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EFFECT OF MEMBRANE PERFORMANCE INCLUDING FOULING ON COST OPTIMIZATION IN BRACKISH WATER DESALINATION PROCESS

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Highlights

- Membrane selection for optimum brackish water desalination operation was presented.
- Verberne Cost Model was employed to evaluate the total cost of the membrane unit.
- Economic and fouling assessments were used to select the most suitable membrane.
- High permeability membrane did not guarantee energy consumption and cost savings.
- Impacts of different membranes and recovery rates on water costs were investigated.

ABSTRACT

Membrane selection is a crucial step that will affect the economic feasibility of the membrane water treatment process. A comprehensive evaluation consisting of Verberne Cost Model, assessment of membrane performance and fouling propensity, osmotic pressure differential (OPD) and specific energy consumption (SEC) was employed to determine the potential of nanofiltration (NF 270, NF 90 and TS 80) and low pressure reverse osmosis (XLE) membranes to be used in brackish water desalination process. The aim was to save costs by replacing the typical brackish water reverse osmosis (BW 30) membrane. Verberne Cost Model showed that higher flux NF membranes resulted in lower overall costs. However, after assessing the membrane performance, NF 270 and TS 80 were excluded due to their high fouling propensity and their failure to reduce total dissolved solids (TDS) in the solution. Instead, NF 90 membrane which produced water with acceptable TDS and has moderate permeability ended up to be more cost competitive compared to BW 30 membrane, with 17%-21% lower total costs and 13%-17% lower water costs. Apart from this, OPD and SEC were applied to justify the selection of optimal membrane recovery rate based on the water costs calculated. It was determined that the optimal recovery rate was 80% where the SEC and water costs were close to available water treatment plants. Overall, this study showed that the selection of membrane can be carried out by using Verberne Cost Model assisted by assessment of membrane performance and fouling propensity, OPD and SEC.

Keywords: Membrane Process; Brackish Water Desalination; Cost Model; Economic Evaluation; Membrane Fouling

1. INTRODUCTION

Membrane desalination is an energy intensive process where most of the energy is being consumed to supply the necessary operating pressure. It has been reported that high pressure pumps are responsible for more than 40 % of the total expenditures of membrane desalination plant [1]. In terms of power consumption, pumps consumed as much as 80 % of the overall electricity supplied to desalination plant [1]. However, technological advancement in desalination process such as energy recovery devices, efficient design and operation of desalination plant managed to cut down the energy consumption from 30 kWh/m^3 in 1979 to around 3.9 kWh/m^3 today. Furthermore, with the most recent developments, it has been demonstrated that the energy consumption by seawater reverse osmosis (SWRO) desalination process can be reduced to roughly 2.0 kWh/m^3 [2]. Although the energy consumption of SWRO desalination has been substantially reduced, it is still considerably higher than conventional surface water treatment technologies. Since reducing the energy consumption is critical for lowering the desalination water costs, consideration should be given in using brackish water with lower osmotic pressure or in selecting membrane with higher rejection but at lower operating pressure. [2-3].

Brackish water contains much lesser dissolved mineral salts which indicate the operating pressure for the membrane process can be lowered down significantly, as compared to seawater desalination process. This opens the opportunity for nanofiltration (NF) membrane to be used in brackish water treatment process, since NF membrane offers higher water production (permeate) while operating at lower pressure compared to reverse osmosis (RO) membrane. Higher production rate also reflects the chance to reduce energy consumption and increase the economic values of the desalination plant. Theoretically, the application of NF membrane in brackish water desalination process would be favourable due to its advantages over RO membrane as aforementioned [4]. However, this comes at the expense of lower membrane salt rejection capability as high permeability membrane normally has high salt permeability too [5]. In other words, further treatment is required to get the permeate TDS concentration down to the recommended range. This might incur extra costs for the additional treatment process and offset the benefit of high flux performance. The cost comparison study among different types of NF and RO membranes for brackish water desalination has been limited and this information is required to know to what extent the NF membrane is economically preferable than RO membrane. Such comparison is also important to clarify the arguments about the use of high permeability membrane will not result in significant energy and cost savings [5].

Economic evaluation can provide the necessary cost comparison among different membranes and also the required information before decision of new investment on the membrane treatment plant can be made [6,7]. The cost of membrane water treatment plant varies and is dependent on the production capacity, type of treatment involved, design criteria, climate condition, characteristics of land and building, etc. Membrane flux or the production capacity is the most important aspect for the design of membrane filtration plant as it is a direct measure of productivity, operating pressure (energy requirements) and amount of membrane required (membrane area) [8–10]. Hence, a cost model which utilizes simple experimental results such as flux and rejection yet capable to provide acceptable cost estimation is desirable for the selection of appropriate membrane of a new water treatment plant.

Various cost models have been developed to provide estimation of total costs for planning, initial screening purposes and to better understand the impacts of different designs and operating conditions on membrane treatment costs [6,10–14]. Among the available models, Verberne Cost Model will be of particular interest since the equations involved were based on project practical data and it has been successfully employed in estimating the cost of membrane water treatment process based on simple experimental results [6,14]. However, Verberne Cost Model does not take membrane fouling propensity into consideration. It is widely known that fouling plays a vital role in affecting the overall membrane performance and this will have significant impact on the capital and operating costs. Furthermore, as mentioned above, energy consumption and desalinated water costs are particularly important where both are heavily dependent on membrane recovery rate and difficult to be justified by Verberne Cost Model. Considering Verberne Cost Model alone is not enough to provide a comprehensive evaluation of membrane treatment process, the membrane performance, fouling propensity and energy consumption also have to be included during the selection of membrane for the water treatment plant.

This study attempted to utilize Verberne Cost Model in predicting the total costs of membrane brackish water desalination processes using different NF and RO membranes. Economic evaluation from the cost model will be combined with membrane fouling propensity and performance for the selection of appropriate membrane to replace the typical brackish water RO membrane with the aim to save costs. In addition, energy consumption, represented by osmotic pressure differential (OPD) and specific energy consumption (SEC) was adopted to assist in the determination of optimal membrane recovery rate based on the water cost calculated from Verberne Cost Model. Overall, a comprehensive evaluation including Verberne Cost Model supported by membrane performance, fouling propensity and energy consumption will be carried out to assess and decide which membrane performs the best and suitable for this brackish water desalination process. The rationales behind the use of high permeability membrane and its impact on energy and cost savings will also be evaluated.

2. MATERIALS AND METHODS

2.1 Chemicals and Membranes

All chemicals used are analytical grade, unless stated otherwise. Humic acid (HA), ferric chloride ($FeCl_3$), kaolin, calcium chloride ($CaCl_2.2H_2O$), sodium bicarbonate ($NaHCO_3$), and sodium chloride (NaCl) were purchased from Sigma Aldrich (Malaysia). Ultrapure (UP) water with a quality of 18 $M\Omega cm^{-1}$ was used for all solution preparation. Membrane used in this study can be divided into two categories; NF membranes (NF 270, NF 90 and TS 80) and RO membranes (XLE and BW 30). All of the membranes were purchased from Dow Filmtec (USA) except for TS 80 membrane which was purchased from Trisep (USA). The characteristics of the membranes are shown in Table 1. BW 30 membrane will be the control membrane in this study where its performance will be the benchmark for other membranes. NF 270 membrane is known to be a high flux NF membrane while NF 90 has high salt rejection capability with moderate flux. TS 80 is a NF membrane that has slightly higher flux over BW 30. XLE is a RO membrane which was specifically fabricated to be operated with lower energy consumption (lower operating pressure and higher flux compared to typical RO membrane).

