Cost of fouling in full-scale reverse osmosis and nanofiltration installations in the Netherlands

M. Jafari [a,b,*], M. Vanoppen [b,c], J.M.C. van Agtmaal [d], E.R. Cornelissen [b,c,e], J.
S. Vrouwenvelder [a,f], A. Verliefde [b,c], M.C.M. van Loosdrecht [a], C. Picioreanu [a]

[a] Department of Biotechnology, Faculty of Applied Sciences, Delft University of Technology, Van der Maasweg 9, 2629, HZ, Delft, the Netherlands
[b] Particle and Interfacial Technology Group, Faculty of Bioscience Engineering, Ghent University, Coupure Links 653, B-9000 Ghent, Belgium
[c] Centre for Advanced Process Technology for Urban Resource Recovery (CAPTURE), 9000 Ghent, Belgium
[d] BCF Systems Separation Processes Ltd, Olmendreef 2a, 4651 RP Steenbergen, the Netherlands
[e] KWR Water Cycle Research Institute, Groningenhaven 7, 3433, PE, Nieuwegein, the Netherlands
[f] King Abdullah University of Science and Technology (KAUST), Water Desalination and Reuse Center (WDRC), Division of Biological and Environmental Science and
Engineering (BESE), Thuwal 23955-6900, Saudi Arabia

HIGHLIGHTS

- Fouling cost as fraction of OPEX is ~24% for surface water RO and ~11% for anoxic NF
- Anoxic NF systems can be regarded as a baseline with minimum fouling cost.
- Main fouling cost factor is the membrane replacement, followed by the energy cost.
- Chemical cleaning has a minor contribution to the overall cost of fouling.
- The down-time cost matters mainly for the plants with frequent cleaning events.

GRAPHICAL ABSTRACT
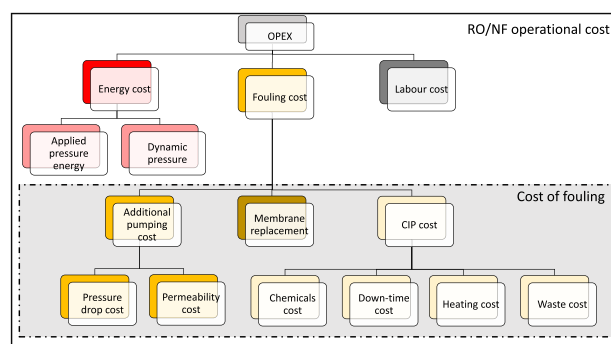
ARTICLE INFO

ABSTRACT

The economic impact of fouling in spiral wound membranes is not yet well explored. There has been an established assumption that the cost of fouling in membrane processes is significant, but this hypothesis has not been thoroughly evaluated. We conducted an economic analysis on seven full-scale installations, four nano-filtration (NF) and three reverse osmosis (RO), to estimate the cost of fouling in industrial plants. The cost of fouling was calculated in detail, including costs of increase in feed channel pressure drop, water permeability reduction, early membrane replacement, and extensive cleaning-in-place (CIP). The estimated cost of fouling was expressed as a fraction of operational expenses (OPEX) for each plant and the major cost factors in fouling and CIP costs were identified.

The selected NF plants were fed with anoxic ground water, while the feed water to RO plants was either surface water or municipal wastewater effluent. All the NF plants produce drinking water, while the RO plants

produce demineralized water for industrial applications. We found that the cost of fouling in the RO plants was around 24% of OPEX, while the fouling related costs in NF cases was only around 11% due to the low biofouling potential of the anoxic ground water. The major factor in the cost of fouling is the early membrane replacement cost, followed by additional energy and with only a minor contribution from the cleaning costs. The down-time cost (caused by the interruption of water production during a CIP event) can be the major CIP cost factor for the plants with frequent cleaning events, while the cost of chemicals dominates in the plants with non-frequent CIP. In case of manual cleaning-in-place, the cost of fouling is increased by around 2% for the RO plants with frequent CIP. The manual execution of CIP cleaning is an attention point to reconsider, as the reviewed plants hold an automated CIP cleaning, providing membrane productivity advantages.

1. Introduction

There has been extensive research on fouling and its impacts on membrane processes specifically nanofiltration (NF) and reverse osmosis (RO). The main focus is usually on evaluation and mitigation of adverse impacts of fouling on process operation. Researchers have vastly studied different types of fouling (e.g., biofouling, scaling, organic and colloidal fouling), fouling formation mechanisms [1–7], fouling properties (composition, (in)organic fractions and structure) [3,8–17], and novel characterization techniques [1,18–23]. Many studies have been dedicated to fouling mitigation [24–26] and control mechanisms [27,28]. Development of cost-effective cleaning-in-place (CIP) [29–32] as well as physical cleaning methods (e.g., back-washing, air-bubble cleaning) [33–35] have attracted significant attention. Besides, there has been a strong effort to develop anti-fouling membranes through surface modification to control fouling formation [24,36]. Moreover, many studies were dedicated to the development of optimal feed spacers [37] and pre-treatment steps [38] to control and mitigate fouling. Mathematical models have been also developed to improve understanding of fouling mechanism and predict the fouling impacts on membrane performance [39–41].

However, the economic impact of fouling on membrane processes (i. e., the extent that fouling causes additional costs on membrane processes) received only limited attention in the technical and scientific literatures. There has been an established assumption that the cost of fouling in the membrane process for water treatment is significant, but this hypothesis (to the best of the authors' knowledge) has not been well documented for RO and neither for NF systems. Some studies have reported that the cost of biofouling in a water production membrane system is around 20 to 30% of operating expenditure (OPEX) [42,43], but they did not provide any detailed calculations or background information. Fouling formation leads to an increase in OPEX through higher energy consumption, the need for early membrane replacement and additional CIP cleaning.

Porcelli et al. [31] studied in detail the cost of chemical cleaning in Ultrafiltration (UF) for potable water production. They suggested an operational scenario (including the chemical cleaning protocols) to reduce the production cost. Ang et al. [44] reported the technical and economic feasibility of brackish water desalination using several commercial RO and NF membranes. They used a common cost model in the economic assessment of water treatment plants (i.e., the Verberne cost model), which uses the practical plant parameters to calculate OPEX and CAPEX of a brackish water desalination plant. Their plant design was based on experimental data using synthetic feed water instead of the real feed water with a more complex composition, which leaves questions about the practicality of the study.

The adverse impact of fouling on RO/NF systems can be translated into: loss in water permeability [1,9], increase in feed channel pressure drop [9,12,20,33], elevated salt passage (i.e., loss of water quality leading to early membrane replacement) [8,29,45] and the necessity of periodic chemical Cleaning-In-Place (CIP) [3,30,46]. These operational problems lead to more energy consumption, capacity loss, regular membrane replacements, and in general, increasing OPEX of the plants [43,45]. In addition, there are some indirect fouling related costs such as CIP waste management, down-time (during CIP events) and CIP labour that also contribute to the overall cost of fouling in membrane water treatment [31].

In this study, we implemented a comprehensive economic analysis to calculate the cost of fouling in spiral wound membrane systems (i.e., RO, NF) which produce demineralized water and drinking water in The Netherlands. To better evaluate the relative economic impact of fouling on the daily cost of each individual plant, the cost of fouling is normalized to OPEX of each plant as suggested by [31,42,43]. When making a comparison of the cost of fouling between different plants, the introduction of a base case (i.e., a plant without any fouling) would not be logical, because fouling is inherently part of any membrane process and it cannot be excluded [47] .Therefore, to enable a fair comparison between the plants, the cost of fouling was normalized with the OPEX of each individual plant.

The aims of this study, conducted on seven full-scale water production installations in The Netherlands, were: i) to derive the total cost of fouling in spiral wound membrane systems (i.e., RO and NF) as a fraction of the OPEX, ii) to make a detailed analysis of cost factors and their contributions to CIP and fouling and iii) to evaluate the impact of manual and automated CIP on the OPEX and cost of fouling.

The novelty of this study consists in its methodology and approach to quantify the economic impact of the fouling in full-scale RO/NF. The study uses historical plant performance parameters and robust non-empirical cost models to calculate the cost of fouling in full-scale RO/NF installations. The results and approach proposed in this work improve the reliability of techno-economic analyses in water-treatment plants, which often either neglect fouling or use empirical models (instead of plant's performance data) to calculate the cost of fouling. Moreover, detailed analysis of fouling cost factors allows practitioners to better target fouling prevention/cleaning strategies. Finally, this study contributes to the scientific literature of fouling characterization by introducing a new indicator (i.e., cost of fouling as fraction of OPEX) to quantify the severity of fouling in membrane systems for water treatment.

2. Methodology

2.1. Plant description and operation

In total, seven water production plants were compared in this study: four using nanofiltration and three using reverse osmosis processes.

2.1.1. Full-scale Nanofiltration (NF) plants for drinking water production

Four NF full-scale installations (NF1, NF2, NF3 and NF4) were selected, which produce drinking water from anoxic groundwater from different groundwater wells in The Netherlands. All these plants operate at constant flux mode by adjusting feed pressure to achieve the desired flux value. All four installations consist of identical pre-treatment steps (10 μm cartridge filter and additional phosphonate-based antiscalant). The high solubility of reduced metal ions (i.e., iron) present in the water under anoxic conditions leads to a much lower fouling potential compared to aerated feed water [48]. In anoxic groundwater, a lower fouling potential is expected compared to aerated water. In anoxic

conditions, the metal ions (i.e., iron and manganese) are in their soluble reduced state (Fe^{2+} and Mn^{2+}) which leads to less fouling by precipitates, expressed by a lower fouling potential indicator. Moreover, the lack of oxygen in anoxic conditions slows down the biofilm formation (i. e., anoxic microbial growth has lower yield compared to aerobic conditions) and consequently lower biofouling potential. Furthermore, the low fouling potential characteristic of anoxic feed water enables a relatively long operational life (>10 years) of the membrane modules (Table 1).

