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Internship Report Evaluation & Optimization of the Brine Filtration at Nobian MEB

EFFICIENT AND COST-EFFECTIVE REMOVAL OF $Fe(OH)_3$

ABDULRAHMAN OMER S5806402

Date: 28/02/2026

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Acknowledgment

I would also like to thank Simon Snijders for his assistance during the onboarding process and for helping me to integrate smoothly into the company and its working environment.

Also, I would like to extend my sincere appreciation to my colleagues, Koen Weijers, Jelle Stavenga and Roy Postmus, in the process engineering department for their continuous support and valuable contributions throughout this project for their technical insights, guidance, and willingness to share their expertise during my internship.

I am also grateful to the maintenance and operations departments at Nobian Meb for their practical support and cooperation, as well as to all others who contributed to this work. This includes the R&D experts and the personnel at the analysis laboratory in Hengelo, whose analytical support and technical input were essential to the completion of this report.

Moreover, I would like to thank my academic supervisors, Dr. Vassilis and Dr. Dulce, for their support, trust and constructive feedback throughout the academic supervision of this internship.

Finally, I would like to express my great gratitude to my supervisor at Nobian, Alessandra Kerremans, for her guidance, encouragement, and continuous support during this internship.

Abstract

The Nobian MEB chlor alakli plant is facing increasing economic and operational pressure due to rising energy costs and aging equipment. Reliable brine purification is critical to protect the ion exchange membranes in the electrolyser and operate it as efficiently as possible. The existing precoat brine filtration system is approaching its end of service life and presents challenges related to operational complexity, waste generation, maintenance costs, and performance variability.

This study evaluates the current brine filtration system and investigates alternative filtration strategies to ensure consistent iron removal while reducing operational risks and expenditure. By examining the iron levels in the salt received it was determined that to be able to bypass the filter the salt received can't have an iron content of higher than 0.1mg Fe/kg salt. Also to keep using the filter costs the plant almost 100K€/year and a capital investment of 780K€. A membrane filtration technology that uses filters made from pPTFE was assessed as a potential replacement technology. The system has demonstrated stable filtration performance, effective iron removal, elimination of α-cellulose usage, and reduce waste generation, simpler operation through automated backwash and chemical cleaning. Although the membrane option requires high upfront capital, it promises to deliver lower operating costs, so that the extra investment can be retrieved in OPEX savings in about 3-8 years.

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1. Introduction

The chemical industry in the Netherlands is increasingly under structural pressure, driven by high energy costs (Dantuma, 2025), elevated regulatory burdens, weak demand, and tough global competition. Dutch chemical producers struggle with energy prices that remain high compared with neighboring countries (NL Times, 2025), squeezing margins and reducing the incentive to invest in new capacity or maintaining existing installations. At the same time, production levels have contracted significantly over recent years (De Nederlandsche Bank, 2025), with output still below previous peaks. These pressures have contributed to plant closures. The chlor alkali industry in the Netherlands, in integral part of the country's broader chemical sector that produces chlorine, caustic soda and hydrogen through highly energy intensive electrolysis process, is specially threatened by these challenges. Together, these pressures and challenges shape a competitive landscape where Dutch chlor-alkali plants must carefully balance between sustainability commitments, regulatory compliance and cost management to remain viable.

1.1. Nobian Delfzijl

The Nobian Delfzijl site consists of three production facilities Salt, Delesto and MEB (membrane electrolysis pant). Salt is the main raw material for the MEB chlor alkali process, so in essence the company manages an integrated value chain, see

Figure 1. From responsible extraction to tailored salt and brine product to secure high-quality feedstock. through washing, screening and recrystallization. Nobian's salt processing and brine purification systems are designed to remove particulates and dissolved impurities. Delesto operates as the central electricity and utilities (Steam, and process water) hub at the Delfzijl site. At the MEB site is where the chlor-alkali process is operated to produce chlorine gas, caustic soda, sodium hypochlorite (bleach) and hydrogen gas, so through this integrated value chain Delesto produces electricity and steam to supply both plants, salt produces the raw material for MEB, and MEB supplies back the hydrogen to Delesto as fuel and then sell the other products to the customers within the chemical park.

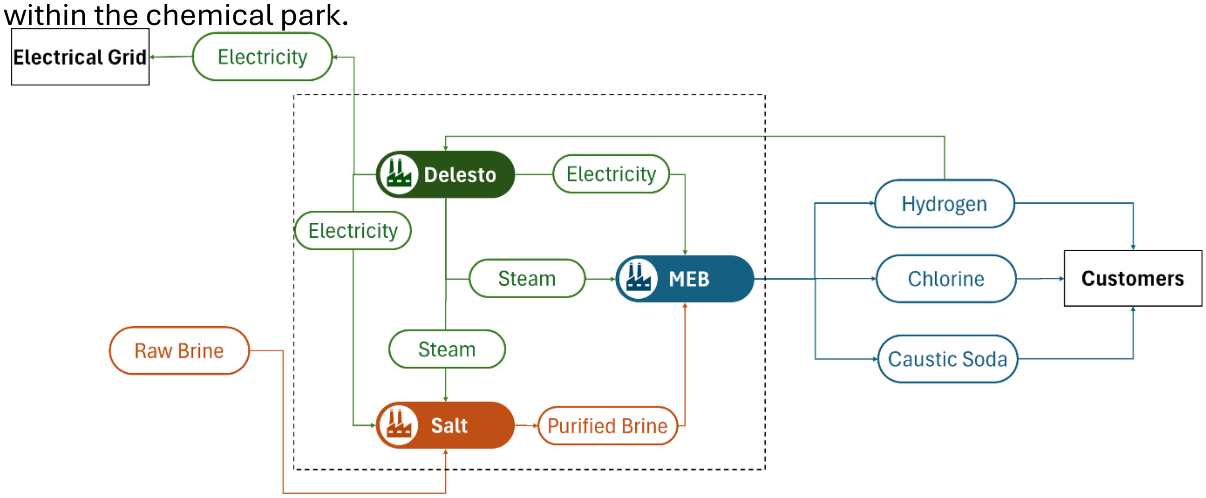

Figure 1: Nobian integrated value chain at Delfzijl

1.2. Chlor Alkali Process

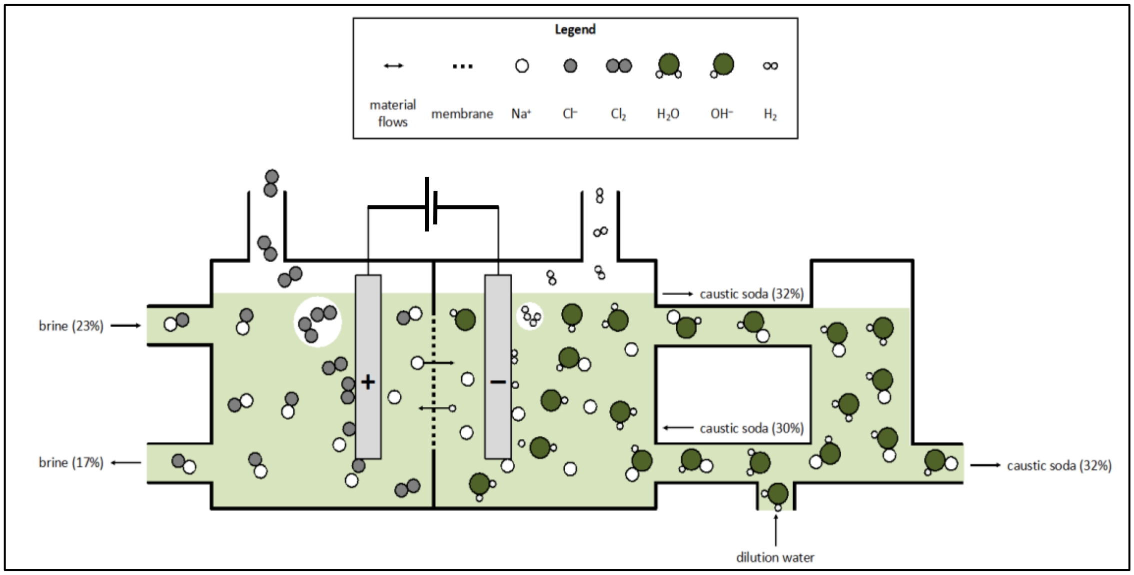

Figure 2: Membrane electrolytic cell

The Chlor-alkali process is used to produce chlorine and caustic soda, both are two essential chemicals used in a variety of industries such as water treatment, plastics and polymers pharmaceuticals and many more. and hydrogen is a byproduct of the process.

