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Cleaning Strategies and Cost Modelling of Experimental Membrane-based Desalination Plants

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Abstract In Project WASTEC, an experimental Reverse Osmosis (RO) desalination system was developed. It serves as a platform for testing new technologies. For this system, we solved two problems, which are described in this paper. Firstly, we developed and investigated strategies for scheduling chemical enhanced backwashing and chemical cleaning and secondly, due to the experimental nature of the project, several new technological developments with respect to materials and methods were integrated into the system and requires tools for evaluating the economic viability of the new technologies. In this task, the economics of membrane-based desalination will be investigated. Baseline systems of reverse osmosis and pretreatment systems (microfiltration and ultrafiltration) will be economically examined and compared for their investments and operational costs. Sensitivity of the different plant and membrane parameters to the cost will be studied. Results show that with respect to costs, for a 200m^3/hr design capacity plant, a volume of water is produced by a MF process at a cost of \$0.494 and at a cost of \$0.486 by an ultrafiltration process microfiltration. The reverse osmosis process cannot be compared directly, but it required \$ 0.49 / m^3 for a plant with 56 m^3/hour design capacity. The values are in line with the costs reported in literature for membrane-based filtration.

Keywords: *operational strategy, membrane filtration, cost estimation, cost sensitivity analysis*

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1. Introduction

Climate change and its consequences on water availability require innovative technologies in water production. Recently, the application of membrane filtration in water treatment is increasingly becoming popular. In this field, several technological innovations are happening. This may be in pre-treatment methods, membrane materials or coatings, energy recovery units etc. In our ICON-Project WASTEC, an experimental Reverse Osmosis (RO) desalination plant was developed with a number of technological innovations in membrane coatings, pretreatment with microfiltration and ultrafiltration and operating strategies. One common phenomenon of membrane filtration is membrane degradation through fouling and scaling, which makes the membranes lose their performance with time and consequently requirement of early membrane replacement and additional operating costs. Furthermore, membrane degradation is associated with increased energy use. One most commonly used solution to this issue is regular backwashing (BW) and chemical cleaning, which requires intelligent scheduling with respect to energy, time, chemicals and cost savings. Cleaning scheduling has been studied in some papers, e.g., [1,2,3,4]. In Thejani et al. [3] an operational strategy for backwashing was developed for a side stream tubular domestic wastewater treatment system. In [2], a cleaning sequence model was developed and used for optimizing cost and energy. Chen et al. [5] developed a model predictive control system to mitigation the effects of chemical cleanings on membranes. In [4], the author investigated the application of air injection into the membrane before the process of backwashing. Their results show clearly that their method brought better performance and that frequency of the air aided backwash cycles was vital to mitigation of membrane degradation. The research in [6] investigates ozone-based Chemical Enhanced Backwash (CEB) for microbial fouling control of ultrafiltration membranes made of ceramic. In case of more compressed cake on the membrane, their results indicates a better performance of about 35% compared to the classical backwash without ozone CEB. For our experimental RO desalination plant, clean-in-place (CIP) operation is replaced by ozone CEB to reduce downtime period, reduce chemical costs as it is shown in [7] that ozone concentration required in CEB is relatively low compared to a CIP, and the filtration downtime resembles that of a classical hydraulic backwash. In this plant, a cleaning cycle is made up of the Filtration period, the classical BW and the ozone CEB. Therefore, we developed and tested several model-based strategies for scheduling membrane cleaning, which include fixed-time schedules, condition-based schedules and self-learning and adaptive schedule optimized for cost, chemicals and energy use. These strategies will be compared and discussed in this paper.

Technological advancements have been confirmed in many studies, for example, Panagopoulos et al. [8] showed an improvement in their RO desalination plant from $2.0/m^3 in 1998 to $0.5/m^3 in 2019. Before new technologies are put on the market, various factors should be considered in the analysis of alternatives. Nevertheless, cost is a vital factor, which should be estimated.

Some studies on cost estimation of membrane-based filtration systems have been conducted [9,10,11]. In the 90s, quite a number of reports on cost estimation in membrane filtration could be found and that is when the first cost models such as WTCost and Deep were introduced. For example, Chellam [11] reported capital and O&M costs for existing MF, UF, and NF plants and Sanz et al. [12] studied the cost of water treatment with RO membranes. He used a number of assumptions to get the capital cost of each component and for the operating costs; he used semi-empirical relationships of operating cost with capacity, feed rate and the number of membrane modules, which are installed at the plant. His estimations for membrane replacement as an annual cost varied from 10 - 25 % of the initial membrane cost. In 1991, Fuquaer et al. [13] collected data through survey of membrane plants in Florida to develop cost models for large RO plants. In their model, they expressed the Capital costs solely as a linear function of the plant design capacity and the Operating costs were expressed as a function of plant size and the feed water quality. Membrane Replacement cost was estimated at 20% of the cost required to replace all of the membranes in the installation. WTCost© [14] is a development by the US Bureau of Reclamation for calculating water treatment costs by membrane filtration. With the software, users can evaluate their water treatment processes.

These models are very good for scaling existing plants, but are not suitable for detailed evaluation of economic viability of alternatives based on new technological developments as in our case, therefore, the development of other type of empirical and parametric cost model for evaluating the economic viability of alternatives are required.

Technological advancements in designing membranes and system integration have decreased the cost to desalinate brackish water by over 50% in the last 20 years [15], e.g., the Texas Water Development Board estimated a range from $0.29 to $0.67 per m^3 total cost for producing drinking water from brackish groundwater [16]. However, a Water Reuse Association study in 2012 showed that cost trends for large Seawater Reverse Osmosis projects appear to have flattened since 2005, but have varied widely in the range of $0.78 to $2.39 per m^3 since then [17]. In a recent study, Panagopoulos [18] reported in China a drop of the cost of desalinated seawater from $0.59-$1.10/m^3, and a drop in the desalination cost of brackish water in the range of $0.29-$0.61/m^3. In Atab et al. [19] a cost of 0.12 £/m^3 for desalinating 24,000 m^3/d of water was reported. The salinity of the input water was measured at 15,000 ppm and the required permeate water quality was at less than 400 ppm. Atab et al. [19] also stipulated that the cost would be further reduced by using the resulting brine water to make salt and extract minerals.

Most of these studies have been done in the Middle East for large desalination plants. In most of these studies, the cost components include construction and buildings, chemical feed systems, control and instrumentation, site work, storage, concentrate disposal, process piping, yard piping, site electric, pretreatment, cleaning, and booster pumps. Some studies include though special components such as aerators. In most studies, labor and maintenance costs, pretreatment costs are estimated as factored capital costs. For example in Bartholomew et al. [20], the authors reported of including direct cost of membrane units, pumps, and pressure exchangers in their study to estimate the capital cost and then they used factors to account for indirect capital, installation, siting, and engineering costs. They also estimated the energy, saline make-up, and membrane replacement operating costs directly and indirectly estimated maintenance & labor and chemical operating costs using factors. Ncube and Inambao analyzed the effects of energy and costs to RO desalination technology, emphasizing on energy recovery with devices such as pressure exchangers and turbine type energy recovery devices [21].

In Judd et al. [22] the present value of construction of low-pressure membrane filtration plants and their operation was determined for the UK. In the study, data was acquired from 15 full-scale plants and the authors used cost curves for their cost calculations as in the Global Water Intelligence Desalination Cost Estimator. They included cost for labor, energy, chemicals and membrane replacement. The present value of the operation costs was found to be more than the capital costs by 3-5 times based on a life cycle of twenty years. They also found out that the difference increased as the design capacities decreased.

There are various commercially available desalination cost models on the market. These include the WTCost©, Global Water Intelligence Desalination SWRO Cost Estimator, Desalination Economic Evaluation Program (DEEP), AUDESSY, WRA model, ROSA, and CH2M HILL's proprietary cost model. Most of these models are factored models, i.e., they model use capital cost estimates for the major equipment, and then adds factors to account for the remainder of the capital costs. This assumes a lot of vendor information. Required for Pilot projects is an empirical and parametric model for studying early stage development.

Therefore, we developed an empirical and parametric model that estimates capital and operating cost of cross-flow membrane filtration based on the relevant plant design parameters. Information gathered from membrane manufacturers was used to empirically derive a regression equation for capital cost. We break down the capital costs into the pipes, pumps cost, and the cost of the membranes and therefore, it is reasonable to assume that the pipes and pumps cost varies according to the number of modules installed in the plant. First, the capacity of the plant must be translated into the number of modules required to provide the design flow. We divide the operating cost into components for energy consumption, chemical dosage, membrane replacement, concentrate disposal, and is determined from attributes of the membrane system.

Our main contributions can be summarized as follows:

- Development and comparison of several model-based strategies for scheduling membrane cleaning (timed, conditioned and self-optimizing schedules)
- Development of an empirical and parametric model for estimating capital and operating cost of cross-flow membrane filtration based on the relevant plant design parameters
- Application of time-dependent permeate flux which considers membrane pore size, operating conditions and flux enhancement
- Examination of the effect of operating parameters and membrane characteristics on capital and operating costs.

2. Methods and Materials

2.1. Strategies for Scheduling Membrane Cleaning

The procedure for cleaning of the experimental membrane filtration system is regular classical hydraulic backwashing and ozone CEB. Three cleaning strategies were investigated, which include fixed-time schedules, condition-based schedules and self-learning and adaptive schedule optimized for cost, chemicals and energy use. To be able to study the effects of strategies for scheduling the cleaning cycle, we collected data from the system about the change in permeate flow over time and developed a permeate flow model parameterizable for different feedwater qualities. We did experiments with the chemical dosing and to find the required cleaning durations for both hydraulic backwash and ozone CEB and to what level the permeate flux is recovered.

Membrane degradation

Before conducting the investigation on the strategies, a model of the characteristics of the flux through the membrane over time was required. It has been shown in many empirical studies that the water permeability of a membrane declines exponentially with time [23,24,25]. It is also known that we can obtain the permeate flow rate by multiplying the water permeability, the transient membrane pressure and the factors for temperature correction and fouling as in Equation 1. Therefore, the permeate flow rate resembles the exponential character of the permeability of water and have the same time constant.

$$J_w = A_w F\left(\Delta P - \Delta\pi\right) \tag{1}$$

where Aw is the water permeability without fouling, F is the fouling factor, t is the operation time.