Periodical CIPs are performed in all NF plants as the feed channel pressure increased by 25–40% compared to the start-up value of the plant [8]. The CIP protocol in all the four installations is similar and consists of two steps: *i*) acid cleaning (circulation with citric acid 2% *w/ w*, 35 °C, 3 h), and *ii*) alkaline cleaning (circulation with NaOH, pH 11–12, 0.01 M, 35 °C, 3 h). The acid-base cycles are repeated three times, and at the end the modules are rinsed with NF permeate water. The NF plants undergo non-frequent CIP routines (once every two years) due to the low fouling potential of their feed water. All the chosen plants are equipped with automated CIP systems. During acid and base cleanings in all the NF plants, the flow rate alternates between low (5 m^3/h per element) and high (10 m^3/h per element) values every half an hour. More details about the NF plants and their feed water characteristics can be found in [8]. The main characteristics of the chosen NF plants are listed in Table 1.

2.1.2. Full-scale Reverse Osmosis (RO) plants for demineralized water production

Three full-scale RO plants (RO1, RO2 and RO3) for demineralized water production in The Netherlands were selected. All plants consist of a two-stage RO system, they are all equipped with automated CIP systems and operated at a constant permeate flux mode. The RO1 plant produces demineralized water with a conductivity below 10 μS/cm from secondary wastewater effluent of a food company. The pre-treatment steps in RO1 consist of coagulation, flocculation and sedimentation processes followed by ultrafiltration (UF). The UF permeate is dosed with antiscalant to lower the scaling potential in the RO step. The RO2 plant produces demineralized water with a conductivity below 0.2 μS/ cm from river water and its pre-treatment steps include 100 μm pore sized strainer, in-line coagulation and UF. The UF permeate is again dosed with antiscalant. The RO3 produces demineralized water from surface water and its pre-treatment steps include a coarse screen, coagulation, flocculation and sedimentation processes followed by UF (no antiscalant dosing).

Periodical CIPs are performed in all the RO plants as the feed channel pressure increase by 15% [29]. The CIP protocol for RO1 is as follows: i) alkaline cleaning (circulation (9 m^3/h) with NaOH with, pH 12, 0.01 M, 35 °C, 1 h), ii) alkaline cleaning (soaking with NaOH, pH 12, 0.01 M,

35 °C, 30 min), iii) repeat the alkaline cleaning cycle 2 more times, iv) rinsing with demineralized water, v) acid cleaning (circulation (9 m^3/h) with HCl, pH 2.1, 35 °C, 1 h), vi) acid cleaning (soaking with HCl, pH 2.1, 35 °C, 30 min), vii) repeat the acid cleaning cycle 2 more times, viii) final rinsing with demineralized water.

The CIP protocol for RO2 includes the following steps: *i*) alkaline cleaning (circulation (9 m^3/h) with NaOH, pH 12, 0.01 M, 35 °C, 2 h), *ii*) alkaline cleaning (soaking with NaOH, pH 12, 0.01 M, 20 °C, ~ 20 h), *iii*) rinsing with demineralized water, *iv*) acid cleaning (circulation (9 m^3/h) with Divos 2 (JohnsonDiversey, UK), pH 1.6, 35 °C, 1.5 h), *vi*) final rinsing with demineralized water.

The CIP protocol for RO3 plant includes the following steps: *i*) soaking with demineralized water for 30 min, *ii*) alkaline cleaning (circulation (9 m^3/h) with sodium bisulphite, pH 10.5, 1–1.5% *v/v*, 35 °C, 1 h), *iii*) alkaline cleaning (soaking with sodium bisulphite, pH 10.5, 1–1.5% v/v, 20 °C, 2 h), *iv*) acid cleaning (circulation (9 m^3/h) with Divos 2 (JohnsonDiversey, UK), pH 2.5, 35 °C, 30 min), *v*) acid cleaning (soaking with Divos 2, pH 2.5, 20 °C, 1 h, *vi*) final rinsing with demineralized water. The RO plants characteristics are listed in Table 1. More details about these RO plants and their feed water can be found in [29].

The CIP efficiency of different plant varies between both plants and seasonally. For instance, the average CIP efficiency in NPD reduction in RO plants is in the range 9–15% while the average CIP efficiency in K_w recovery varies in the range of 3–10%. Similar results was observed for NF plant CIP efficiency where the average CIP efficiency based on NPD was around 10% and on K_w around 3–5%.

2.2. Plant performance data and fouling parameters

For these water production plants, the impact of fouling was estimated using the main operational parameters such as water permeability and feed channel pressure drop. These parameters are routinely monitored to ensure the optimal operation of a plant. The operational parameters are often normalized to the standard conditions to allow an objective comparison between different plants [8]. The normalized specific water permeability, K_w [m s^{-1} kPa^{-1}], is the actual membrane water flux normalized to net driving force (net transmembrane pressure) and corrected for feed water temperature [8,49]. The normalized feed channel pressure drop, *NPD* [kPa], is the actual feed channel pressure drop per membrane element adjusted for feed water temperature and volume as explained by [8,50]. Detailed calculations of the performance parameters are given in **Supplementary Information (SI)**. Performance parameters during operational time for all the seven selected plants were mainly reported in [8,29], unless specified otherwise. To evaluate the effects of fouling on the process parameters, the performance indicators have been considered in average over the operation time (i.e., at least

Table 1
Plant specifications and feed water parameters included in the current study [8,29].

Plant name → Plant characteristics ↓	NF1	NF2	NF3	NF4	RO1	RO2	RO3
Feed water source	Anoxic groundwater	Anoxic groundwater	Anoxic groundwater	Anoxic groundwater	Secondary wastewater effluent	Surface water	Surface water
Plant product	Drinking water	Drinking water	Drinking water	Drinking water	Demi water	Demi water	Demi water
Pre-treatment steps	Cartridge filter 10 μm	Cartridge filter 10 μm	Cartridge filter 10 μm	Cartridge filter 10 μm	UF	UF	UF
Years of operation	10	9	8.8	9	5.5	5	5
Production capacity (m^3/ day)	2880	1785	4608	2880	7680	4800	5000
Water recovery (%)	80	80	78	80	80	80	80
Membrane type	Trisep-8040-TS82	Trisep-8040-TS82	Trisep-8040-TS82	Trisep-8040-TS82	DOWFILMTEC BW30XFR-400/34i	DOWFILMTEC LE-440i	DOWFILMTEC LE-440i
CIP frequency (CIP/year)	0.7	0.6	0.6	0.7	17	17	7
CIP duration (hours/event)	8	8	8	8	8	24	6.5
Membrane area (m^2) (number of modules)	2412 (144)	1809 (89)	4221 (215)	2412 (144)	20,000 (500)	10,000 (250)	10,417 (260)

one round of membrane replacement). The initial values of the performance parameters (K_{w0} and NPD_0) have been recorded without the unstable values obtained during the start-up period (~ 4 days after installation) of each plant. The time-averaged performance parameters of the plants were calculated between two CIP events as explained in [31]. The performance parameters used to calculate the cost of fouling (i.e., additional costs such as additional energy consumption, loss of capacity and early membrane replacement) as fraction of operation costs have been listed in Table 2.

2.3. Economic analysis and cost estimation

A cost calculation spreadsheet has been developed in which all relevant costs caused by fouling in the full-scale installations have been included. Fig. 1 illustrates an overview of the scheme used in this study to calculate the cost of fouling as a fraction of OPEX. The economic impact of fouling on the plant OPEX was considered. CAPEX was not included in these calculations, because CAPEX calculations usually are highly case-dependent and subjective. The economic impacts of fouling on the pre-treatment steps have been considered out of the scope of this study. This is due to the fact that the considered NF plants are all equipped with identical pre-treatment steps (i.e., cartridge filter) and the RO plants have all ultrafiltration (UF) in upstream. Therefore, it is reasonable to compare fouling impact only on the main purification steps (RO, NF) and neglect often similar and typical pre-treatment steps prior to NF/RO units.

The fouling formation causes an increased feed channel pressure drop, reduction in water permeability, and an increase in salt passage. The deterioration of these performance parameters is compensated by additional pumping energy (constant flux operation mode) and consequently higher pumping costs. The replacement of membrane modules occurs normally when their lifetime was exceeded or when salt passage exceeds 5% due to extensive fouling formation (typical values of demineralized water quality) [29]. CIP cost basically consists of chemical costs (acids and bases), down-time cost, CIP solutions heating costs and waste management costs (Fig. 1). As all the plants in this study are equipped with automated CIP, labour cost in the CIP was neglected. In addition, cost of rinsing water (during rinsing steps of CIP events) is neglected as often the amount of water used is negligible (<0.5% permeate production) [51]. The cost of membrane replacement has been considered in OPEX since membrane replacement is often caused by operational issues such as severe fouling [42,52]. In all the investigated plants, the cost of brine management is negligible due to the plants

vicinity to the waterbodies which allows a safe discharge of the brine. Moreover, no contractual penalty (additional costs for delayed water delivery to clients) are considered. Contractual penalties in practise are often countered by additional plant capacity (CAPEX).

