phosphorus. DMF is a novel concept where the aim is to reduce organic matter without biological treatment. To confirm this, a pilot plant has been installed on-site. In the pilot set-up, the wastewater goes initially through a sand trap, coagulation and flocculation step, microsieving, and enters eventually the membrane stage. Currently, this concept results in removals of up to 94% COD and up to 97% phosphorous. DMF has also proven to require less energy than biological processes. Furthermore, the concept has significant potential for increased biogas production as well as removal of micropollutants such as microplastics.

Keywords: Direct membrane filtration; P-removal; wastewater treatment

INTRODUCTION

It is increasingly important to adapt sustainable wastewater treatment methods regarding both energy and space consumption. In contrast to conventional activated sludge processes, direct membrane filtration (DMF) is an abiotic concept requiring low energy and giving the advantage of increased biogas production. Several studies have proven that membrane filtration, e.g. the DMF-concept has several benefits including meeting increasing effluent requirements (Hey et al., 2017, Nascimento et al., 2018, Zhao et al., 2019)

Due to stricter effluent requirements, the FREVAR municipal wastewater treatment plant need to change its treatment process from the current practice. The aim of this research is to investigate whether the direct membrane filtration concept can be a future solution. The DMF operation is so far mainly based on smaller scale research and needs to be optimised to match the wastewater characteristics given at FREVAR. The wastewater characteristics at FREVAR are typical for Norwegian treatment plants; low strength in general and low temperatures during winter.

The DMF process is interesting for FREVAR, as well as other municipal wastewater treatment operators facing BOD, COD and phosphorus removal commonly solved by biological treatment. Furthermore, several studies have found that the DMF process with coagulation is feasible for energy recovery (Guyen et al., 2019, Hey et al., 2017, Jin et al., 2016). Guven et al. (2019) concluded that biogas potential of organically-rich DMF sludge can be several times higher than for conventional activated sludge. Furthermore, the worldwide issue of microplastics in the oceans can be significantly reduced by membrane filtration as part of wastewater treatment, demonstrated by Talvitie et al. (2017).

MATERIALS AND METHOD

The DMF-pilot plant is situated at FREVAR municipal wastewater treatment plant in Fredrikstad, Norway. The pilot wastewater influent is a side stream of the main flow of the treatment plant, separated from the main flow between the influent screen and the sand trap due to iron-chloride addition in the sand trap. Therefore the pilot has its own sand trap with subsequent coagulation (PAX-18) for phosphorous precipitation. Thereafter, flocculation is performed by adding anionic or non-ionic polymer, enhancing floc building. Both coagulation and flocculation process have a retention time of 15 minutes. Thereafter, rotating microsieving is performed with a mesh size of 100 μm for substrate removal.

Filtrate from the microsieve enters the membrane tank containing two submerged membrane modules operated parallel. Both modules contain microfiltration membranes (MFP2, Alfa Laval) with a pore size of 0.2 μm and are operated at trans-membrane-pressure of maximum -0.5 bar (vacuum). The membranes are made of polyvinylidene fluoride (PVDF) and have a hollow sheet module configuration. They have a total surface area of 386 m^2 , and a maximum operating flux of 10 LMH, corresponding to 96 m^3/day , approximately 400 PE. To minimise fouling, the membrane modules are constantly aerated. Membrane cleaning (CIP) is performed at TMP -0.6 bar. Wastewater sampling is conducted by taking accredited samples in the influent and effluent for analyses of COD, BOD, TP, heavy metals, and organic toxins at an accredited laboratory.

This study does not only conduct the operation of the DMF as a proof of concept in larger scale for wastewater treatment, but also aims to confirm higher biomethane potential (BMP) described by earlier studies. A comparison in terms of operation, energy and space consumption to conventional treatment methods will be performed.

RESULTS AND DISCUSSION

Microfiltration has shown to retain all suspended solids present in wastewater. In particular, wastewaters with high fractions of particular organic matter e.g. COD clearly benefits from this kind of concept. This is the case for the wastewater influent at FREVAR, where currently 94% of the incoming COD has been removed, while general removal rates are higher than 85% in the full-scale plant. However, new effluent limits require a COD removal of at least 75% or maximum concentration of 125 mg/L , which can be easily fulfilled with DMF.

Furthermore, phosphorous removal of 90% is required at FREVAR depending on the inlet and outlet-concentration. This requirement is particularly challenging since a substantial part of the sewers leading to the treatment plant are combined, and not separated. During high hydraulic loads, to obtain 90% P-removal, mathematically the effluent concentration must be close to zero due to low influent concentration. Coagulation and flocculation are ensuring sufficient carbon and phosphorous removal and in addition creating stable flocs. During initial testing, high removal rates of carbon and phosphorus (>90 %) have been achieved.

CONCLUSION

The pilot plant provides satisfactory results, meeting the future COD and phosphorous treatment limits. Thus, the DMF process seems to be a feasible solution for the new treatment process at FREVAR wastewater treatment plant. Due to its low energy consumption and additional benefits e.g. separation of microplastics, this concept can be an interesting solution for other treatment plants facing similar challenges. Another advantage of the DMF process is

the organic-rich sludge suitable for biogas production to increase the overall energy efficiency of the process even further.