Table 1
Membrane characteristics

Membrane	**Water permeability ($Lm^{-2}h^{-1}bar^{-1}$)[a]**	**Salt Rejection (%)[b]**	**Surface roughness (nm)**	**Zeta potential (mV)[c]**	**Contact Angle (^{o})[d]**
NF 270	14.5	>97	9.0 ± 4.2[c]	-41.3	15.0 ± 2.2
NF 90	11.5	>97	129.5 ± 23.4[c]	-37	96.5 ± 5.0
TS 80	6.5	99	79.4[e]	-32[e]	29.8 ± 3.4
XLE	7.5	99	142.8 ± 9.6[c]	-27.8	93.5 ± 1.5
BW 30	5.0	99.5	68.3 ± 12.5[c]	-10.1	90.3 ± 2.7

[a] Water permeability for each membrane was obtained from the slope of membrane flux vs operating pressure (2, 4, 6, 8 and 10 bar) graph using UP water at 25^{o}C
[b] Salt rejection as provided by membrane manufacturers at operating conditions: 25^{o}C, 2000 ppm $MgSO_4$ (NF 270 at 4.8 bar; NF 90 at 4.8 bar; TS 80 at 7.6 bar), 2000 ppm NaCl (XLE at 6.9 bar; BW 30 at 15.5 bar)
[c] Zeta potential value (at pH 9) and surface roughness were taken from [15]
[d] Contact angle was measured in the lab
[e] Surface roughness and zeta potential (at pH 9) of TS 80 were taken from [16]

2.2 Synthetic Test Waters

Synthetically prepared waters with fixed turbidity were used for this work. The HA concentration for each batch of run was 20 ppm. Suitable amount of kaolin was added into the synthetic water to adjust its turbidity to 30 ± 0.5 NTU. The pH of all the synthetic water prior to coagulation process was adjusted to 7 by using sodium hydroxide (NaOH) and hydrochloric acid (HCl). The composition of the synthetic test waters was shown in Table 2. W1 test water was prepared by referring to water composition in River Oise, France and Lake Mead, Nevada published by The Dow Chemical Company where the TDS concentration was categorized as low TDS brackish water [17]. On the other hand, W2 test water was prepared based on the composition of Tan Tan brackish water published by Dach (2008). This concentration was known as moderate TDS brackish water. Since the test waters were prepared from NaCl, $CaCl_2.2H_2O$ and $NaHCO_3$ salts, the composition of the solutions would be different from the real waters because precise control of elemental concentration was not possible. By referring to those references, it can be ensured that the test waters prepared were within the brackish water TDS range. The composition of the solutions was slightly adjusted as this study was the continuity of our previous scaling study [19].

Table 2
Characteristics of the synthetic test waters

Label	Salts Composition
W1	400 ppm mixture (80 ppm NaCl, 200 ppm $CaCl_2.2H_2O$, and 170 ppm $NaHCO_3$)
W2	4000 ppm mixture (2800 ppm NaCl, 1100 ppm $CaCl_2.2H_2O$, and 400 ppm $NaHCO_3$)

2.3 Jar Test Coagulation and Cross-flow Process Setup

Coagulation pretreatment prior to NF/RO membrane process was carried out in a conventional jar test apparatus (Model ZR4-6, Zhongrun Water, China). The coagulation procedures consisted of three steps: vigorous stirring after the addition of coagulant (100 rpm for 1 minute), mild stirring (30 rpm for 29 minutes), and settling (30 minutes). The dosage of $FeCl_3$ coagulant was varied in order to obtain the optimal dosage which removed most of the turbidity and HA. The supernatant from the coagulation process with optimal dosage was then used as the feed water for membrane experiment.

Bench-scale cross-flow membrane experimental setup with recycle loop as shown in Fig. 1 was used for this experiment. The cross-flow system will be conducted in total recycle mode where the permeate and retentate will be recycled back into the feed tank. The membrane test cell (CF 042, Sterlitech, USA) has 0.0042 m^2 membrane effective filtration areas. The supernatant water from the coagulation process (3.5 L) will be used as the feed for the cross-flow system. The operating conditions for temperature, pressure, and cross-flow velocity were 27 ^oC, 10 bars, and 42 cm/s respectively. This membrane filtration experiment was conducted for 5 hours with all the operating conditions being controlled and maintained at the values aforementioned. The performance of the membrane process was assessed and presented as flux versus time and salt rejection versus time. Extent of membrane flux decline and salt rejection capability decline was calculated by using the equation below:

$$Decline = \frac{M_i - M_f}{M_i} \times 100\% \quad (1)$$

where M_i denotes the initial flux/salt rejection and M_f represents the final flux/salt rejection.

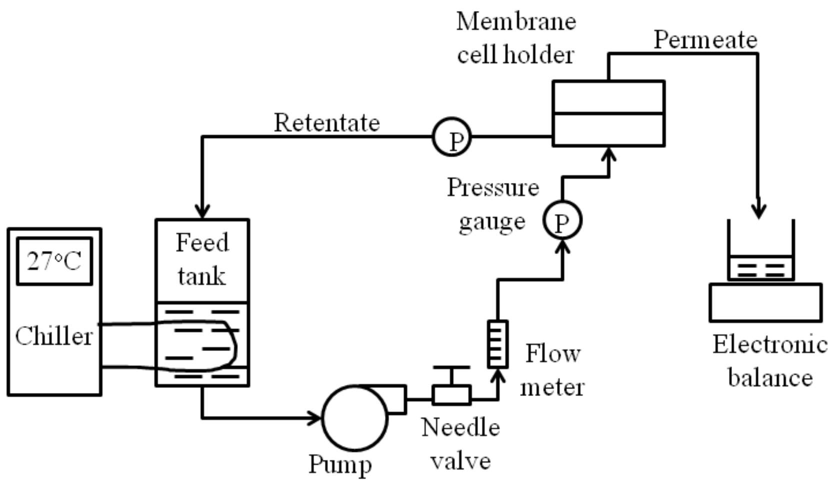


Fig. 1. Schematic diagram of bench-scale cross-flow filtration process

2.4 Verberne Cost Model

In this study, Verberne Cost Model was used to estimate the overall cost for the whole membrane filtration process [10-11]. Basically, cost assessment using Verberne Cost Model was estimated based on the feed flow and operating pressure of the membrane filtration process. The details regarding Verberne Cost Model were presented in Table 3. First, the desired production capacity, feed flow and recovery percentage of the membrane process were assumed. Then, the number of membrane modules needed for this production capacity was calculated based on the experimental data (operating pressure and membrane permeate flux). It has to be noted that the permeate flux used here was the experimental steady flux after the initial drop in flux. Next, total investment and operating costs on a yearly base were predicted using the equations in Table 3. The operating costs per cubic meter permeate were calculated and compared among the different membranes. The desired capacity of the membrane plant was assumed to be 2000 m^3/h, corresponding to a large plant for 250000 residents [6]. The feed flow will be 2500 m^3/h with the plant being operated at a recovery of 0.8. Membrane surface for one module was estimated to be 30 m^2. The permeate flow was taken from the steady membrane flux obtained from experiments as tabulated in Table 6.

2.5 Osmotic Pressure Differential (OPD) and Specific Energy Consumption (SEC)

Osmotic pressure differential (OPD) represents the concentration difference across the membrane in the desalination process. Operating at lower OPD indicates lower pressure operation which will not only reduce the pumping, maintenance and fixed costs, but also reduce the fouling in the membrane stage. Furthermore, specific energy consumption (SEC), which translates to the amount of energy required to produce a certain amount of water, can be lowered down due to its dependency on OPD. The calculation of OPD and SEC was shown in Table 4 [20,21]. These estimations will be used to assist Verberne Cost Model in the selection of optimal membrane recovery rate for economic benefits.

2.6 Analytical Methods

HA absorbance was measured using UV/Vis spectrophotometer (Lambda 35, PerkinElmer, USA) at the wavelength of 254 nm. Zeta-Sizer (Malvern, UK) was used to measure stability of the suspensions. Contact angle measurement was carried out using Drop Shape Analysis System Goniometer (Model DSA100, KrussGmbH, Germany). Turbidity of the water was measured using 2100 N Laboratory Turbidimeter (Hanna, USA). Conductivity and pH of the solution were measured using HI 2550 Benchtop Meter (Hanna, USA). The fouled membranes were characterized (surface view and cross sectional view) using field emission scanning electronic microscopy-energy dispersive X-ray (Merlin Compact, Zeiss, Germany).