To determine the cost of fouling (ϵ_f) as a fraction of the OPEX (ϵ_{OPEX}) in each case study per year, the costs of additional pumping energy due to the fouling (ϵ_p), membrane replacement costs (ϵ_{mem}), and CIP costs (ϵ_{CIP}) have been calculated using Eq. (1). The OPEX of each plant consists of costs of energy (ϵ_{energy}), membrane replacement (ϵ_{mem}), cleaning-in-place (ϵ_{CIP}) and labour for operation of the plant (ϵ_{labour}) as stated in Eq. (2) . The additional pumping energy cost is caused by both permeability reduction cost (ϵ_{perm}) and pressure drop increase costs due to fouling deposition (ϵ_{NPD}) according to eq. (3). The CIP cost (ϵ_{CIP}) includes the costs of chemicals ($\epsilon_{chemical}$), down-time ($\epsilon_{down-time}$), heating cost ($\epsilon_{heating}$) and waste management (ϵ_{waste}) according to Eq. (4).

$$\epsilon_f = \epsilon_p + \epsilon_{mem} + \epsilon_{CIP} \tag{1}$$

$$\epsilon_{OPEX} = \epsilon_{energy} + \epsilon_{mem} + \epsilon_{CIP} + \epsilon_{labour} \tag{2}$$

$$\epsilon_p = \epsilon_{perm} + \epsilon_{NPD} \tag{3}$$

$$\epsilon_{CIP} = \epsilon_{chemical} + \epsilon_{down-time} + \epsilon_{heating} + \epsilon_{waste} \tag{4}$$

The total cost of energy consumption, ϵ_{energy} [€/year], is calculated based on the average applied pressure ($P_{f\ avg}$) (including average transmembrane pressure (TMP_{avg}), axial pressure drop due to constant flux mode and dynamic pressure), flow rate of the water, operation time and amount of down-time of the plant due to CIP cleaning (much lower pressure applied during CIP events) using Eq. (5) [31],

$$\epsilon_{Energy} = \frac{Cf_e}{\eta} P_{f_{avg}} \left(Q_f \times (t - N_{CIP} t_{CIP}) \right) \tag{5}$$

where Cf_e is the unit cost of electrical energy in The Netherlands, η is the typical electrical power conversion efficiency (Table 3), Q_f is flow rate, t is operational time (i.e., one year), N_{CIP} is the number of CIP events during the time frame, and t_{CIP} is duration of each CIP event.

Impact of fouling on membrane replacement has been considered by calculating cost of replacing the modules during the operation time according to Eq. (6):

$$\epsilon_{mem} = \frac{Cf_{module}}{t} \cdot N_{module} \tag{6}$$

where Cf_{module} is unit cost of one membrane module (depending on

Table 2

Performance parameters and clean-in-place (CIP) information for the investigated full-scale installations, used to calculate the cost of fouling as fraction of OPEX [8,29]. Performance data are averaged over the entire operational time of plants working with their original membrane modules (see Table 1 and Fig. S1 about average performance parameters and operational period).

Variables	Symbol	Units	NF1	NF2	NF3	NF4	RO1	RO2	RO3
Performance parameters									
Initial normalized pressure drop	NPD_0	kPa	275	282	251	265	168	125	123
Average normalized pressure drop	NPD_{avg}	kPa	359	359	293	344	216	173	175
Initial water permeability	K_{w0}	$\times 10^{-8}$ m s^{-1} kPa^{-1}	1.3	1.3	1.3	1.1	1	1.6	1
Average water permeability	$K_{w\ avg}$	$\times 10^{-8}$ m s^{-1} kPa^{-1}	1.2	1	~1.3	~1.1	0.8	1.1	0.8
Design flux	J_{design}	L m^{-2}/h	22	22	22	22	20	20	20
Feed flow rate	Q_f	m^3 d^{-1}	3600	2225	5908	3600	9600	6000	6250
permeate flow rate	Q_p	m^3 d^{-1}	2880	1785	4608	2880	7680	4800	5000
Average transmembrane pressure	TMP_{avg}	bar	6.3	6.3	6.3	6.3	13	13	13
CIP info									
Acid solution	pH_{acid}	[−]	4.7	4.7	4.7	4.7	2.1	1.6	2.5
Alkaline solution	pH_{base}	[−]	12	12	12	12	12	12	10.5
Acid volume/event	V_{acid}	m^3	12	7.5	17	12	20	10	10.5
Base volume/event	V_{base}	m^3	12	7.5	17	12	20	10	10.5
Acid temperature	T_{acid}	°C	35	35	35	35	35	35	20[a]-35[b]
Base temperature	T_{base}	°C	35	35	35	35	35	20[a]-35[b]	20[a]-35[b]

[a] Soaking temperature.
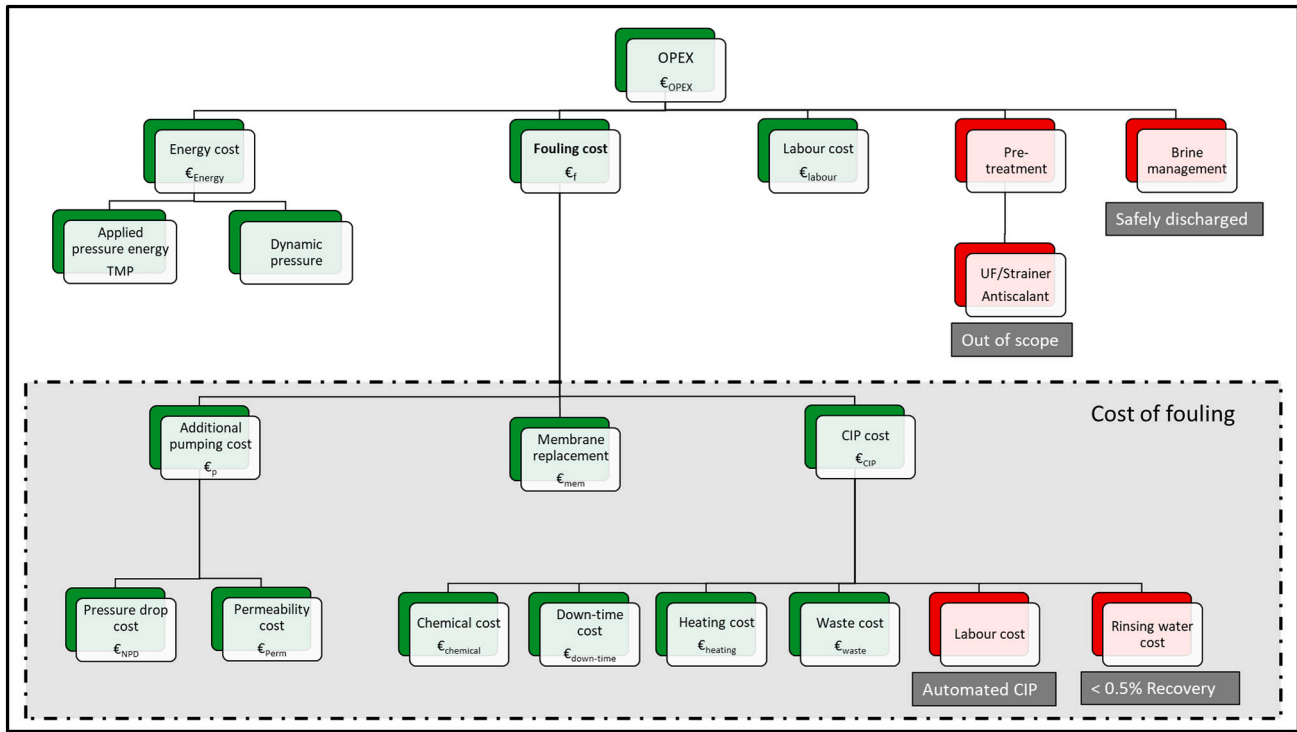[b] Circulation temperature.

Fig. 1. Schematic overview of the cost of fouling calculation as a fraction of OPEX, used in this study to compare seven full-scale RO and NF installations in The Netherlands. Cost of fouling is considered using cost factors inside the dot-dashed box. Only the cost factors in the green rectangles are considered, while the cost factors in the red rectangles are either negligible or assumed to be out of scope. The justifications for omission are provided below the red boxes. CIP pumping cost is less than 0.01% of total CIP cost.

Table 3
Cost factors for cost of fouling calculation as fraction of OPEX.

Cost factors	Symbol (unit)	Value
Power supply efficiency	η	0.6 [31]
Electricity cost	Cf_e (€/kWh)	0.10 [54]
Product's margin	Cf_w (€/m³)	0.1
Labour cost	Cf_{labour} (€/year)	50,000 [55]
Waste management cost	Cf_{waste} (€/m³)	2.6
Membrane module price	Cf_{module} (€/module)	500[b]-750[c]
Chemical cleaning cost	Cf_c	
Caustic soda (50%)[a]	(€/m³)	700 [56]
Citric acid (50%)	(€/m³)	1500 [56]
Hydrochloric acid (30%)	(€/m³)	700 [56]
Divos 2	(€/m³)	2000 [56]
Sodium bisulphite (10%)	(€/m³)	2000 [56]

[a] Solution concentration.
[b] Average Nanofiltration (NF) module price.
[c] Average Reverse osmosis (RO) module average price.

membrane type), N_{module} is the number of modules in each installation. The membranes are replaced at the end of their life time, as mentioned in Table 1 ("Operation time"). The membrane replacement costs are divided over the operation time to normalize the annual cost of membrane replacement.