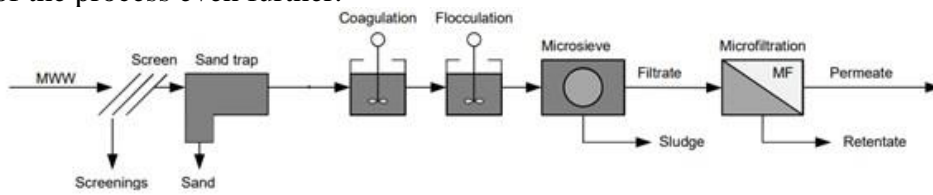


Figure 1.1 Simplified process flow chart (Hey et al. 2017)



Figure 1.2-1.4 The pilot installation at FREVAR

Figure 1.2 Sand trap (left) and coagulation tank (right)

Figure 1.3 Flocculation tank (left) and microsieve (right)

Figure 1.4 Membrane tank



Figure 1.5 Water from the coagulation tank (left), clear DMF effluent (right)

REFERENCES

- Guven, H., Dereli, R. K., Ozgun, H., Ersahin, M. E. & Ozturk, I. 2019. Towards sustainable and energy efficient municipal wastewater treatment by up-concentration of organics. *Progress in Energy and Combustion Science*, 70, 145-168.
- Hey, T., Väänänen, J., Heinen, N., La Cour Jansen, J. & Jönsson, K. 2017. Potential of combining mechanical and physicochemical municipal wastewater pre-treatment with direct membrane filtration. *Environmental Technology (United Kingdom)*, 38, 108-115.
- Jin, Z., Gong, H., Temmink, H., Nie, H., Wu, J., Zuo, J. & Wang, K. 2016. Efficient sewage pre-concentration with combined coagulation microfiltration for organic matter recovery. *Chemical Engineering Journal*, 292, 130138.
- Nascimento, T. A., Fdz-Polanco, F. & Peña, M. 2018. Membrane-Based Technologies for the Up-Concentration of Municipal Wastewater: A Review of Pretreatment Intensification. *Separation & Purification Reviews*, 1-19.
- Talvitie, J., Mikola, A., Koistinen, A. & Setälä, O. 2017. Solutions to microplastic pollution – Removal of microplastics from wastewater effluent with advanced wastewater treatment technologies. *Water Research*, 123, 401407.
- Zhao, Y. X., Li, P., Li, R. H. & Li, X. Y. 2019. Direct filtration for the treatment of the coagulated domestic sewage using flat-sheet ceramic membranes. *Chemosphere*, 223, 383-390.

Appendices

Appendix A: Datasheet PAX-18 (In Norwegian)

Appendix B: Datasheet Superfloc N-100

Appendix C: Measurements of inflow, flux, and temperature from the pilot plant

Appendix D: Script for extraction of values from pilot plant data

Appendix E: Graphical calculation of equalisation volume

Appendix F: Equations and associated assumptions used in DMF mass balance

Appendix G: Example of DMF mass balance using $Q_{\max\text{dim}}$

Appendix H: Equations and associated assumptions used in the MBBR mass balance

Appendix A

kemira

Where water
meets chemistry™

Technical Data Sheet

1 (1)

18.02.2016

Kemira PAX-18

KEMIRA PAX-18, polyaluminiumklorid med middels basisitet, er en effektiv koagulant for behandling av både drikkevann og avløpsvann. KEMIRA PAX-18 er basert på høyladet aluminium, slik at man trenger mindre av produktet for å gjøre mer. Dette resulterer i lavere doser og derfor reduseres slamvolum og behov for justering av pH. KEMIRA PAX-18 er også mer effektiv til å fjerne partikler og/eller fosfor i forhold til tradisjonelle koagulanter.

Produktspesifikasjon

Typiske analyser

Utseende	Gulaktig væske	Aktive substanser	Ca. 3,3 mol/kg
Aluminium (Al ³⁺)	9,0 ± 0,3 %	Jern (Fe _{tot})	<0,01 %
Al ₂ O ₃	17,0 ± 0,6 %	Klorider (Cl ⁻)	21 ± 1 %
Basisitet	42 ± 3 %	Viskositet (20°C)	35 ± 10 mPas
Tetthet	1,37 ± 0,03 g/cm ³	pH (20°C)	<1
		Krystalliseres ved	-20°C

Kvalitet

KEMIRA PAX-18 er godkjent av Mattilsynet til bruk som koagulant ved drikkevannrensing. Kemira Chemicals AS er sertifisert iht ISO 9001:2008 og 14001:2004.

Dosering

Dosering med membran doseringspumper av ikke-korroderende materiale er best egnet. KEMIRA PAX-18 bør doseres uten fortykning.

Lagring

Lagertanker og rørsystemer skal bygges av egnet ikke-korroderende materiale slik som glassfiberarmert polyester eller tværbundet polyeten. KEMIRA PAX-18 er sterkt etsende og kontakt med utstyr må unngås. KEMIRA PAX-18 har en anbefalt holdbarhet på 6 måneder. Som med ethvert kjemikalie, anbefales det å rengjøre tanken hvert år. Den første leveransen av et produkt skal gjøres til en ren oppbevaringsinnretning for å sikre optimal ytelse og lagring. Ved utendørs lagring, bør tank og rør isoleres og varmebeskyttes. Temperaturen på produktet bør opprettholdes over 0 °C med en -30 °C omgivelsestemperatur

Sikker håndtering

Håndteringen av ethvert kjemikalie krever forsiktighet. Enhver ansvarlig for bruk eller håndtering av KEMIRA PAX-18 bør gjøre seg kjent med fullstendige sikkerhetstiltak beskrevet i vårt HMS-datablad.