Table 3
List of equations for Verberne Cost Model

Capital cost items	Operating cost items
(i) Civil investments: buildings where the installation is housed. Depreciation period for these investments is 30 years. $C_{civil} = 862Q_F + 1239n$	(a) Depreciation costs: depreciation rate upo cost. The investments are linearly dep interest is neglected. $C_{deprec} = \frac{C_{civil}}{30} + \frac{C_{mech} + C_{electro}}{15} + \frac{C_{membrane}}{5}$
(ii) Mechanical engineering: costs for pumps, filters, piping, etc. Depreciation period is 15 years. $C_{mech} = 3608Q_F^{0.85} + 908n$	(b) Consumption costs: can be divided into chemicals costs Energy costs: energy required to pump the feed membrane system. It is assumed that membrane sy Wh/m^3 for each m^3 feed and feed pressure (bar). cost is estimated as 0.05 €/kWh. $C_{energy} = 40PQ_F \times \frac{0.05}{1000} \times 24 \times 365$ Chemicals: cost for chemical used in the process is be 0.025 €/m^3 of filtrate $C_{chemical} = 0.0225Q_P \times 24 \times 365$ $C_{consump} = C_{energy} + C_{chemical}$
(iii) Electrotechnical investments: costs for energy supply, control engineering and all electronic components. Depreciation period is 15 years. $C_{electro} = 1.4 \times 10^6 + 54PQ_F$	(c) Maintenance costs: 2% of the total investme $C_{maint} = 0.02C_{invest}$

(iv) Membrane investments: cost for membrane installation. The membrane lifetime is taken as 5 years (depreciation period). It was assumed that one membrane module costs about 1000 €.	(d) Specific operation costs: quality control and installation (2% of the total investment cost
$C_{membrane} = 1000\,n$	$C_{spec} = 0.04 C_{invest}$
Total investment costs:	Total operating costs:
$C_{invest} = C_{civil} + C_{mech} + C_{electro} + C_{membrane}$	$C_{operating} = C_{deprec} + C_{consump} + C_{maint} + C_{spec}$

Table 4
List of equations for OPD and SEC

OPD (bar)	$OPD = \frac{K(C_f - C_0)}{1 - Y}$ where K is the coefficient in the linear relationship between the concentratio pressure (0.801 L.bar.g^{-1}), C_f is the feed TDS (g/L), C_0 is the permeate TDS (g/L) and Y is the wat
SEC (kWh/m^3)	$SEC = \frac{K(C_f - C_0)}{Y(1 - Y)}$ where K is the coefficient in the linear relationship between the concentratio pressure (0.0223 kWh.L.m^{-3}.g^{-1}), C_f is the feed TDS (g/L), C_0 is the permeate TDS (g/L) and recovery

3. RESULTS AND DISCUSSION

3.1 Assessment of Membrane Performance and Fouling Propensity

Consistency in membrane performance and fouling are two important aspects to consider when operating a membrane filtration plant. Table 5 presents the changes in solution pH, zeta potential, UV_{254} absorbance and turbidity after the coagulation process for W1 and W2 solutions. It can be seen that the particles zeta potential in both solutions increased moderately after coagulation processes. Such increment might affect the membrane fouling propensity as charge repulsion is one of the important mechanisms that prevent the deposition of foulant onto the membrane surface [22]. Based on the turbidity and UV_{254} absorbance values in W1 and W2 supernatant solutions, it can be assumed that both contain similar amount of residual organic foulant. The coagulation mechanism involved in the removal of foulant has been reported in our previous publication [23].

Table 5
Characteristics of the test solutions before and after coagulation process

Solutions	W1		W2	
	Before	After	Before	After
pH	7.0	7.1	7.0	7.4
Zeta potential (mV)	-22	-15	-19	-13
UV_{254} (cm^{-1})	0.65 ± 0.05	0.05 ± 0.02	0.66 ± 0.05	0.05 ± 0.01
Turbidity (NTU)	30 ± 0.5	0.68 ± 0.1	30 ± 0.5	0.53 ± 0.1

For low TDS W1 solution, the salt rejection capabilities of the entire membranes remained constant, except NF 270 which underwent 12% decrement (Fig. 2 and Table 6). However, permeate flux decreased at an average 5% for all the membranes. The degradation in membrane performance can be attributed to fouling phenomena. Basically, foulant in the solution formed a cake layer on the membrane surface, affecting the permeation of water and rejection of salt ions [18-22]. Images from FESEM (Fig. 3) indicated that fouling occurred for NF 270 was the worst, as not only organic foulant blocked the membrane surface, but calcium complex precipitated on the membrane surface too, as supported by the EDX result in Fig. 3b. Hence, the performance of the highest flux membrane (NF 270) was the least consistent. The high adsorption rate of foulant on the surface of NF 270 membrane was probably due to its high permeation drag [25]. Since it has the highest flux compared to the rest of the membranes, back diffusion of salts and foulant to bulk solution was hindered.

Consequently, calcium ions accumulated at the membrane surface started to get precipitated into calcium scales when its concentration exceeded its solubility level. Such phenomenon was not observed on other membranes (Fig. 3c-3f) where their permeation drag was lower.

Table 6
Initial and final membrane permeate flux and conductivity rejection

Membranes	**Flux (LMH)**		**Conductivity Rejection (%)**	
	400 ppm	**4000 ppm**	**400 ppm**	**4000 ppm**
NF 270	131/126	92/66	35/31	21/18
NF 90	101/94	60/56	93/93	91/91
TS 80	56/55	47/46	78/78	62/60
XLE	79/75	47/42	92/92	91/91
BW 30	49/48	33/31	97/97	96/96

*values in the table indicate initial/final

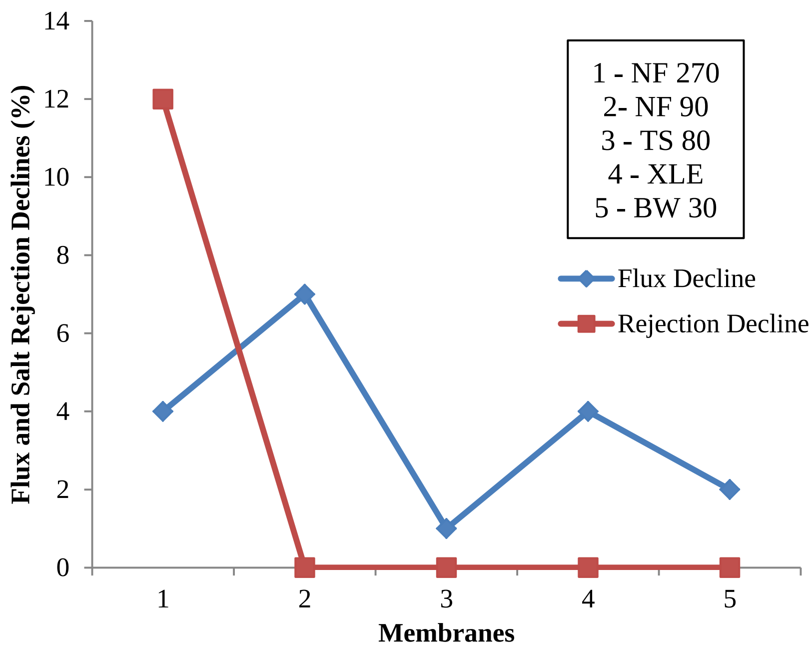


Fig. 2. Membrane performance and analysis for W1 brackish water: Flux decline and salt rejection capability decline for each membrane process