Labour cost, €$_{labour}$ [€/year], as part of OPEX is calculated based on number of operators ($N_{labour} = 2$) (Personal communication with Evides Industriewater B.V.) and their annual wages (Cf_{labour}) as stated in Eq. (7)

$$\text{€}_{labour} = Cf_{labour} \cdot N_{labour} \tag{7}$$

The cost of feed channel pressure drop increase due to fouling, €$_{NPD}$ [€/year], is calculated based on the changes in normalized pressure drop (NPD) averaged during operation time (t) as stated in Eq. (8) [31]:

$$\text{€}_{NPD} = \frac{Cf_e}{\eta} \cdot t \cdot \left(NPD_{avg} - NPD_0 \right) \cdot Q_f \tag{8}$$

where NPD_{avg} is the average NPD during operation time, NPD_0 is the initial NPD in the beginning of the operation (prior to fouling).

The reduction in permeability caused by fouling is compensated by pressurizing the feed water and consequently higher energy costs. The permeability costs, €$_{perm}$ [€/year], is related to the change in the water permeability (K_w) during operation as mentioned in Eq. (9) [31]:

$$\text{€}_{perm} = \frac{Cf_e}{\eta} \cdot t \cdot J_{design} \cdot \left(\frac{1}{K_{w_{avg}}} - \frac{1}{K_{w_0}} \right) \cdot Q_f \tag{9}$$

where K_{wavg} is the average water permeability during operation time, K_{w0} is the initial permeability in the beginning of the operation (prior to fouling).

The CIP cost and its constituent cost factors are stated in eq. (4). The chemical cost in CIP events, €$_{chemical}$ [€/year], is calculated based on the volume of chemical solutions used in one cleaning event (V_{acid}, V_{base}), concentration (C_{acid}, C_{base}), number of CIP events per year (N_{CIP}), and unit cost of chemicals (Cf_c) according to Eq. (10) [31],

$$\text{€}_{chemical} = Cf_c \cdot N_{CIP} \cdot (C_{acid} \cdot V_{acid} + C_{base} \cdot V_{base}) \tag{10}$$

The down-time cost caused by disruption in water production due to each CIP event is calculated according to Eq. (11) [31].

$$\text{€}_t = Q_p \cdot Cf_w \cdot N_{CIP} \cdot t_{CIP} \tag{11}$$

where Q_p is water production (m³/h), Cf_w is the margin of unit of product water €/m³ and N_{CIP} is the number of CIP events and t_{CIP} is the duration of each CIP event.

As CIP solutions need to be heated up before each CIP events, heating cost of CIP solutions are calculated based on the required energy to heat up the chemical solutions from ambient temperature ($T_{amb} = 10$ °C) to

cleaning temperature (T_{CIP}) Eq. (12) [31].

$$\epsilon_h = \frac{Cf_e}{\eta} \cdot \rho \cdot C_p \cdot (T_{amb} - T_{CIP}) \cdot V_{CIP} \tag{12}$$

where C_p is specific heat capacity (4.2 kJ/kg/K), and ρ is the density of the chemical.

The chemical waste produced during each CIP routine is usually neutralized and the cost of waste management is calculated as suggested by [53] according to Eq. (13)

$$\epsilon_w = Cf_{waste} \cdot V_{CIP} \cdot N_{CIP} \tag{13}$$

The relevant cost factors for cost calculations are listed in Table 3. Other performance parameters can be found in Table 2.

2.3.1. Manual and automated CIP cleaning

Although all the selected plants in this study are equipped with automated CIP, manual CIP practice is still widespread in many full-scale RO/NF plants. The annual cost of manual CIP, ϵ_{manual} [€/year], is calculated based on number of operators considering safety measures ($N_{operator}$), their annual wages (Cf_{labour}) and number of CIP events per year (N_{CIP}) according to Eq. (14). Five-day work week has been considered for operators and each CIP event has been count as whole working day.

$$\epsilon_{manual} = \frac{Cf_{labour} \cdot N_{labour} \cdot N_{CIP}}{270} \tag{14}$$

This cost calculation only takes into account the "direct cost" of automated and manual CIP. All other related costs for automated CIP (e. g., maintenance and capital investments for the automation) and manual CIP (e.g., safety and incident costs) are neglected. As all the plants under investigation are equipped with automated CIP, this was considered as the baseline and the additional cost of manual CIP was evaluated. A summary of all the equations for cost calculation are presented in Table 4.

3. Results

3.1. Cost of fouling and CIP costs in relation to plant OPEX

A comprehensive economic analysis has been carried out on seven full-scale RO and NF installations to evaluate cost of fouling as fraction of the OPEX in each plant. The cost of fouling is calculated based on all

Table 4
Summary of cost calculation equations used in this study.

Equation	Number	Explanation
$\epsilon_f = \epsilon_p + \epsilon_{mem} + \epsilon_{CIP}$	(1)	Cost of fouling
$\epsilon_{OPEX} = \epsilon_{energy} + \epsilon_{mem} + \epsilon_{CIP} + \epsilon_{labour}$	(2)	Operational Expenses (OPEX)
$\epsilon_p = \epsilon_{perm} + \epsilon_{NPD}$	(3)	Pumping energy cost
$\epsilon_{CIP} = \epsilon_{chemical} + \epsilon_{down-time} + \epsilon_{heating} + \epsilon_{waste}$	(4)	CIP cost
$\epsilon_{Energy} = \frac{Cf_e}{\eta} P_{f_{avg}} \left(Q_f \times (t - N_{CIP} t_{CIP}) \right)$	(5)	Total cost of energy consumption
$\epsilon_{mem} = \frac{Cf_{module}}{t} \cdot N_{module}$	(6)	Membrane replacement cost
$\epsilon_{labour} = Cf_{labour} \cdot N_{labour}$	(7)	Labour cost
$\epsilon_{NPD} = \frac{Cf_e}{\eta} \cdot t \cdot (NPD_{avg} - NPD_0) \cdot Q_f$	(8)	Cost of feed channel pressure drop
$\epsilon_{perm} = \frac{Cf_e}{\eta} \cdot t \cdot J_{design} \cdot \left(\frac{1}{K_{w_{avg}}} - \frac{1}{K_{w0}} \right) \cdot Q_f$	(9)	Permeability reduction cost
$\epsilon_{chemical} = Cf_c \cdot N_{CIP} \cdot (C_{acid} \cdot V_{acid} + C_{base} \cdot V_{base})$	(10)	Chemical cost
$\epsilon_t = Q_p \cdot Cf_w \cdot N_{CIP} \cdot t_{CIP}$	(11)	Down-time cost due to CIP
$\epsilon_h = \frac{Cf_e}{\eta} \cdot \rho \cdot C_p \cdot (T_{amb} - T_{CIP}) \cdot V_{CIP}$	(12)	CIP solution heating cost
$\epsilon_w = Cf_{waste} \cdot V_{CIP} \cdot N_{CIP}$	(13)	CIP waste disposal cost
$\epsilon_{manual} = \frac{Cf_{labour} \cdot N_{labour} \cdot N_{CIP}}{270}$	(14)	Manual CIP cost

the considered additional costs caused by fouling (i.e., energy, membrane replacement, and chemical cleaning) as depicted in Fig. 1 and described in detail in the methodology section. An estimation of the cost of fouling allows to better quantify the negative impact of fouling in the plants under study. However, to be able to fairly compare the economic impact of fouling among different plants, the cost of fouling is normalized to the OPEX of each plant. By choosing to normalize the of cost of fouling to OPEX, a comparison can be made regardless of plant size.

A detailed analysis on the costs of fouling and of CIP was made to identify the major contribution to the cost factors. In addition, the automated and manual CIP costs were compared to evaluate the impact of CIP automation on fouling cost.

The economic evaluation revealed that the average cost of fouling is around 11% of OPEX for the NF cases, while considerably higher (24 ± 3%) for the RO plants. The fraction ($\frac{\epsilon_f}{\epsilon_{OPEX}} \times 100$) is compared for the seven full-scale installations in Fig. 2. The significant difference observed between the cost of fouling in RO and NF plants is caused by a more frequent membrane replacement in RO plants as compared to NF plants, higher energy consumption by the RO and more numerous CIP events (thus, a longer down-time) all caused by higher fouling potential in the ROs operating mainly with surface water compared to NFs that are operating with anoxic ground water.

In order to understand these notable differences between NF and RO, a more detailed cost calculation follows.

The cleaning-in-place (CIP) costs as a fraction of the total OPEX (as one of the non-operational costs of fouling) is shown in Fig. 3. The CIP cost as fraction of the total OPEX is relatively low, only ~0.5% for NF installations compared to ~2% for the RO cases. The CIP cost is higher in RO as these plants underwent more frequent CIP events compared to the NF ones.