Kemira stiller disse opplysningene til rådighet som en tjeneste overfor sine kunder, og det er utelukkende en veiledning for kundene når de skal vurdere produktene. Du må prøve våre produkter for å avgjøre om de egner seg til din bruk, både fra et helse-, sikkerhets- og miljøperspektiv. Du må også underrette ansatte, fullektige, leverandører, kunder eller eventuelle tredjeparter som kan bli eksponert for produktene, om alle aktuelle forholdsregler. Alle opplysninger og all teknisk bistand gis garantert, eller garantien kan endres uten varsel. Du påtar deg det fulle ansvar for å overholde alle opplysninger og forholdsregler samt offentlige lover og forskrifter som gjelder behandling, transport, levering, lossing, avvikling, håndtering, salg og bruk av hvert produkt. Ingenting i dette dokumentet skal tolkes som en anbefaling til å bruke produktet i strid med patenter for eventuelle materialer eller bruken av disse. Kemira PAX-18 er varemerker eller registrerte varemerker som tilhører Kemira Oyj eller dets datterselskaper

Kemira Oyj

P.O.Box 330 (Porkkalankatu 3)
FI-00101 Helsinki
Finland

www.kemira.com

Europe, Middle-East and Africa
Tel: +358 10 8611

Asia-Pacific
Tel: +86 21 6037 5999

Americas
North America
Tel: +1 770 436 1542
South America
Tel: +55 11 2189 4900

Appendix B

Kemira

Where water
meets chemistry™

Technical Data Sheet

1 (2)

June 18

SUPERFLOC® N-100

Nonionic Dry PAMs

Superfloc N-100 is one of Kemira's highly effective nonionic flocculants. They condition solids for dewatering operations and aid water clarification processes in various industries. They show exceptional performance in liquid-solid separations in a wide range of conditions.

Applications

This product may be beneficial in any liquid-solid separation processes. They are especially recommended for:

- Dissolved air flotation
- Filtration
- Gravity settling
- Mechanical dewatering
- Water Clarification.

Benefits

- Dry product minimizes storage requirements
- Economical to use – effective at low dosage levels
- Effective high solids removal
- Effective over a wide range of pH; does not alter the system pH
- Eliminates or reduces use of inorganic salts
- High storage stability
- Fast settling rates.

Health and Safety

Before handling these materials read the corresponding Kemira Safety Data Sheets (SDS) for safety and health data.

For chemical inventory regulatory control listing information, see the SDS.

Handling and Storage

Solutions are no more corrosive than water. Recommended materials of construction include stainless steel, fibreglass, plastic, and glass or epoxy-lined vessels. Do not use iron, copper or aluminum.

The shelf life of these products is 24 months when stored in unopened packages in a dry atmosphere at temperatures no higher than 40°C/104°F.

Spilled polymer is very slippery and should be absorbed onto an inert material and collected prior to thoroughly flushing with water.

Product Addition

Stock solutions can be prepared up to 0.5 % concentration via an automated make-up unit or on a batch basis. Solutions should be aged 30-60 minutes for maximum effectiveness. High quality make up water should be used. Secondary dilution water should be added to the stock solution prior to the addition point at a ratio up to 10:1. Centrifugal pumps should be avoided for polymer transfer.

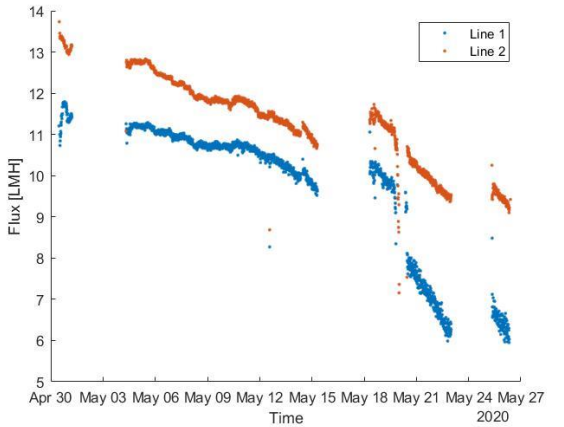
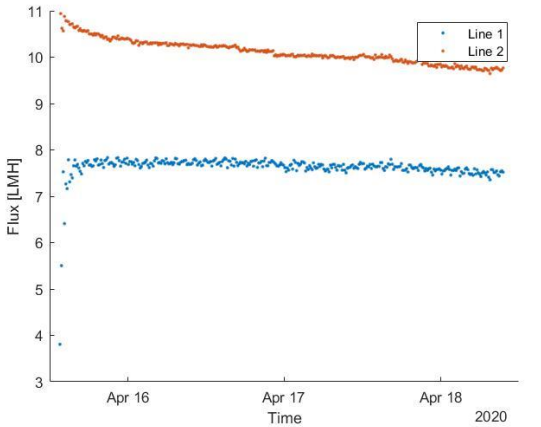
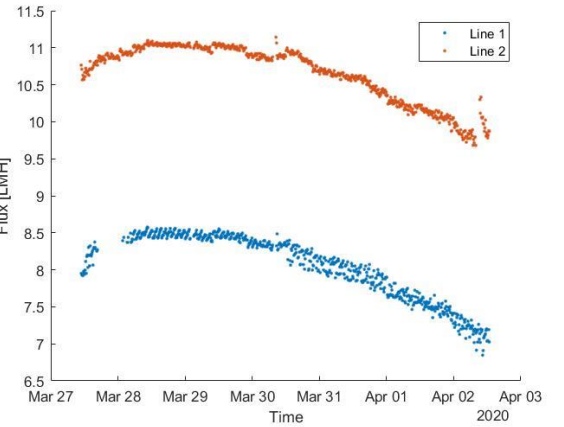
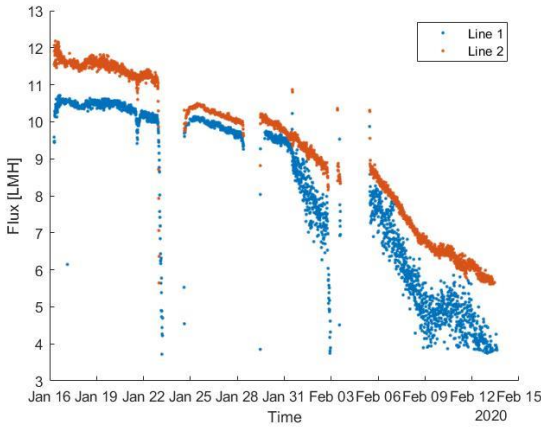
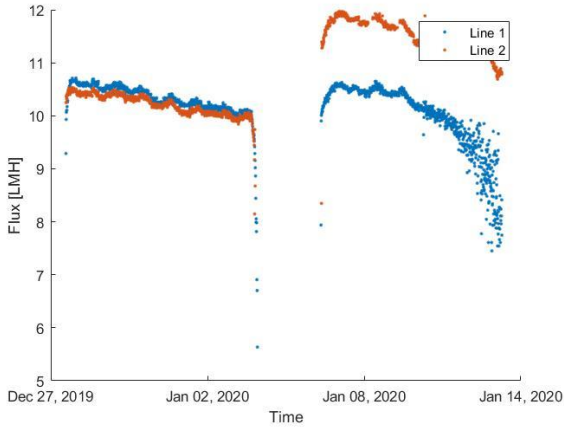
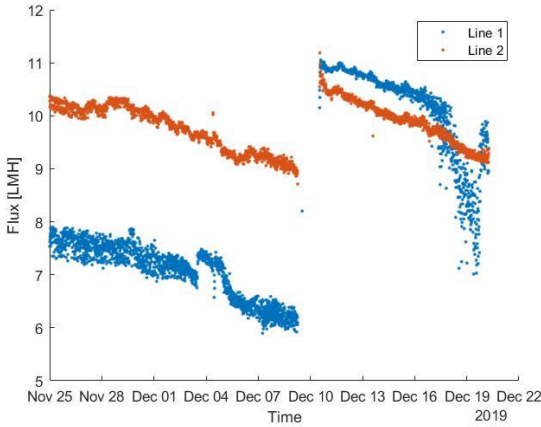
Kemira Oyj
P.O.Box 330
FI-00101 Helsinki
Finland
www.kemira.com