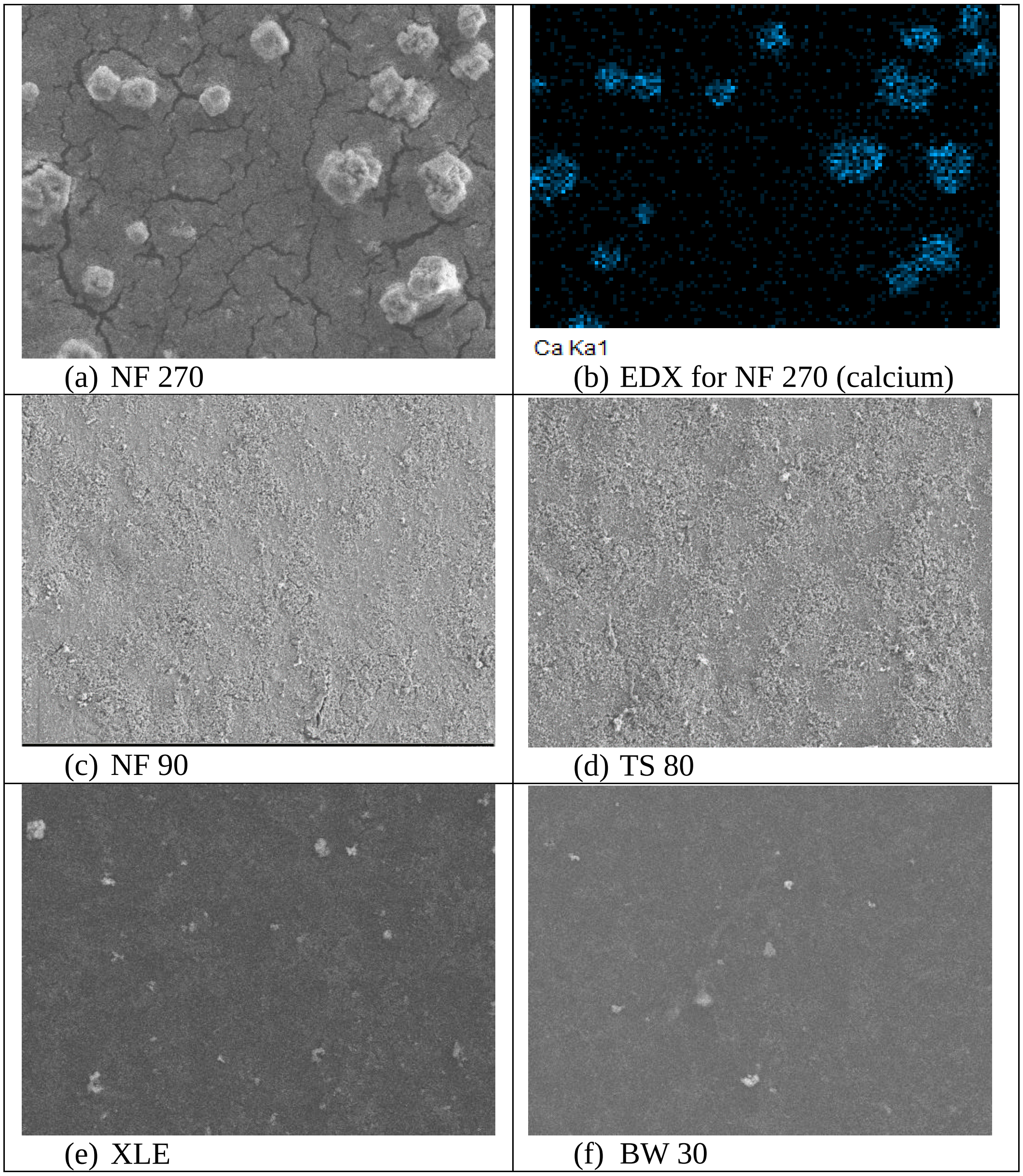


Fig. 3. Membrane FESEM images at 1000x magnification for W1 solution

Besides the permeation drag factor, membrane surface charge may as well play a role here. Among the membranes used in this study, NF 270 has the most negative charge [15]. Even though its surface has been shielded by a foulant layer, calcium cations still have a higher affinity towards the membrane via opposite charge interaction. Hence, opposite charge interaction and high permeation drag resulted in elevation of calcium ions concentration for NF 270 process which eventually led to the drop in salt rejection capability and flux. On the other hand, even though the rest of the membranes have rougher surface and lower hydrophilicity, their lower permeation drag and less negative surface charge made up for the compensation of their inferior membrane characteristics. The foulant layers were relatively thinner compared to NF 270 membrane with no sign of calcium scales observed (Fig. 4A). High salt rejection shown by NF 90, XLE and BW 30 did not encounter scaling issues which could be attributed to their lower permeation drag and weaker opposite charge interaction.

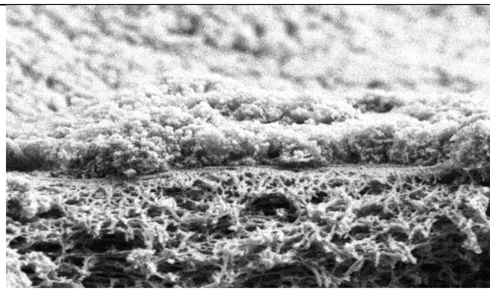
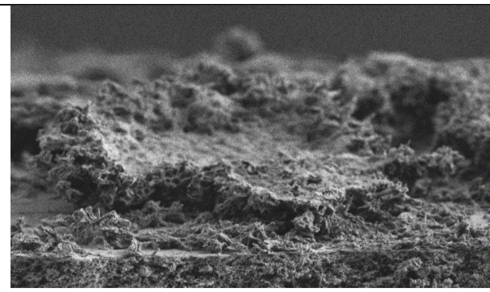
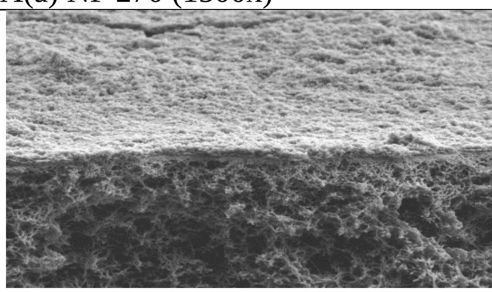
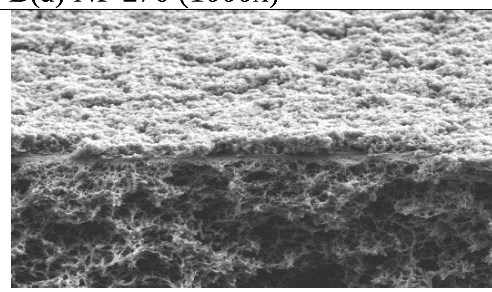
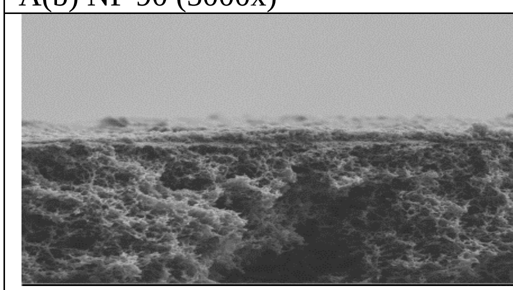
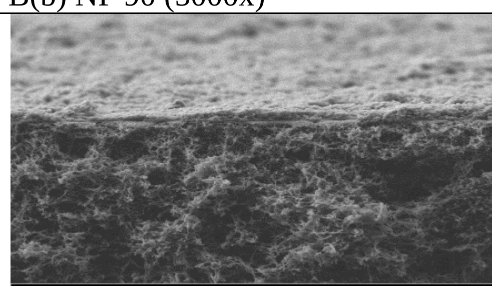
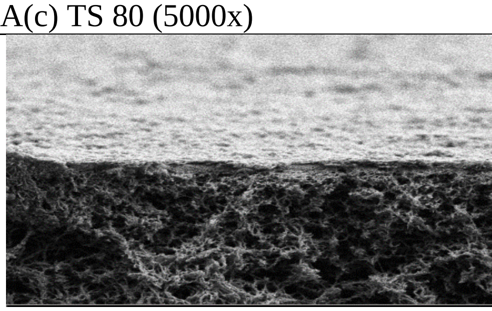
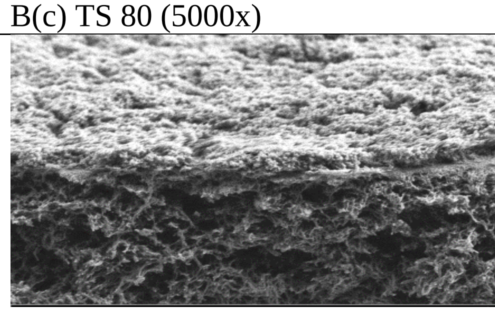
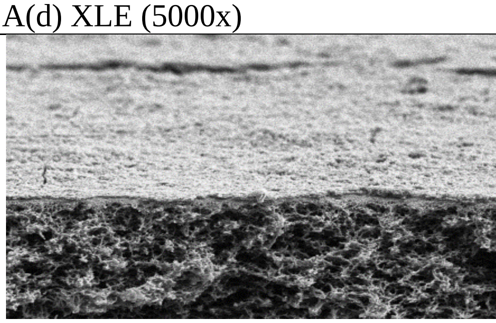
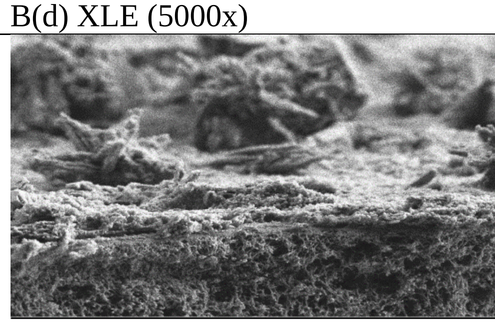
 A(a) NF 270 (1500x)	 B(a) NF 270 (1000x)
 A(b) NF 90 (5000x)	 B(b) NF 90 (5000x)
 A(c) TS 80 (5000x)	 B(c) TS 80 (5000x)
 A(d) XLE (5000x)	 B(d) XLE (5000x)
 A(e) BW 30 (5000x)	 B(e) BW 30 (5000x)