3.2. Cost factors in fouling and CIP

The cost factors contributing to the total cost of fouling are detailed in Fig. 4, these include the costs of feed channel pressure drop (ϵ_{NPD}), water permeability (ϵ_{perm}), membrane replacement (ϵ_{mem}), and CIP (ϵ_{CIP}) for all installations in this study. Clearly, the main contributor to the total cost of fouling in all cases is the cost of membrane replacement ($\epsilon_{mem} = \sim 40$ to 65%). The energy cost to compensate for the increased feed channel pressure drop (ϵ_{NPD}) was around 30% of the total cost of fouling for the NF cases, while only ~9% for RO cases in this study. The least significant in the total cost of fouling was the CIP cost, amounting
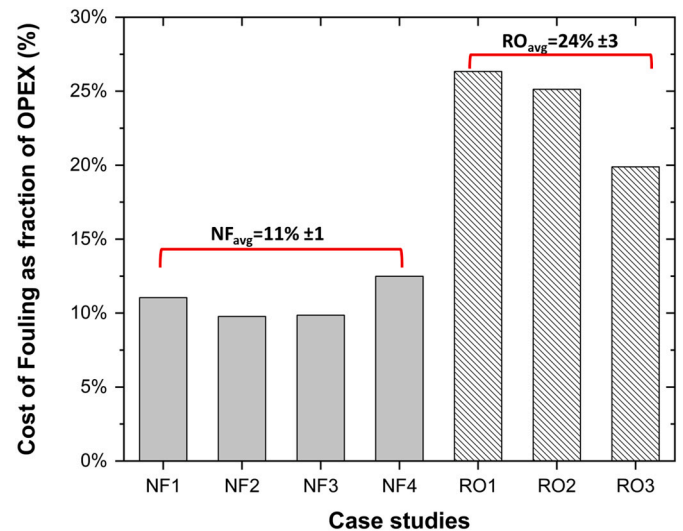

Fig. 2. Cost of fouling as a fraction of OPEX for four NF and three RO full-scale installations.
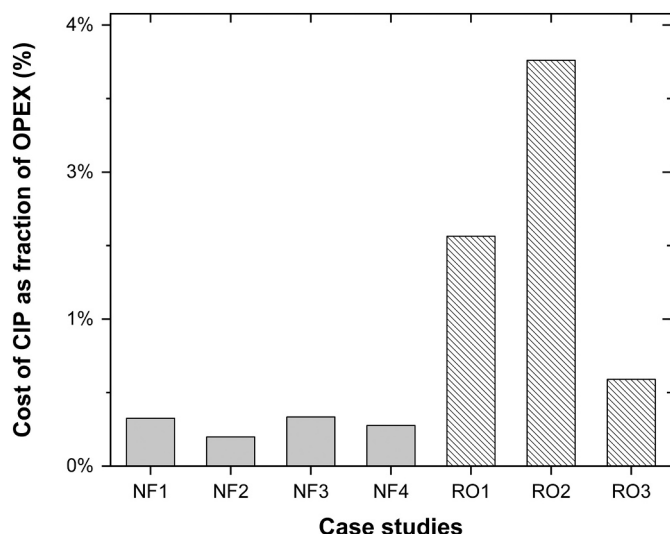
Fig. 3. Cost of CIP as fraction of OPEX for all the four NF and three RO full-scale installations.
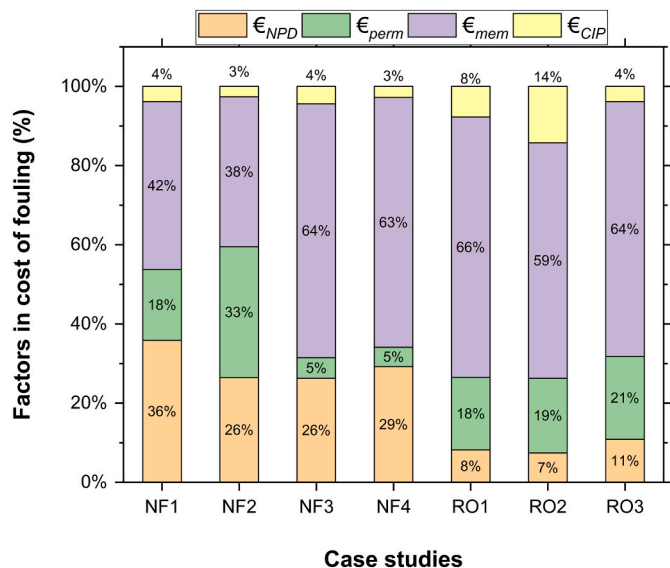

Fig. 4. Factors affecting the cost of fouling for four NF and three RO full-scale installations. Cost factors include additional pumping energy due to an increase feed channel pressure drop ($€_{NPD}$) and decreased water permeability ($€_{perm}$), as well as costs for the membrane replacement ($€_{mem}$) and cleaning-in-place ($€_{CIP}$).

from ~4% for non-frequent CIP plants (NF1–4 and RO3) to ~10% of total cost of fouling for plants that undergo frequent CIP cleaning (RO1 and RO2).

The CIP cost was furthermore also broken down into several cost factors as shown in Fig. 5. CIP cost factors were analysed in detail, as CIP cost could be optimized more easily compared to other cost factors of fouling (membrane cost and energy cost) from standpoint of utility companies (out the sphere of control of water companies).

For the NF cases, the dominant cost contributor in CIP is by far that of the chemicals (74%), followed by heating (14%), down-time cost (10%) and negligible (<2%) waste management costs (Fig. 5). Because the CIP protocols in all NF plants were very similar, the cost factor distribution was also almost identical in all NF cases. However, the RO plants had a totally different CIP cost factors distribution. The loss of revenues during down-time ranks as the most significant in all the RO cases (~ 40–70%), followed by heating costs. The cost of CIP waste disposal is negligible
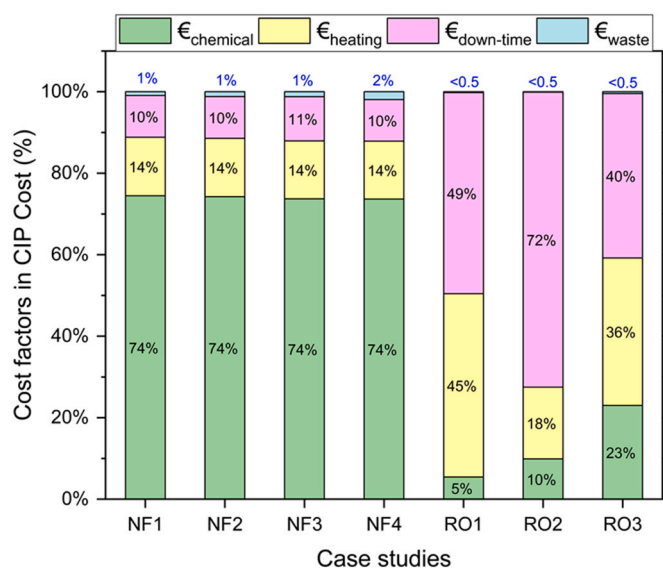

Fig. 5. The CIP cost factors for four NF and three RO full-scale installations, including the cost of chemicals ($€_{chemical}$), of heating the CIP solution ($€_{heating}$), down-time cost during CIP events ($€_{down-time}$) and waste management cost ($€_{waste}$).

(~0.2% of the total CIP cost) due to relatively low amount of CIP solution used (~ 40 L/module) and low frequency of CIP events.

3.3. Impact of manual and automated CIP on the costs of fouling and cleaning

Although CIP cleaning was automated in all the case studies, manual CIP cleaning is still widespread in many existing NF/RO full-scale installations. Therefore, we evaluated the potential cost increase in case manual CIP would be performed for all the case studies. Fig. 6a shows that the automated cleaning would not increase the cost of fouling for NF plants (<0.1%), while for RO plants up to ~2% increase in total cost of fouling is estimated because of the higher CIP frequency. The impact of manual versus automated CIP can be seen more clearly in the cost of CIP (Fig. 6b). The observed results clearly suggest that CIP automation would lead to a direct cost saving for the plants with frequent CIP events (i.e., RO plants). However, it may be possible that considering some indirect CIP costs for both manual (e.g., incident and safety costs) and automated CIP (e.g., maintenance costs) could lead to a different conclusion.

In the RO cases, with more CIP events per year, the cost of CIP as a fraction of OPEX would decrease in case of automation by a factor of 1.5 to 2 compared to the manual CIP.