Europe, Middle-East and Africa
Tel +358 10 8611
North America
Tel +1 770 436 1542

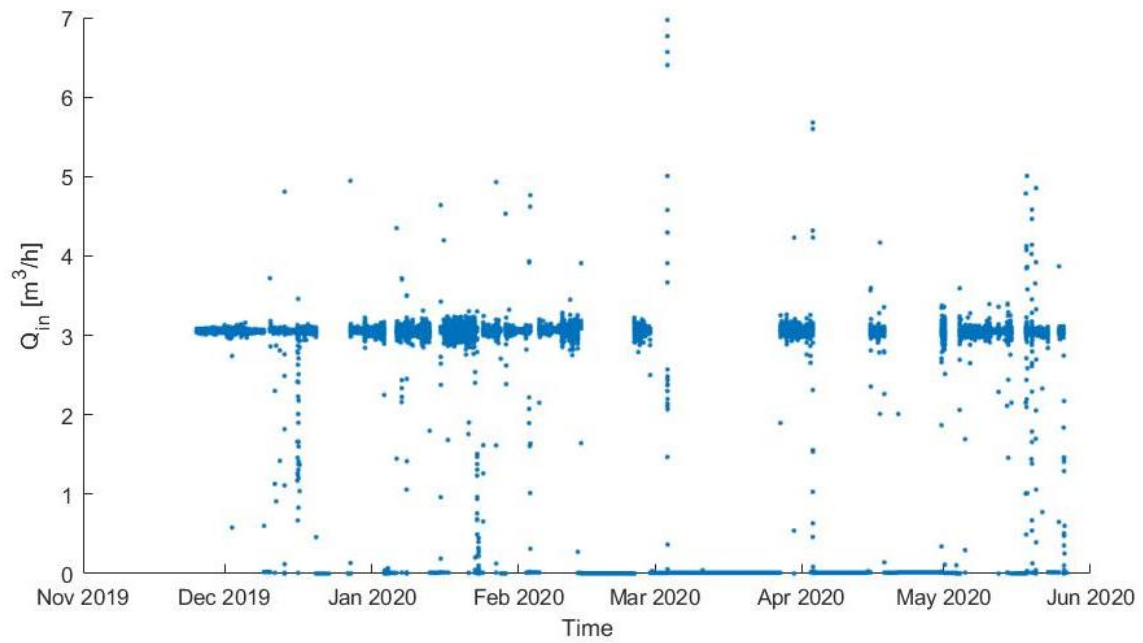
South America
Tel +55 11 2189 4900
Asia-Pacific
Tel +86 21 6037 5999