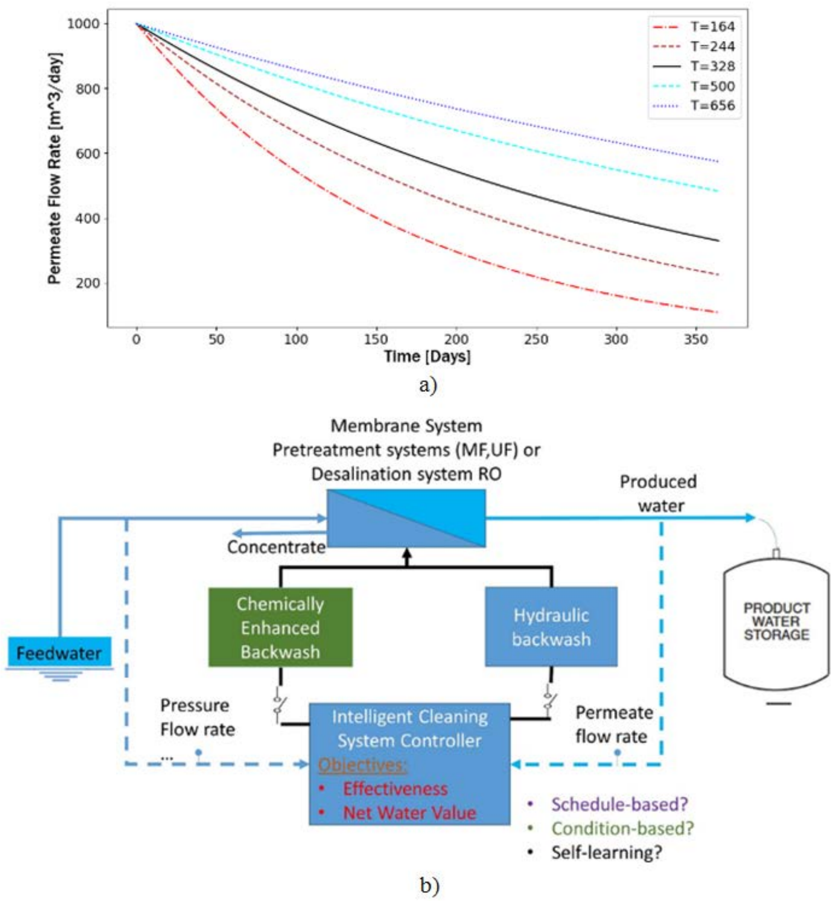

Figure 1. (a) The behavior of the permeate flow rate for a cubic meter per day water production system without membrane regeneration. The water fed into the system varied in composition and therefore the fouling potential for the three cases. (b) Setup for studying cleaning strategies

Therefore, for a one cubic meter per day water production system without membrane regeneration, the permeate flow rate was observed in a long-term study for a period of two years. Water of different quality and fouling potential was fed into the system. The collected permeate flux data was normalized and used to develop a flux decline model as shown in Figure 1 (a). We then use the developed model for investigating the different physical and chemical cleaning strategies as illustrated in Figure 1(b).

Other empirical studies were conducted to determine the dependency of the fouling rate proxied by the decrease in Transmembrane pressure (TMP) on the effort put in cleaning. The cleaning effort is defined by Equation 2 as the ratio of the cleaning dose (i.e. product of the CEB duration and chemical concentration) to CEB frequency.

$$C_{effort} = \frac{CEB\ Duration[\mathrm{min}] * Chem\ Conc.[ppm]}{CEB\ frequency[hr]} \quad (2)$$

An exponential decay dependency as presented in Equation 3 and plotted in Figure 2 was observed between fouling rate and cleaning effort.

$$y = ae^{-kx} \quad (3)$$

A rapid decrease of the fouling rate as the cleaning effort increases can be seen. Further increase of the cleaning effort after 20min-ppm/hr as well as extending the CEB duration beyond 5 minutes does not bring significant improvements and observable benefits. For the feed water conditions in the investigation, the recommended operating range is between 10 and 20 min-ppm/hr, using a CEB cleaning duration of 2 to 5 minutes.

Membrane cleaning strategies

Two technologies were scheduled for cleaning, water only BW and CEB replacing cleaning-in-place (CIP). During BW, water is flushed on top of the membrane and a backwash flux (J_{bw}) using some of the product water is applied at a pressure P_{bw} from the bottom. During this process, a specific energy E_{bw} kWh/m^3 is required. With backwash, the water production drop of the membrane is restored to r_{bw}%. In the case of CEB, a chemical dose is applied to the water during the backwash. It is assumed that water production drop of the membrane recovers by r_c% after a CEB. During CEB, a water flux J_{ceb} from the clean permeate and a specific energy E_{ceb} kWh/m^3 are required.

Several logical combinations of the BW and CEB were tested and evaluated in a **time-based schedule** and **condition-based schedules** over a year for a 1000 liter per day system and different empirical time constants T of 164, 328 and 656 days. For the **time-based schedule**, the results of two combinations for a decay constant of 328 days are shown below in comparison to one another and to the one without maintenance. In the first strategy, Case 1: BW was run daily basis and CEB on a monthly basis. In the second strategy, Case2: BW was run on a weekly basis and the rate of chemical cleaning CEB was increased from monthly to weekly. The metrics used for comparison were the permeate flow rate, the net water production and the cumulative water produced.

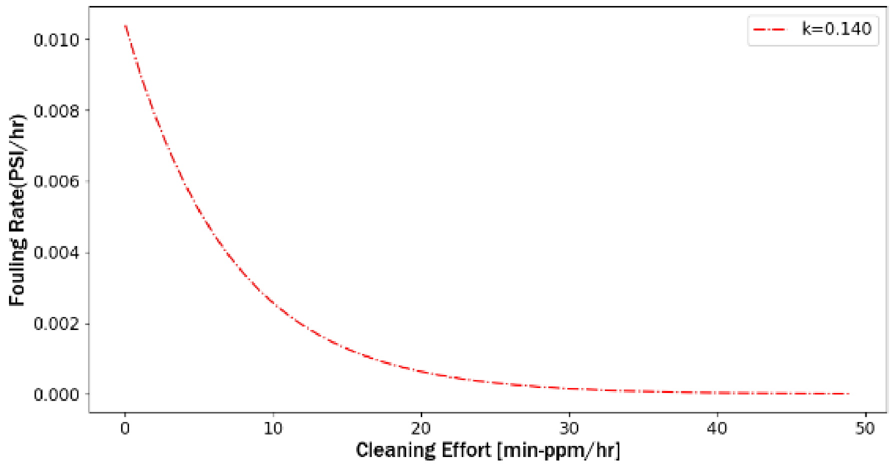

Figure 2. Fouling rate proxy transmembrane pressure as a function of the cleaning effort is an exponential decay curve (a = 0.0104, k =0.140)

The condition-based cleaning works well under assumptions, but operating conditions change continuously and the long-term behavior of the system does not remain constant in a desalination plant. There are unknown parameters in the system, i.e., the flux recovered after a BW (r_b) and the flux recovered after a CEB cycle (r_c). The normalized drop in permeate water production at which BW (d_b) and CEB (d_c) are triggered are the set points, which need to be determined. Therefore, the control system need to continuously evaluate the cleaning effectiveness by measuring r_b and r_c to adapt the set points d_b and d_c so that the cumulative water production is maximized. Simultaneously, the condition-based maintenance also affects the cleaning effectiveness and thus continuous parameter estimation and system learning is required. The cleaning effectiveness is influenced by man parameters, including the time of exposure, the type of chemicals used and the condition at which the membrane was cleaned. Therefore, it is of great importance that the optimal duration of the BW and the CEB cycles and the corresponding chemical dosage used are empirically determined as shown in Figure 2. The algorithm for the **self-learning condition-based cleaning** takes the feed water conditions represented by the time constant T, the membrane degradation models and parameters. The drop in normalized permeate flow rate d_b % at which the BW is triggered, the drop in normalized permeate flow rate d_{ceb} % at which the CEB is triggered, the water production recovery by BW r_b % and the the water production recovery by CEB r_{ceb} % are initialized and the cumulative water produced Qp_{cum} calculated. The BW and CEB are triggered based on d_{ceb} and d_b setpoints and the errors in r_b, r_{ceb} and Qp_{cum} are calculated. d_{ceb} and d_b are varied in small increments of δ_{ceb} and δ_b, respectively, r_{ceb} and r_b are measured and corresponding cumulative water produced Qp_{cum} calculated. d_{ceb}. In addition, d_b are set using r_{ceb} and r_b such that Qp_{cum} is maximum. The algorithm terminate when the criterion, which includes chemical costs, is satisfied.

2.2. Cost Estimation

For the cost estimation of the WASTEC pilot plant, we look at the CAPEX and the OPEX. The model encompasses capital costs (plant construction expenditures), annual operation, and maintenance (O&M) costs. It is based on the cost calculations for the individual cost components as per volume of water produced.

2.2.1. Capital Cost Modeling (CAPEX)

The CAPEX or the capital cost, CC* is the sum of the plant construction expenditures, which are made up of the cost of building the plant, C_p and the initial cost of buy the membranes, C_m. The first part of the cost, C_p is composed of all the costs of the "supporting equipment" (e.g., pipes, pumps, housing, automation and control devices, BW and CEB systems, monitoring equipment), which facilitate the filtration process. Typically, quotes from vendors and historical data are used to determine capital costs for a plant. However, such an approach locks in underlying design assumptions that may affect capital costs and distorts scalability and cost estimates for pilot/ experimental plants for which there is no design experience. The model of calculating the cost should be scalable and useful for pilot projects. Therefore, we calculate the cost of the supporting equipment and the initial membrane costs based on the design area of membranes required, A_{mem} and consequently the number of modules required $nmod$ to supply this membrane area. The area of membranes required to provide the required design plant water capacity and the corresponding number of membrane modules required is basically a function of permeate flux across the membrane. As discussed in the membrane degradation section 2.1.1, permeate flux is time-variant due to fouling, but sometimes for analyzing costs it might be helpful to use the average permeate flux, which is a constant. However, in our case we also want to calculate the cost of BW and CEB, therefore, we have to consider a time-variant permeate flux. As seen in the previous sections, the flux enhancement methods (BW and CEB) restores the permeate flux to some extent, producing a periodic increase and decline in flux through each cleaning cycle. Therefore, the time-dependence of the permeate flux and the cost of flux enhancement are of interest in evaluating membrane costs.

