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Abstract

In this research, the effect of operating parameters on the fresh water production cost of hybrid Multi Effect Distillation (MED) and Reverse Osmosis (RO) system is investigated. To achieve this, an earlier comprehensive model developed by the authors for MED+RO system is combined with two full-scale cost models of MED and RO processes collected from the literature. Using the economic model, the variation of the overall fresh water cost with respect to some operating conditions, namely steam temperature and steam flow rate for the MED process and inlet pressure and flow rate for the RO process, is accurately investigated. Then, the hybrid process model is incorporated into a single-objective non-linear optimisation framework to minimise the fresh water cost by finding the optimal values of the above operating conditions. The optimisation results confirm the economic feasibility of the proposed hybrid seawater desalination plant.

***Keywords*:** Seawater desalination; MED+RO hybrid system; Sensitivity analysis; Optimisation; Fresh water cost

1. Introduction

The seawater desalination technology has undergone a significant progress since the 1960s to overcome the issue of fresh water shortage caused by limited available resources (Sadri et al., 2017). However, the research on desalination technologies is still required to enhance the process's economy and efficiency, in a way to reduce the fresh water cost. Basically, the desalination system can be of a thermal type, using heat to evaporate and distillate the seawater, or membrane type, where electrical power is required to pump the seawater through the membranes. Nowadays, Multi Stage Flash (MSF) is the most used thermal desalination process around the world and especially in the Arabian Gulf region. However, the high energy demand and high propensity of fouling due to scale formation are the main concerns for the MSF thermal process (Hawaidi and Mujtaba, 2010; Alsadaie and Mujtaba, 2017). This in turn has added more challenges to investigate the more energy-efficient desalination technology besides the progressive demand for fresh water. In this respect, Multi Effect Distillation (MED) is a well-known and reliable technology to produce fresh water at low operating temperature and pressure (compared to MSF) with very low product salinity at large capacities (Almulla et al., 2005). Recently, MED gained more attention than other thermal processes due to its high effectiveness, straightforward operation and maintenance and feasible economic characteristics. This is particularly true in the case of low temperature MED process (LT-MED), which can achieve high performance together with few fouling/scaling problems, negligible heat losses and reduced need for thermal insulation (Al-Shammiri et al., 1999). The Reverse Osmosis (RO) process has demonstrated to be a practical technology, which is characterised by a significant reduction in energy consumption. This process offers several advantages beyond the conventional thermal

water treatment techniques such as MSF process. For instance, the RO membrane is a flexible process, ease to operate, liable and compact, which can significantly be used as an economically profitable separation process (Al-Obaidi et al., 2018). Specifically, the RO process is characterised by handling different plant configurations and capacities in addition to high salt rejection (99%) and up to 40% of recovery rate. This in turn enables the RO membrane technology to be extensively used to produce fresh water from surface water resources (Goh et al., 2016). In this respect, the combination of MED thermal process with RO membrane technology was confirmed by several researchers to be an energy-efficient desalination process. Helal et al. (2003), (2004a) and (2004b) demonstrated the hybridization of MSF and RO processes as the preferred technology for seawater desalination with improvement the cost of desalted water. Indeed, the low-temperature MED process proved to be more appropriate to be coupled with the RO process, which aids to apply low temperature steam (Mahbub et al., 2009). Mahbub et al. (2009) presented the concept of combined cycle power (CC) plant with MSF, MED and RO (standalone), or with hybrid MSF+RO and MED+RO. This confirmed that the hybrid CC+MED+RO system can reduce the energy consumption by around 17%, compared to CC+MSF+RO system. This also includes the estimation of water production cost that showed the lowest value of 1.09 $/m^3 for the proposed hybridisation of MED+RO system. Therefore, it is fair to expect that the optimisation of the hybrid system of MED+RO processes can lead to a significant reduction of fresh water production cost, especially in an optimised hybrid configuration. In this respect, the hybridization of MED and RO technologies is reported by the latest published research of the authors (Filippini et al., 2018). To investigate the feasibility of several configurations of MED and RO hybrid system, Filippini et al. (2018) evaluated four different ways to connect the two processes, concluding that the best overall configuration is the

one presented in Fig. 1, where the RO process is placed as an upstream process. The choice was made considering the quantity and quality of fresh water produced, the energy consumption and the recovery ratio as the process performance indicators. The proposed design has made it possible to obtain different advantages compared to other configurations, including a better recovery ratio for seawater salinity under 41000 ppm, a fresh water salinity consistently lower than 200 ppm, and a low overall energy consumption.

1.1 Novelty and contribution of this work

Filippini et al. (2018) studied a hybrid MED_TVC (thermal vapor compression) +RO desalination system, confirming the advantages of placing the RO process upstream in a full hybridized configuration. Up to authors' knowledge, an economic assessment and consequent optimisation of this kind of hybrid process where RO is fully hybridized with MED have not yet been explored. In Section 2, the proposed hybrid system is described, while in Section 3, the authors provide the details of the mathematical models of both process including the cost model to evaluate the economic performance of the plant. Then, in Section 4, the attention is on the impact of process parameters on the fresh water cost via a sensitivity analysis. Finally, a non-linear single objective optimisation is carried out in Section 5 to investigate the lowest fresh water cost by manipulating the operating conditions of the hybrid system within specified constraint bounds. Therefore, the current research is a complementary part of the previous presented research.

2. Description of the RO upstream of MED+RO hybrid system

Fig. 1 shows the proposed configuration of the hybrid MED+RO system under investigation. The MED process consists of several effects where the feed saline water is sprayed on a horizontal tubular heat exchanger where steam flows, and partially evaporated. The vapor is partially used to pre-heat the feed, and the rest is directly sent to the next effect. In this respect, the MED process is sometimes conjugated with TVC section in order to reuse part of the steam produced in the last effect as a motive steam. In the present work, the TVC section is deactivated and the thermal process is operated as a low-temperature MED without steam upgrading. This is because it has been demonstrated that the installation of TVC is not convenient from an economic point of view (see Section 4.1). Specifically, the MED process contains 10 effects and a final condenser, where each effect includes an evaporator, a pre-heater for the feed and a flashing box. The motive steam for the first effects is generated from an external utility. The specification and operation conditions of the MED process are given in Table A.1 in the Appendix A.

The multistage RO process comprises three stages connected in series where the retentate of the first stage is reprocessing in the second stage and then the retentate of the second stage is reprocessing in the last stage (retentate reprocessing design). Stages 1, 2, and 3 contain 20, 15, and 8 pressure vessels, respectively, connected in parallel and operating under the same operating conditions. Also, each pressure vessel holds eight identical spiral wound modules connected in series with 37.2 m^2 of an effective area of a commercial thin film composite membrane (type TM820M-400/ SWRO from Toray). The permeates of the three stages are combined to form the product stream with a low salinity. The technical characteristics of the membrane with the lower and upper bounds of operating conditions are given in Table A.1 in the Appendix A.

The permeate stream of the RO process, which has a salinity in the order of some hundreds of ppm depending on the operative conditions, is blended with the pure distillate of the MED process, to obtain the final fresh water stream with a low salinity. The international standards of the World Health Organization (WHO) has regulated the salinity of good quality drinking water around 300 ppm. However, the salinity of the most tap water should be lower than 200 ppm (WHO, 2011).