Three main technologies are used in the chlor-alkali industry, all depend on electrolysis. The mercury cell process, the diaphragm cell process and the most recent being the membrane cell process. All three processes rely on the electrolysis of brine feed to separate the Sodium and chloride ions from the solution by using electricity, therefore requiring substantial electrical energy. Although all three processes produce the same produces and have the same feedstock, The cell design and technology used lead to differences in energy consumption. Product purity, and environmental impact see Table 1the membrane process being the most energy efficient and the least impactful on the environment (Industrial Emissions Directive , 2014). In the membrane cell process, seen in Figure 2, A cation exchange membrane separates the anolyte and catholyte, only allowing the Na^+ to migrate from the anolyte side to the catholyte side, and simultaneously preventing any OH^- migration through it. Thus allowing for chloride ions to oxidize at the anode releasing chlorine gas (Cl_2), and electrons that pass through wire to the cathode, and the Na^+ combines with the OH^- in the catholyte compartment forming caustic soda NaOH, and the water splits at the cathode forming more OH^- in the catholyte and protons (H^+) which then reduce at the cathode to form hydrogen gas. See reactions (1) to (3)

$$2Cl^- \rightleftharpoons Cl_2 + 2e^- \qquad (E^o = 1.36\,V) \qquad (1) \quad \text{Anode half reaction}$$

$$2Na^+ + H_2O + 2e^- \rightleftharpoons 2H_2 + OH^- \quad (E^o = -0.83\,V) \qquad (2) \quad \text{Cathode half reaction}$$

$$\boldsymbol{2NaCl_{(aq)} + H_2O_{(l)} \leftrightarrow 2NaOH_{(aq)} + Cl_{2(g)} + 2H_{2(g)}} \qquad \textbf{(3)} \quad \textbf{(Overall reaction)}$$

The cell operation and production efficiency will depend on the cell voltage, current density and overall current efficiency. Where the current density (*J*) represents the amount of current applied to the electrochemical cell expressed as current per electrode area with directly correlates to the amount of product produced. The cell voltage (V_{cell}) is heavily dependent on *J*, the cell standard potential (V_o) and the resistances of the electrochemical cell (K), this relation is expressed in equation #

$$V_{cell} = V_o + (K \times J)$$

The current efficiency in the electrochemical cell describes the ratio between the actual amount of product, caustic soda, to the theoretical amount that should've been produced as predicted by Faraday's law.

Table 1: Comparison between different Chlor Alkali technologies (Nouryon, 2019)

	Mercury Cell	Diaphragm Cell	Membrane Cell
Caustic soda quality	High purity; very low NaCl content (<30 ppm). Mercury content present before treatment (5–150 µg Hg/L), requiring downstream purification.	Lower purity; contains 1.0–1.5 wt% $_3$. Not suitable for all applications without further treatment.	High purity; very low NaCl content (<50 ppm). Suitable for most demanding applications.
Caustic soda concentration	~50% directly from the process.	~12%; requires concentration to ~50% for many applications, increasing energy use.	32–33%; requires concentration to ~50% for some applications.
Chlorine quality	High quality; very low oxygen (<0.1%) and hydrogen content.	Lower quality; oxygen content typically 1.5–2.5%.	High quality; oxygen content typically 0.5–2%, depending on process conditions.
Brine quality requirements	Moderate brine purification required; tolerance to some impurities is relatively high.	Moderate brine purification required; impurity tolerance higher than membrane cells.	Very high brine purity required; trace impurities can significantly affect membrane performance.
Electrical energy use (ACkWh/t Cl_2)	High 3360 at 10kA/m^2	Medium 2720 at 1.7kA/m^2	Lowest 2650 at 5kA/m^2
Environmental impact	High environmental impact due to the use of mercury, risk of emissions to air, water, and waste streams.	Moderate environmental impact The diaphragm contains asbestos	Lowest environmental impact; no mercury or asbestos, lowest energy consumption.

1.3. Brine Circulation

The brine circulation at Delfzijl begins with the dichlorination of the anolyte (chlorinated brine), as chlorine can damage the equipment downstream. Chlorine is present as dissolved $Cl_{2(aq)}$ which is stripped by a 500Nm^3/h air stream over a packed bed column. Chlorine can also be present as HClO which is acidified with HCl to pH of around 2 to shift the reaction to the left and Cl_2 can be stripped out in the column.

$$Cl_2 + H_2O \rightleftharpoons HCl + HClO \quad (4)$$

residual active chlorine leaving the dichlorination column is then chemically reduced with sodium sulphite.

$$Na_2SO_3 + NaOCl \rightleftharpoons Na_2SO_4 + NaCl \quad (5)$$

In order not to let the levels of chlorate and sulphate become high, brine must be purged.

After dichlorination the brine pH is raised to about 9 with NaOH to precipitate ferric chloride as $Fe(OH)_3$

$$FeCl_3 + 3NaOH \rightarrow Fe(OH)_{3\downarrow} + 3NaCl \quad (6)$$

The iron hydroxide precipitate can then be removed in the precoat filter. The brine is then concentrated in the salt dissolver to the target salt concentration 305 mg/L. The clarified, alkaline brine then passes through ion-exchange columns to reduce $Ca2^+$, $Mg2^+$ and $Sr2^+$ to ppb levels. After ion exchange, the brine's pH is adjusted down to ~2 to meet electrolyser feed requirements, and the treated brine is ready for introduction into the electrolysers. (Nobian MEB)

1.4. Brine filtration

The brine filter is designed to remove precipitated metal hydroxides and other solids in the brine. The filter vessel is filled with candles and utilizes the use of filter aid in the form of an α-cellulose precoat. The cellulose forms a porous cake over the candles to capture all the precipitates and fine solids. The use of filter aid protects the filter candles, improves retention of colloids, and makes cleaning easier. Also, while in operation, a continuous dosing of α–cellulose called body feed. Which helps to replenish and reinforce the cake under higher or variable solids loads, extending service life between full recoats, however it increases the solids sent to waste. When the maximum differential pressure is reached or 2 weeks have passed, a recoat is needed. First the used cake slurry is sent to the waste brine filter press where the α-cellulose is removed and collected for landfill disposal, and waste brine is discharged into the sewer. (Nobian MEB)

2. Project: Challenges, Scope & Objectives

2.1. Challenges with current brine filtration process

The current brine precoat filtration system presents several operational, reliability and cost challenges. While the units in the precoat brine filtration system are reaching their end of life and it is apparent on their growing maintenance costs such as precoat tank, waste brine tank, and an aging filter press. Moreover, the ebonite (HRL) lining of the filter vessel will have to be changed, adding greatly to the CAPEX and the filter vessel will not be available for use for more than three months. Which poses a risk on the brine filtration capabilities for that period. Moreover, α-cellulose can migrate into the ion exchange tanks causing resin contamination (Vries, 2016) and performance loss, and if the precoat/body feed or cleaning sequence is not tightly controlled the α cellulose can foul or physically damage the filter candles (Pall Corporation, 2014), shortening their service life and risking unplanned downtime.

2.2. Project Scope & Objective

The project aims to identify a simplified, reliable, and low-waste filtration solution to ensure consistent brine quality. The objective is to select and validate an alternative filtration technology that meets the strict iron-removal specifications while reducing operational risk, minimizing manual intervention, lowering waste generation and lifecycle costs, and eliminating reliance on aging infrastructure, all without causing unacceptable interruptions to electrolyser operation.