For the permeate flux over time, $J(t)$, the membrane area A_{mem} required to produce a design flow of water can be expressed as:

$$A_{mem} = \frac{Q_f\left(R^* t_0 + t_{ff}\right)}{\vartheta - J_{bf} t_{bf}} \tag{4}$$

in which $Q_f = f(Q_{reqd})$ is the feed flow, R^* is the set-point recovery defined by the continuous-waste flow of concentrate, t_0 is the time between flux enhancement cycles, Q_{reqd} is the plant design capacity, J_{bf} denotes the permeation rate of clean water through the membrane during backflush, t_{bf} is the time required for one backflush, and ϑ represents the integral of flux over operating time such that $\vartheta = \int_0^{t_0} J(t)dt$, t_{ff} is the time required to fastflush.

$$t_0 = \frac{t_{ceb} - n\left(t_{bf} + t_{ff}\right)}{n} \tag{5}$$

Where n is the number of backflushs and fastflushs per cleaning cycle, t_{ceb} is the time required for CEB. The time for one entire operating and flux enhancement cycle is equal to t_{tot} can be expressed as

$$t_{tot} = n\left(t_0 + t_{bf} + t_{ff}\right) + t_{ceb} \tag{6}$$

The trues system water recovery R accounts for water used for BW and CEB and hence is given by Equation 7.

$$R = \left(R^* t_0 + t_{ff}\right)\left[\frac{\vartheta - J_{bf} t_{bf} - \frac{J_{ceb} t_{ceb}}{n}}{\left(\vartheta + J_{ff} t_{ff}\right)\left(t_0 + t_{ff}\right)}\right] \tag{7}$$

Equation 8 is applied to determine the feed low rate, Q_f, required to obtain the design plant capacity Q_{reqd}.

$$Q_f = \frac{Q_{reqd} t_{tot}}{nR\left(t_0 + t_{ff}\right)} \tag{8}$$

Equation 4 to Equation 8 allow the calculation of the required membrane area by dividing the required plant design capacity by the net amount of permeate flux. The net amount of permeate flux is calculated from the difference of the volume of clean water used for BW and CEB a unit area of membrane and the volume of water produced per area of membrane between cleaning cycles while the filtration is running. As a result, the membrane area required for a plant of a given capacity is adjusted to account for the clean product water that is utilized by the BW and CEB.

Lastly, the number of membrane modules, n_{mod} required required to produce a flow at least as large as Q_{reqd} with a membrane area per module, A_{mod}, is calculated by dividing A_{mem} by A_{mod} and rounding to the next highest integer , since fractions of modules have no significance, as given by Equation 9.

$$n_{\mathrm{mod}} = CEIL\left[\frac{A_{mem}}{A_{\mathrm{mod}}} + 0.5\right] \tag{9}$$

The total cost of the membranes C_m can be calculated as expressed by Equation 10 based on the number of membrane modules required, calculated by Equation 9, and the cost of each membrane module, c_{mod}

$$C_m = c_{\mathrm{mod}} n_{\mathrm{mod}} \tag{10}$$

As in the project "WASTEC", new technologies for improving the membranes such as Atomic Layer Deposition (ALD) coating was introduced, some additional membrane cost might incur due to this. In such a case, the cost of the ALD coating per square meter was determined and included in the initial membrane costs.

As described previously, the "Supporting Equipment" typically include all the construction and site work, chemical feed systems, booster pumps, control and instrumentation, storage, concentrate disposal, process piping, yard piping, electric, and pretreatment. If we break down the capital cost CC* into the initial cost of the membranes C_m and the cost of the "supporting equipments" C_p, it is logical that the cost of the "supporting equipments" is assumed to vary according to the number of modules installed in the plant. To be able to express the capital cost of "supporting equipments", C_p in terms of the number of installed modules, n_{mod} , the capacity of the plant must be converted to the number of installed modules required to provide the design flow, Q_{reqd} in m^3/hr. These relationships are given in the cost estimates given in Montgomery, 1990. The equation for the number of modules as a function of capacity, from a least-squares curve fit, is:

$$n_{\mathrm{mod}} = 2.0176 + 0.22135 Q_{reqd} \tag{11}$$

and the cost may be expressed as a logarithmic function. The exponent in equation 12 is under one, indicating that economics of scale should be realized as plant capacity increases.

$$C_p = 150,037.56 n_{\mathrm{mod}}^{0.74678} \tag{12}$$

Finally, the total capital costs CC* can be expressed as in Equation 13

$$CC^* = C_p + C_m \tag{13}$$

To express total capital costs in cost per unit volume of water, the total capital cost, CC*, must first be amortized over the design life of the plant. In this way, total capital cost is converted to annual costs. The amortized capital cost is then divided by the design flow rate, Q_{reqd} to express capital cost per unit volume of water treated,

$$CC = \frac{(C_p + C_m)(A/P)}{Q_{reqd}} \tag{14}$$

where CC is the capital cost per volume water produced and (A/P) is the ratio of annualized cost to the present single cost, calculated as,

$$\frac{A}{P} = \frac{i_c(1+i_c)^{DL}}{(1+i_c)^{DL} - 1} \tag{15}$$

where DL is the time over which the capital costs are amortized, expressed in years, and i_c is the discount rate for capital investments, expressed as a percentage.

2.2.2. Operating Cost

For the operating cost for the membrane filtration plant, we account for all individual cost components, which include the energy, chemicals, membrane replacement, and disposal of concentrate costs.

Membrane replacement cost

Membranes are replaced at fixed intervals, based on manufacturer estimates for membrane life. This cost, which occurs once every membrane life cycle throughout the lifetime of the treatment facility, is transformed into an annual cost, and then to a cost per volume of water produced, by Equation 16.

$$C_{mr} = \frac{c_{\mathrm{mod}} n_{\mathrm{mod}} (A/F)}{Q_{reqd}} \tag{16}$$

where C_{mr} is the membrane replacement cost per volume water produced, c_{mod} is the cost per membrane module, and

$$A/F = \frac{i_f}{(i_f + 1)^{ML-1}} \tag{17}$$

where i_f is the discount rate for membrane replacement, expressed as a percentage, and ML is the membrane life. By annualizing the cost of the membrane modules over one replacement period, the membrane replacement cost is included each year.

If ALD coating is applied to the membrane, the replacement costs are also affected. It was observed that ALD coating extended the membrane life by over 80%

Energy cost

For our feed through system, the energy cost has three components resulting from 1) the quantity of energy required to pump feed water, 2) the energy required to recycle water, and the energy require for backflush water at specified pressures. The sum of these three energy requirements, multiplied by the cost of energy, yields the total energy cost for the membrane unit.

The energy required to pump feed water depends upon the feed flow, Q_f is given by Equation 18. The required Q_f is calculated as given by Equation 8.

$$E_1 = \frac{P_1 Q_f}{\left[\frac{\eta_1}{100}\right]} \tag{18}$$

where E_1 is the energy consumption of the feed pump, P_1 is the feed pressure, and η_1 is the efficiency of the feed pump [%].

In a feed through system, the recycle flow rate can be expressed as in the following relationship:

$$Q_r = Q_t - Q_p - Q_w \tag{19}$$

$$Q_t = A\bar{u} n_{\mathrm{mod}} n_f \tag{20}$$

$$Q_p = \bar{J} a_{\mathrm{mod}} n_{\mathrm{mod}} \tag{21}$$

$$Q_w = \left[1 - \frac{R}{100}\right] Q_f \tag{22}$$

where Q_r is the recycle flow rate, Q_t is the flow rate entering the membrane module, Q_p is the product flow rate, and Q_w is the brine flow rate, A is the cross-sectional area of hollow-fiber or tube, a_{mod} is the membrane area of one module.

Finally, the energy requirement E_2 of the recycle pump may be expressed as in equation 23

$$E_2 = \frac{P_2 Q_r}{\left[\frac{\eta_2}{100}\right]} \tag{23}$$

where η_2 is the efficiency of recycle pump, $P_2 = \frac{\delta P_2}{\delta L} L$ is the absolute value of incremental pressure drop δP_2 over the length of the membrane tube is multiplied by the length L of the membrane element to estimate the total pressure drop across the element.

Similarly, the energy needed for the flux enhancement cycle can be obtained. However, the flux recovery is composed of the fastflush and the backflush. A fastflush is powered by the feed pump for the fastflush flow, Q_{ff}, therefore, its energy requirement can be written as,

$$E_{ff} = \frac{P_1 Q_{ff}}{\left[\frac{\eta_1}{100}\right]} \tag{24}$$

$$Q_{ff} = u_{ff} A n_{\text{mod}} n_f \tag{25}$$

where u_{ff} is the velocity of fastflush water. A separate backflush pump supplies the energy for the backflush component of flux enhancement, but the energy requirement is calculated similarly as:

$$E_{bf} = \frac{P_{bf} Q_{bf}}{\left[\frac{\eta_3}{100}\right]} \tag{26}$$

where P_{bf} is the backflush pressure, specified by user, Q_{bf} is the flow during backflush; and η_3 is the efficiency of the backflush pump. The backflush flow is calculated by:

$$Q_{bf} = J_{bf} A_{\text{mod}} n_{\text{mod}} \tag{27}$$

where J_{bf} is the flux of product water through the membrane during backflushing. The flux enhancement is not operated continuously, therefore, Equations 24 and 26 are corrected in equation 28, where E_3 is scaled by the percent of the total operating time spent fastflushing t_{ff} and backflushing t_{bf}.

$$E_3 = \frac{E_{ff} t_{ff}}{t_{tot}} + \frac{E_{bf} t_{bf}}{t_{tot}} \tag{28}$$

where t_{tot} is the total operating period between cleaningcycles. Finaly, the total cost of energy required by the plant can be obtained as expressed in Equation 29.

$$C_e = \frac{c_{kw}\left(E_1 + E_2 + E_3\right)}{Q_{reqd}} \tag{29}$$

where C_e is the cost of energy per volume of water produced, c_{kw} is the cost per kWh electricity used, and Q_{reqd} is the capacity of plant.

Chemical cost

The chemical costs consists of the cost of treating the feed water with chemicals added to the feed flow at the head of the plant and the cost of chemicals used in CEB. These individual costs are calculated from the quantity of the chemical dosage and the bulk cost of the chemical. The feed flow Q_f calculated from Equation 8 is carried into equation 30 for chemical cost:

$$C_c = \frac{Q_f c c_b}{Q_{reqd}} \tag{30}$$

where C_c is the chemical cost per volume of water produced, c is the coagulant dose, and c_b is the cost of bulk coagulant.

The costs associated with the addition of other chemicals can be included by summing the doses of each chemical to be added and calculating a weighted average of the costs of the chemicals.

Concentrate disposal cost

As in [9], the disposal costs can be incorporated into the cost analysis as multiples of the energy and chemical costs. The wastewater fIow, Q_w calculated in equation 22, is used to calculate the cost of concentrate disposal per volume of filtered water produced,

$$C_d = \frac{\left[c_{kw} \frac{P_1 Q_w}{\eta_1}\right] + \left(Q_w c c_b\right)}{Q_{reqd}} \tag{31}$$

where C_d is the cost of concentrate disposal per volume of filtered water produced.

The time-dependent permeate flux require in the previous equations was modelled empirically as described in the following section.

2.3. Modelling Time-dependent Permeate Flux

Several experiments were conducted on the pilot project for data collection on permeate flux variability. Different membranes with different pore sizes were used and the type of coagulant used in the pretreatment was varied. Several profiles of the permeate flux decline with time under different conditions were obtained. The following typical structure can be seen in every profile. The permeate flux start at an initial level J_o and declines with time to a certain level until a backflush cycle is started. After the back flash, the permeate flux is restored to a certain percentage of J_o and it starts to decline again until another backflush and this structure repeats itself. Because the permeate flux is not restored fully after each and every backflush, a slow but sure irreparable damages happens to the membrane, reflected in the long-term decline.