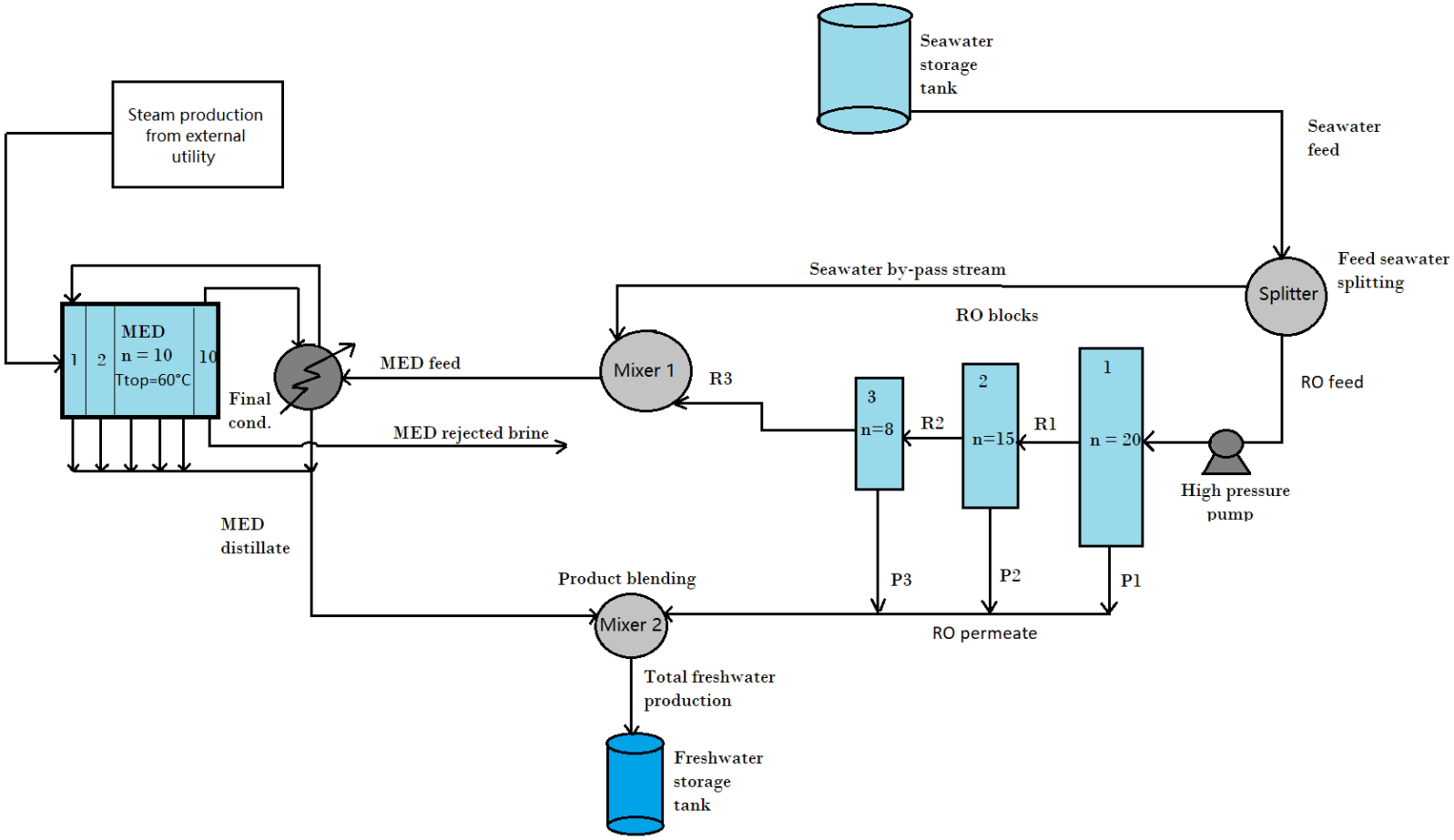

Fig. 1. Schematic diagram of full MED+RO system, with RO process placed upstream (Adapted from Filippini et al. (2018))

3. The hybrid MED+RO process and cost models

More recently, Filippini et al. (2018) have developed a comprehensive model to predict the performance of hybrid MED+RO system used for seawater desalination. Specifically, two separate models were developed for the individual RO and MED processes and were validated against actual experimental data from the literature to confirm their consistency. Those models where combined in order to describe the hybrid plant shown in Fig. 1. For the convenience of the reader, the details of the proposed models of MED, RO, and hybrid system are given in Tables A.2, A.3 and A.4, respectively, in the Appendix A. The cost evaluation of MED process is reported by several researchers such as García-Rodríguez et al. (1999), Sayyaadi et al. (2010), and Druetta et al. (2014). Also, the cost evaluation of RO seawater desalination includes the total annualised cost and operating and maintenance cost with optimisation water production cost have been considered by several researchers such as Malek et al. (1996), Marcovecchio et al. (2005), and Sassi (2012). The following sections describe the economic models for the individual processes besides the illustration of the specific economic parameters and the developed correlation of the total fresh water cost of the hybrid plant.

3.1 Economic model of MED-TVC process

The fresh water cost (FWC) of the MED_TVC process is the total annual production cost divided by the total annual productivity of the thermal process. The total annual cost (TAC) of the seawater desalination MED process comprises the total capital cost (TCC) and annual operational cost (AOC). Basically, the total capital cost includes the equipment, installation, and indirect costs. The operational and maintenance cost comprise several costs such as the steam

cost, chemicals cost, labor, and other related costs. The model developed by Druetta et al. (2014) will be considered to calculate the cost parameters of MED_TVC process. In this respect, Table 1 gives the economic model equations, while Table 2 presents the economic parameters used in this model.

Table 1. Equations of economic model for MED-TVC process (Druetta et al, 2014)

No.	Title	Unit	Equation
1	Fresh water cost	($/m^3)	$FWC_{MED} = \frac{TAC}{M_{fresh,MED}\, THY\ 3600}$
2	Total Annual Cost	($/yr)	$TAC = AOC + CRF\ x\ TCC$
3	Total Capital Cost	($)	$TCC = CAPEX_{dir} + CAPEX_{indir}$
4	Direct CAPEX	($)	$CAPEX_{dir} = CAPEX_{equipment} + CAPEX_{civil\_work}$
5	Indirect CAPEX	($)	$CAPEX_{indir} = 0.25\ CAPEX_{dir}$
6	Equipment cost	($)	$CAPEX_{equipment} = C_{intake} + C_{MED} + C_{cond} + C_{TVC}$
7	Civil work cost	($)	$CAPEX_{civil\_work} = 0.15\ CAPEX_{equipment}$
8	Seawater intake and pre-treatment cost	($)	$C_{intake} = \frac{K_{intake}\ 24\ 3600\ M_{seawater,MED}}{\rho}$
9	MED plant cost	($)	$C_{MED} = K_{MED}\ C_{mat\_MED}\ A_{MED}^{0.64}$
10	Final condenser cost	($)	$C_{cond} = K_{cond}\ C_{mat\_cond}\ A_{cond}^{0.8}$
11	Cost of TVC section	($)	$C_{TVC} = 7912\ M_{ev} \left(\frac{T_{v,n}}{P_{v,n}}\right)^{0.005} P_{steam}^{0.75}$
12	Annual operating cost	($/yr)	$AOC = AOC_{chem} + AOC_{lab} + AOC_{pow} + AOC_{man} + AOC_{steam}$
13	Cost of chemical treatment	($/yr)	$AOC_{chem} = \frac{C_{chem}\, THY\ 3600\ M_{seawater,MED}}{\rho}$
14	Cost of human labor	($/yr)	$AOC_{lab} = \frac{C_{lab} THY\ 3600\ M_{fresh,MED}}{\rho}$
15	Cost of power for pumps	($/yr)	$AOC_{pow} = \frac{C_{pow} THY\ 100\ M_{fresh,MED}}{\rho\ \eta} f(\Delta P)$
16	Cost of manutention	($/yr)	$AOC_{man} = 0.002\ TCC$
17	Cost of external steam	($/yr)	$AOC_{steam} = \frac{C_{steam} THY\ (T_s - 40)\ M_{steam}}{80} + 0.005\ TCC$
18	Capital recovery factor	(1/yr)	$\boldsymbol{CRF = \frac{Ir(1+Ir)^{life}}{(1+Ir)^{life}-1}}$