3. Theory

3.1. Ion Exchange Membrane

The cation exchange membrane itself is central to the membrane cell process, where it facilitates the sodium ion transport into the catholyte, while preventing hydroxide ions from migrating back into the analyte. A breakthrough in the membrane technology came with the invention of perfluorinated ion-exchange membranes, Such as Nafion®, which are derived from Teflon chemistry (Thomas F. O'Brien, 2005). They exhibited high chemical and thermal stability, which made them a suitable alternative in the chlor alkali industry. Over time, membrane efficiency improved dramatically: from producing 10–15% caustic at low efficiency to today's capability of generating around 32% caustic at up to 97% efficiency. Research currently aims to achieve direct production of 50% caustic. The membrane mainly consists of a polymeric matrix containing a significant concentration of covalently bonded fixed ionic groups. In a cation exchange membrane (CEM) the fixed sites can be sulfonate (SO_3^-) or carboxylate (COO^-) acid groups. The perfluorinated membranes developed for the chlor alkali cell are designed to exhibit super selectivity, high thermal and chemical stability. (Nouryon, 2019)

In chlor-alkali cells, perfluorinated polymers form a phase-separated structure with hydrophobic (fluorocarbon) and hydrophilic (ionic) domains. Water-swollen ionic clusters create channels that allow cation transport. Various structural models have been proposed, but the Gierke Cluster-Network Model is the most widely accepted. It describes the membrane as having spherical ionic clusters (~3–5 nm) connected by narrow channels (~1 nm) (Sutton, 2016). Cluster

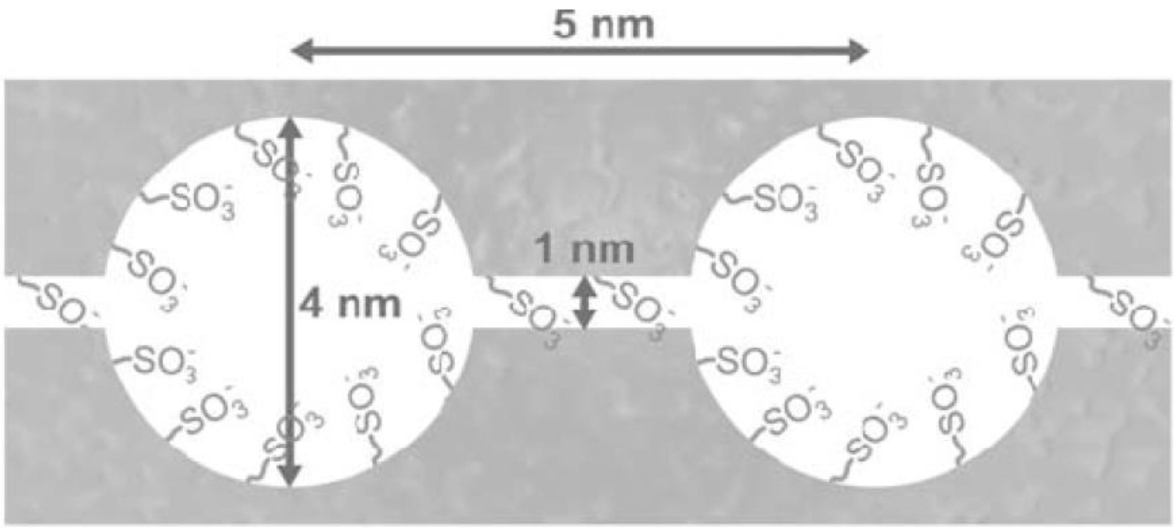

Figure 3: Gierke cluster model (Sutton, 2016)

size depends on equivalent weight, ion type, water content, and temperature. The carboxyl ionomer gives superior current efficiency and facilitates making strong caustic (32%) in the electrolyser (Thomas F. O'Brien, 2005). However, due to its greater cost, lower conductivity, and susceptibility to become nonconductive at lower pH, the carboxylate ionomer is generally used with the sulfonate ionomer in a bilayer membrane, with the perfluoro sulfonate

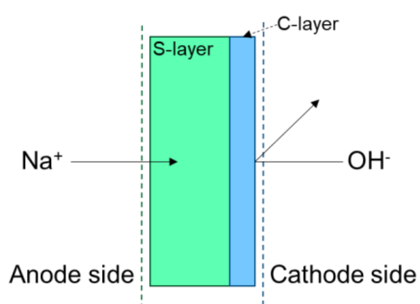

Figure 4: Illustration of CEM layers

(S-layer) layer facing the anode and the perfluoro carboxylate (C-layer) facing the cathode see Figure 4 Where the S-layer facilitates the movement of the sodium ions to the catholyte, and the C-layer prevents the hydroxide ions from moving int the anolyte side due to its higher electrical resistance, thus significantly improving the cell efficiency see Table 2

Table 2: Properties of S& C layers of CEM

Sulfonic (S-layer)	**Carboxylic (C-layer)**
Lower electrical resistance	Relatively higher resistance
Lower selectivity for cations	Higher selectivity for cations
Resistant to anolyte (Cl_2)	Resistant to catholyte (NaOH)

3.1.1. Effect of brine impurities

Impurities in the brine reach and affect the ion-exchange membrane through different transport mechanisms. Cations are driven by the applied electric field, neutral species accompany the water flow, and anions are also carried by water flow but are retarded by the membrane's negative charge. These impurities can either accumulate on the surface, get entrapped in the membrane matrix, or pass into the product stream depending on their charge and solubility. Precipitate salts, hydroxide based are the most common, depending on their solubility in water will either accumulate on the membrane's surface if they're poorly soluble, or get entrapped and crystalize in the membrane Figure 5 where they can cause severe increase in resistances and hence the cell voltage. Impurities that deposit near the cathode can reduce the current efficiency effects are cumulative, so trace amounts over the membrane's life can produce significant degradation (Ewart, 2007). Precipitates can block Na+ transport, alter electrode reactions and increase overpotentials, contaminate the product, and cause mechanical damage, leading to lower throughput, higher energy use, poorer product quality and shorter membrane life. Therefore, upstream removal and pH control, in the brine treatment are important mitigation techniques.

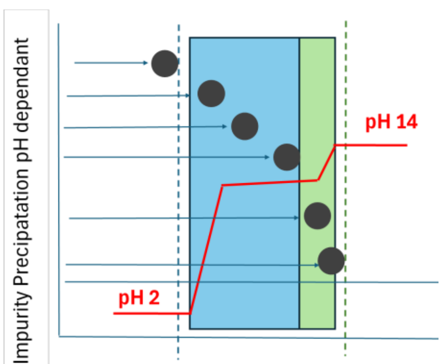

Figure 5: Precipitation through the membrane (Ewart, 2007)

3.2. Precoat filtration

In the brine filtration process, the removal target is ferric hydroxide, $Fe(OH)_3$, which forms a colloidal, gel-like precipitate with an effective particle size typically well below 1 µm and a strong tendency to deform and blind, blocking the pores of conventional filter media (Trevor Sparks, 2015). To manage these unstable solids, the process utilizes dead-end ceramic candle filtration, in which the entire volumetric flow is directed orthogonally toward the filter surface, forcing all suspended solids toward the medium and promoting progressive cake formation; as the cycle proceeds, increasing cake thickness becomes the dominant resistance element and principal filtration layer. Because $Fe(OH)_3$ alone produces an impermeable, gelatinous mass rather than a porous cake, the candles are first coated with an α -cellulose precoat, which provides a highly permeable structure with an effective particle retention capability of approximately 0.5 µm, preventing immediate blinding of the ceramic substrate. A continuous body-feed of cellulose fibers is introduced into the brine to mechanically condition and "structure" the gel precipitate, enabling agglomeration, improved permeability, and enhanced interaction, and local adsorption within the depth of the forming cake. During filtration, the cellulose/$Fe(OH)_3$ cake evolves into a robust, three-dimensional depth-filtration matrix that effectively retains ultrafine iron species while maintaining acceptable hydraulic conductivity under the imposed pressure gradient. The integration of dead-end operating mode, ceramic support structure, α-cellulose precoat, and ongoing body feed conditioning allows the system to consistently polish brine to <20 ppb total iron, even when handling highly colloidal, deformable ferric hydroxide precipitates with otherwise unfilterable characteristics.