Appendix C

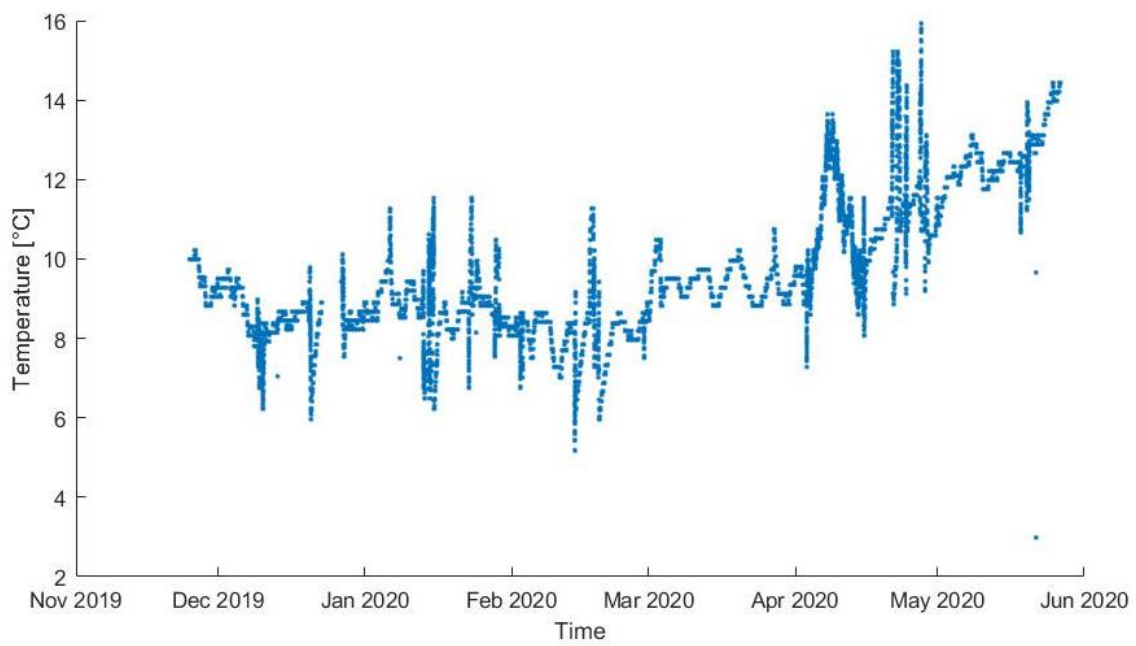
The flux of membranes in the pilot plant. These are standardised to a temperature of 15°C, in accordance with (Van Nieuwenhuijzen et al., 2000). The dates of these figures correspond to the dates of Figure 3.19-3.24.



The inflow of the pilot plant, 25.11.2019-26.05.2020.



The temperature in the membrane tank of the pilot plant, 25.11.2019-26.05.2020



Appendix D

The script used for extraction of time values from operational data for the membranes is presented in this appendix. Permeate flow for each of the membranes was used as a foundation for finding the same point in the operational cycle. Another option could be to use the inflow; however, the membranes' cycles are not always parallel, so the permeate flows were used in order to distinguish between the two.

Running the script in matlab gives a new string of times of which the membrane cycle his running, and not relaxing. The script does not provide a perfect result, but removes the majority of the "noise" in the data, making graphical representation clearer and easier to interpret.

The script was slightly adapted for different cases as it was observed that the flux, i.e. outflow from the membranes was varying over time.

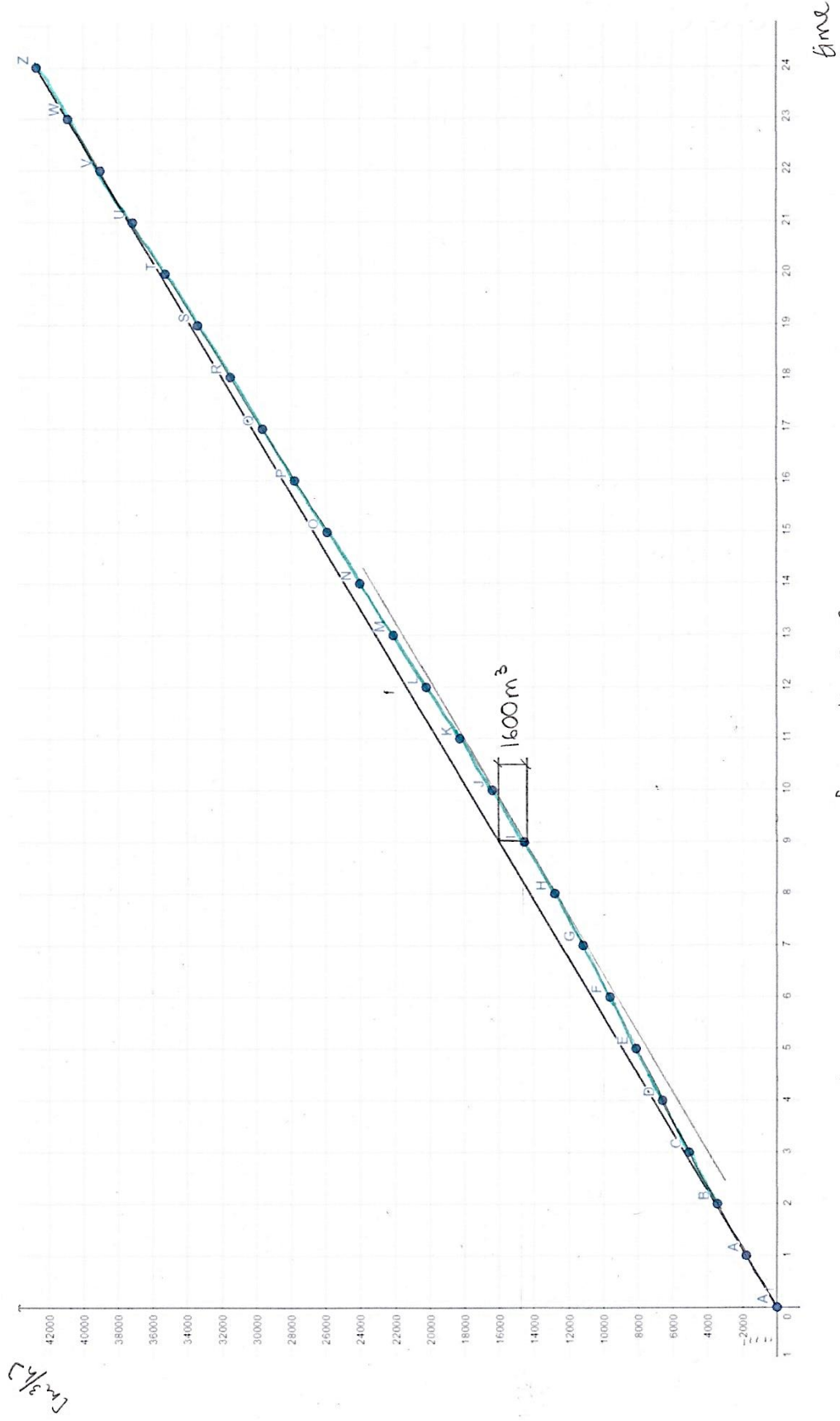
Below is the script for membrane line 2, hence the variable name Q2.

```
n=length(Q2);
i=1;
q=1;

% Make empty string to put in dates
dates2=strings(n,1);
% Make empty matrix to put in Q values

for i=1:length(Q2)          % Check all the Q-values
    if Q2(i)<0.9 && Q2(i+1)<0.9 && Q2(i+2)>0.9 && Q2(i+3)>0.9 &&
Q2(i+4)>0.9          % Check values in cycle
        % Extract the time value and add to dates if correct place in cycle
        dates2(q)=Time(i+4);
        q=q+1;
    end
end
```