Table 2. Parameters used in the economic model of MED-TVC

Parameter	Description	Value	Unit	Parameter	Description	Value	Unit
THY	Total hour per year	8760	(hr/yr)	C_{steam}	External steam	0.004	($/kg)
K_{intake}	Seawater intake	50	($ day/m^3)	C_{mat_MED}	Material of MED	3644	($/$m^2$)
K_{MED}	Coeff. for MED	1.4	(-)	C_{mat_cond}	Material of condenser	500	($/$m^2$)
K_{cond}	Coeff. for condenser	2.8	(-)	Ir	Interest rate	0.07	(-)
C_{chem}	Chemical treatment	0.024	($ / m^3)	life	Life of the plant	25	(year)
C_{lab}	Labour	0.05	($ / m^3)	f(ΔP)	Pressure losses	3571	(-)
$\mathbf{C_{pow}}$	Power	0.09	($/**kWh**)	**η**	Efficiency of power generation	0.75	(-)

3.2 Economic model of RO process

The fresh water cost (FWC) of the RO process is the total annual production cost divided by the total annual productivity of the membrane process. The total annual cost (TAC) is the sum of total capital cost (TCC) and the annual operating cost (AOC). The capital cost comprises equipment, installation, and indirect costs. However, the operational and maintenance cost includes several costs such as high-pressure pump cost, chemicals cost, labour, and other related costs. This study is basically based on the work of Malek et al. (1996), Marcovecchio et al. (2005), Koroneos et al. (2007), Al-Obaidani et al. (2008), Lu et al. (2012), and El-Emam and Dincer (2014). The final cost model equations of RO process are given in Table 3. Also, Table 4 shows the economic parameters of this model.

Table 3. Equations of economic model for RO process

No.	Title	Unit	Equation/ References and notes
19	Total annual cost	($/yr)	$TAC = TCC + AOC$ (Koroneos et al., 2007)
20	Total capital cost	($/yr)	$TCC = \left[\left(C_{wip} + C_{Pump} + C_{me}\right) SD\ CC\right]$
21	Total annual operating cost	($/yr)	$AOC = OC_{Pu} + OC_{sc} + OC_{ch} + OC_{me} + OC_{lab} + OC_{maint} + OC_{bd}$
22	Water intake and pre-treatment cost	($)	$C_{wip} = 996\ (86400\ Q_{f(plant)})^{0.8}$ (Malek et al., 1996)
23	Capital cost of high-pressure pump	($)	$C_{Pump} = \left[52\ (3600\ Q_{f(plant)}\ \left(P_{f(plant)}\ 0.101325\right))^{0.96}\right]$ (Lu et al., 2012)
24	Membrane module and pressure vessel capital cost	($)	$C_{me} = N_s\ N_{PV}(C_{ele}\ N_{ele} + C_{PV})$ (Calculated based on the current membrane and pressure vessel prices)
25	Pumping operating cost	($/yr)	$OC_{pu} = 365x24\ \left[\left(\frac{\left(3600\ \left(P_{f(plant)}\ 0.101325\right)\right)\ Q_{f(plant)}\ )}{3.6\ \varepsilon_{pump}\ \varepsilon_{motor}}\right)\right] E_c\ L_f$ (Lu et al., 2012)
26	Annual operating spares cost	($/yr)	$OC_{sc} = 3600x24x365\ C_{cf}\ Q_{p(plant)}\ L_f$
27	Effluents disposal cost	($/yr)	$OC_{bd} = 3600x24x365\ C_{bd}\ Q_{p(plant)}\ L_f$
28	Annual chemical treatment cost	($/yr)	$OC_{ch} = 3600x24x365\ C_{ct}\ Q_{f(plant)}\ L_f$
			Marcovecchio *et al.* (2005), El-Emam and Dincer (2014) and Al-Obaidani *et al.* (2008).
29	Annual membrane replacement cost	($/yr)	$OC_{me} = 0.2\ C_{me}$
30	Annual labour cost	($/yr)	$OC_{lab} = C_{lab}\ 3600x24x365\ Q_{p(plant)}$ (Koroneos et al., 2007)
31	Fresh water cost	($/m³)	$FWC_{RO} = \frac{\left(\frac{TCC}{CCRF}\right)+AOC}{3600x\ 24x\ 365\ Q_{p(plant)}}$
32	Capital cost recovery factor	(yr)	$CCRF = \left[\frac{(i+1)^n-1}{i\ (i+1)^n}\right]$
			Koroneos et al., 2007)
33	Annual maintenance costs	($/yr)	$OC_{main} = 0.02\ PUC\ 3600x24x365\ Q_{p(plant)}$

34	Specific energy consumption	(kWh/m³)	$SEC = \frac{\left(P_{f(plant)}\, x101325\right)\, Q_{f(plant)}}{Q_{p(Total)}\, \varepsilon_{pump}} / 36x10^5$

Table 4. Parameters used in the economic model of RO process

Parameter	Description	Unit	Value	Reference
SD	Site development and indirect costs	(-)	1.411	Malek *et al.* (1996)
CC	Capital charge rate per annum	(-)	0.08	Malek *et al.* (1996)
C_{ele}	Membrane element cost	($)	1000	Email contact with the supplier (Toray membrane)
C_{PV}	Pressure vessel cost	($)	100	
N_s	Stages number	(-)	3	The proposed RO process design presented in Fig. 1
N_{PV}	Pressure vessel number	(-)	43	
N_{ele}	Membrane elements number	(-)	344	
E_c	Electricity unit cost	($/kWh)	0.09	Marcovecchio et al. (2005), Lu et al. (2006) and Valladares Linares et al. (2016).
L_f	Plant load factor per annum	(-)	0.85	
C_{cf}	Cost of cartridge filters replacement (the replacement rate)	($/m³)	0.033	
C_{ct}	Cost of chemical treatment	($/m³)	0.018	
C_{bd}	Cost of effluents disposal	($/m³)	0.0015	
C_{lab}	Labour cost	($/m³)	0.02	Koroneos et al. (2007)
i	Discount rate	(%)	8	
n	The plant life	(yr)	25	Marcovecchio et al. (2005)
ε_{pump}	Pump efficiency	(%)	85	
$\boldsymbol{\varepsilon_{motor}}$	Motor efficiency	(%)	98	