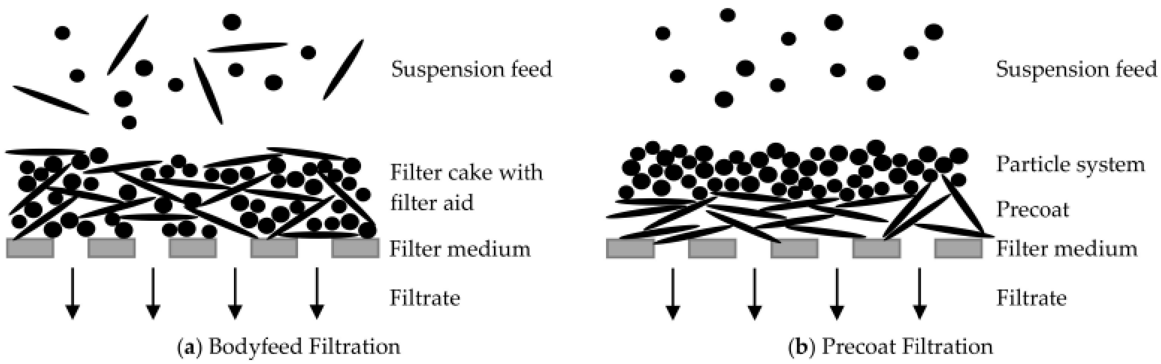

Figure 6: illustration highlighting the difference between bodyfeed & precoat filtration (Volker Bächle, 2021)

3.3. Ion exchange resin

An ion exchange (IOX) system consists mainly of using a packed bed of chelating resin that is effective in removing divalent cations, namely Ca_2^+, Mg_2^+ & Sr_2^+. The ion exchange process involves the reversible ion swap with a liquid without any permanent structural change. Typically, the resin bound alkali ions (Na^+) are replaced by divalent alkaline earth ions from the solution. The main parameter to govern the exchange process is the operating capacity, the number of exchangeable ions a resin can remove under specific conditions, depending on factors such as brine flowrate, pH, temperature & inlet brine ion concentration. After all the available ions are exchanged, when operational capacity is reached, the resin can be regenerated by an acid cycle that replaces the bonded ions with H^+ which is then washed with caustic to bring the resin back to its active form.

The objective of the ion exchanger is to bring down the divalent ion concentration in the brine to below 20 ppb. The main reason for maintaining a pure brine is to prevent anode and membrane fouling, accelerated membrane degradation, rising cell voltage and decreased cell efficiency. The two main types of chelating resins used are Iminodiacetic acid (IDA)and Aminomethylphosphonic acid (AMPA). The latter usually have 20% more capacity under ideal conditions (Korbacs, 2020). However, it is more sensitive to iron positioning, where it can irreversibly bind to iron in the brine.

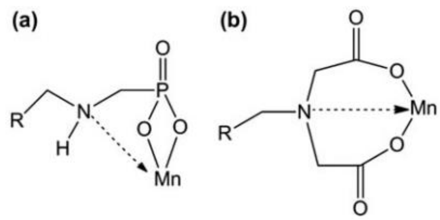

Figure 7: Removal mechanism of Mn a. IDA b. AMPA (Gulay Coskun, 2015)

4. Methodology

The methodology adopted in this study combines current process assessment, mass balance calculations, targeted brine sampling and cost evaluation to determine the current operating conditions both technically and economically. Each methodological step was designed to quantify the performance limitations of the current filtration system, identify the allowable impurity levels in the salt feed, and evaluate both technical and economic feasibility of potential improvement measures. By integrating process data, laboratory analyses, and operational constraints, this methodology provides a structured and framework for recommending the most reliable and cost-effective solution for long-term brine quality control. and as to serve as a baseline for comparisons to proposed solutions

4.1. Current Situation Assessment

The brine filtration system operates with 444 filter elements providing a total filtration surface area of 83.5 m^2. Under standard conditions, the incoming brine should contain no more than 1 ppm of suspended solids, with a maximum particle size of 50 µm. When filter aid is applied, the filtration step improves brine clarity further, achieving less than 0.5 ppm for particles larger than 0.3 µm. At an operational flow rate of 160 m^3/h, with a filtration flux of approximately 1.92 $m^3/h.m^2$ based on the available surface area. The filter vessel is engineered for demanding process conditions, with a design pressure of 9 bar, brine pH between 9 and 10, a high salinity level of around 305 g/L, and a maximum allowable flowrate of 170 m^3/h (Pall Corporation, 2006). Maintaining this level of filtration efficiency is essential to protect the downstream ion exchange resin, which is particularly sensitive to iron (Fe) and relies on effective upstream solid removal to maintain long-term performance.

4.2. Mass Balance

A mass balance was carried out to estimate the iron concentration in the salt that would still allow the brine to meet specification limits in the electrolyser feed. This assessment also provides an approximation of the operational iron-removal capacity of the brine filter. Because no permanent sampling point exists to directly measure the Fe concentration in the brine entering the filter, the mass balance was constructed around the salt dissolver tank (inlet) and the brine filter outlet (product) streams.

Between 2017 and 2025, the salt used on site showed an average iron content of 0.2 mg Fe/kg salt, with peak values up to 0.3 mg Fe/kg salt. The recycled brine entering the dissolver contains approximately 205 g/L salt, which is then concentrated to 305 g/L in the salt dissolver. The data for the mass balance Table 3 were obtained from samples collected between 3–7 December 2025, and the corresponding flow rates were extracted from Trendminer®. Figure 8 illustrates the mass flows included in the balance.

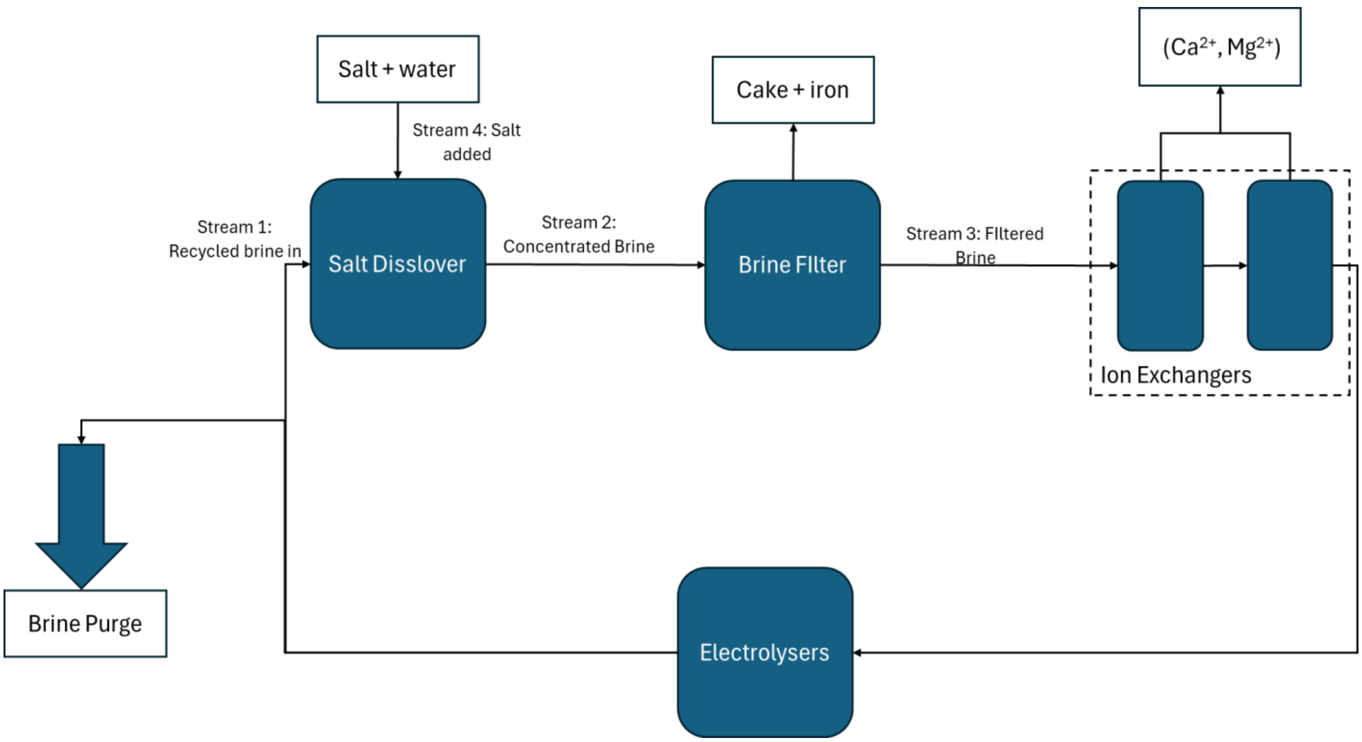

Figure 8: Brine cycle PFD with streams intended for the mass balance

Table 3: mass balance table

	Stream 1	Stream 2	Stream 3	Stream 4 (salt)
Flowrate ton/h	93.0	108.2	108.2	
Salt Concentration g/l	170.5	305.0	305.0	
Fe in Brine mg/l	0.006	0.037	0.006	
Fe in Salt mg/kg				0.2
Brine Density kg/m^3	1170	1170	1170	
Salt amount kg/h	13553	28206	28206	14653
Fe amount mg/h	477	3408	555	2931

Based on the mass balance results, using salt containing 0.2 mg Fe/kg salt increases the iron concentration in the electrolyser feed brine to approximately 37 µg/L, which exceeds the site's critical limit of 20 µg/L. To maintain the brine within specification, the maximum allowable iron content in the salt should therefore not exceed 0.1 mg Fe/kg salt.