The profile is characterized by a short and a long-term decay behavior, therefore the data collected were fit to decay model structures using a linear least-squares method and it was found out that the best model which describe the short-term behavior, i.e., the flux decline between two backflush cycles was a second order decay equation linearized to the form expressed in equation 32.

$$\frac{1}{J} = \frac{1}{J_0} + k_2 t \tag{32}$$

where J is the permeate flux, J_o denotes the permeate flux immediately after the most recent backflush, k_2 is a second order rate coefficient of flux decline in hr m^2/Lmin, and t is the time of operation since the last backflush.

The long-term behavior is characterized by a first order decay, which we can express in a linearized form as in equation 33.

$$\ln J_0 = \ln J_i - k_1 t_{lb} \tag{33}$$

where Ji is the initial permeate flux at start of operation, k_1 is a first order rate coefficient of irreversible flux decline in (L/min), and t_{lb} is the start time of the current membrane operating period.

If ALD coating is applied to the membrane, the flux is also affected. It was observed that ALD coating improved the flux recovery ratio by 12.5%.

3. Results and Discussions

3.1. Comparison of Cleaning Strategies

The comparison of no maintenance system and the maintained system based on the two cleaning strategies after accounting for the water used for the BW and CEB illustrated in Figure 3a and Figure 3b shows the superiority of scheduled maintenance. For the results of the cumulative water production illustrated in Figure 3c, the two cases were compared to the ideal situation, which is when a system without flux decline is assumed. The figure show that implementing the two time-based maintenance schedules improves the system compared to the one without any maintenance.

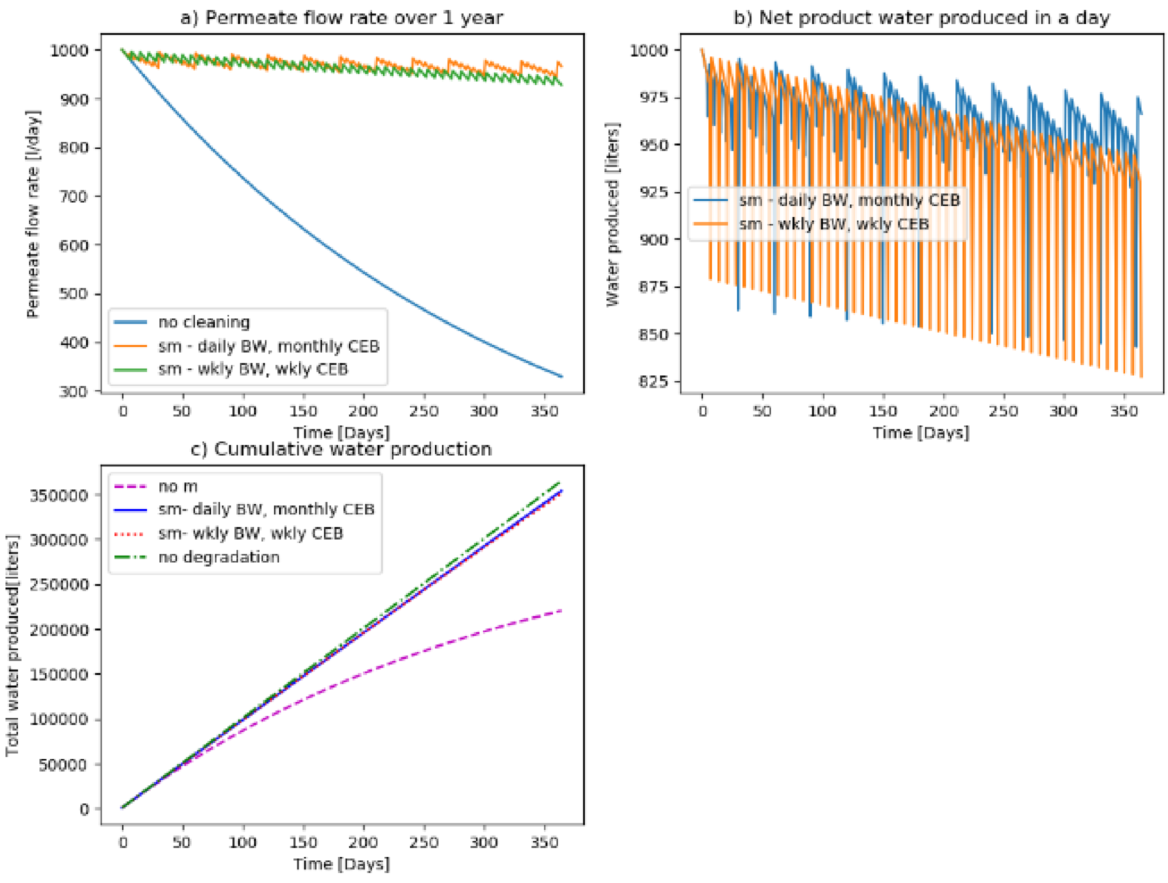

Figure 3. a) Permeate flow rate over one year, b) Net water produced in a day during first year after accounting for water used for BW flushing and CEB, c)Cumulative water produced over a year by an ideal system, system without maintenance and systems with different schedules of BW and CEB

Table 1. Results of the comparison between the two schedule-based maintenance strategies for 1000L per day production system

	Case 1: BW was run daily basis and CEB on a monthly basis			Case 2: BW was run on a weekly basis and CEB on a weekly basis		
r_b [%]	70			70		
r_c [%]	95			95		
T[days]	164	328	656	164	328	656
Total water production for a year						
No degradation [m3]	365	365	365	365	365	365
BW&CEB [m3]	338.258	344.704	348.382	341.400	346.589	350.071
No BW&CEB [m3]	146.734	220.543	280.150	146.734	220.543	280.150
Increase [%]	**130.524**	**56.298**	**24.356**	**132.666**	**57.152**	**24.958**

The two scheduling strategies were also applied to the system with decay constant of 164 and 656 simulating the quality of feed water and the results are compiled in Table 1. A general statement can be given that all time-based maintenance schedules improve the system in terms of the total produced water, but the total costs as will be discussed later, increases significantly with the frequency of CEB. The results also show that the improvement in the amount of water produced is more significant if the feed water quality is bad, as implied by shorter flux decay constants.

Two strategies for condition-based maintenance for a decay constant, T of 328 days for a 1000 liter per day system are discussed in the following. The conditions for applying the BW and CEB are in Case1: BW at 5% drop in normalized permeate flow (NPF) and CEB at 10% drop in NPF, and Case 2: BW at 7.5% drop in NPF and CEB at 15% drop in NPF.

The comparison of no maintenance system and the maintained system based on the two condition-based strategies after accounting for the water used for the BW and CEB illustrated in Figure 4a and Figure 4b shows the superiority of the condition-based maintenance. For the results of the cumulative water production illustrated in Figure 4c, the two cases were compared to the ideal situation, which is when a system without flux decline is assumed. The figure show that implementing the two condition-based maintenance schedules improves the system compared to the one without any maintenance.

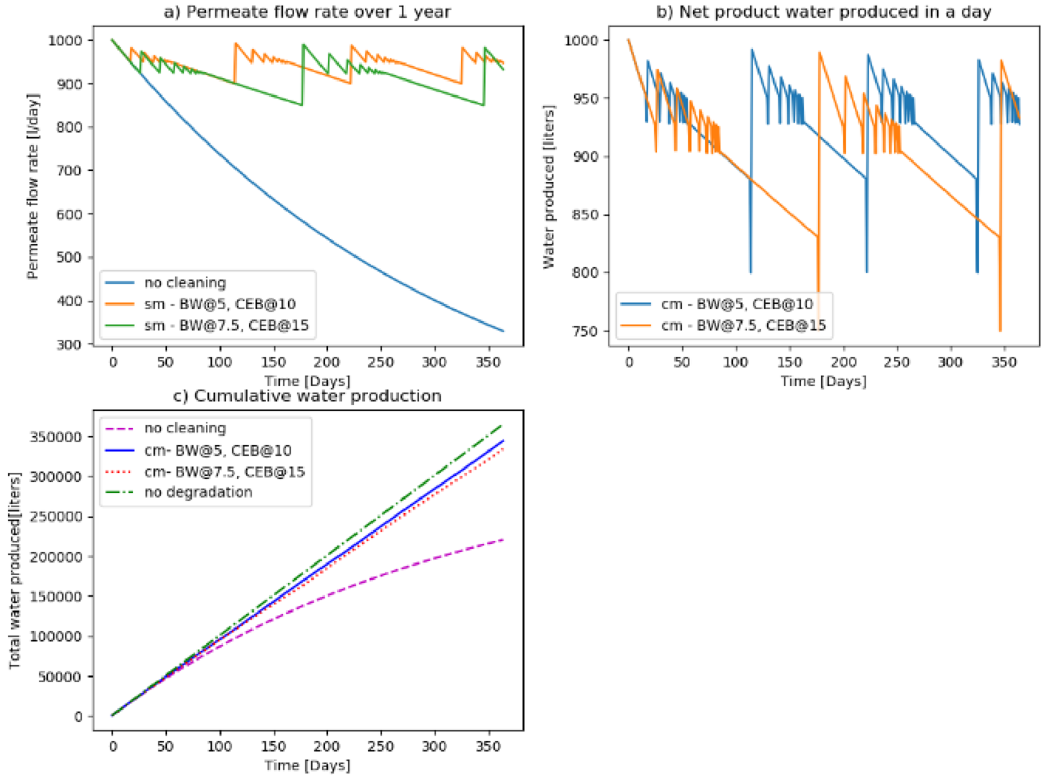

Figure 4. a) Permeate flow rate over one year, b) Net water produced in a day during first year after accounting for water used for BW flushing and CEB, c) Cumulative water produced over a year by an ideal system, system without maintenance and systems with different schedules of BW and CEB

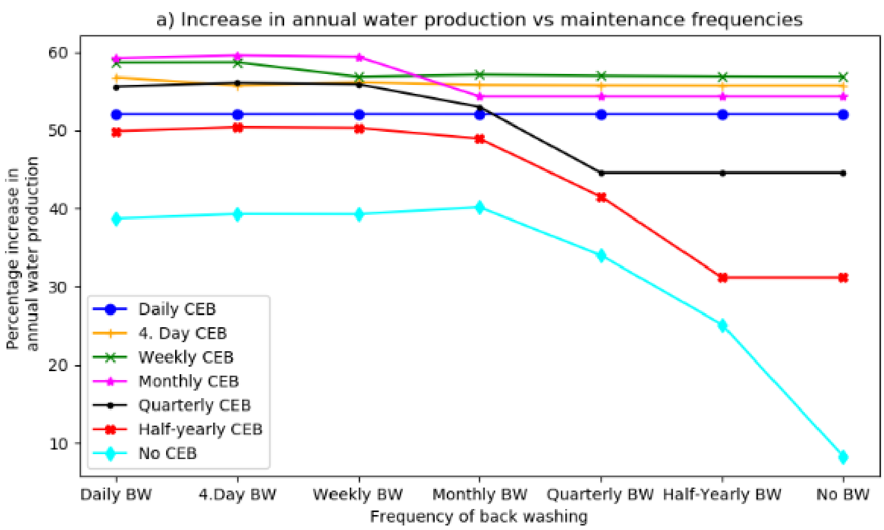

Figure 5. Performance of different time-scheduled maintenance strategies compared with the baseline without maintenance for increase in annual water production

The experiments of using different combination of schedules for BW and CEB produced the results shown in Figure 5 and can be used to find the optimal time schedule for BW and CEB based on net water produced as the optimal criterion and to find the most cost effective plan using the model from section 2.2. The overall percentage increase in water production for all maintenance strategies compared to the baseline can be seen. The differences in the strategies can also be seen and by considering the net annual water production as the optimal criterion, it can be seen that the monthly CEB and semi-weekly BW is the optimal time-based maintenance strategy. It is the strategy that gives over a 59.3% increase in water production over a baseline system. Similar results are also obtained if the frequency of CEB is set to monthly, the BW set to daily or weekly. The results also show that when the number of CEB cycles decreases, BW takes a greater effect on increasing the annual water production and needs to be done more frequently. The case without any CEB reveals that an increase of about 40% can be achieved with only BW. Even though 40% is significantly less than the results of the optimal case, it indicates that BW can be useful in fouling reduction if chemicals are not available.