Fig. 4. Membrane FESEM cross-sectional view images for A. W1 solution and B. W2 solution

The performance of NF 270 was further aggravated when the TDS of the solution increased to 4000 ppm. It can be seen from Fig. 5 that its flux and salt rejection capability decreased around 28% and 12%, respectively. Images from FESEM (Fig. 6a & 6b) confirmed that scaling was the main reason for this degradation. The scales were formed first before the foulant layer. This might indicate that the opposite charge interaction was more prominent as the concentration of calcium ions was much higher compared to 400 ppm TDS solution. As a result of this scale layer, flux dropped sharply and permeation of salt was enhanced [26]. Besides that, the fluxes of XLE and BW 30 membrane underwent higher decrement, at around 10% and 7% compared to 4% and 2%, respectively, when treating 400 ppm TDS solution. This probably could be attributed to the higher salt concentration.

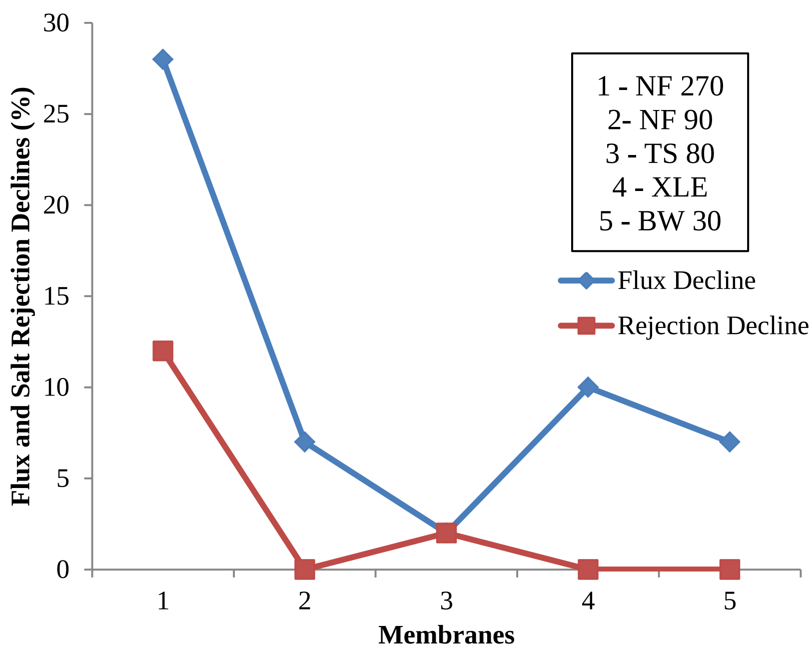
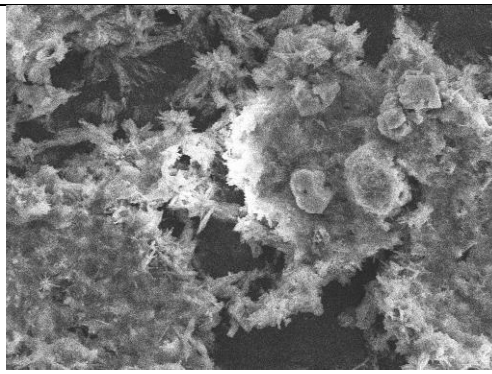
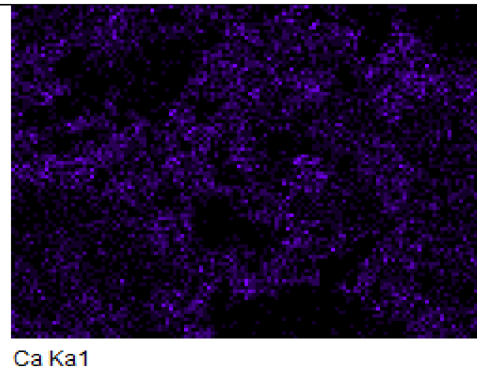


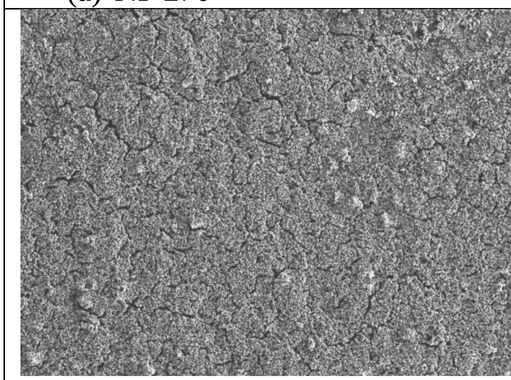
Fig. 5. Membrane performance and analysis for W2 brackish water: Flux decline and salt rejection capability decline for each membrane process