In line with the above economic models, the fresh water cost of the hybrid MED+RO system presented in Fig. 1 is calculated as

$$FWC_{Hybrid} = \frac{\left(FWC_{MED}\ M_{fresh,MED}\right)+\left(FWC_{RO}\ Q_{p(plant)}\right)}{Q_{Hybrid}} \quad (1)$$

$$Q_{Hybrid} = Q_{p(plant)} + M_{fresh,MED}$$

(2)

Q_{Hybrid} (m^3/s) denotes the total fresh water production of the hybrid MED+RO, evaluated as the sum of the distillate from the thermal process $M_{fresh,MED}$ (m^3/s) and the total permeate from the RO process $Q_{p(plant)}$ (m^3/s) as presented in Eq. (2).

4. Fresh water cost variation with respect to some operating conditions

Firstly, the economic advantages of only running the MED process without the TVC section are presented in this section. Secondly, a sensitivity analysis of the hybrid MED+RO system (RO upstream design, Fig. 1) is performed to evaluate the fresh water cost via simulation and using the cost models presented in the previous section. Specifically, the hybrid process performance in terms of the final cost of produced fresh water will be tested against the variation of operating pressure and feed flow rate of the RO process, and steam flow rate and temperature of the MED process. For the sensitivity analysis, seawater feed concentration and temperature are fixed at 39000 ppm and 25 °C, respectively, as well as the cost of electricity and cost of steam, at 0.09 $/kWh and 0.0042 $/kg, respectively.

4.1 Economic feasibility of TVC section

Undoubtedly, the main advantages of instilling TVC section together with the MED process is to produce a higher quantity of distillate using less stem from an external utility. This is possible because part of the distillate from the last effect is entrained by the TVC section and "upgraded" to being re-used as a motive steam (Dessouky et al, 2002). As a result, the performance ratio, defined as the quantity of distillate produced divided by the quantity of external steam provided

to the MED process, significantly increases. The model equations of the TVC section are reported in Table A.5 in the Appendix A.

More importantly, the capital cost of the thermal compressor (Eq. (11) in Table 1), can be relevant. Also, the cost of external steam (Eq. (17) in Table 1), can be higher because of its increased temperature even if its flow rate is reduced. Table 5 presents a comparison between some economical parameters of MED_TVC and MED standalone. As expected, the performance ratio drops by 44% when TCV is not installed. This is attributed to a significant increase of around 36% in the required external steam to generate the same amount of distillate. Moreover, it is not complicated to notice the increase of TCC by around 17% as a result to considering the TVC section installation cost. This in turn has increased the AOC by around 55%. Basically, running the TVC requires a high-pressure steam, which is at 1500 kPa in this simulation, that interprets the increase of AOC. Therefore, this simulation shows that the steam at high temperature of around 200 °C results in a substantial increase in the total fresh water cost that evaluated by Eq. (17) of Table 1. Therefore, the TVC section will be disabled in this study to obtain the minimum cost of fresh water based on the reasons described above.

Table 5. Economical comparison between MED with TVC and MED without TVC section.
($T_{s(MED)}$= 70 °C, $M_{s(MED)}$= 8 kg/s, $Q_{f(RO)}$= 0.058 m^3/s, $P_{(RO)}$ = 50 atm)

Calculated Parameter	MED_TVC	MED
Fresh water cost from MED ($/$m^3$)	1.02	0.78
TCC (M$)	7.29	6.22
AOC (M$/yr)	1.89	1.22
PR	12.17	8.45
External steam flow rate (kg/s)	5.87	8

4.2 Sensitivity analysis: Impact of operating pressure of the RO process

Fig. 2 shows the impact of operating pressure variation of the RO process on the fresh water cost and recovery ratio of the hybrid MED+RO system. It is obvious that increasing the operating pressure from 40 atm to 80 atm (within the permissible manufacturers' limits) has a positive and significant impact on the fresh water cost. Statistically, the variation of operating pressure alone can reduce the fresh water cost of around 15.7%. This can be attributed to the increase of fresh water permeation through the membranes as a response to increasing the applied pressure, which increases the product flow rate of about 12%. This is already evidenced in the increase of total recovery ratio as a response to increasing the operating pressure (Fig. 2). Specifically, any significant increase of product flow rate would serve the reduction of fresh water cost based on Eq. (31) presented in Table 3. Another interesting conclusion that can be drawn from Fig. 2 is that the fresh water cost decreases fast up to an inlet RO pressure of 70 atm, then the reduction is less significant. Based on Eq. (31) in Table 3, it can be noted that the fresh water cost is mainly dependent on both the total capital cost, total operating cost, and total production flow rate. It seems that the progress of product flow rate is slightly lower at high operating pressures (above 71 atm) compared to its behaviour at the range of 40 to 70 atm. Also, it can be said that running the process at high pressures requires higher energy consumption to operate the pumps, which in turn elevates both the total capital cost and operating cost.

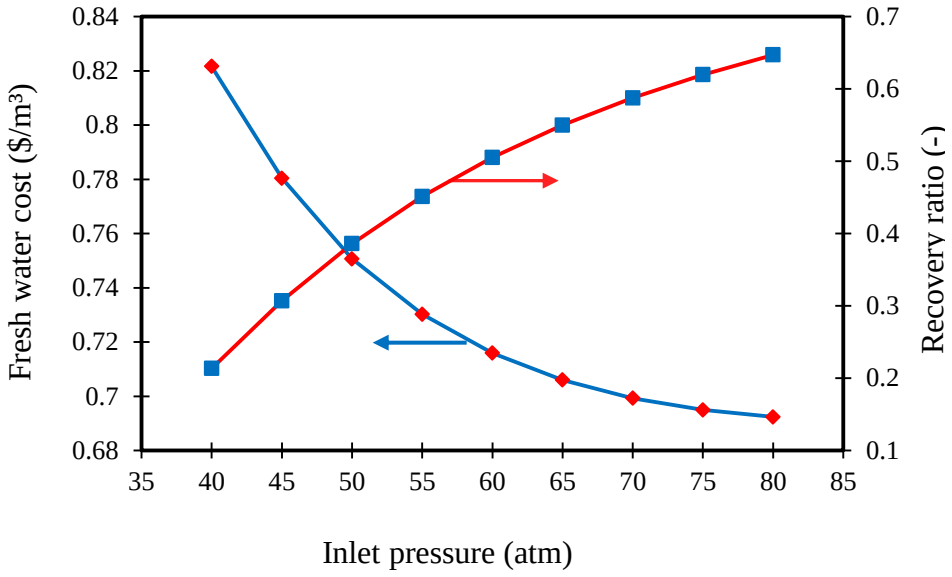

Fig. 2. Fresh water cost and recovery rate against inlet pressure of the RO process

($T_{s(MED)}$= 70 °C, $M_{s(MED)}$= 8 kg/s, $Q_{f(RO)}$= 0.058 m^3/s)