Furthermore, the mass balance demonstrates that the brine filter can remove a substantial portion of the dissolved iron, keeping the brine within acceptable limits under current operating conditions. Any alternative filtration or pre-treatment system would need to achieve an iron concentration removal from about 60 µg/L down to below 10 µg/L match or exceed the performance indicated in this assessment.

4.3. Addition of a sampling point before the filter

To further evaluate the performance and efficiency of the brine filter, an additional sampling point was installed directly before the filter (F8201). This allowed a direct comparison between the brine quality entering the filter and the brine quality after filtration. By doing so, it became possible not only to quantify the filter's iron-removal efficiency but also to observe fluctuations in iron loading over time (e.g., resulting from cavern production, brine quality variation, or filter jacket condition).

Brine samples were collected over the period 28 January – 16 February, directly before and after the filter. The iron concentrations and calculated removal efficiency are summarized in Table 4:Fe concentrations before & after filter The data show that the filter is capable of removing iron effectively (80–92%) when the inlet concentration is sufficiently high. However, on several days the removal efficiency drops significantly (<25%), indicating periods of low incoming iron concentration, possible partial filter loading, or variation in filter performance.

Table 4:Fe concentrations before & after filter

Date	Fe in raw brine	Fe in filtered brine	Fe removed	Filter Efficiency
Units	mg/l	mg/l	Mg/l	
28/01	0.113	0.011	0.102	90%
30/01	0.069	0.013	0.056	81%
02/02	0.382	0.029	0.353	92%
04/02	0.035	0.032	0.003	9%
06/02	0.022	0.017	0.005	23%
09/02		0.012		
11/02	0.092	0.083	0.009	10%
12/02	0.088	0.011	0.077	88%
16/02	0.016	0.013	0.003	19%

Analytical Method Used

The analytical procedure followed the method described in the laboratory report "Fe analysis brine before filter F8201".

The key steps were:

Sample pretreatment

- 32.5 mL of brine taken into a 50 mL digestion tube
- 1 mL of 70% HNO_3 added
- Samples acidified and boiled using a Digi prep block
- Dilution to 50 mL with metal free water, then further diluted (10×, 100×, 1000×)

Elemental analysis

- Measurement performed using ICP OES according to AMH 4649
- Instrument: Agilent 5110 ICP OES

This method ensures complete dissolution of any particulate or undissolved iron present in the samples, especially in pre filter brine sample where iron may be partly present as solids (Koppleman, 2026).

Discrepancy Between This Report and Operational Sample Values

A noticeable discrepancy was found between the iron values reported in the laboratory's "after-filter" sample set and the concentrations normally recorded in STARLIMS®. This difference can arise from the sample pretreatment methods: for the report, all samples were acidified and boiled, where as the standard involves acidification only. still requires further investigation to whether which pretreatment method is more accurate, and whether the boiling step will need to be adapted to all brine samples taking. Moreover, on several days the filter exhibited quite low performance, the filter operating conditions and regeneration times, in this period were assessed in Trendminer® to examine if this filter related, but the trends show nothing out of the ordinary.

4.4. Expenditure Analysis for the current brine filtration system

This section presents a comprehensive economic assessment of the costs associated with the relining and refurbishment of the existing brine filtration system. The analysis covers the major units listed below and their auxiliaries

1. The brine filter vessel (F-8201)
2. The precoat tank V-8203
3. The Waste brine Tank V-8204
4. The Filter press S-8206

The analysis includes both the capital expenditure of refurbishment and operational expenditure (OPEX) for routine maintenance cellulose consumption & filter replacement.

4.4.1. Annual maintenance Expenditure

Retrieving the maintenance data for the past 10 years from SAP® estimates the average annual costs for the brine filtration system to be around €54,000 see Table 5 for breakdown.

Table 5: brine filtration maintenance costs last 10 years

Unit	Total Maintenance Cost per Year (€)
F-8201	314,767.34
V-8204	65,930.05
V-8203	64,041.21
S-8205	95,907.02
Total	540,645.62
Total costs/yr (10 year span)	54,064.56

4.4.2. CAPEX for the refurbishment of the brine filtration system

Since the brine filter is nearing its end of life, and per advice from the last inspection the system will require to be refurbished, including relining the brine filter vessel with new ebonite (HRL) lining, and the filter press will require to be replaced see Table 6 for breakdown. All the values were obtained from Nobian RDM preliminary study to upgrade their brine filtration system

Table 6: expected CAPEX costs for refurbishing to filter system

Unit	CAPEX Refurbishing (€)
F-8201	100,000
V-8204	154,000
V-8203	154,000
S-8205	325,000
Civil Work	50,000
Total CAPEX	783,000

4.4.3. Operating Costs Related to Cellulose Consumption

α-cellulose is consumed during each filtration cycle as both precoat and bodyfeed material. The unit cost breakdown is:

- Precoat: 61.25 kilo per cycle
- Bodyfeed: €8.75 kilo per cycle
- Number of cycles per year: 26
- Total annual cellulose cost: **€ 12,432.45**
- Cost of cellulose per kilo: 6.81 €/kilo
- Cost per cycle: € 477.00 (including handling overheads)

4.4.4. OPEX for Filter Replacement (8-Year Interval)

Filter internals and associated media require renewal approximately once every eight years. The associated periodic OPEX is:

Filter replacement cost of 444 filter elements €244,22 (includes purchase and installation costs)

This cost can be calculated for budgeting:

Table 7: Projected OPEX for keeping the current filter system

Cost Category	Value (€ / year)
Maintenance (10-year average)	54,064.56
Cellulose OPEX	12,432.45
Filter Replacement OPEX (every 8 years)	30,525.00
Total OPEX per Year	97,022.01

4.4.5. Economic results

The relining and refurbishment of the current brine filtration system requires an upfront CAPEX investment of €k783, supplemented by a predictable annual OPEX of approximately €k97 per year. The primary cost drivers are the maintenance of the filter units—particularly F-8201—while cellulose consumption constitutes a minor recurring cost.

Both the CAPEX and OPEX will serve as a reference in decision making in choosing a different alternative.

4.5. Evaluation Criteria for Solution Assessment

To ensure a clear and objective comparison of the proposed solutions, the following evaluation criteria have been defined based on the project's technical, operational, and financial priorities. Each criterion reflects a key performance area and is weighted according to its relative importance in the final decision. Table 8 shows the criteria with their assigned weights

Table 8: important criteria to look after when evaluating

Criterion	Weight	What to Evaluate
Cost & ROI	40%	ROI and payback period will be based benefits of reduced OPEX (compared to baseline)
Filtration Effectiveness	30%	Ability to meet or exceed filtration requirements (filtration area, filtration flux, lab analysis, pilot studies)
Reliability & Risk Mitigation	10%	reduction of operational risks (downtime, contamination, variability). Robustness under off-spec conditions.
Integration & Installation	15%	Ease of fit with current equipment, piping, controls, and utilities
Waste Reduction	5%	Reduction in brine and dry cake wastes

4.5.1. Cost & ROI

The cost is the primary decision driver, as it would measure the total cost of ownership of each alternative compared to the baseline expenditure. The ROI would be assessed on operational cost benefit against keeping the current filter, with an expected payback period of 3 years only. And this metric would help quantify the comparison between the different alternatives.