Appendix E



Appendix F

To pretreatment

(1)

$$Q_1 \quad Q_{dim}$$

$$TBOD_1 \quad BODload$$

$$SBOD_1 \quad TBOD_1 - (SS_1 \times 75\% \times 0.56)$$

$$SS_1 \quad SS \text{ load based on PE}$$

To microsieve (2)

$$Q_2 \quad IF (Q_1 + Q_8 + Q_{14}) < 45216,$$

$$Q_2 = Q_1 + Q_8 + Q_{14},$$

$$ELSE Q_2 = 45216$$

$$TBOD_2 \quad \frac{Q_2}{Q_1 + Q_8 + Q_{14}} \times (TBOD_1 + BOD_8 + BOD_{14})$$

$$SBOD_2 \quad TBOD_2 - (SS_2 \times 75\% \times 0.56)$$

$$SS_2 \quad \frac{Q_2}{Q_1 + Q_8 + Q_{14}} \times (SS_1 + SS_8 + SS_{14})$$

To sedimentation
(stormwater
treatment) (3)

$$Q_3 \quad IF Q_2 < 45216,$$

$$Q_3 = 0,$$

$$ELSE Q_3 = Q_1 + Q_8 + Q_{14} - Q_2$$

$$TBOD_3 \quad \frac{Q_3}{Q_1 + Q_8 + Q_{14}} (TBOD_1 + BOD_8 + BOD_{14})$$

$$SBOD_3 \quad TBOD_3 - (SS_3 \times 75\% \times 0.56)$$

$$SS_3 \quad \frac{Q_3}{Q_1 + Q_8 + Q_{14}} \times (SS_1 + SS_8 + SS_{14})$$

To membranes (4)

$$Q_4 \quad Q_3 - Q_6$$

$$TBOD_4 \quad 85\% \times TBOD_3$$

$$SBOD_4 \quad TBOD_4 - (SS_4 \times 75\% \times 0.56)$$

$$SS_4 \quad 75\% \times S_3$$

Effluent (5)

$$Q_5 \quad Q_4 - Q_{10} + Q_3 - Q_9$$

Comments

By varying this, and the loads, the impact on the removal can be investigated.

SS is assumed to contain 75% VSS, and it is assumed to be 0.56 mg VSS/mg BOD

With 45216 being Qdim.

SS is assumed to contain 75% VSS, and it is assumed to be 0.56 mg VSS/mg BOD

If the flow is less than Q_{dim} , the sedimentation is unused. If Q_{dim} is exceeded, the sedimentation step will be in use.

SS is assumed to contain 75% VSS, and it is assumed to be 0.56 mg VSS/mg BOD

SS is assumed to contain 75% VSS, and it is assumed to be 0.56 mg VSS/mg BOD

TBOD ₅	$TBOD_2 \times 10\% + TBOD_3 \times 30\%$	Adding the effluent from membranes and sedimentation.
SBOD ₅	$TBOD_5 - (SS_5 \times 75\% \times 0.56)$	SS is assumed to contain 75% VSS, and it is assumed to be 0.56 mg VSS/mg BOD
SS ₅	$SS_2 \times 5\% + SS_3 \times 20\%$	Adding the effluent from membranes and sedimentation.
Sludge from microsieve (6)		
TS ₆	<i>Based on pilot plant measurements</i>	
SS ₆	$25\% \times SS_2$	
Q ₆	$\frac{SS_6}{TS_6}$	
Out of thickener (7)		
TS ₇	<i>Assumed</i>	
SS ₇	$95\% \times SS_2$	95% solid capture.
Q ₇	$\frac{SS_7}{TS_7}$	
Reject from thickener (8)		
BOD ₈	$SS_8 \times 75\% \times 0.56$	SS is assumed to contain 75% VSS, and it is assumed to be 0.56 mg VSS/mg BOD
SS ₈	$SS_6 - SS_7$	
Q ₈	$Q_6 - Q_7$	
Sludge from sedimentation (9)		
TS ₉	<i>Assumed</i>	
SS ₉	$80\% \times SS_3$	
Q ₉	$\frac{SS_9}{TS_9}$	
Sludge from membranes (10)		
TS ₁₀	<i>Based on pilot plant measurements</i>	
SS ₁₀	$SS_4 + SS_{chem} - SS_5$	SS _{chem} is the chemical sludge production.
Q ₁₀	$\frac{SS_{10}}{TS_{10}}$	
Buffer tank (11)		
TS ₁₁	$\frac{TS_7 \times Q_7 + TS_9 \times Q_9 + TS_{10} \times Q_{10}}{Q_7 + Q_9 + Q_{10}}$	
SS ₁₁	$SS_7 + SS_9 + SS_{10}$	
Q ₁₁	$\frac{SS_{11}}{TS_{11}}$	