4.3 Impact of feed flow rate of the RO process

The impact of feed flow rate variation of the RO process on the fresh water cost and total permeate flow rate of the hybrid MED+RO system is reported in Fig. 3. In this respect, the variation of feed flow rate from 0.04 to 0.11 m³/s at fixed other operating parameters causes an intensive reduction of fresh water cost beyond 0.076 m³/s, which is subsequent with a noticeable exponentially increase after 0.076 m³/s of feed flow rate. It is fair to say that increasing the feed flow rate of the RO process would increase the bulk velocity inside all the modules operating at retentate reprocessing design. This in turn aids to reduce the concentration polarisation inside each module that lifts the mass transfer coefficient and albeit slightly improves the water flux through the membranes. However, the incremental increase of permeate flow rate occurred as a result to increasing the inlet feed flow rate would not be comparable with the progressive increase of inlet feed flow rate, which in turn causes a continuous reduction of the total water

recovery. Consequently, it is rational to expect a slight increase of total permeate flow rate of the RO process due to increasing the operating flow rate, which in turn reduces the fresh water cost beyond 0.076 m³/s (Fig. 3). However, any further increase of feed flow rate over 0.076 m³/s would causes a slight reduction of total permeate flow rate besides the progressive increase in total capital cost and operating cost due to increasing feed flow rate (Fig. 4). The simulation results at feed flow rate above 0.076 m³/s represent a significant increase of the water intake and pre-treatment cost, the capital cost of high-pressure pump, which are function of feed flow rate. This in turn causes a remarkable increase of fresh water cost (Fig. 3) albeit a noticeable an optimum feed flow rate, which ensures the optimum fresh water cost. More importantly, increasing inlet feed flow rate of the RO process would increase the total retentate flow rate that combine the seawater stream to form the feed stream of the MED process.

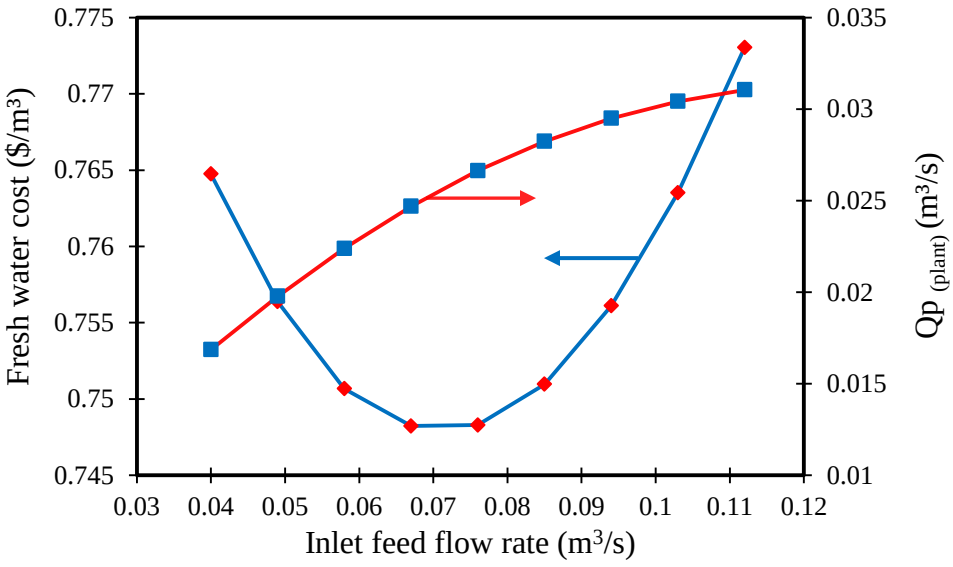

Fig. 3. Fresh water cost and total permeate flow rate against inlet feed flow rate of the RO process ($T_{s(MED)}$= 70 °C, $M_{s(MED)}$= 8 kg/s, $P_{f\,(RO)}$ = 50 atm)

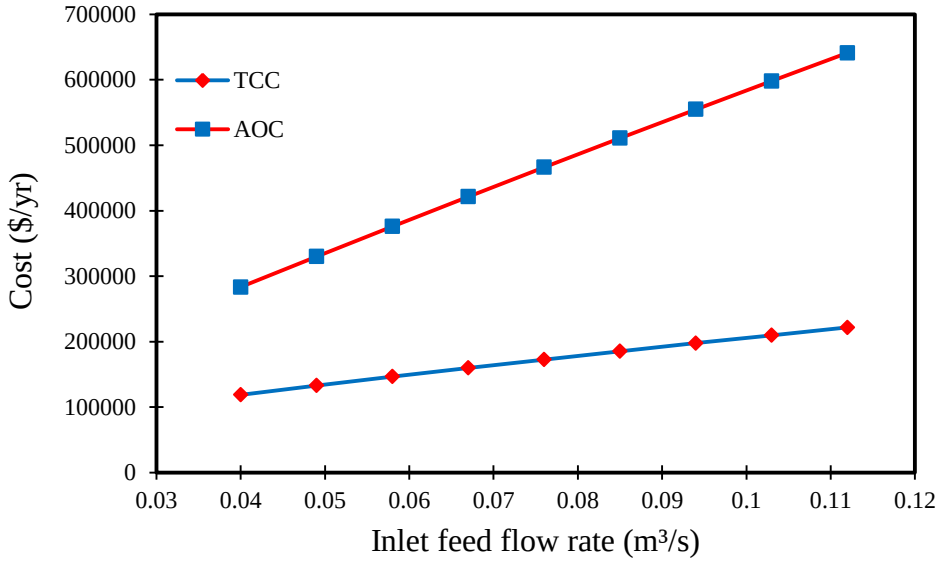

Fig. 4. Total capital cost and annual operating cost against inlet feed flow rate of the RO process

($T_{s(MED)}$= 70 °C, $M_{s(MED)}$= 8 kg/s, $P_{f(RO)}$= 50 atm)

To summarise the above simulation results, it can be said that the plant total water recovery rate and specifically the clean water produced from the total feed flow rate can be considered as the key parameter describing the hybrid process performance besides its significant effect on the fresh water production cost as a relevant parameter. Basically, it can be affirmed that any significant improvement of the total plant production flow rate will improve the total fresh water production cost. This in turn elucidates the high contribution of production rate that influences the economic viability of the hybrid process. However, the capital cost of high-pressure pump in the RO process is considered as the key component of total capital cost that significantly affect the price of treatment. Note, the fresh water cost is mainly related to the total capital cost and annual operating cost of the RO process, which are readily related to the operating flow rate as given in Eq. (31) in Table 3.

4.4 Impact of steam temperature of the MED process

The impact of steam temperature of the MED process on the fresh water cost of the hybrid MED+RO system is given in Fig. 5. Running the MED process at low temperatures with steam below 70 °C would increase the performance ratio of the process. This in turn would improve the economy by reducing the operating cost related to the external steam (less steam is required to produce a certain amount of fresh water). Operating with low temperature steam means also a lower cost for the utility, in accordance to Eq. (17) in Table 1. On the other hand, if the MED process is forced to operate in a smaller temperature window, the TCC associated with the plant construction increases. More specifically, operating the MED process at low steam temperatures means lower temperature difference available for heat exchange (Dessouky et al, 2001). This leads to much higher area of exchange required and obviously more expenses related to the material and construction costs. In this regard, it is expected to find an optimal value of steam temperature that collaborates a minimum overall fresh water cost. Basically, this would offer the best compromise between performance and capital expenses. Fig. 5 demonstrates the optimal value of steam temperature to be around 70 °C.