The two condition-based scheduling strategies were also applied to the system with decay constant of 164 and 656 simulating the quality of feed water and the results are compiled in Table 2. It is shown that Case 1 with BW at 5%, and CEB at 10% NPF drop yield more total produced water for all three decay constants. Unfortunately, it requires more CEB than the second case, which will again affect the system lifetime costs as will be seen in the cost discussion later.

Again as in the time-based maintenance strategy, different combinations of activation time of BW and CEB for the condition-based strategies were compared as shown in Figure 6 based on the net water production and to find the most cost effective plan. The best condition-based maintenance strategy under the given conditions and assumptions outlined triggers a CEB at a 5% drop in normalized permeate flow (NPF) and a BW at a 2.5% drop in NPF. Trigger at these points produces over 10% increase in cumulative annual water production compared to scheduled quarterly maintenance cycle which is commonly used and a 57.7% increase in water production.

We also compared the performance of the best time-based maintenance strategy and the best condition-based maintenance strategies. For the assumptions stated, the results were similar with 59.3 % of time-based (semi-weekly BW and monthly CEB) and 58.2 % of condition-based (BW@2.5% and CEB@5%). This is not surprising because the operating conditions and the long-term behavior of the system is assumed to remain constant. However, in reality this is not the case and the operating parameters change throughout the day and therefore, comparing the time-based scheduling and the condition-based maintenance shows that condition-based maintenance is more desirable than using time schedule-based maintenance, due to the fact that seasonal and stochastic changes in feed water quality and operating conditions are quite unpredictable. Therefore, the timing of maintenance cannot be predicted accurately too, which is why schedule-based maintenance will certainly not work well compared to condition-based maintenance.

Table 2.

	Case 1: BW at 5%, and CEB at 10% in NPF drop			Case 2: BW at 7.5% and CEB at 15% drop in NPF		
d_b [%]	5			7.5		
d_c [%]	10			15		
T[days]	164	328	656	164	328	656
Total water production for a year						
No degradation [m^3]	365	365	365	365	365	365
BW&CEB [m^3]	333.655	338.467	338.422	325.742	327.848	332.927
No BW&CEB [m^3]	146.734	220.543	280.150	146.734	220.543	280.150
Increase [%]	**127.387**	**53.470**	**20.800**	**121.994**	**48.655**	**18.839**

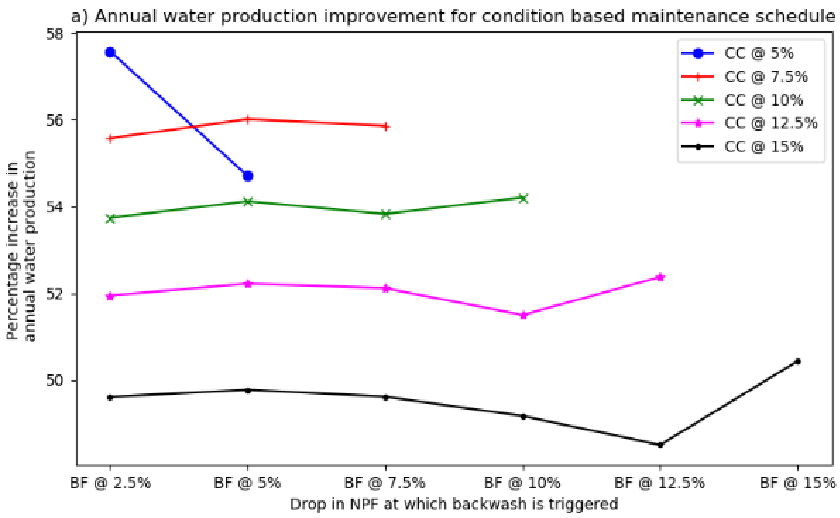

Figure 6. Performance of different condition-based maintenance strategies compared with the baseline without maintenance for annual water production

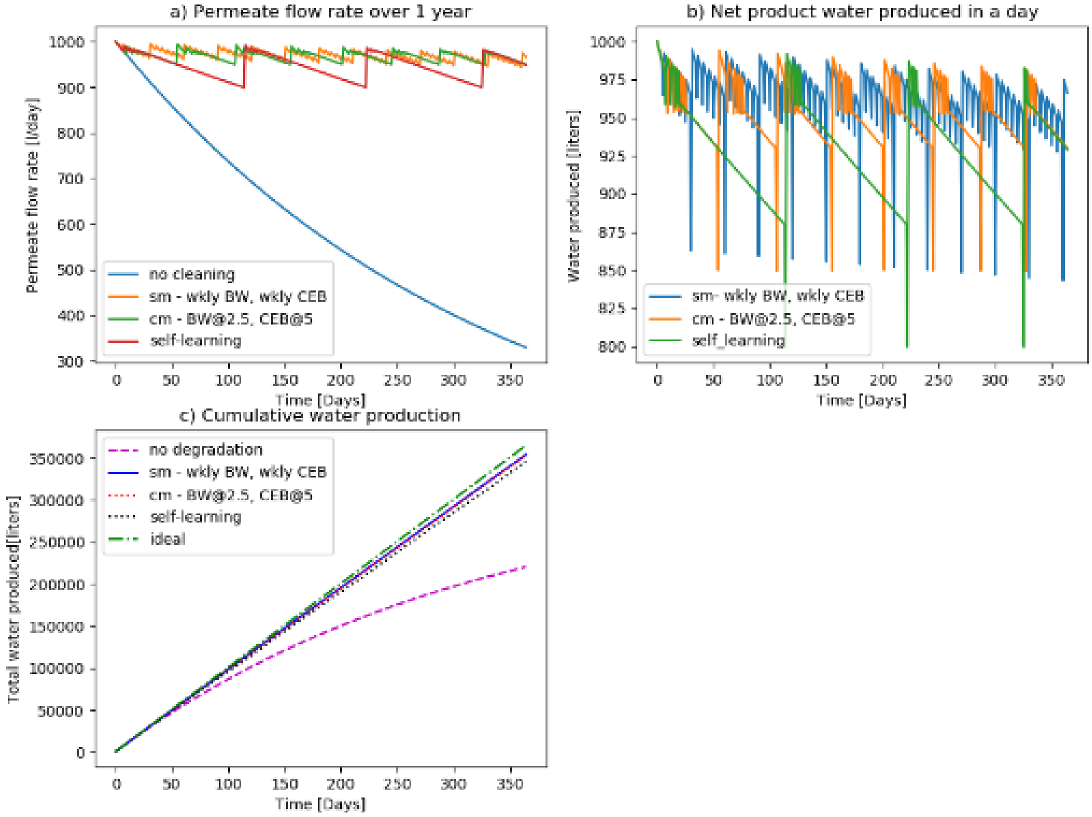

Figure 7. a) Permeate flow rate over one year, b) Net water produced in a day during first year after accounting for water used for BW flushing and CEB, c) Cumulative water produced over a year by an ideal system, system without maintenance and the results of the self-learning condition-based cleaning algorithm

The results of the self-learning condition-based cleaning algorithm are compared to the best scheduled-based cleaning and condition-based cleaning and the results are shown in Figure 7. As expected the CEB is triggered with very low frequency by the self-learning condition-based cleaning algorithm due to chemical costs.

The developed model for time-dependent permeate flux and flux enhancement by backwashing can be used to evaluate the effects of operating conditions, membrane pore size, backwash frequency and duration on capital and operating costs.

Using the cost model, we wanted to know which scheduled cleaning strategy was the most cost effective. Assuming that a chemical cleaning cycle cost $7.50, the tradeoff between the cost of CEB and the rise in water production can be calculated. The criterion used for comparing the strategies is the net value of water produced C in \$/m^3, which is the total amount of water produced multiplied by the average cost of water and then subtracting the total annual cost of chemicals as expressed by equation 34.

$$C = A_{cc}C_{cc} - (W_G - W_L)c_w \tag{34}$$

Where c_w is the life-cycle amortized specific cost of clean permeate water in \$/m^3, A_{cc} is the amount of cleaning chemicals used and c_{cc} is the cost of cleaning chemicals in \$/m^3.

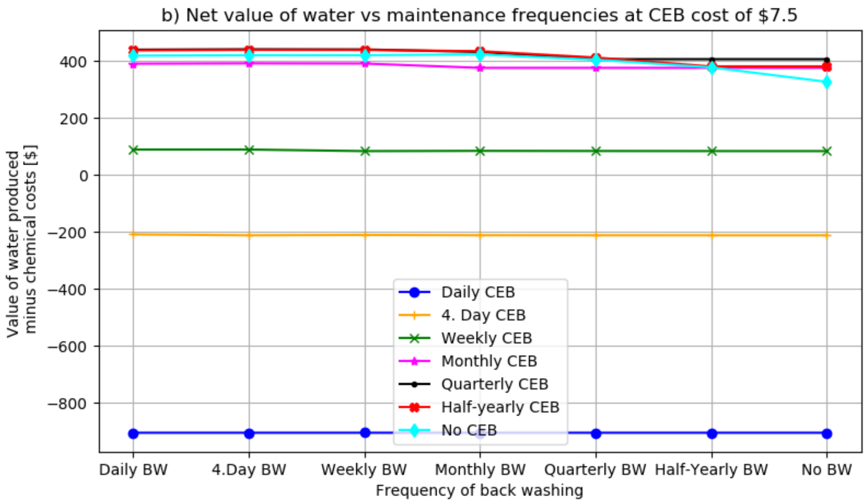

Figure 8. Cost comparison of the of schedule-based cleaning strategies based on the net value of water produced in one year after accounting for cleaning chemicals

The results are shown in Figure 8. The cost effective schedules are the CEB@quarterly cycles in combination with BW@semi-weekly and weekly. They show a high net value of water produced in a year of about \$441 compared to that of the system with no cleaning at \$302.14 an increase of the value of water produced per year by 46%.