4. Discussion

Fouling, as the main bottleneck of membrane processes in water treatment, is generally associated with additional costs (i.e., due to early membrane replacement, additional energy consumption and extensive cleaning) [43,57–59]. The impact of fouling on the total costs can be in the form of higher operational cost (e.g., higher energy consumption and cleaning) [31,42,44,59,60] and in a higher investment costs (e.g., additional pre-treatment steps and over-sized design) [25,26,61,62]. Fouling costs depend on plant design, operational parameters and feed water quality [57]. To calculate the cost of fouling as a fraction of operating cost (OPEX), several process and design parameters (Table 1 and Table 2) as well as the costs of consumables (Table 3) are required. In this study we analysed the cost of fouling for several full-scale NF and RO installations in The Netherlands.
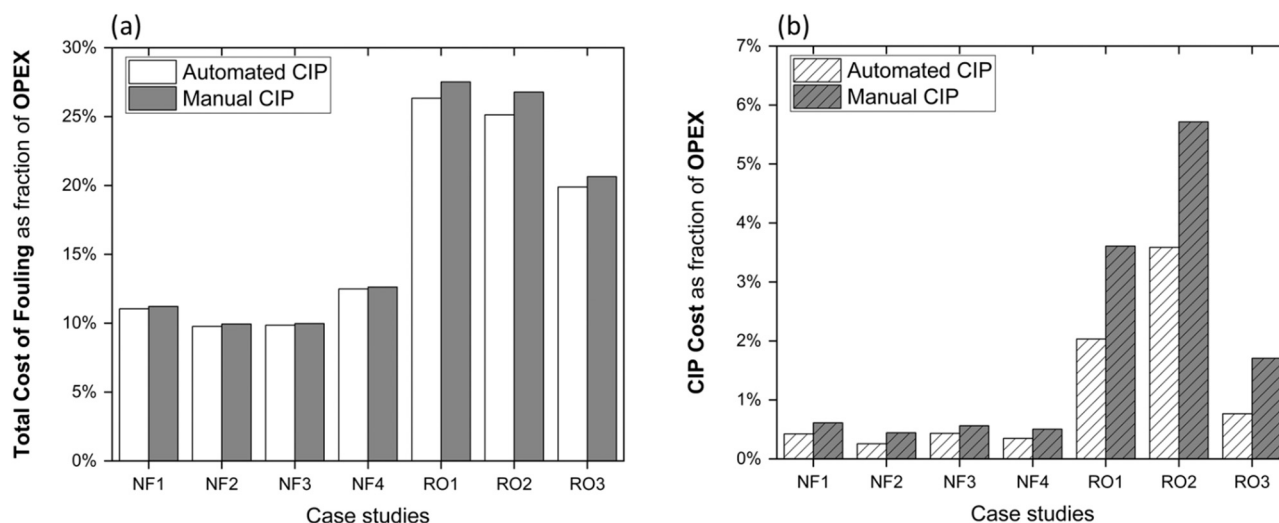
Fig. 6. A comparison between manual and automated CIP on four NF and three RO full-scale installations. Impact of manual and automated CIP on (a) total cost of fouling as fraction of OPEX (b) CIP cost as fraction of OPEX. All the cases in this study are equipped to automated CIP. Manual CIP for all the cases were calculated based on a CIP protocol involving two operators during the CIP events.

4.1. Cost of Fouling and CIP as fraction of OPEX

The calculated cost of fouling as a fraction of OPEX was around 24% for RO and 11% for NF systems with automatic CIP systems (Fig. 2). The cost of fouling in RO systems observed in this study is in the range of other reported data, such as [42,43] who estimated the cost of fouling around 20–30% of the operating costs, however, without providing detailed information on the basis for their estimates. In general, the cost of fouling is mainly correlated to the feed water type, pre-treatment steps, plant design and operational parameters. The higher cost of fouling in RO1 (~27% of OPEX) than in other RO plants (Fig. 2) can be explained by the feed water type used in this plant (wastewater treatment plant effluent)(Table 1). This also agrees with other observations of higher fouling severity in plants using wastewater treatment effluent compared to those using surface water as feed source [29]. The higher fouling cost of RO2 compared to RO3, despite their similar type of feed water, is attributed to the longer CIP cleanings for RO2 (Fig. 3). The cost of fouling in RO1 and RO2 is very similar despite the fact that RO1 is fed with municipal wastewater treatment effluent compared to RO2 which is fed with surface water. Although the RO1 is expected to be more prone to fouling compared to RO2, membrane autopsy results suggested that RO1 suffers from organic fouling while in RO2 biofouling dominates. This explains the same number of CIP events for RO1 and RO2 (Supplementary Information Fig. S2).

On the other hand, the cost of fouling in all NF plants was very similar (~ 11% of OPEX) (Fig. 2) due to almost identical feed water type, plant design, and operational conditions (Table 1). The low fouling potential in the studied NF plants is mainly associated to their feed source (anoxic ground water). The higher solubility of reduced iron ions under anoxic conditions was found to create less gel and colloidal deposits on the membrane, resulting in lower fouling compared to aerated feed water [8,48]. Moreover, in anoxic conditions, biofilm growth is slower leading to lower biomass production compared to aerobic conditions [6].

Another important issue for practitioners is the choice of frequent CIP (extensive) versus early membrane replacement. The maximum number of CIP event per years (CIP frequency) before the cost of CIP surpasses annual membrane replacement cost was calculated. Fig. 7 shows the maximum number of CIP events per year until it is economically favourable to replace the membrane rather than perform more CIP. In RO installations many more cleaning events than in the NF plants can be performed until the CIP cost exceeds the membrane replacement
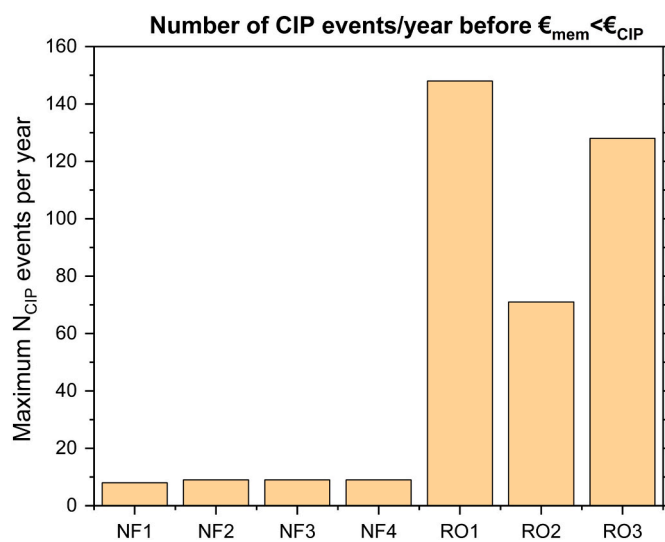

Fig. 7. Maximum number of CIP events per year until the annual cost of CIP is higher than annual cost of membrane replacement (economically viable conditions).

cost: in RO ~1–3 CIP per week, while the maximum economically viable CIP frequency in the NFs under study is around 1 CIP per month. RO2 showed a lower number of economically feasible CIP cycles (~70 CIP events per year) than the other RO plants under study. This is due to much longer down-time (soaking overnight step) in the CIP protocols of the RO2 plant.

Although the cost of fouling cannot be estimated in reference to a base case without fouling (due to the inherent fouling in all membrane processes treating water), the NF cases considered here can be regarded as being very close to a base case ("minimum" fouling costs) due to their extremely low fouling potential. On the other hand, it is expected the cost of fouling in large Sea Water Reverse Osmosis (SWRO) plants would be higher than in surface water RO plants (this study) due to significant adverse impact of scaling and cake enhanced concentration polarization (as reported by [63]).

One should note that the cost of fouling as a fraction of OPEX can change depending on what is included or neglected in both OPEX and cost of fouling calculations. For example, the cost of fouling can increase

in case of any binding contractual costs (e.g., penalty for water delivery delay) and for any limitation on CIP waste management (e.g., local discharge limits and taxes). On the other hand, extensive brine management can increase the operating costs [64,65] leading to a decrease in the cost of fouling as fraction of OPEX. Therefore, the numbers derived in this study should not be seen as absolute values, but rather as estimation criteria in comparing different plants.

4.2. Factors in CIP and fouling costs

Analysing the factors involved in the cost of fouling shows, for all seven plants in this study (NF and RO), that membrane replacement causes the largest costs (ϵ_{mem}) (Fig. 4). Similarly, Pearce [58] reported that UF membrane replacement as the biggest OPEX cost factors among four full-scale UF plants under their study. The second largest contributor in fouling costs comes from pumping ($\epsilon_{perm} + \epsilon_{NPD}$). In all RO plants, the energy required to compensate permeability reduction (ϵ_{perm}) is greater than the energy required to compensate pressure drop (ϵ_{NPD}) (Fig. 4). This is in-line with previous reports showing that maintaining the designed flux value in a fouled membrane needs more energy compared to pressure drop compensation [18].

For plants using the CIP with a low frequency, i.e., all the NF cases in this paper, the cost of chemicals ($\epsilon_{chemical}$) is the main cost factor in the CIP cost while for the plants that undergo frequent CIP events (ROs), the down-time cost ($\epsilon_{down-time}$) is the main factor in their CIP cost. This is especially visible for RO2, where the plant undergoes 17 CIP events per year and each event takes around 24 h ($\epsilon_{down-time=} \sim 70\%$ total CIP cost). Pearce [58] also reported a significant of down-time cost for plants for UF plants with frequent CIP cleaning. In contrast, waste management costs are a negligible fraction of the total CIP costs (<0.2%) for all plants (Fig. 5). However, the waste management cost (ϵ_{waste}) can be significantly higher in case of any local discharge limitation leading to additional transport costs [58]. A combined handling of CIP solutions and brine (i.e., merging CIP waste with the brine stream) could lower the waste management costs.

The information shown in Fig. 4 and Fig. 5 could be used to better target parameters to optimize CIP protocols, as described by [29–31]. For example, in RO2, the CIP protocol could be optimized by shortening the soaking time, which is the main contributor of the significant down-time cost in RO2 (Fig. 5).

4.3. Manual versus automated CIP

The impact of CIP automation has been evaluated as part of the OPEX for all selected plants, as shown in Fig. 6. Automation leads to negligible saving in OPEX for processes not needing frequent CIP, such as the NF, while CIP automation would save around 2% in OPEX for the plants that undergo frequent CIP cleanings (such as the ROs). Although CIP automation might not have a direct economic impact on OPEX for low frequency CIP plants, it could still significantly improve plant safety as reported by [66]. Even though CIP automation presents some clear technological and economic advantages, it is not yet widespread among many membrane water treatment plants. This could be perhaps due to the economic viability of manual cleaning (i.e., operator low wages often in developing countries) and other societal considerations in some countries (e.g., to secure employment). Others suggested that the labour cost (e.g., number of operators, wages) are heavily dependent on the philosophy of owner [58].