In this respect, Fig. 6 highlights the reduction of TCC for higher steam temperatures, which is counter-balanced by an increase of AOC.

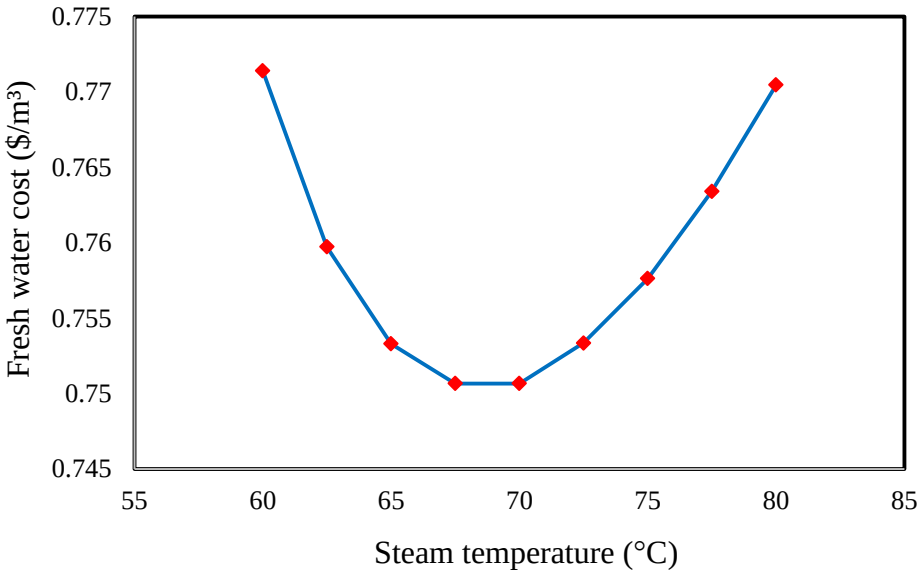

Fig. 5. Fresh water cost against steam temperature of the MED process

($M_{s(MED)}$= 8 kg/s, $Q_{f(RO)}$= 0.058 m³/s, $P_{f(RO)}$= 50 atm)

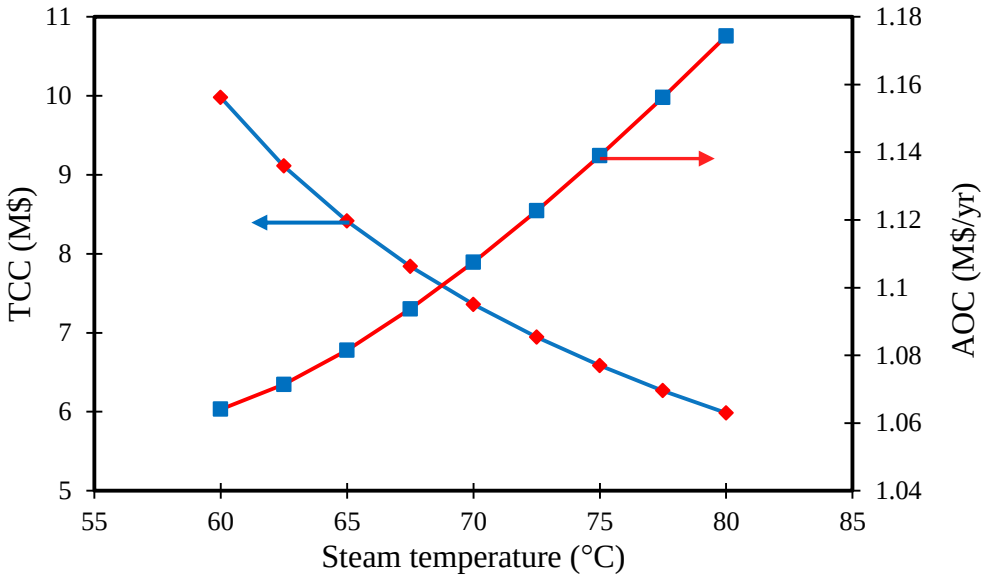

Fig. 6. Total capital cost and annual operating cost against steam temperature of the MED process

($M_{s(MED)}$= 8 kg/s, $Q_{f(RO)}$= 0.058 m³/s, $P_{f(RO)}$= 50 atm)

4.5 Impact of steam flow rate of the MED process

The impact of steam flow rate of the MED process on the fresh water cost and distillate is plotted in Fig. 7. This shows a slight reduction of fresh water cost as a response to increasing the steam flow rate. Also, it can be noticed that there is a weal relationship between the fresh water cost and steam flow rate. This insignificant reduction can be attributed to the increased productivity of the thermal process (Fig. 7). Indeed, the quantity of distillate produced by the MED process has a strong linear relationship with the quantity of steam provided. Specifically, this increases the AOC linked with steam supply despite the reduction of the specific cost of fresh water due to a bigger quantity of product is obtained.

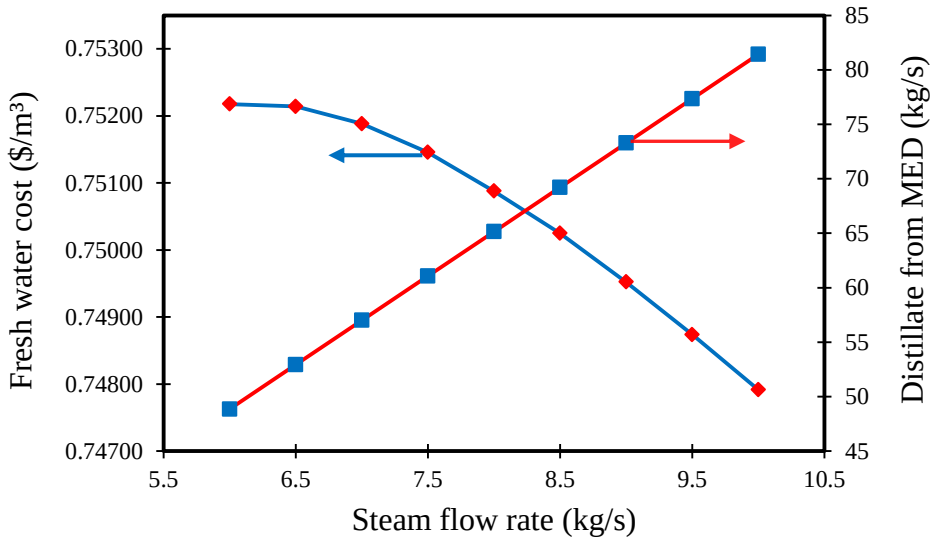

Fig. 7. Fresh water cost versus steam flow rate of the MED process ($T_{s(MED)}$= 70 °C, $Q_{f(RO)}$= 0.058 m³/s, $P_{f(RO)}$= 50 atm)

4.6 Joint impact of RO properties on fresh water cost

From a practical aspect, it is fair to expect the occurring of a simultaneously step change in two operating parameters of the RO process. Therefore, this section aims to understand the impact of a simultaneously change in the feed pressure and feed flow rate of the RO process on the fresh water cost of the hybrid system.