Sludge treatment
(12)

SSin ₁₂	SS ₁₁	
SSout ₁₂	SSin ₁₂ - (SSin ₁₂ × VSScontent ₁₂ × VSSreduced ₁₂)	SSin ₁₂ × VSScontent ₁₂ × VSSreduced ₁₂ being the VSS reduced in the sludge handling.
VSScontent ₁₂	Assumed	
VSSreduced ₁₂	Assumed	
Biogas yield	Assumed	
Methane	Biogas yield × 65%	
Biogas production	Biogas yield × (SSin ₁₁ - SSout ₁₁) × VSScontent ₁₂	
Methane production	Biogas production × 65%	
TSout ₁₂	$\frac{SSout_{12}}{Q_{11}}$	

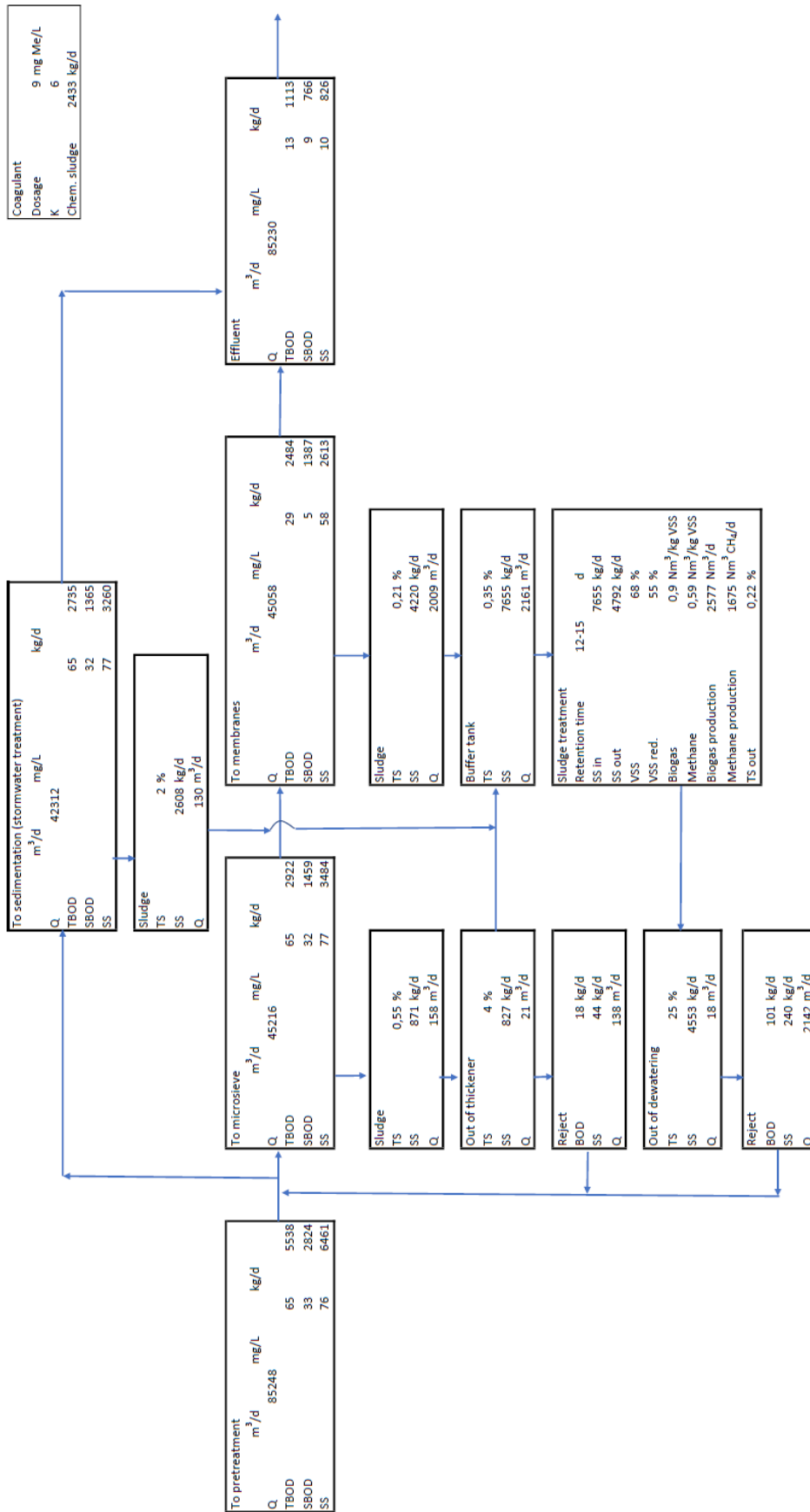
Out of dewatering
(13)

TS ₁₃	Assumed	
SS ₁₃	95% × SSout ₁₂	95% solid capture.
Q ₁₃	$\frac{SS_{13}}{TS_{13}}$	