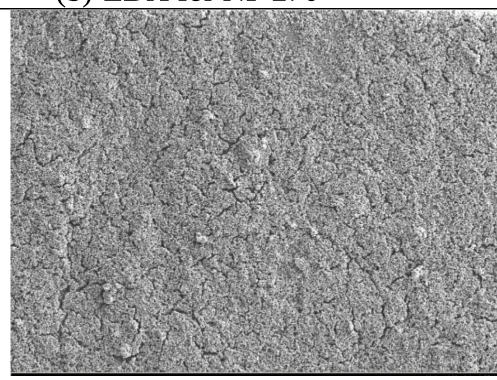
(a) NF 270



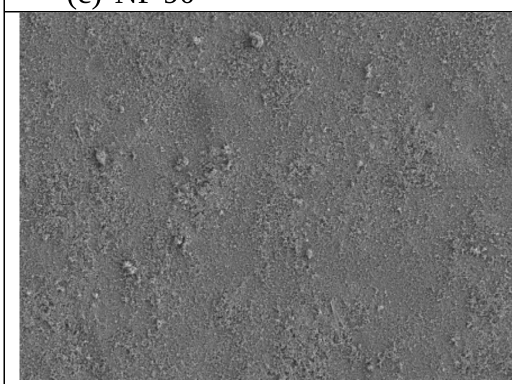
(b) EDX for NF 270



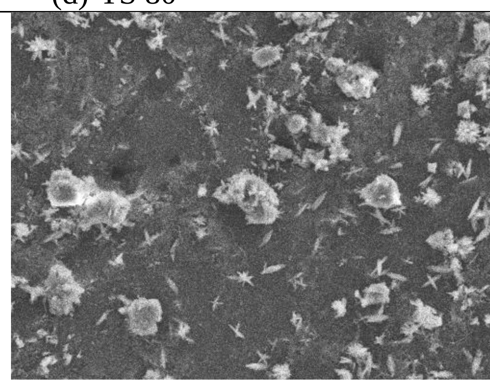
(c) NF 90



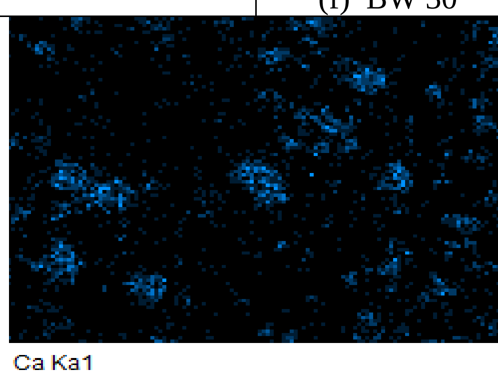
(d) TS 80



(e) XLE



(f) BW 30



(g) EDX for BW 30

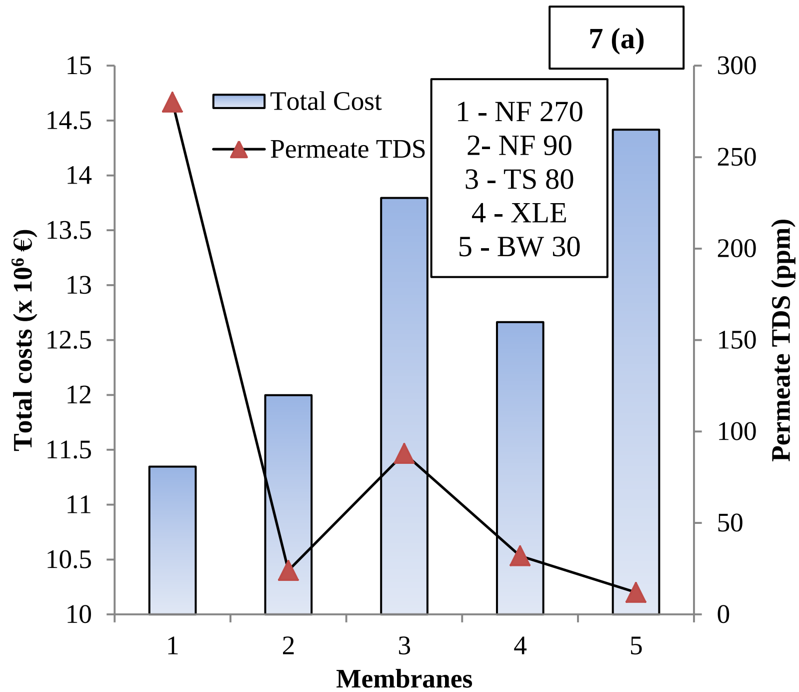
Fig. 6. Membrane FESEM images at 1000x magnification for W2 4000 ppm solution

It was observed that the cake layers formed for all membranes were thicker when treating W2 solution (Fig. 4B). Dissolved salts will get accumulated at the membrane surface due to the membrane high rejection capability. Consequently, the surface charge of foulant will be suppressed by the presence of salts [27,28]. Eventually, weaker charge repulsion between the foulants-foulants and foulants-membrane resulted in higher tendency to deposit onto the membrane surface, as can be seen by the thicker foulant layers formed [23]. On the other hand, calcium scales were seen to appear on BW 30 membrane (Fig. 6f & 6g). This may be attributed to the high salts rejection which resulted in calcium concentration exceeded its solubility level. The high salts rejection capability induced a more severe concentration polarization effect. Calcium concentration accumulated on the membrane surface increased over time and eventually precipitated out as scales. The formation of calcium scales in these cases was different from NF 270, where in the latter case charge interaction between calcium cations and negatively charged NF 270 was the reason contributing to scaling issue.

3.2 Total Investment and Capital Costs

As discussed in the previous section, NF 270 was found to be unfavourable due to its severe fouling and scaling issues. Hence, for the following economic evaluation, focus will be given to other membranes. Membrane performance and fouling propensity evaluations have helped to screen out the unsuitable membrane. The raw information (membrane flux) required for Verberne Cost Model was tabulated in Table 6.

The findings from Verberne Cost Model were tabulated in Table 7 to Table 11. Estimation from Verberne Cost Model (Table 10 and Fig. 7) shows that NF and XLE membranes which have higher permeate have lower operating and investment costs compared to the BW 30 membrane. The operating costs per cubic meter filtrate for NF and XLE membranes (Table 11) were considerably lower than BW 30 as well. However, the amount of total costs among the NF and XLE membranes were different for both TDS solutions, as displayed in Fig. 7. For 400 ppm solution, it was discovered that TS 80, as a NF membrane had lower flux over XLE membrane, resulting in its operating/investment costs to be higher. When the TDS of the solution was increased to 4000 ppm, the trend was reversed. The drop in permeate flux for XLE membrane was more significantly affected by the TDS, probably due to its high salt rejection capability. Our comparison study between NF and RO membranes indicated that not necessarily NF membrane will always give lower plant expenditure, as some new generation of improved low operating pressure RO membrane such as XLE can be quite competitive too. Nonetheless, TDS of the solution has to be taken into consideration as well, as the increase in TDS reduces the filtrate of XLE due to its high salt rejection capability.



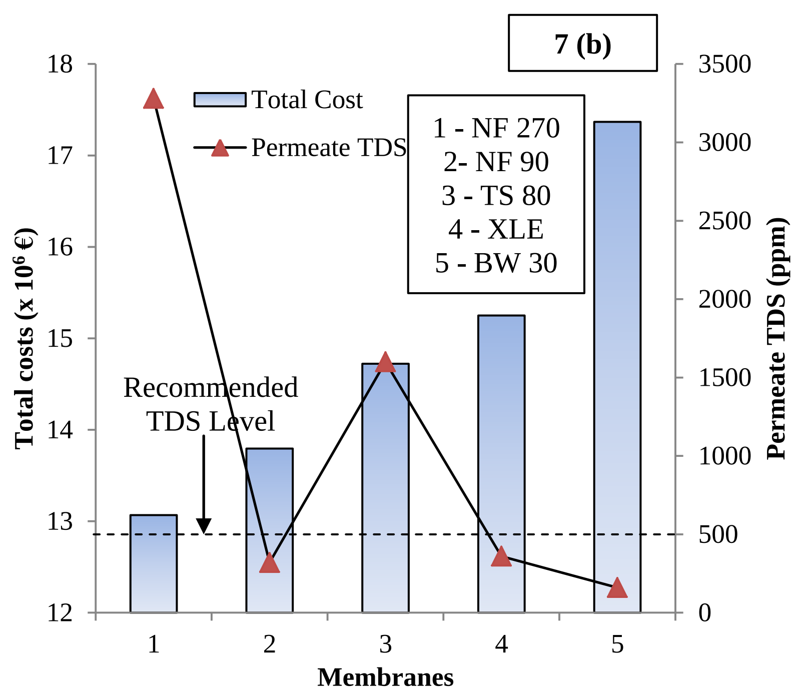


Fig.7. Total investment/operating costs and permeate TDS for each membrane process of (a) W1 solution (b) W2 solution

In addition to total operating and investment costs, the quality of the permeate is one of the determining factors in deciding on a membrane filtration plant. World Health Organization has recommended the TDS of the drinking water to be lower than 500 ppm. Hence, in our study, the permeate TDS that needs to be further reduced was W2 4000 ppm solution. As portrayed in Fig. 7b, NF 270 and TS 80 failed to reduce the TDS of the solution to below 500 ppm. Such failures might indicate that these membranes were not suitable for higher TDS water treatment process. It could also be concluded that one pass membrane filtration was not enough to produce the desired quality permeate [6]. Instead, further modification to the design such as two pass membrane process configuration is required, where the additional membrane modules will definitely increase the total operating/investment costs considerably. Thus, for higher TDS solution, NF 90 and XLE are two alternative membranes that can replace BW 30, although the latter membrane produced much better quality permeate. Nevertheless, the TDS of permeate from NF 90 and XLE membranes were within the recommended range. Further analysis with respect to specific energy consumption and osmotic pressure differential for membrane desalination process will be carried out using NF 90, XLE and BW 30 for W2 solution. Such analysis can provide more specific details regarding the influence of membrane recovery rate on energy consumption and operating costs per cubic meter treated water.