4.5.2. Filtration Effectiveness

This criterion measures how well each solution meets or exceeds the required filtration performance. It includes the system's ability to consistently achieve target quality specifications under normal and variable operating conditions. Key considerations include particle removal efficiency, fouling tendencies, stability of performance over time, and the impact on downstream process quality. Because performance consistency is critical, the preferred solution should be capable of meeting filtration requirements most or all of the time with minimal variance.

4.5.3. Integration & Installation Complexity

This area assesses how easily the proposed solution can be integrated into the existing process environment. It considers installation time, required downtime, compatibility with existing equipment, piping, controls, and utilities, as well as the need for additional engineering or hardware modifications. Training and ease of adoption for operators also factor into this criterion. A solution that installs quickly and interfaces smoothly with the current infrastructure is favored because it minimizes disruption and implementation cost.

4.5.4. Reliability & Risk Mitigation

This criterion evaluates the long-term dependability of each option, including equipment reliability, maintenance requirements, and the supplier's ability to provide ongoing support. It also considers how well the solution reduces operational risks such as unexpected downtime, contamination, performance instability, or safety hazards. A more reliable system reduces operational complexity and mitigates risk by ensuring stable, predictable performance under normal and off-spec conditions.

4.5.5. Waste Reduction

This measures the extent to which each solution reduces the volume and frequency of waste generated during operation. It includes reductions in disposable filters, consumable materials, and associated handling or disposal costs. While this is the least critical of the evaluation criteria, waste reduction is still valuable because it supports sustainability goals and reduces the environmental footprint of the process.

5. Proposed Solutions

Three scenarios will be evaluated to determine the most suitable solution. Scenario 1 considers installing a new membrane filter. Scenario 2, which serves as the baseline case, involves refurbishing the existing precoat filter. Scenario 3 evaluates the option of eliminating the filtration process entirely.

5.1. Back pulse membrane filtration

The existing filter relies on filter aids, such as α-cellulose, because the filter pores are larger than some of the solid particles in the brine. Hence a filter aid layer must first be formed to capture the finer particles in the brine and prevent blinding of the filter surface. A viable alternative is back pulse membrane filters as they eliminate this complexity by using media with uniformly small enough pore sizes to retain solids directly on the membrane surface without the need for a precoat or a filter aid. Newer membrane filters have the advantage of providing a clear filtrate from the beginning of operation. Only forming a thin cake layer on the membrane surface, resulting in minimal cake compression and easy removal of solids through backwashing (back pulse). This technology has the potential to go beyond replacing polishing filters. Ultimately, it may enable direct filtration of treated brine in a single step, combining the roles of clarifiers, primary filters, and polishing filters into one membrane filtration unit. (Thomas F. O'Brien, 2005)

Back pulse filtration was evaluated as an option to replace current precoat filtration, The Profilco® membrane will be evaluated in this study. The filters are made from PTFE and do exhibit a high filtration area per volume. During operation, $Fe(OH)_3$ forms a layer on the membrane, increasing the pressure drop. When either the pressure drop reaches a defined limit or a maximum filtration time is met, the system automatically initiates a backflow sequence to remove the built-up sludge. The system will be fully automated, controlling valves and sequences, including chemical cleaning with 10% HCl and the necessary pre- and post-flushes.

The current MEB plant has a max filtration capacity of 170 m^3/h of brine feed based on current precoat filter design and operation (Pall Corporation, 2006)The proposed membrane filters can be retrofitted into the existing 148 assembly heads, and each candle is fitted with 13 filter tubes of 15mm diameter and length of 1850 mm (Profilco) having an active filtration area of 1.13 m^2 adding up to a total filtration area of 167m^2 almost double of the precoat filter. And a filtration flux of around 1 $m^3/h \cdot m^2$.

The results indicate that the back-pulse membrane filtration system significantly outperforms the existing precoat filter in terms of filtration flux and stability.
While the precoat system exhibits an average filtration flux of approximately 1.92 $m^3/h \cdot m^2$, its performance is highly dependent on cake formation, cellulose quality, and operational conditions. In contrast, the membrane system provides a more consistent and predictable flux of around 1.0 $m^3/h \cdot m^2$ for the same filter size, but benefits from twice the available filtration surface area, leading to a higher total throughput and improved overall stability.

Moreover, the membrane filter offers constant filtration performance from startup, unlike the precoat filter, which requires a full precoat cycle to reach effective retention levels. Sampling results of the existing system show high variability in removal efficiency (as low as 9% on certain days), indicating sensitivity to inlet conditions and filter age. Lab scale (J.B. Westerink, 2024) testing of the back-pulse membrane system. demonstrated effective iron removal efficiency, even with higher iron content with reliable filtrate clarity and no dependency on precoat buildup.

One of the most notable advantages of the membrane system is the elimination of operational steps tied to cellulose use.
The back-pulse filter does not require a precoat tank, does not require a filter press, and does

not rely on cellulose consumption—removing key sources of manual intervention, equipment failures, and waste generation. The cleaning, back-pulse cycles, and chemical regeneration are fully automated, reducing operator workload and minimizing process disruption.

However, the membrane elements must be replaced more frequently; every three years, for the Profilco® filters, compared to the eight-year interval of the precoat filter candles. Despite the higher replacement frequency, the overall waste volume remains considerably lower because no cellulose waste is generated, and sludge production is reduced.

While the membrane system fits into the existing vessel footprint, its installation does require two additional utilities:

1. A hydrochloric acid (HCl) line for automated chemical cleaning (10% HCl).
2. A pressurized air line to execute the back-pulse cycle.

These additions represent minor modifications to the current brine filtration infrastructure but are necessary for proper operation. Integration complexity remains moderate and does not require long downtime, in contrast to the precoat system refurbishment, which would keep the vessel offline for more than three months.

This alternative demonstrates clear differences in technical performance, operational complexity, risk profile, and economic viability. The results from brine sampling, mass balance calculations, pilot trial insights, and cost modelling provide a comprehensive foundation for comparing these alternatives.

5.2. Economical Assessment

The Profilco® filter alternative represents a replacement solution with a total installed capital cost of €997,890. In terms of operational expenditure, the Profilco® system incurs low recurring costs, driven mainly by the scheduled replacement of filter elements every three years, averaged to €69,066.67 per year, along with minor chemical costs for HCl . This results in a total annual OPEX of €71,464.72, see Table 9 for the breakdown, positioning the Profilco® option as a high-CAPEX but simplified operating solution. The economic profile suggests a system designed for long equipment life and reduced operational complexity, with most lifecycle costs concentrated in upfront investment rather than ongoing maintenance.

Table 9: Economical evaluation of installing Profilco filters

CAPEX		
Tubesheet	€	50,000
Filter vessel	€	180,000
Filter Elements MT22/1700	€	185,000
Piping	€	40,000
Civil	€	15,000
Misc for vessel	€	61,500.00
Automation	€	22,883.33
Engineering	€	138,595.83
Project Mngmt	€	55,438.33
Contingency	€	83,157.50
Installation	€	166,315.00
Total	**€**	**997,890.00**
OPEX		
HCl cost	€	1,336.79
DI water cost	€	1,061.26
Filter elements (changed every 3 years)	€	69,066.67
Total/yr	**€**	**71,464.72**

5.2.1. Sensitivity to capital expenditure and payback potential

The option to implement a back-pulse membrane filter is highly sensitive to CAPEX, given its upfront cost of approximately €998k. However, the system offers lower OPEX due to no cellulose consumption, no requirement for a filter press, lower waste disposal costs, faster regeneration.

the back-pulse membrane options all reduce annual operating cost by €25,557 (primarily by eliminating precoat/press steps and related maintenance), but their business case is highly sensitive to the capital envelope Table 10.