The cost results for the condition-based cleaning are shown in Figure 9. It can be seen that the strategies with CEB at 12.5% drop in NPF and BW at a 5% or a 7.5% drop in NPF are the most cost effective. They all result in a net value of water produced of \$444 in a year, which designate an increase of about 47% over the value of water produced by a system with no cleaning at \$302/year.

3.2. Cost Sensitivity Studies for Membrane Desalination

In section 2.2, the cost model has been developed. In this section, the model will be used to study the influences of the different plant and membrane parameters to the resulting costs of the membrane-based filtration. Specifically, the plant parameters in the study included the plant capacity, the recovery rate, feed pressure, permeate flux and operating temperature and in the membrane parameters, the membrane dimensions and number of modules, average cross-flow velocity and expected membrane life were considered.

As a common procedure in sensitivity analysis, we vary one parameter, for example the plant capacity, while keeping all the others constant for each set of simulations. Cost of borrowing money is also included in the model so we assume a 20 year project period for the purpose of capital costs amortization.

The results of some selected parameters will be shown in the following starting from the design plant capacity.

3.2.1. Plant capacity and Percent Recovery

The plant capacity was varied from 2 m^3/hr to 500 m^3/hr and the results are shown in Figure 10a. A sudden drop in the cost of water produced from \$1.842/m^3 at 2m^3/hr to \$0.437/m^3 at 100 m^3/hr can be seen and then a slow decrease until it flattens out. The results are as expected because the cost of supporting materials, pipes and pumps decrease with increasing plant design capacity. Although recovery is an important parameter to consider in optimizing the performance of a membrane, Figure 10b shows that recovery has virtually no effect on the unit cost of water produced by MF. This effect is due largely to the low cost of concentrate disposal assumed in these calculations.

3.2.2. Permeate Flux

As shown in Figure 11, the total costs goes down rapidly as the permeate flux increase, until at about 100l/m^2hr) with the total cost of \$1.639/m^3. The total cost approaches its minimum at the permeate flux of 300l/m^2hr, at \$0.4911m^3. All components of the cost of the filtration system are functions of the permeate flux. It is obvious that as the permeate flow per membrane area declines, the number of membrane modules necessary to produce the required flow increases and since more membrane modules are needed to produce the same design flow when membranes with low permeation rates are used, the capital cost of membranes increases. Furthermore, as the cost of pumps and pipes are functions of the number of modules, the capital cost of pipes and pumps also increases as the permeation rate decreases. This large number of modules needed to yield a given design flow also requires a large number of modules replacement, which is reflected in the increase in membrane replacement cost. With the decreasing, permeate flow per surface area of membrane, the operating costs also increases.

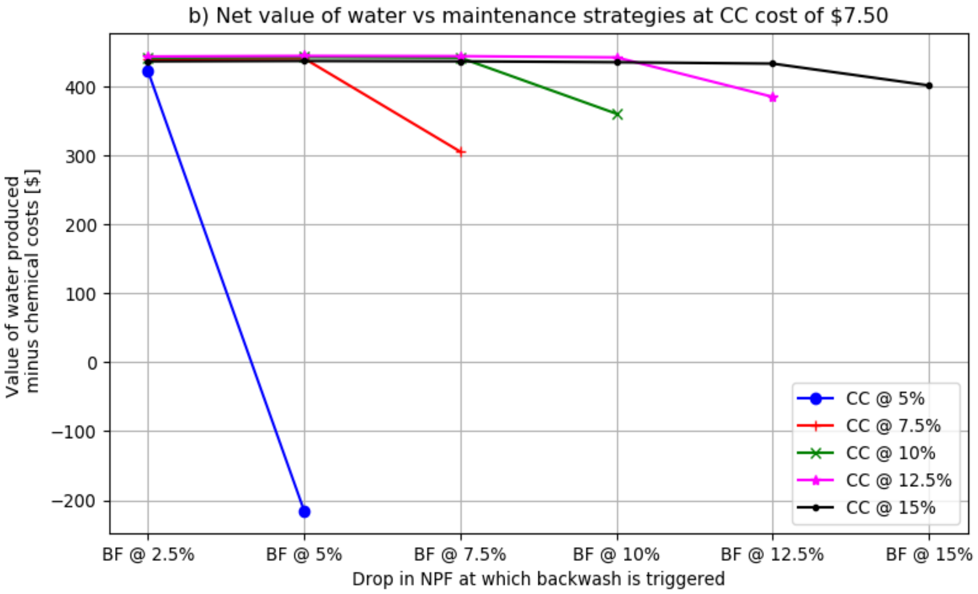

Figure 9. Cost comparison of the of condition-based cleaning strategies based on the net value of water produced in one year after accounting for cleaning chemicals

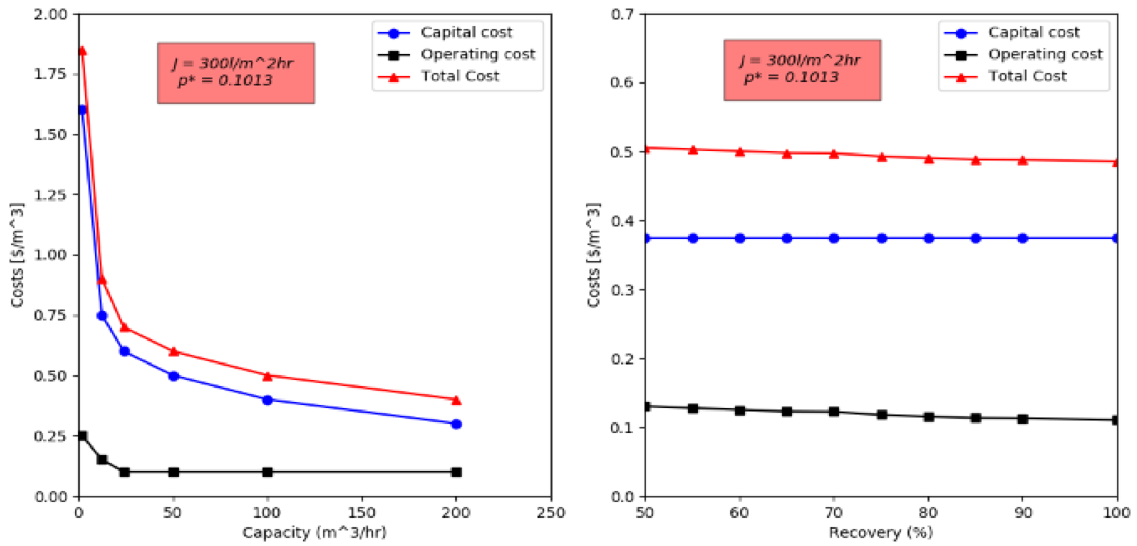

Figure 10. Cost sensitivity to a) plant capacity and b) percent recovery

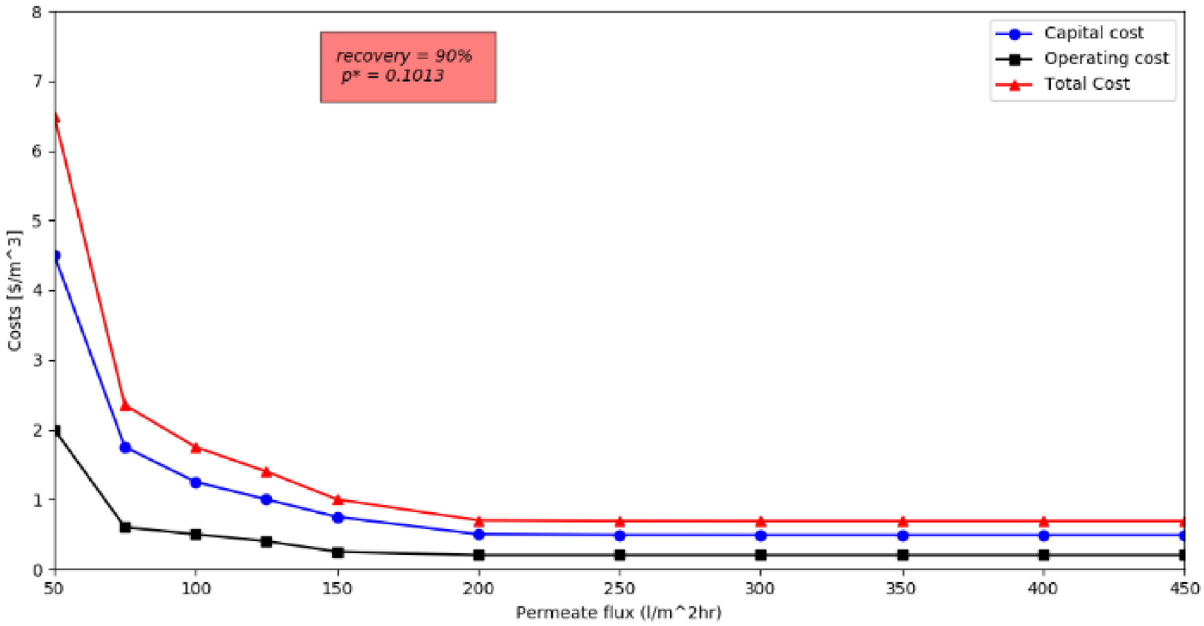

Figure 11. Cost sensitivity to permeate flux for membrane filtration at 90% recovery

3.2.3. Effect of Feed Pressure, Temperature

As expected, the cost to pump the feed water up to the required feed pressure increases linearly as the feed pressure increases. It can be seen in Figure 12 that as the fed pressure goes down, the energy cost of both the recycle loop flow and the flux enhancement cycle decrease. This is explained by that the number of required modules get less as the permeate flux increases with feed pressure and followingly as the number of modules required to yield the design capacity get less, the amount of water that is recycled and the water for the backflush and fastflush decreases. Therefore, the energy consumption of the pumps which drive these cycles also decreases.

With increasing temperature, both the capital cost and operating cost decrease. At 10°C, capital cost is \$0.483/m^3 and declines steadily to \$0.309/m^3 at 30°C, a decrease of 36%. This can be explained by the fact that the permeate flux increases with increasing temperature.

3.2.4. Membrane Life

The membrane lifetime has a very large influence on the total cost of the system, as we can see in Figure 13. Mainly, the effects of membrane life on cost is reflected in the membrane replacement cost component. It can be seen that if we increase the membrane life from 2 to 4 years the total cost of membrane filtration is reduced by 28%, from \$0.834/m^3 to \$ 0.605/m^3. However, if we increase the membrane life further to 8 years, we only get an additional decrease in cost of 19%. As a general effect, one can implicate that any attempt to increase membrane life saves cost, but one should also know that as the membrane lifetime increases, the percent reductions in cost decrease.