Table 7

Number of modules required for each testing solution and membrane to produce permeate of 2000 m^3/h at 27

Membranes	W1		W2	
	Permeate flux (L/m^2.h)	Number of modules required	Permeate flux (L/m^2.h)	Number of m
NF 270	126	527	66	1000
NF 90	94	706	56	1200
TS 80	56	1200	46	1455
XLE	75	889	41	1600
BW 30	48	1371	30	2182

Table 8

Investment costs (x 10^6 €)

Items	W1					W2			
	NF 270	NF 90	TS 80	XLE	BW 30	NF 270	NF 90	TS 80	XLE
C_{civil}	2.81	3.03	3.64	3.26	3.85	3.39	3.64	3.96	4.14
C_{mech}	3.27	3.43	3.88	3.60	4.03	3.70	3.88	4.11	4.24
$C_{electro}$	2.75	2.75	2.75	2.75	2.75	2.75	2.75	2.75	2.75
$C_{membrane}$	0.53	0.71	1.20	0.89	1.37	1.00	1.20	1.46	1.60
C_{invest}	9.35	9.92	11.47	10.49	12.01	10.84	11.47	12.27	12.73

Table 9
Operating costs on a yearly base (x 10^6 €)

Items	W1					W2			
	NF 270	NF 90	TS 80	XLE	BW 30	NF 270	NF 90	TS 80	XLE
C_{deprec}	0.60	0.65	0.80	0.71	0.85	0.74	0.80	0.88	0.92
$C_{consump}$	0.83	0.83	0.83	0.83	0.83	0.83	0.83	0.83	0.83
C_{maint}	0.19	0.20	0.23	0.21	0.24	0.22	0.23	0.25	0.25
C_{spec}	0.37	0.40	0.46	0.42	0.48	0.43	0.46	0.49	0.51
$C_{operating}$	1.99	2.08	2.32	2.17	2.41	2.23	2.32	2.45	2.52

Table 10
Total investment and operating costs (x 10^6 €)

Membrane	W1					W2			
	NF 270	NF 90	TS 80	XLE	BW 30	NF 270	NF 90	TS 80	XLE
Total	11.35	12.00	13.79	12.66	14.42	13.07	13.79	14.72	15.25

Table 11
Operating costs per cubic meter permeate (€)

Membrane	W1					W2			
	NF 270	NF 90	TS 80	XLE	BW 30	NF 270	NF 90	TS 80	XLE
Cost	0.114	0.119	0.133	0.124	0.137	0.127	0.133	0.140	0.144

3.3 Effect of Overall Recovery on OPD and SEC

Fig. 8 and Fig. 9 depict the membrane desalination processes for W2 brackish solution based on the overall recovery and associated SEC and OPD. In order to present a clearer observation, only recovery from 50% to 90% will be covered. Fig. 8 and Fig. 9 indicate that NF 90 and XLE can achieve the same recovery rate at a slightly lower OPD and SEC relative to BW 30. In general, the differences between those membranes were not that significant as the TDS concentration of the synthetic brackish water was not high as compared to seawater.

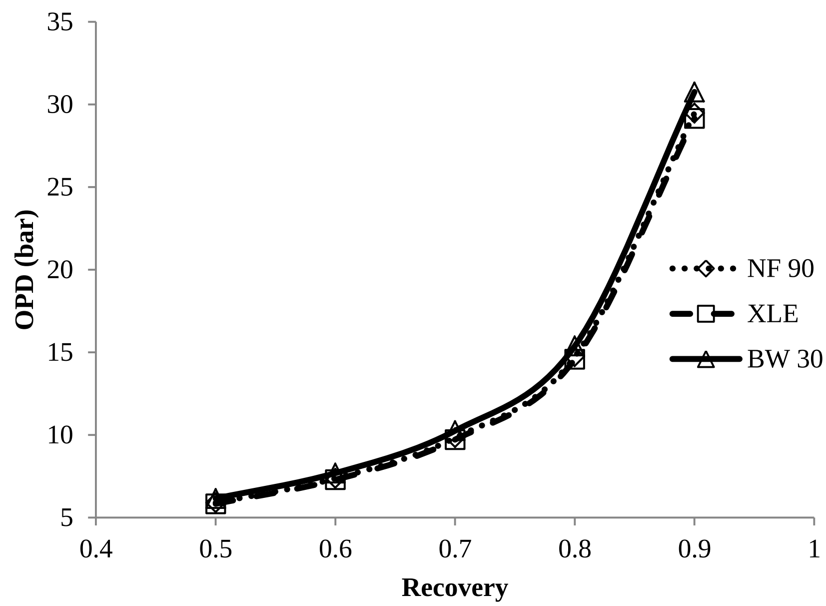


Fig. 8. Predictions of OPD at different recoveries

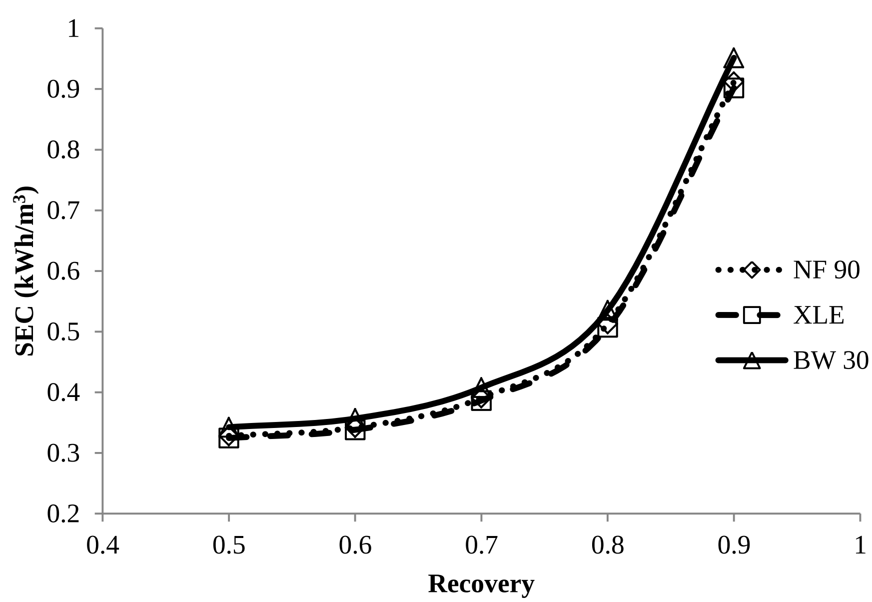


Fig. 9. Predictions of SEC at different recoveries

The reported SEC of conventional drinking water supply from surface or groundwater treatment processes was less than 0.5 kWh/m^3 [15-16]. From Fig. 9, it can be seen that the SEC of the membrane processes is close to this range when operated at 80 % recovery rate. Hence, it can be presumed that the membrane processes are as competitive as conventional drinking water treatment plants when the recovery rate is at 80 %. Nonetheless, the SEC of the membrane processes was lower than 0.5 kWh/m^3 when operated at lower recovery rate. This is expected as the operating pressure was lower and resulted in lower energy consumption. However, this was achieved with the sacrifice of higher operating costs per cubic meter treated water, as shown in Table 12 where the water cost was higher when operated at lower recovery rate. On the other hand, when the membrane process was operated at 90 % recovery rate, the OPD and SEC increased tremendously (Fig. 8 & 9) even though its operating costs per cubic meter treated water was the lowest (Table 12). Thus, there appears to be a trade-off recovery where the SEC and water cost were acceptable without increasing much of the total investment and operating costs. From this study, membrane process operated at 80 % recovery was deemed to be the optimal operating condition based on the arguments aforementioned. Such observation indicates that the assumption of 80 % recovery made in this Verberne Cost Model was acceptable. It has been shown that OPD and SEC results can be utilized to support Verberne Cost Model in the selection of optimal recovery rate to further increase its reliability. After all, this prediction did not directly take fouling issue into consideration and the selection between NF 90 and XLE membranes have to take into account the fouling tendency and membrane performance assessments carried out in Section 3.1.