Table 10: expected ROI with different filter vessel material

	CAPEX	OPEX benefit	ROI in years
Refurbish	€ 783,000.00	Baseline	
Profilco CS/HRL	€ 997,890.00	€ 25,557.29	8.4
Profilco FRP	897,890.00	€ 25,557.29	4.5
Profilco retrofit	867,890.00	€ 25,557.29	3.3

The economic modelling shows that the ROI and payback period become highly attractive when alternative construction materials are considered. It is possible to reduce payback period significantly (from 8 to 3 years) just by altering different vessel material. Using vinyl ester-lined FRP vessels or retrofitting the existing filter vessel for membrane installation can substantially reduce capital costs Figure 9shifting the financial balance in favor of the membrane solution. These modifications leverage a strong OPEX reduction to improve payback performance compared to the baseline option of refurbishing the precoat system.

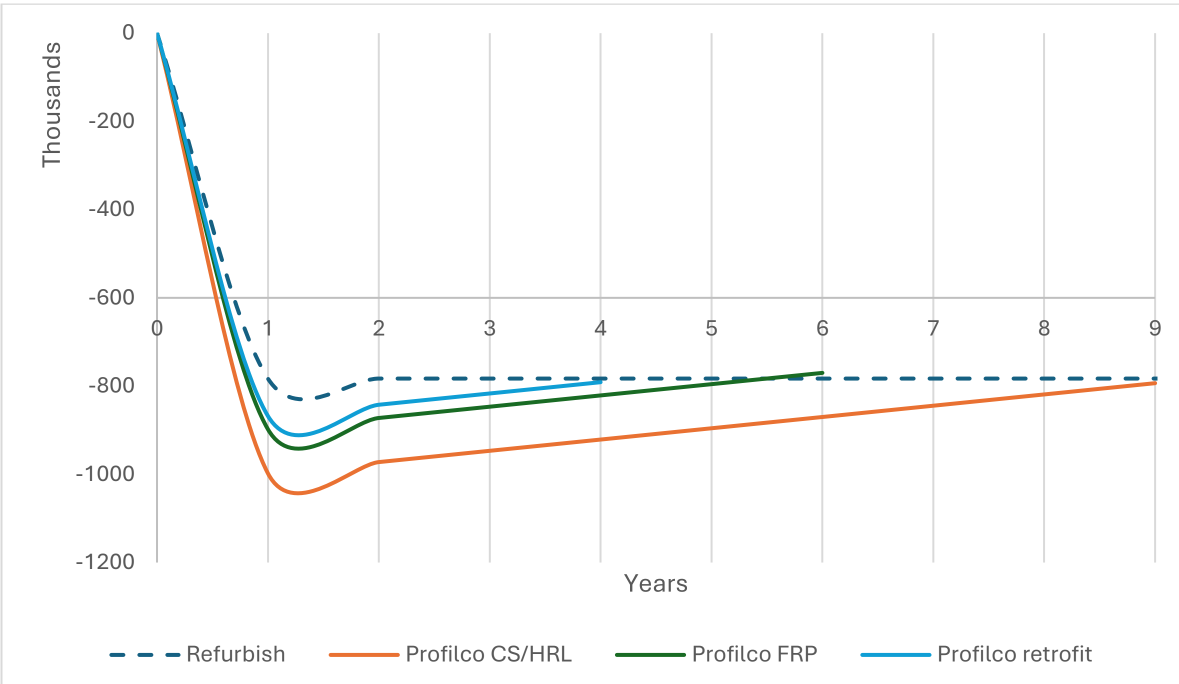

Figure 9: Payback period chart

6. Results and Discussion

The proposed scenarios were evaluated in this project, which are

- Installation of a new back pulse membrane filtration system
- Refurbishing of the existing precoat brine filtration system – baseline
- Bypassing the brine filtration step entirely

6.1. Scenario 1: Installation of the New Back-Pulse Membrane Filter

Installing the new back-pulse membrane filtration system represents a high-CAPEX investment, but it also brings substantial operational and technical advantages that position it as a strong long-term solution. The system offers a noticeably lower OPEX due to the elimination of cellulose consumption, reduced maintenance requirements, and the absence of a filter press and precoat infrastructure. Its filtration performance is both consistent and immediate, with stable clarity achieved from the start of each filtration cycle, unlike the precoat system, which requires time to form a functional cake layer. The pilot trial demonstrated promising results, confirming the technology's reliability, its excellent iron-removal efficiency, and its suitability as current Best Available Technology (BAT) for brine treatment. In addition, the back-pulse filter produces significantly less waste because no cellulose is used and sludge volumes are lower, improving both environmental footprint and handling costs. The system's regeneration time is shorter and fully automated, simplifying operation and reducing the dependency on manual process steps. However, the technology does require a slight modification of the existing setup, including the installation of a hydrochloric acid line for chemical cleaning and a pressurized air line to conduct the back-pulse sequence, and the membrane elements need to be replaced more frequently; every three years. Despite these considerations, the membrane system offers clear advantages in process stability, cleanliness, and long-term operability.

6.2. Scenario 2: Refurbishing of existing precoat filtration system

Refurbishing the existing precoat filtration system presents the advantage of requiring the lowest CAPEX among all equipment-replacement options, while relying on a technology that is already well understood, widely used in chlor alkali industry (Thomas F. O'Brien, 2005), and proven capable of filtering $Fe(OH)_3$ and suspended solids under normal operating conditions. However, this option also carries several limitations. The filtration performance during startup is often inconsistent due to the time required for the precoat layer to form properly, and sampling results have shown that removal efficiency can be highly variable, with several days exhibiting less than 25% iron removal. Operational complexity remains high, as the system depends on multiple auxiliary units, frequent manual intervention, and the handling and dosing of α-cellulose, which additionally poses the risk of resin contamination in the ion-exchange units. Recurring costs associated with cellulose consumption, routine maintenance, and filter press operation are significant, and the system's aging infrastructure means that refurbishment would come with an extended downtime, particularly for relining the filter vessel and replacing the filter press.

6.3. Scenario 3: Bypassing the Filter Entirely

Bypassing the brine filter altogether presents the most economically attractive option, as it requires no major capital expenditure and results in dramatically lower operating costs. This approach has been temporarily applied in the past during filter breakdowns Appendix , with no immediate adverse effects observed on the overall process, provided that the incoming salt supply contains extremely low impurity levels. Indeed, when the salt received contains iron concentrations up to approximately 0.1 mg Fe/kg, filtration may not be strictly necessary. However, this scenario is highly dependent on salt quality, which is not consistently guaranteed, and therefore introduces a substantial degree of operational risk. Over time, even trace levels of iron can accumulate within the system and potentially reduce efficiency and promote a higher risk of increasing the cell voltage of electrolyser membranes. The current resin type used in the ion exchange softeners is particularly sensitive to iron, meaning that bypassing filtration could shorten resin life, reduce Ca/Mg removal capacity, and increase regeneration frequency and replacement costs. Although no immediate operational disruptions may be visible, iron contamination tends to build up slowly, potentially compromising downstream processes and affecting membrane performance in the electrolyser. Therefore, while this option minimizes both CAPEX and OPEX, it significantly increases the risk of long-term damage caused by impurities and is heavily reliant on consistently high salt purity, which makes it the least robust option from a process-reliability standpoint.

6.4. Overall Discussion & Criteria Matrix

The results clearly show that the back-pulse membrane filtration system provides the strongest combination of performance, operational simplicity, and long-term cost reduction. It improves filtration flux, stabilizes performance, eliminates cellulose-related issues, and drastically simplifies operations. Supported by the highest score in the criteria matrix (4.15) Table 11. Although the initial investment is high, strategic reductions in CAPEX—through material selection or vessel retrofit—substantially enhance the ROI and payback period. Refurbishing the existing precoat system offers a lower initial cost but fails to resolve the underlying issues of variability, manual workload, waste generation, and operational complexity. Bypassing the filter offers compelling short-term economic benefits but exposes the process to unacceptable long-term risks related to iron contamination and resin degradation.

Table 11: Criteria matrix scores

		Score 1-5		
Criterion	Weight	Scenario 1	Scenario 2	Scenario 3
Cost & ROI	40%	3	4	5
Filtration Effectiveness	30%	5	3	1
Reliability & Risk Mitigation	15%	5	4	1
Integration & Installation	10%	5	5	5
Waste Reduction	5%	4	1	5
Weighted Average	**100%**	**4.15**	**3.65**	**3.20**

7. Conclusions & recommendations

After evaluating the performance, reliability, and economics viability of the brine filtration system at Nobian MEB. Through focusing on ensuring a consistent brine quality for the membrane electrolyser, while minimizing operational risks and expenditures. The research confirms that effective upstream removal of iron and suspended solids is essential to protect the ion exchange membrane in the electrolyser and prevent long term performance degradation.