4.4. Practical implications and future studies

There has been extensive research on fouling and its impacts on membrane processes. Several techno-economic analyses of RO/NF installations have been conducted using empirical relations without considering the economic impact of fouling. Thus, the results obtained in this study will help improving the techno-economic analyses by

taking into account also the economic impacts of fouling. Considering the cost of fouling as a fraction of OPEX (~ 25%), other costs factors can still be optimized like the plant automation (leading to less labour cost) or "smart" energy consumption (adjusting water production to available excess energy) as described by [51].

We estimate higher fouling severity for seawater desalination plants compared to fresh water membrane plants (due to higher energy consumption and concentration polarization). Considering different fouling criteria for seawater RO (e.g., higher NPD threshold to perform a CIP), further research is required to calculate the cost of fouling for the seawater desalination plants.

Fouling potential in RO/NF installations is often mitigated using either extensive pre-treatment or cleaning [26]. These fouling mitigation strategies lead to additional costs, however, extensive pre-treatments are often labelled as CAPEX, while cleaning costs are included in OPEX. Thus, the plant designer should take into account several financial considerations to decide their fouling mitigation strategies. For example, cost analysis of the current study illustrates that membrane replacement cost is the biggest cost factor in the fouling cost. To reduce the membrane replacement cost, factors leading to early membrane replacement, such as CIP cleaning, should be minimized. Therefore, the non-invasive physical cleanings such as air-water cleaning and hydraulic cleaning could be the effective and economic alternatives [33,34,67].

5. Conclusions

An economic study on the cost of fouling on full-scale spiral wound membrane systems has been carried out in seven full-scale RO and NF installations in the Netherlands, using plant-wide performance data. The cost of fouling as a fraction of the plant OPEX was evaluated, together with the factors contributing to the fouling and the cleaning-in-place (CIP) costs. It has been found that:

- The cost of fouling as fraction of OPEX is around 24% for the RO installations, while only 11% for anoxic NF cases, due to the low biofouling potential of the anoxic ground water used. The cost of fouling in the anoxic NF systems can be considered as a "*minimum*" possible in a full-scale water treatment installation because of the overall low fouling propensity of anoxic groundwater;
- The most important factor in the cost of fouling is the early membrane replacement cost, followed closely by additional energy cost. CIP costs have a minor contribution to the overall cost of fouling;
- The down-time cost is the most important CIP cost factor for the plants with frequent CIP events, while the cost of chemicals dominates in the plants with non-frequent CIP;
- CIP automation could save up to 3% of OPEX for the plants with frequent cleanings, while automation would provide only negligible direct cost savings for plants with non-frequent cleaning.

Author statement

M. Jafari: Investigation, Data acquisition, Writing- Original draft preparation, **M. Vanoppen**: Investigation, **J.M.C. van Agtmaal**: Investigation, **E.R. Cornelissen**: Supervision, **J.S. Vrouwenvelder**: Investigation, **A. Verliefde**: Investigation, Supervision **M.C.M. van Loosdrecht**: Supervision, Funding, **C. Picioreanu**: Supervision, Funding.

Declaration of competing interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

Acknowledgements

Appendix A. Supplementary data

Supplementary data to this article can be found online at https://doi.org/10.1016/j.desal.2020.114865.

References

[1] A. Al Ashhab, A. Sweity, B. Bayramoglu, M. Herzberg, O. Gillor, Biofouling of reverse osmosis membranes: effects of cleaning on biofilm microbial communities, membrane performance, and adherence of extracellular polymeric substances, Biofouling 33 (2017) 397–409.
[2] C. Dreszer, A.D. Wexler, S. Drusova, T. Overdijk, A. Zwijnenburg, H.C. Flemming, J.C. Kruithof, J.S. Vrouwenvelder, In-situ biofilm characterization in membrane systems using optical coherence tomography: formation, structure, detachment and impact of flux change, Water Res. 67 (2014) 243–254.
[3] Q. Li, M. Elimelech, Organic fouling and chemical cleaning of Nanofiltration membranes: measurements and mechanisms, Environ. Sci. Technol. (2009) 4393–4398.
[4] N.M. Farhat, L. Javier, M.C.M. Van Loosdrecht, J.C. Kruithof, J.S. Vrouwenvelder, Role of feed water biodegradable substrate concentration on biofouling: biofilm characteristics, membrane performance and cleanability, Water Res. 150 (2019) 1–11.
[5] Y. Kim, S. Li, N. Ghaffour, Evaluation of different cleaning strategies for different types of forward osmosis membrane fouling and scaling, Journal of Membrane Science 596 (2020).
[6] J.V.P.P. Hiemstra, B. Rietman, J. Verdouw, Aerobic versus anaerobic nanofiltration: fouling of membranes, in: AWWA Membrane Conference, 1999. Long Beach, CA.
[7] H.S. Vrouwenvelder, J.A.M. van Paassen, H.C. Folmer, J.A.M.H. Hofman, M. M. Nederlof, D. van der Kooij, Biofouling of membranes for drinking water production, Desalination 118 (1998) 157–166.
[8] F. Beyer, B.M. Rietman, A. Zwijnenburg, P. van den Brink, J.S. Vrouwenvelder, M. Jarzembowska, J. Laurinonyte, A.J.M. Stams, C.M. Plugge, Long-term performance and fouling analysis of full-scale direct nanofiltration (NF) installations treating anoxic groundwater, J. Membr. Sci. 468 (2014) 339–348.
[9] N.M. Farhat, L. Javier, M.C.M. Van Loosdrecht, J.C. Kruithof, J.S. Vrouwenvelder, Role of feed water biodegradable substrate concentration on biofouling: biofilm characteristics, membrane performance and cleanability, Water Res. 150 (2019) 1–11.
[10] H. Sanawar, N.M.F.I. Pinel, S.Z.S. Bucs, J. Zlopasa, J.C. Kruithof, G.J.M.C.M.V. L. Witkamp, J.S. Vrouwenvelder, Enhanced biofilm solubilization by urea in reverse osmosis membrane systems, Water Res 1 (2018).
[11] D.J. Miller, P.A. Araujo, P.B. Correia, M.M. Ramsey, J.C. Kruithof, M.C. van Loosdrecht, B.D. Freeman, D.R. Paul, M. Whiteley, J.S. Vrouwenvelder, Short-term adhesion and long-term biofouling testing of polydopamine and poly(ethylene glycol) surface modifications of membranes and feed spacers for biofouling control, Water Res. 46 (2012) 3737–3753.
[12] J.S. Vrouwenvelder, J. Buiter, M. Riviere, W.G.J. van der Meer, M.C.M. van Loosdrecht, J.C. Kruithof, Impact of flow regime on pressure drop increase and biomass accumulation and morphology in membrane systems, Water Res. 44 (2010) 689–702.
[13] J.S. Vrouwenvelder, C. Hinrichs, W.G.J. Van der Meer, M.C.M. Van Loosdrecht, J. C. Kruithof, Pressure drop increase by biofilm accumulation in spiral wound RO and NF membrane systems: role of substrate concentration, flow velocity, substrate load and flow direction, Biofouling 25 (2009) 543–555.
[14] W. Guo, H.H. Ngo, J. Li, A mini-review on membrane fouling, Bioresour. Technol. 122 (2012) 27–34.
[15] Q. Li, M. Elimelech, Organic fouling and chemical cleaning of Nanofiltration membranes: measurements and mechanisms, Environ. Sci. Technol. 38 (2004) 4683–4693.
[16] F. Meng, S.R. Chae, A. Drews, M. Kraume, H.S. Shin, F. Yang, Recent advances in membrane bioreactors (MBRs): membrane fouling and membrane material, Water Res. 43 (2009) 1489–1512.
[17] T. Yu, L. Meng, Q.B. Zhao, Y. Shi, H.Y. Hu, Y. Lu, Effects of chemical cleaning on RO membrane inorganic, organic and microbial foulant removal in a full-scale plant for municipal wastewater reclamation, Water Res. 113 (2017) 1–10.
[18] M. Jafari, N. Derlon, P. Desmond, M.C.M. van Loosdrecht, E. Morgenroth, C. Picioreanu, Biofilm compressibility in ultrafiltration: a relation between biofilm morphology, mechanics and hydraulic resistance, Water Res. 157 (2019) 335–345.
[19] M. Jafari, P. Desmond, M.C.M. van Loosdrecht, N. Derlon, E. Morgenroth, C. Picioreanu, Effect of biofilm structural deformation on hydraulic resistance during ultrafiltration: a numerical and experimental study, Water Res. 145 (2018) 375–387.
[20] W.A. Hijnen, E.R. Cornelissen, D. van der Kooij, Threshold concentrations of biomass and iron for pressure drop increase in spiral-wound membrane elements, Water Res. 45 (2011) 1607–1616.
[21] P. Desmond, J.P. Best, E. Morgenroth, N. Derlon, Linking composition of extracellular polymeric substances (EPS) to the physical structure and hydraulic resistance of membrane biofilms, Water Res. 132 (2018) 211–221.
[22] R.B. Mohle, T. Langemann, M. Haesner, W. Augustin, S. Scholl, T.R. Neu, D. C. Hempel, H. Horn, Structure and shear strength of microbial biofilms as determined with confocal laser scanning microscopy and fluid dynamic gauging using a novel rotating disc biofilm reactor, Biotechnol. Bioeng. 98 (2007) 747–755.
[23] J.S. Vrouwenvelder, J.A.M. van Paassen, L.P. Wessels, A.F. van Dam, S.M. Bakker, The membrane fouling simulator: a practical tool for fouling prediction and control, J. Membr. Sci. 281 (2006) 316–324.
[24] S. Bucs, R. Valladares Linares, A. Siddiqui, A. Matin, Z. Khan, M.C.M. van Loosdrecht, R. Yang, M. Wang, K.K. Gleason, J.C. Kruithof, J.S. Vrouwenvelder, Coating of reverse osmosis membranes with amphiphilic copolymers for biofouling control, Desalin. Water Treat. 68 (2017) 11.
[25] W. Gao, H. Liang, J. Ma, M. Han, Z.-l. Chen, Z.-s. Han, G.-b. Li, Membrane fouling control in ultrafiltration technology for drinking water production: a review, Desalination 272 (2011) 1–8.
[26] S. Jiang, Y. Li, B.P. Ladewig, A review of reverse osmosis membrane fouling and control strategies, Sci. Total Environ. 595 (2017) 567–583.
[27] J.S. Vrouwenvelder, F. Beyer, K. Dahmani, N. Hasan, G. Galjaard, J.C. Kruithof, M. C. Van Loosdrecht, Phosphate limitation to control biofouling, Water Res. 44 (2010) 3454–3466.
[28] N. Derlon, J. Mimoso, T. Klein, S. Koetzsch, E. Morgenroth, Presence of biofilms on ultrafiltration membrane surfaces increases the quality of permeate produced during ultra-low pressure gravity-driven membrane filtration, Water Res. 60 (2014) 164–173.
[29] F. Beyer, J. Laurinonyte, A. Zwijnenburg, A.J.M. Stams, C.M. Plugge, Membrane fouling and chemical cleaning in three full-scale reverse osmosis plants producing demineralized water, J. Eng. 2017 (2017) 1–14.
[30] S. Madaeni, Y. Mansourpanah, Chemical cleaning of reverse osmosis membranes fouled by whey, Desalination 161 (2004) 13–24.
[31] N. Porcelli, S. Judd, Chemical cleaning of potable water membranes: the cost benefit of optimisation, Water Res. 44 (2010) 1389–1398.
[32] N. Porcelli, S. Judd, Chemical cleaning of potable water membranes: a review, Sep. Purif. Technol. 71 (2010) 137–143.
[33] E.R. Cornelissen, J.S. Vrouwenvelder, S.G.J. Heijman, X.D. Viallefont, D. Van Der Kooij, L.P. Wessels, Periodic air/water cleaning for control of biofouling in spiral wound membrane elements, J. Membr. Sci. 287 (2007) 94–101.
[34] Y. Wibisono, E.R. Cornelissen, A.J.B. Kemperman, W.G.J. van der Meer, K. Nijmeijer, Two-phase flow in membrane processes: a technology with a future, J. Membr. Sci. 453 (2014) 566–602.
[35] O.D. Basu, Backwashing, in: E. Drioli, L. Giorno (Eds.), Encyclopedia of Membranes, Springer, Berlin Heidelberg, Berlin, Heidelberg, 2015, pp. 1–3.
[36] D. Rana, T. Matsuura, Surface modifications for antifouling membranes, Chem. Rev. 110 (2010) 2448–2471.
[37] H.S. Abid, D.J. Johnson, R. Hashaikeh, N. Hilal, A review of efforts to reduce membrane fouling by control of feed spacer characteristics, Desalination 420 (2017) 384–402.
[38] N. Prihasto, Q.-F. Liu, S.-H. Kim, Pre-treatment strategies for seawater desalination by reverse osmosis system, Desalination 249 (2009) 308–316.
[39] A.I. Radu, M.S.H. van Steen, J.S. Vrouwenvelder, M.C.M. van Loosdrecht, C. Picioreanu, Spacer geometry and particle deposition in spiral wound membrane feed channels, Water Res. 64 (2014) 160–176.
[40] A.I. Radu, J.S. Vrouwenvelder, M.C.M. van Loosdrecht, C. Picioreanu, Modeling the effect of biofilm formation on reverse osmosis performance: flux, feed channel pressure drop and solute passage, J. Membr. Sci. 365 (2010) 1–15.
[41] C. Picioreanu, J.S. Vrouwenvelder, M.C.M. van Loosdrecht, Three-dimensional modeling of biofouling and fluid dynamics in feed spacer channels of membrane devices, J. Membr. Sci. 345 (2009) 340–354.
[42] J. Johnson, Effect Of Fouling On The Operating Expense (Opex) Of Reverse Osmosis Systems, AMTA Conference, Las Vegas, USA, 2014.
[43] H.-C. Flemming, Biofouling and me: my Stockholm syndrome with biofilms, Water Res. 115576 (2020).
[44] W.L. Ang, D. Nordin, A.W. Mohammad, A. Benamor, N. Hilal, Effect of membrane performance including fouling on cost optimization in brackish water desalination process, Chem. Eng. Res. Des. 117 (2017) 401–413.
[45] A. Jiang, H. Wang, Y. Lin, W. Cheng, J. Wang, A study on optimal schedule of membrane cleaning and replacement for spiral-wound SWRO system, Desalination 404 (2017) 259–269.
[46] W.A. Hijnen, C. Castillo, A.H. Brouwer-Hanzens, D.J. Harmsen, E.R. Cornelissen, D. van der Kooij, Quantitative assessment of the efficacy of spiral-wound membrane cleaning procedures to remove biofilms, Water Res. 46 (2012) 6369–6381.
[47] S.A.S.R.D. Noble, Membrane Separations Technology: Principles and Applications, Elsevier Science, 1995.