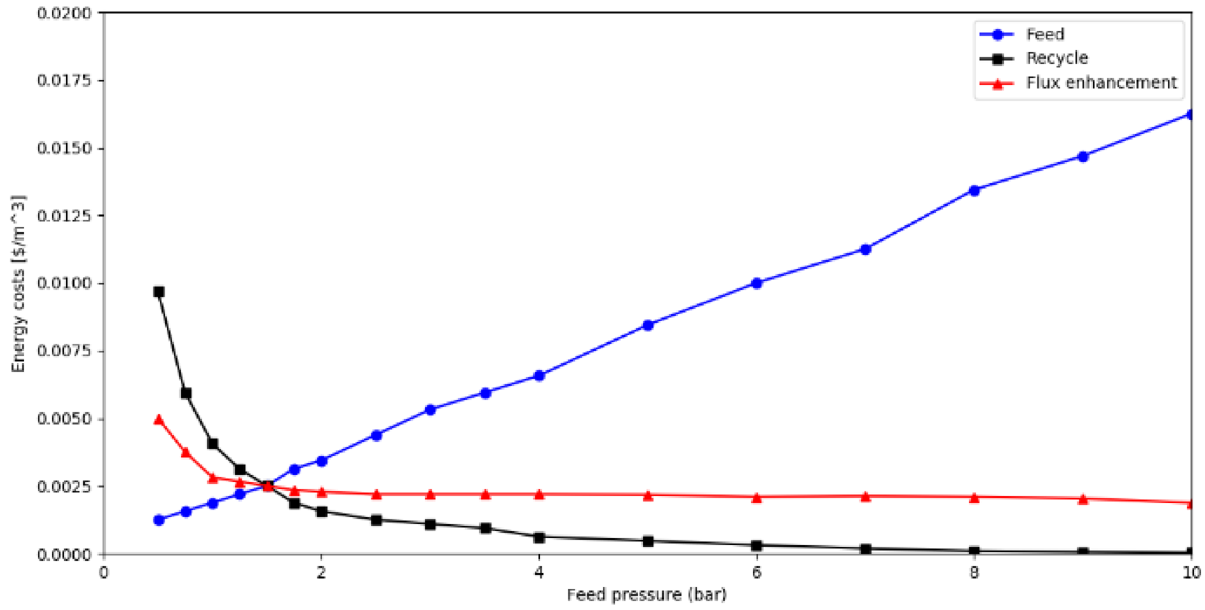

Figure 12. Effect of feed pressure on energy cost

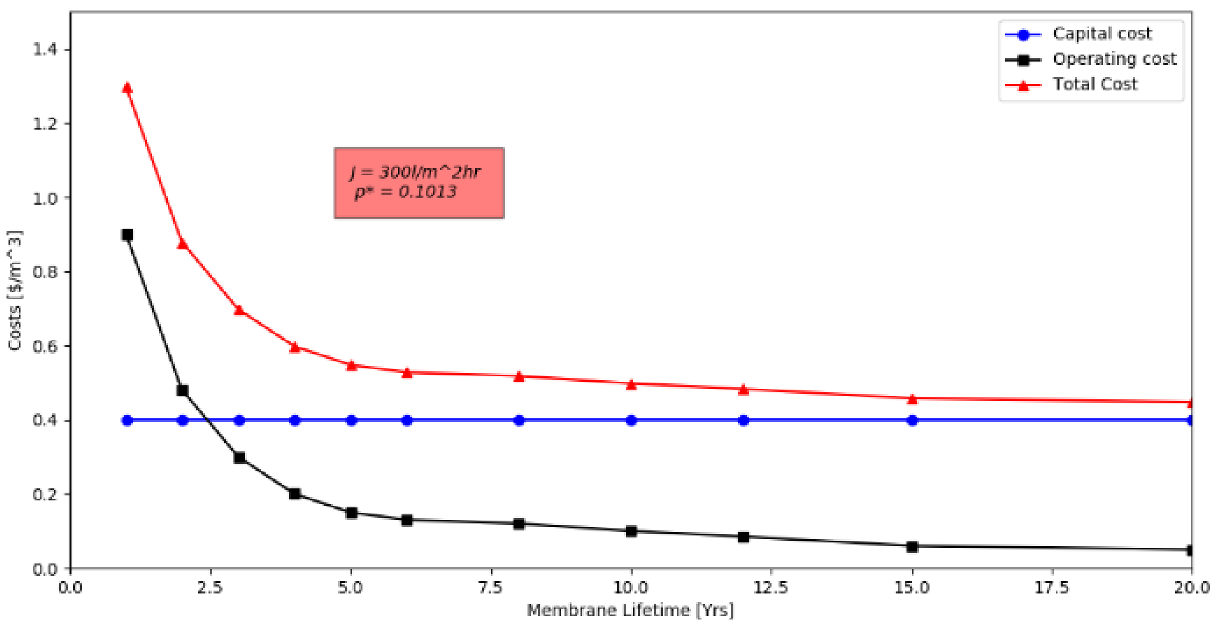

Figure 13. Cost sensitivity to membrane life

3.2.5. Backflush Frequency and Duration

We varied the backflush frequency from 15 seconds to 20 minutes and the results are shown in Figure 14. A minimum backflush frequency at 3 minutes can be identified for both the operating and capital cost. The existence of a minimum backflush frequency can be easily explained by that high frequency in backflushing yields a higher average permeate flux, however, it will also consume more of the finished water produced during each operating period. On the other hand, if the number of backflushes are reduced, less percentage of the total filtered water is used, but the permeate flux declines to a lower point prior to each backflush. Hence, the minimum at the point where these two effects are balanced. The cost with related to the duration of the backflush is shown in Figure 14. The Figure shows that the cost increase with the duration of the backflush. This is as expected, because the net permeate production decreases as more finished water is consumed by the backflush.

Sensitivity of net value of water produced to the price of chemical cleaning

This section analyses the price sensitivity of net value of water produced to the cost of the cleaning chemicals. Figure 15 shows the net value of water for schedule-based cleaning at a CEB cycle cost of $1.50 and $13.00. It can be seen that, the schedule above a monthly period show similar performance at a low cost of a CEB cycle of $2.50. At a CEB cost of $13.00, the half yearly, quarterly and no chemical cleaning schedules produce similar cost results. We can deduct from the results that as the cleaning chemicals price increases, the CEB cleaning schedules with higher frequency get more expensive and therefore have a net value over a one-year period which is lower.

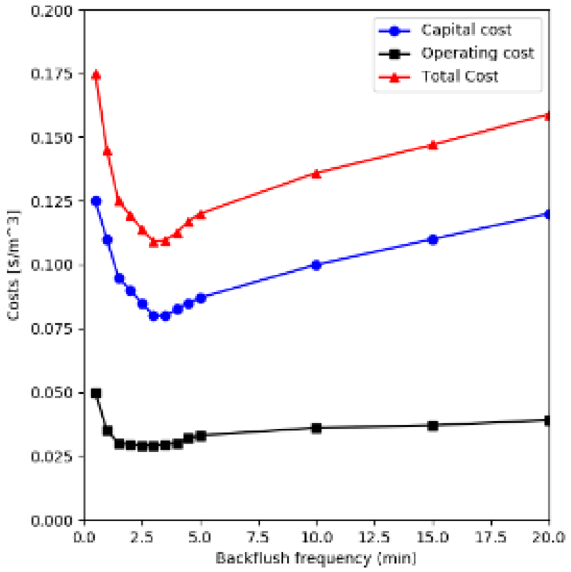

Figure 14. Cost sensitivity to backwash frequency and duration

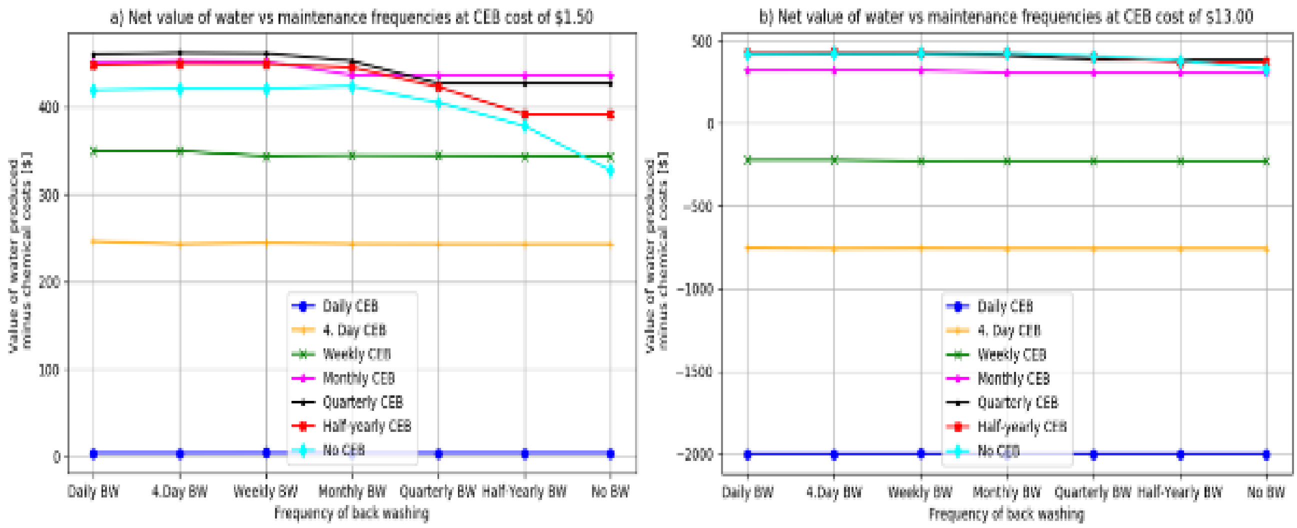

Figure 15. Sensitivity of net value of water produced in one year after accounting for cleaning chemicals of schedule-based cleaning strategies with a CEB cost of a) $1.50 and b) $13.00

3.3. Process Comparison

We considered MF, UF and RO filtration in our studies. Every technology has its advantages and disadvantages with respect to material and functionality i.e., size of material retained [26]. However, costs have to be studied for each. We used the developed model to study the costs of the baseline systems of MF, UF and RO Filtration for plant of different capacities and the results are shown in Figures 16.

As shown in the methodology part, the capital cost is derived from the unit cost of a membrane module and the size of the required membrane area. Microfiltration is associated with higher permeate flux than ultrafiltration, therefore, it requires less membrane area to produce the same flow. Less required membrane area mean less investments cost for membranes for MF, but this is not the full picture, because the membrane costs just take a small portion of the capital costs. On the other hand, as the number of required support components such as pumps and pipes increases with the number of membrane modules and consequently the cost. In the baseline systems, the membrane area for the UF was larger which makes the required number of modules for UF less and therefore the capital cost for UF is less than that of MF as the cost of pipes and pumps for each system takes the larger portion of the capital costs.