Fig. 8 shows the influence of joint parameters of the RO in determining the fresh water cost of the hybrid system at fixed operating conditions of MED process. This in turn shows a strong relationship between the fresh water cost and both the RO pressure and feed flow rate. The most common conclusion that can be made is that increasing the productivity of the RO would be convenient to obtain an overall reduction of fresh water cost. This is readily occurred at high pressures and flow rates.

The simulation results of Fig. 8 demonstrate the lowest fresh water production cost as 0.65 $ per cubic meter of fresh water, when the operating conditions for pressure and flow rate are 80 atm and 0.112 m^3/s, respectively. Corresponding recovery ratio of the RO process at those conditions is 60 %.

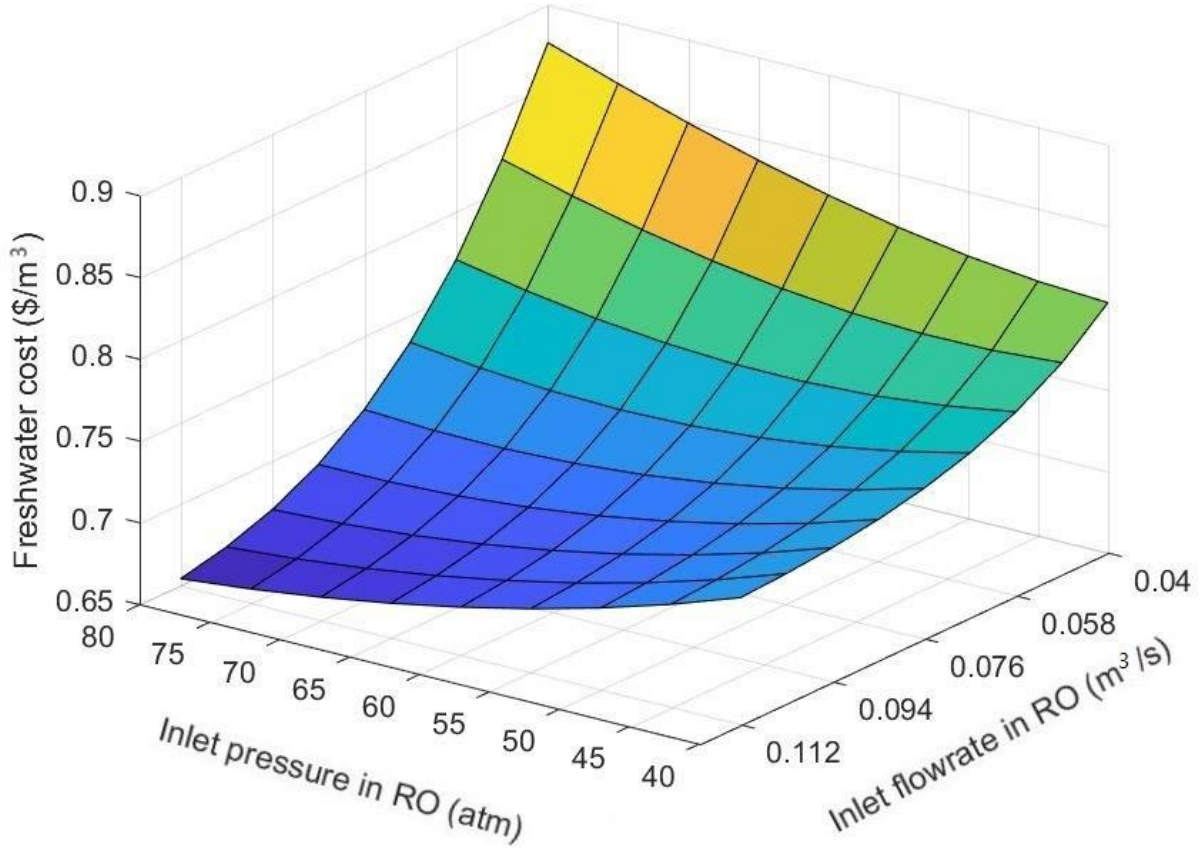

Fig. 8. Fresh water cost against feed pressure and flow rate of the RO process. MED steam temperature fixed to 70 °C and MED steam flow rate fixed to 8 kg/s.

4.7 Joint impact of MED properties on fresh water cost

Steam temperature and steam flow rate have been simultaneously changed to understand the impact of joint parameters of MED on the fresh water cost of the hybrid system.

Interestingly, Fig. 9 confirms the advantages of controlling the temperature of steam fed to the thermal process to be within 65° and 70° C temperature window, which in turn attains the lowest fresh water cost. Indeed, the fresh water cost raises outside of this optimal range of steam temperature. However, fresh water cost is insignificantly impacted by the temperature and flow

rate of the steam fed to the MED process. This means that the margin for optimisation of the thermal process is more restricted compared to the one for the membrane process.