Table 12
Total investment/operating costs (x 10^6 €) and operating costs per cubic meter permeate (€) for NF 90, XLE and BW 30 at different water recovery (W2 brackish water)

Recovery	Costs (€)	NF 90	XLE	BW 30
60 %	Investment and Operating	12.16	13.79	15.38
	Costs per cubic meter permeate	0.166	0.177	0.193
70 %	Investment and Operating	13.25	14.52	16.37
	Costs per cubic meter permeate	0.147	0.158	0.174
80 %	Investment and Operating	13.79	15.25	17.37
	Costs per cubic meter permeate	0.133	0.144	0.160
90 %	Investment and Operating	14.34	15.98	18.36
	Costs per cubic meter permeate	0.122	0.133	0.149

3.4 Validation of economic evaluation

Throughout this comparison study, it can be seen that the selection of membrane for a membrane filtration plant is a very crucial step that must take many factors into consideration. Cost factor should not be the only consideration aspect as membrane performance and permeate quality are important as well. For 400 ppm TDS solution, NF 90 and XLE were deemed as the alternative membranes to replace BW 30 due to their lower operating/investment costs, consistent performance and controllable fouling. The total costs/operating costs per cubic meter permeate for NF 90 and XLE were 17%/13% and 12%/9%, respectively lower than BW 30. Even though XLE is a RO membrane, its production rate is higher than NF membrane (TS 80). This indicated that with the rapid advancement of membrane technologies, lower operating pressure RO membrane can be as competitive as NF membrane in treating low TDS solution. However, for 4000 ppm, performance of XLE degraded to a larger extent compared to NF 90. This made NF 90 the more appropriate membrane to be used for 4000 ppm solution. The use of NF 90 can save up to 21% and 17% of total investment/operating costs and operating costs per cubic meter permeate, respectively.

In addition, the outcomes from Verberne Cost Model indicated that membrane costs for 4000 ppm TDS solution range from around 8 % to 13 % in ascending order; NF 270, NF 90, TS 80, XLE and BW 30. NF 270 membrane represented high permeability membrane compared to BW 30 which had the lowest permeability. The implementation of high flux membrane did not significantly reduce the overall operating/investment costs. Furthermore, as discussed before, the performance of NF 270 was not satisfactory where additional treatment process has to be added to further treat the permeate. This definitely will incur extra capital/operation expenses and energy consumption. Hence, the use of high permeability membrane did not guarantee energy and cost savings as consistency of performance has larger impact on the desalination system and selection of membrane.

Instead, membrane with permeability moderately higher than BW 30 membrane (such as NF 90 and XLE) might be more practical for this brackish water desalination process, as compared to extremely high permeability NF 270 membrane. Even though energy consumption and total expenditure were only slightly lower than

BW 30 membrane, its controllable fouling depicts that it can be operated for a long period with consistent performance. Thus, it is safe to state that high permeability membrane does not necessary reduce the energy consumption and total expenditure of a desalination plant. One might argue that operating flux can be lowered down to reduce its fouling tendency, but other side effects may arise which will definitely offset the benefit of such low operating flux. Operation at low pressure will require more membrane areas/modules for a desired production, not to forget the much lower salt rejection where the permeate requires more thorough further treatment. All these lead to additional energy and costs to be spent for this desalination plant. Current high permeability membranes could not really achieve the expectation of cost savings due to the lower salt rejection capability. Unless there is a breakthrough where membrane with both desired characteristics is developed, then only energy consumption and total expenditure can be reduced.

In order to validate the outcomes from the economic evaluation, water costs for NF 90 membrane were compared with other reported values. According to the economic evaluation carried out by Van der Bruggen et al. (2001) with similar clean water production capacity (48000 m^3/d), the calculated operating costs per cubic meter treated water was 0.126 €. This result was close to the first year operating costs of the Mery-sur-Oise plant (nanofiltration membrane river water treatment plant with clean water production capacity of 140000 m^3/d) with operating costs of 0.12 €/m^3 [31]. For this study, the calculated operating costs per cubic meter treated water was 0.119/0.133 € for NF 90 400/4000 ppm solutions. The operating cost for low salinity solution was close to the values from other evaluation and real membrane water treatment plant. However, for higher salinity W2 solution, the higher operating cost could be attributed to its higher operating pressure due to higher salt content. In addition, lower production capacity (smaller plant) also resulted in higher water cost where it can be seen that the production capacity of nanofiltration water treatment plant was approximately 3 times larger than our case. To support this postulation, the production capacity used in Verberne Cost Model was increased to 140000 m^3/d, which was similar to the nanofiltration water treatment plant. Surprisingly, the operating costs per cubic meter treated water for W2 solution has reduced to 0.092 €, lower than the water price reported by real membrane water treatment plant. Since this estimation was based on lab-scale study, the water cost will be multiplied by

20%-40% to compensate the disparity from large-scale plant. The new estimated water price ranged from 0.11-0.13 €/m^3 and still close to the water price of real treatment plant. This showed that the economic evaluation employed in this study was considered acceptable provided systematic experimental works were conducted.

CONCLUSION

This study compared the economic feasibility of NF and RO membranes for brackish water desalination process to save costs and replace the typical brackish water RO membrane. It can be seen that Verberne Cost Model assisted by SEC, OPD and assessment of membrane performance and fouling propensity can be used to determine the appropriate membrane in brackish water desalination process. With that, NF 90 and XLE membranes can be used as alternative membrane for low TDS (400 ppm) brackish water since both encountered low fouling issues and consumed lesser energy compared to BW 30 membrane. NF 90 and XLE resulted in 17% and 12% reduction in total investment and operating costs, respectively over BW 30 membrane. The operating costs per cubic meter permeate were 13% and 9% lower than BW 30 when NF 90 and XLE were employed. However, for medium TDS (4000 ppm) brackish water, NF 90 membrane was shown to be the preferable candidate compared to XLE membrane due to its more consistent performance. The use of NF 90 for this solution can result in 21% and 17% cost savings for total costs and operating costs per cubic meter treated water. Besides that, trade-off between water production cost, total investment/operating costs and SEC have to be inspected thoroughly to determine the optimal operating recovery rate. By using the prediction from OPD and SEC calculations, the decided optimal recovery rate for 4000 ppm brackish water using NF 90 membrane was 80 % since the SEC was comparable to conventional drinking water treatment plant and the total investment/operating costs were just slightly higher than lower operating recovery rate. Outcome from this study also showed that using high permeability membrane did not result in energy consumption and cost savings. Instead, membrane with moderate flux and consistent performance was more suitable to hit these targets. Overall, this study showed that it is essential to select an appropriate membrane that produces water at low cost but without compromising the total investment/operating costs and consistency of membrane performance.

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Nomenclature for Verberne Cost Model

$C_{chemical}$	Chemicals costs, €
C_{civil}	Civil investments, €
$C_{consump}$	Consumption costs, €
C_{deprec}	Depreciation costs, €
$C_{electro}$	Electrotechnical investments, €
C_{energy}	Energy costs, €
C_{invest}	Total investments costs, €
C_{maint}	Maintenance costs, €
C_{mech}	Mechanical investments, €
$C_{membrane}$	Membrane investments, €
$C_{operating}$	Total operating costs, €
C_{spec}	Specific operation costs, €
n	Number of membrane modules
P	Operating pressure, bar
Q_F	Feed flow, m^3/h
Q_P	Permeate flow, m^3/h

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