The membrane life of the UF system is less than that of the MF, which makes the membrane costs dominants everything and makes the operational costs of the UF higher (Figure 17). On the other hand, just accounting for the day-to-day component costs shows another picture that would make the operating cost of UF significantly less than the others would. Furthermore, figure 17 shows that the operating cost of all the filtration systems is not very sensitive to the different plant design capacities, which can be easily explained in that the increased operational costs from the individual components are covered by the increased plant capacity.

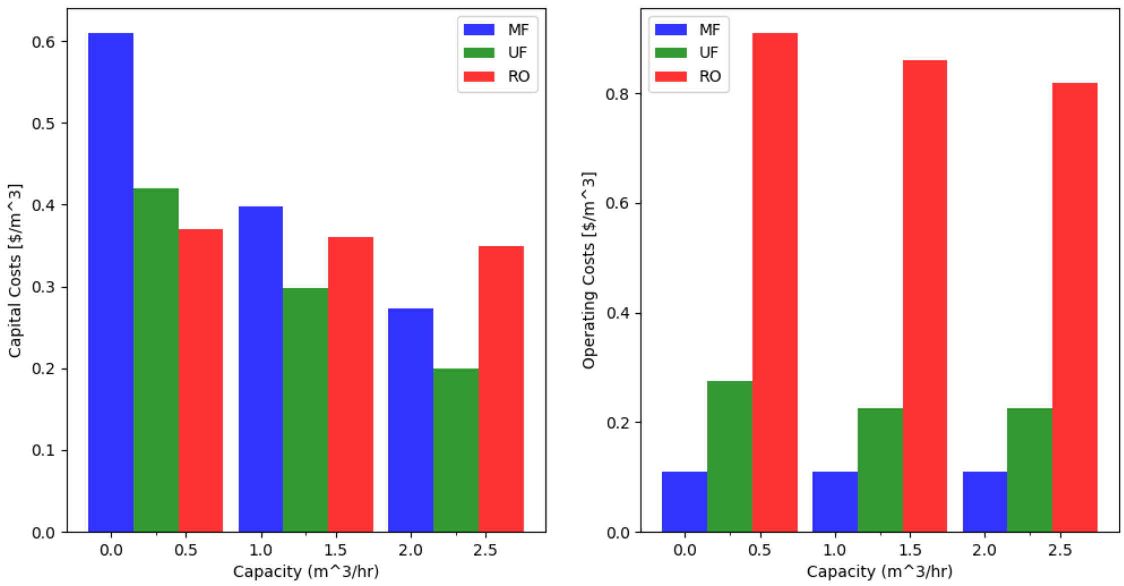

Figure 16. Capital costs of MF, UF and RO Filtration at different capacities

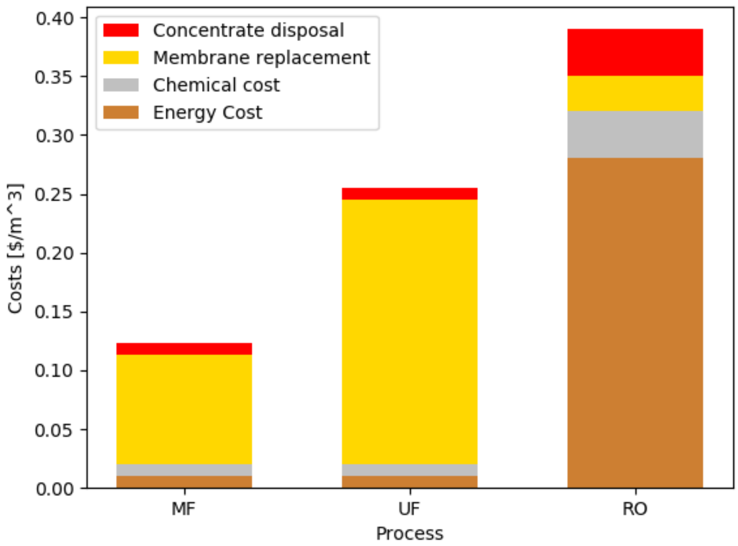

Figure 17. Operating cost components of MF, UF and RO Filtration decrypted

4. Conclusions

The Pilot projects in the field of membrane based water treatment requires empirical and parametric models for cost estimation. Therefore, in this paper a cost model for membrane-based filtration was introduced. The main aim was to develop models, which can be used to economically evaluate the viability of new developments in membrane desalination technology. This include different processes such as MF, UF and strategies such as for chemical and hydrodynamic cleaning and membrane materials and coatings. The model calculates every cost component as per volume of water produced and uses time-dependent permeate flux to determine the number of membrane modules necessary for a design flow. The number of modules is the driving force for all major components of the total cost such as the capital cost of pipes and pumps and membranes and the membrane replacement component of operating cost. Furthermore, the effects of plant and membrane parameters to costs were also studied. The modeling results for microfiltration and ultrafiltration indicates that, due to high permeate productivity of microfiltration is a cost-competitive alternative to ultrafiltration. For a 200m^3/hr design capacity plant, a volume of water is produced by a MF process at a cost of $0.494 and at a cost of $0.486 by an ultrafiltration process. Despite the high energy costs and capital costs of microfiltration to those for UF systems, it is accompanied by lower membrane replacement costs due to greater life expectancy of the membranes. The reverse osmosis process cannot be compared directly, but it required $ 0.49 / m^3 for a plant with 56 m^3/hour design capacity. The values are in line with the costs reported in literature for membrane-based filtration.

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Statement of Competing Interests

The author declare that there are no competing interests.

References

[1] B. Blankert, Short to medium term optimization of ultrafiltration, Chapter 5, July 2007, Ph.D. Thesis, University of Groningen, The Netherlands.

[2] E. Zondervan, B. Roffel, Dynamic optimization of chemical cleaning in dead-end ultrafiltration, J. Membr. Sci. (2007).

[3] Thejani Nilusha, R.; Wang, T.; Wang, H.; Yu, D.; Zhang, J.; Wei, Y. Optimization of In Situ Backwashing Frequency for Stable Operation of Anaerobic Ceramic Membrane Bioreactor. Processes 2020, 8, 545.

[4] Cordier, C.; Eljaddi, T.; Ibouroihim, N.; Stavrakakis, C.; Sauvade, P.; Coelho, F.; Moulin, P. Optimization of Air Backwash Frequency during the Ultrafiltration of Seawater. Membranes 2020, 10, 78.

[5] Chen, J. & Kim, S.L & Ting, Yen-Peng. (2003). Optimization of membrane physical and chemical cleaning by a statistically designed approach. Journal of Membrane Science. 219. 27-45.

[6] Venne, S.; Basu, O.D.; Barbeau, B. Ozone Chemically Enhanced Backwash for Ceramic Membrane Fouling Control in Cyanobacteria-Laden Water. Membranes 2020, 10, 213.

[7] Chang, H.; Liang, H.; Qu, F.; Liu, B.; Yu, H.; Du, X.; Li, G.; Snyder, S.A. Hydraulic backwashing for low-pressure membranes in drinking water treatment: A review. J. Membr. Sci. 2017, 540, 362-380.

[8] Panagopoulos, Argyris; Haralambous, Katherine-Joanne; Loizidou, Maria (November 25, 2019). "Desalination brine disposal methods and treatment technologies - A review". The Science of the Total Environment. 693: 133545.

[9] Malek, A.; Hawlader, M. N. A.; Ho, J. C. Design and economics of RO seawater desalination. Desalination 1996, 105 (3), 245– 261

[10] Ling, C., Wang, Y., Min, C. et al. Economic evaluation of reverse osmosis desalination system coupled with tidal energy. Front. Energy 12, 297-304 (2018).

[11] Chellam, Shankar & Serra, Christophe & Wiesner, Mark. (1998). Estimating Costs for integrated membrane systems. Journal American Water Works Association - J AMER WATER WORK ASSN. 90. 96-104.

[12] Sanz M.A., Bonnélye V., Cremer G., Fujairah reverse osmosis plant: 2 years of operation, Desalination, 203 (2007) 91-99.

[13] T. Fuqua, R, Ortiz, P. Bowen, and R. Creighton. (1991). Membrane treatment and the use of the Floridan Aquifer in South Florida. Proceedings of the American Water Works Association Membrane Process Conference. Orlando, FL, Mar. 10-13, 1991.

[14] United States Bureau of Reclamation (USBR), WTCostII, Report No. 130, 2008.

[15] Noreddine Ghaffour, Thomas M. Missimer, Gary L. Amy. "Technical review and evaluation of the economics of water desalination: Current and future challenges for better water supply sustainability." Water Desalination and Reuse Center KAUST, October 2012.

[16] Jorge Arroyo, Saqib Shirazi. "Cost of Brackish Groundwater Desalination in Texas," September 2012.

[17] Water Reuse Association. "Seawater Desalination Costs," January 2012.

[18] Panagopoulos, Argyris (December 1, 2020). "A comparative study on minimum and actual energy consumption for the treatment of desalination brine". Energy. 212: 118733.

[19] M. Sarai Atab, A.J. Smallbone, A.P. Roskilly, An operational and economic study of a reverse osmosis desalination system for potable water and land irrigation, Desalination, Volume 397, 2016, Pages 174-184, ISSN 0011-9164.

[20] Timothy V. Bartholomew, Nicholas S. Siefert, and Meagan S. Mauter, Cost Optimization of Osmotically Assisted Reverse Osmosis, Environmental Science & Technology 2018 52 (20), 11813-11821.

[21] Randy Ncube and Freddie Inambao, Sea water Reverse Osmosis Desalination: Energy and Economic Analysis. International Journal of Mechanical engineering and Technology 10(12), 2019, pp. 716-731.

[22] Judd, Simon & Carra, Irene. (2021). Low-pressure membrane technology for potable water filtration: true costs. Water Research. 191.

[23] Zhu, P. D. Christofides and Y. Cohen, "Effect of Thermodynamic Restriction on Energy Cost Optimization of RO Membrane Water Desalination," Industrial & Engineering Chemistry Research, vol. 48, no. 13, pp. 6010-6021, 2009.

[24] Sassi, K. And Mujtaba, I.M., 2010, Simulation and optimization of full scale reverseosmosis desalination plant, 20th European Symposium on Computer Aided Process Engineering, 28, 895-900.

[25] Jones, Edward; Qadir, Manzoor; van Vliet, Michelle T. H.; Smakhtin, Vladimir; Kang, Seong-mu (March 20, 2019). "The state of desalination and brine production: A global outlook". Science of the Total Environment. 657: 1343-1356.

[26] Dye, J. Nay, E. Linton, (2017). "The Art of Retrofitting UF/MF Systems: A Comparison of Strategies, Costs, and Results," Proceedings of the AWWA/AMTA Membrane Technology Conference & Expo, Long Beach, CA.