Reject from dewatering (14)

BOD ₁₄	SS ₁₄ × 75% × 0.56	SS is assumed to contain 75% VSS, and it is assumed to be 0.56 mg VSS/mg BOD
SS ₁₄	SSout ₁₂ - SS ₁₃	
Q ₁₄	Q ₁₁ - Q ₁₃	

Appendix G



Appendix H

		Comments
To pre-treatment (1)		
Q ₁	<i>Q_{dim}</i>	
TBOD ₁	<i>BOD_{load}</i>	
SBOD ₁	$TBOD_1 - (SS_1 \times 75\% \times 0.56)$	SS is assumed to contain 75% VSS, and it is assumed to be 0.56 mg VSS/mg BOD
SS ₁	<i>SS load based on PE</i>	
To pre-settling (2)		
Q ₂	$Q_1 + Q_8 + Q_{13}$	
TBOD ₂	$TBOD_1 + BOD_8 + BOD_{13}$	
SBOD ₂	$TBOD_2 - (SS_2 \times 75\% \times 0.56)$	SS is assumed to contain 75% VSS, and it is assumed to be 0.56 mg VSS/mg BOD
SS ₂	$SS_1 + SS_8 + SS_{13}$	
To bioreactor (3)		
Q ₃	$Q_2 - Q_6$	
TBOD ₃	$80\% \times TBOD_2$	Assuming 20% removal of BOD in the pre-settling
SBOD ₃	$TBOD_3 - (SS_3 \times 75\% \times 0.56)$	SS is assumed to contain 75% VSS, and it is assumed to be 0.56 mg VSS/mg BOD
SS ₃	$50\% \times SS_2$	Assuming 50% removal of SS in the pre-settling.
To flotation (4)		
Q ₄	Q_3	
TBOD ₄	$(SS_4 \times 75\% \times 0.56) + SBOD_4$	SS is assumed to contain 75% VSS, and it is assumed to be 0.56 mg VSS/mg BOD.
SBOD ₄	<i>Assumed</i>	
SS ₄	$0.9 \times (TBOD_1 - TBOD_5) \times 0.6 \times \left(\frac{SS_3}{TBOD_3} + 1 \right) - \frac{0.72 \times 1.072^{T-15}}{B_{A,BOD} + 1.33 \times 1.072^{T-15}}$	
Effluent (5)		
Q ₅	$Q_4 - Q_9$	
TBOD ₅	$10\% \times TBOD_1$	Assuming 90% removal of BOD in the process.
SBOD ₅	$TBOD_5 - (SS_5 \times 75\% \times 0.56)$	SS is assumed to contain 75% VSS, and it is assumed to be 0.56 mg VSS/mg BOD.
SS ₅	$10\% \times SS_1$	Assuming 90% removal of SS in the process

Sludge from pre-settling (6)

$$\begin{aligned} TS_6 & \text{ Assumed} \\ SS_6 & 50\% \times SS_2 \\ Q_6 & \frac{SS_6}{TS_6} \end{aligned}$$

Out of thickener (7)

$$\begin{aligned} TS_7 & \text{ Assumed} \\ SS_7 & 95\% \times SS_6 & \text{95\% solid capture.} \\ Q_7 & \frac{SS_7}{TS_7} \end{aligned}$$

Reject from thickener (8)

$$\begin{aligned} BOD_8 & SS_8 \times 75\% \times 0.56 & \text{SS is assumed to contain 75\% VSS, and it is assumed to be 0.56 mg VSS/mg BOD} \\ SS_8 & SS_6 - SS_7 \\ Q_8 & Q_6 - Q_7 \end{aligned}$$

Sludge from flotation (9)

$$\begin{aligned} TS_9 & \text{ Assumed} \\ SS_9 & SS_4 - SS_5 + SS_{chem} & \text{SS}_{chem} \text{ is the chemical sludge production.} \\ Q_9 & \frac{SS_9}{TS_9} \end{aligned}$$

Buffer tank (10)

$$\begin{aligned} TS_{10} & \frac{TS_9 \times Q_9 + SS_7 \times Q_7}{Q_7 + Q_9} \\ SS_{10} & SS_7 + SS_9 \\ Q_{10} & \frac{SS_{10}}{TS_{10}} \end{aligned}$$

Sludge treatment (11)

$$\begin{aligned} SS_{in11} & SS_{10} \\ SS_{out11} & SS_{in11} - (SS_{in11} \times VSS_{content11} \times VSS_{reduced11}) & \text{SS}_{in11} \times VSS_{content11} \times VSS_{reduced11} \text{ being the VSS reduced in the sludge handling.} \\ VSS_{content11} & \text{ Assumed} \\ VSS_{reduced11} & \text{ Assumed} \\ \text{Biogas yield} & \text{ Assumed} \\ \text{Methane production} & \text{ Biogas yield} \times 65\% \\ \text{Methane production} & \text{ Biogas production} \times 65\% \end{aligned}$$

TS _{out11}	$\frac{SS_{out11}}{Q_{10}}$	
Out of dewatering (12)		
TS ₁₂	<i>Assumed</i>	
SS ₁₂	$95\% \times SS_{out11}$	95% solid capture.
Q ₁₂	$\frac{SS_{12}}{TS_{12}}$	
Reject from dewatering (13)		
BOD ₁₃		SS is assumed to contain 75% VSS, and it is assumed to be 0.56 mg VSS/mg BOD
SS ₁₃	$SS_{13} \times 75\% \times 0.56$	
Q ₁₃	$SS_{out11} - SS_{12}$	
	$Q_{10} - Q_{12}$	