Assessment of the existing α-cellulose precoat filtration system showed that, although it is capable of achieving the required iron removal under stable conditions, its performance is variable and sensitive to inlet brine quality. Examining samples before and after the filter demonstrated iron removal efficiencies ranging from around 10% to over 90%, indicating inconsistent filtration behavior. In addition, the aging equipment of the system is going to require a substantial capital investment (780K€) to keep running plus increasing maintenance costs. Refurbishment of the existing system would restore functionality but would not address these fundamental limitations.

The evaluation of the membrane filter system demonstrated clear technical and operational advantages. The membrane filter provides stable filtration performance from startup, achieves reliable iron removal, and significantly reduces waste generation and operational complexity While the membrane solution requires a higher initial investment, the elimination of cellulose consumption, reduced maintenance, and simplified process configuration result in lower annual operating costs. Economic analysis showed that, with optimized vessel design or retrofit options, the membrane system can achieve an attractive payback period and represents a financially viable long-term solution. Bypassing the brine filtration step entirely was shown to be economically attractive in the short term but introduces unacceptable long-term risks. Although acceptable brine quality can be maintained temporarily when salt impurity levels are very low, sustained operation without filtration increases the likelihood of iron accumulation, ion-exchange resin degradation, and gradual deterioration of electrolyser membrane performance. This option therefore lacks robustness and is highly dependent on consistently exceptional salt quality, which cannot be guaranteed.

In conclusion, back-pulse membrane filtration emerges as the most robust and future-proof solution for brine purification at Nobian MEB. It offers superior filtration stability, reduced operational risk, lower waste generation, and improved long-term economic performance compared to refurbishing the existing precoat filtration system. Adoption of this technology supports reliable electrolyser operation, enhances process sustainability, and aligns with the site's objective of reducing operational complexity while maintaining strict brine quality specifications.

For future work, given the proposed membrane filters provide nearly double the filtration area within the existing vessel footprint, a detailed engineering and financial assessment should be conducted to evaluate the feasibility of installing two smaller parallel filter vessels instead of a single large unit. The configuration would allow continuous operation with one filter on standby, significantly reducing the need to bypass the filter during maintenance or unplanned breakdowns. Moreover, a more in-depth material study is recommended to identify suitable alternatives to traditional ebonite lined carbon steel, Such as the lighter FRP with vinyl ester lining. Materials that offer lower capital investment, faster and easier installations, while keeping chemical stability and resistance.

Finally, a systemic analytical study is recommended to resolve the observed discrepancies between the iron concentration measurements obtained using different sample pretreatment methods. This study should compare the acidification only and acidification plus boiling procedures to determine which method most accurately reflects the iron species in the brine, thereby establishing a consistent and reliable analytical basis for future operational decision making.

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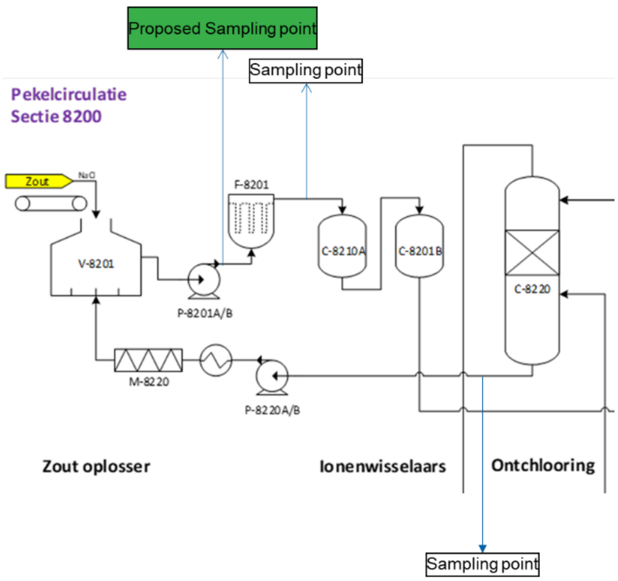
9. Appendix

9.1. Appendix A: Precoat filtration steps

9.2. Appendix B: Adding the new sampling point



The PFD above show the different brine sampling points and the one in green was the one being added, and the one used to send samples to Hengelo for analysis

9.3. Appendix C: CAPEX & OPEX Calculations

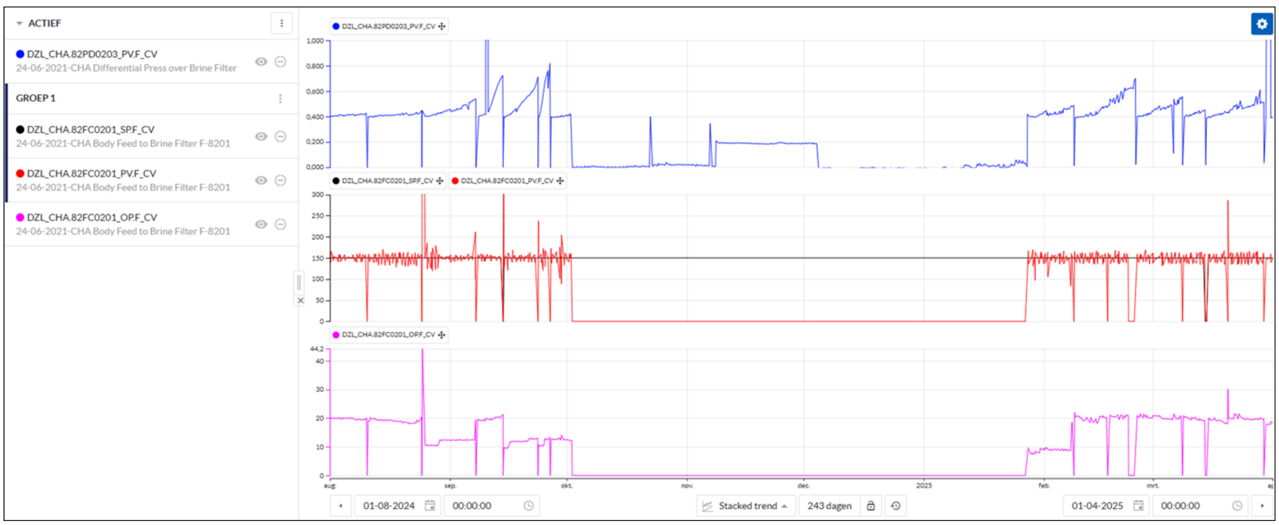
Maintenance Costs since 2015 (summary)

Unit Section	Unit	Maintenance	Project	Total
F-8201	F8201	36,074.13	278,693.21	314,767.34
V-8204	V8204	7,772.70	464.00	8,236.70
V-8204	M8204	23,513.07	-	23,513.07
V-8204	P8204	21,318.28	12,862.00	34,180.28
V-8203	V8203	17,119.17	-	17,119.17
V-8203	P8202	13,815.07	13,484.00	27,299.07
V-8203	M8205	1,210.16	-	1,210.16
V-8203	M8203	5,960.12	-	5,960.12
V-8203	S8206	11,936.69	-	11,936.69
V-8203	A8206	516.00	-	516.00
S-8205	S8205	87,549.72	8,357.30	95,907.02

Projected Capital expenditure of Refurbishing the Brine filter system

Cost of refurbishing	Cost
F-8201 Relining	100,000.00
P-8204 Refurb	3,000.00
P-8203 Refurb	1,000.00
P-8202 Refurb	1,000.00
A-8206 Rep	1,000.00
M-8205 Refurb	1,000.00
M-8203 Refurb	1,000.00
Civil	50,000.00
New Filter Press	325,000.00
Tanks 8203/4 Refurb	300,000.00
total	783,000.00

9.4. Appendix D: Case Study



Between Oct. 24 an Jan 25 (period of 4 months) the filter was being bypassed as seen in the Trendminer® screenshot above.
comparing the Fe concentration data from the lab samples for the same period, it is observed that all results from the "after the filter" & " after the ion exchanger" results were consistently below 10 µg/l