[48] J.A.M. van Paassen, J.C. Kruithof, S.M. Bakker, F.S. Kegel, Integrated multi-objective membrane systems for surface water treatment: pre-treatment of nanofiltration by riverbank filtration and conventional ground water treatment, Desalination 118 (1998) 239–248.

[49] C.A.C. van de Lisdonk, B.M. Rietman, S.G.J. Heijman, G.R. Sterk, J.C. Schippers, Prediction of supersaturation and monitoring of scaling in reverse osmosis and nanofiltration membrane systems, Desalination 138 (2001) 259–270.

[50] W.B.P. van den Broek, M.J. Boorsma, H. Huiting, M.G. Dusamos, S. van Agtmaal, Prevention of Biofouling in Industrial RO Systems: Experiences with Peracetic Acid, Water Practice and Technology, 5, 2010.

[51] G.W.I. (GWI), 'Big data desalination' approach pays off for ACWA Power, in: GWI Magazine, 2018, pp. 12–13.

[52] J. Mendret, A. Azais, T. Favier, S. Brosillon, Urban wastewater reuse using a coupling between nanofiltration and ozonation: techno-economic assessment, Chem. Eng. Res. Des. 145 (2019) 19–28.

[53] E. Zondervan, B.H.L. Betlem, B. Blankert, B. Roffel, Modeling and optimization of a sequence of chemical cleaning cycles in dead-end ultrafiltration, J. Membr. Sci. 308 (2008) 207–217.

[54] Statistica, Energy Cost in EU Countries, 2019.

[55] Statistica, Salary Indicator in EU Countries, 2019.

[56] S. Chauveau, in: M. Jafari (Ed.), Cost of Biofouling in Reverse Osmosis, 2020.

[57] H.C. Flemming, Biofouling in water systems – cases, causes and countermeasures, Appl. Microbiol. Biotechnol. 59 (2002) 629–640.

[58] G.K. Pearce, A cost optimization study of flux and fouling rate for UF in the water industry, Water Supply 8 (2008) 113–120.

[59] A.I. Schäfer, A.G. Fane, T.D. Waite, Cost factors and chemical pretreatment effects in the membrane filtration of waters containing natural organic matter, Water Res. 35 (2001) 1509–1517.

[60] E.J. Nya, Biofilms and biofouling - Cost and effect on drinking water quality for human development.pdf, Journal of Global Biosciences 4 (7) (2015).

[61] S.H. Ebrahim, M.M. Abdel-Jawad, M. Safar, Conventional pretreatment system for the Doha reverse osmosis plant: technical and economic assessment, Desalination 102 (1995) 179–187.

[62] N.M. Farhat, C. Christodoulou, P. Placotas, B. Blankert, O. Sallangos, J. S. Vrouwenvelder, Cartridge filter selection and replacement: optimization of produced water quantity, quality, and cost, Desalination 473 (2020) 114172.

[63] S. Kim, S. Lee, E. Lee, S. Sarper, C.-H. Kim, J. Cho, Enhanced or reduced concentration polarization by membrane fouling in seawater reverse osmosis (SWRO) processes, Desalination 247 (2009) 162–168.

[64] J. Morillo, J. Usero, D. Rosado, H. El Bakouri, A. Riaza, F.-J. Bernaola, Comparative study of brine management technologies for desalination plants, Desalination 336 (2014) 32–49.

[65] A. Zhu, A. Rahardianto, P.D. Christofides, Y. Cohen, Reverse osmosis desalination with high permeability membranes — cost optimization and research needs, Desalin. Water Treat. 15 (2010) 256–266.

[66] W. Labs, Automated Cleaning and Sanitation Saves Time, Produces Better Results, in, Food Engineering, 2014.

[67] R. Wilson, M. Fazel, S. Jarrige, S. Chesters, Air Bubbles Enchance Membrane Cleaning: A Future Perspective, in: The International Desalination Association World Congress on Desalination and Water Reuse Tianjin, China, 2013.