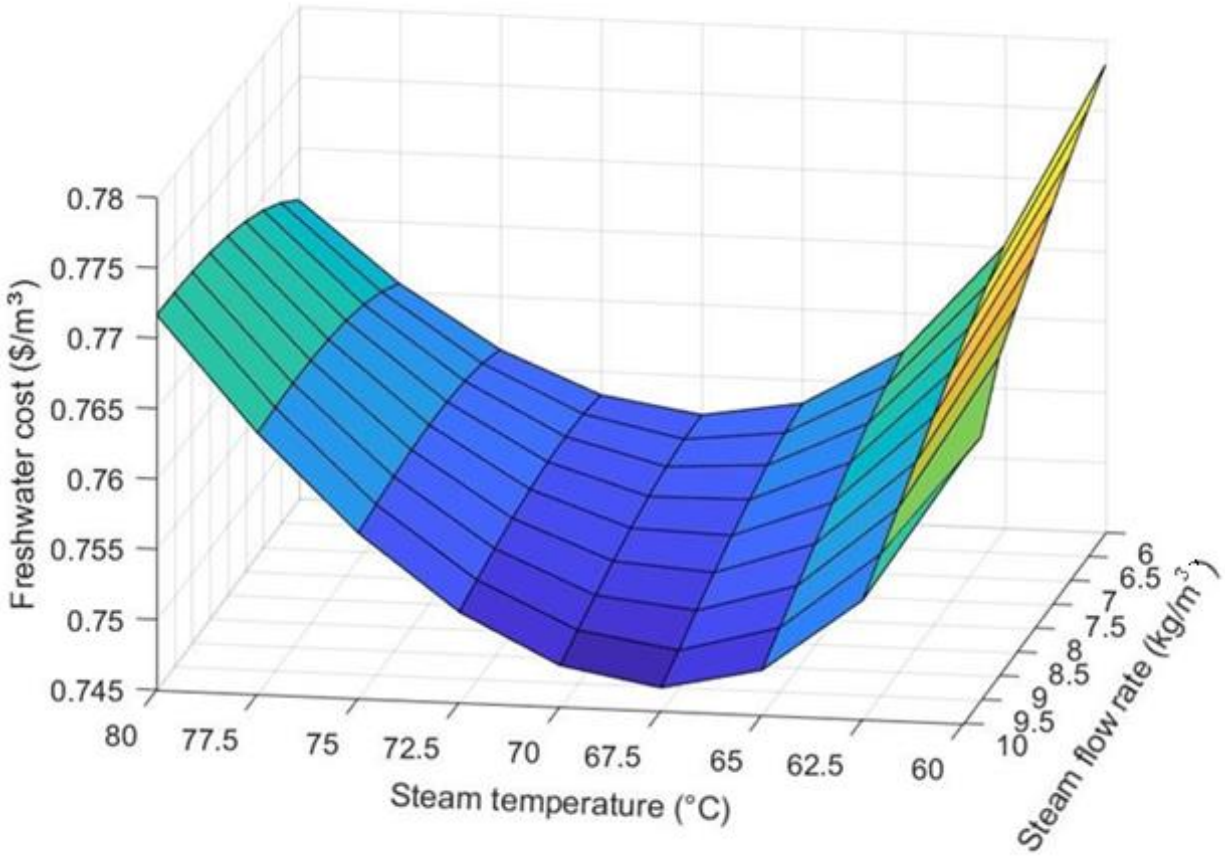

Fig. 9. Fresh water cost against steam temperature and flow rate of MED process. The operating conditions of the RO process are fixed at 50 atm and 0.0588 m³/s of pressure and feed flow rate, respectively

5. Optimisation of fresh water cost of hybrid MED+RO system

The sensitivity analysis presented above has provided a full understanding of the impact of operating conditions of the hybrid process on the fresh water production cost. However, the authors believe that there is a necessity to conduct a rigorous optimisation study to investigate the possibility of minimising the fresh water production cost while respecting some operative

and qualitative constrains. Therefore, the single-object optimisation of water production cost of the proposed hybrid system of MED+RO was performed using the gPROMS software as discussed in the next section.

Basically, the optimisation methodology is characterised by solving purely algebraic model equations. This also includes the minimisation or maximization of the selected nonlinear objective function with implementing a set of decision variables that varied throughout upper and lower limits of operation. Also, the optimisation is already subjected to a few nonlinear constraints to maintain the process requirements. Therefore, the nonlinear algebraic equations of the hybrid MED+RO system can be written in the following compact form:

$f(x, u, v) = 0$

x is the set of all algebraic variables, u is the set of all decision variables (to be optimised) and v indicates the constant parameters of the process. The function f is assumed to be continuously differentiable with respect to all their arguments.

5.1 Description of a single-objective optimisation methodology

The single objective function optimisation has been performed in gPROMS in order to minimise the specific cost of the produced fresh water for the proposed hybrid MED+RO system shown in Fig. 1. In this respect, the optimisation problem is formulated as a Non-Linear Programming (NLP) Problem with process and module constraints. The optimisation variables include the operating pressure and feed flow rate of the RO process, the steam temperature and flow rate of the MED process. However, the seawater properties include salinity and temperature have been considered as environmental variables, and therefore fixed values were assumed. Consequently, seawater salinity was fixed at 39000 ppm and seawater temperature at 25 °C. Furthermore, the

design of MED and RO processes has been assumed constant, which includes the number of effects of MED process (10) and the number of pressure vessels (43) in RO process, as designed by Filippini et al. (2018) and shown in Fig. 1.

The optimisation methodology has considered the upper and lower bounds of operating pressure and flow rate for the RO design process of 20 pressure vessels in the first stage and based on the manufacturers' specifications of a single membrane type TM820M-400/SWRO. Moreover, motive steam temperature is limited in the reasonable range for a low-temperature MED process. External steam flow rate is allowed within a 25% variation around its original value, to remain compliant with a fixed dimension of MED evaporators. Further constrains must be imposed on other process variables to ensure the correct operation of the whole hybrid system. Specifically, it is necessary to guarantee a good fresh water purity by imposing a salinity lower than 200 ppm (WHO, 2011). Also, a maximum salinity in the MED feed coming from RO of 45000 ppm is constrained to control the inlet salinity for the thermal process. This is imposed to maintain the process at a plausible overall recovery ratio that mitigates any significant increase in the quantity of rejected brine. Consequently, a minimum value of 30% of the overall recovery ratio is constrained to maintain the advantage of the RO upstream configuration and fulfill the lowest industrial recovery.

The non-linear optimisation solution used to optimise the fresh water cost of the hybrid MED+RO system considering the limits of operation of individual processes and the constraints of fresh water salinity is described below.

Given:

- Seawater concentration (39000 ppm), and temperature (25 °C), module specifications, membrane elements and pressure vessels number of the first stage of RO process (20), number of effects (10), and the rejected brine salinity of MED process (60000 ppm).

Determine:

- Optimal feed pressure and feed flow rate of the RO process (continuous variables).
- Steam flow rate and temperature of MED process (continuous variables).

So as to:

- Minimise: The fresh water cost of the hybrid MED+RO system.

Subject to:

- Equality (hybrid MED+RO process model) and inequality constraints including the operational parameters of the RO plant and each membrane element (linear bounds of optimisation variables). Also, the constrains of inlet seawater salinity of the MED process and the fresh water salinity of the Hybrid MED+RO system are considered.

The optimisation problem can therefore be mathematically written as follows:

$$\underset{P_{f(plant)}, Q_{f(plant)}, Ts, Ms}{\text{Min}} \quad FWC_{Hybrid}$$

Subject to:

Equality constraints:

Process Model: $f(x, u, v) = 0$

Inequality constraints:

$$(40 \text{ atm}) \quad P_{f(plant)}{}^{L} \leq P_{f(plant)} \leq P_{f(plant)}{}^{U} \ (81 \text{ atm})$$

$$(0.04 \text{ m}^3/\text{s}) \quad Q_{f(plant)}{}^{L} \leq Q_{f(plant)} \leq Q_{f(plant)}{}^{U} \ (0.2 \text{ m}^3/\text{s})$$

$$(60\ °C) \quad T_s^L \leq T_s \leq T_s^U \quad (80\ °C)$$

$$6\ \text{kg/s} \quad M_s^L \leq \quad M_s \quad \leq M_s^U \quad 10\ \text{kg/s}$$

End-point constrain: $RR \geq 30\%$

$$xf \leq 45000\ ppm$$

$$x_{freshwater} \leq 200\ ppm$$

$$(0.001\ \text{m}^3/\text{s}) \quad Q_{f(membrane)}{}^L \leq Q_{f(membrane)} \leq Q_{f(membrane)}{}^U \quad (0.005\ \text{m}^3/\text{s})$$

L and U are the lower and upper limits, respectively.

5.2 Optimisation results and discussion

The non-optimised operating conditions of the hybrid MED+RO system proposed by Filippini et al. (2018) and the optimised ones generated after the single objective optimisation problem are given in Table 6. In this respect, the temperature of the motive steam for the MED is reduced to 68.1 °C compared to the initial value of 70 °C, which highlights the importance of operating with low temperatures when aiming to obtain an optimal performance for the thermal process. Also, the quantity of the external steam to be provided to MED is increased up to 9.7 kg/s, which means a higher production of distillate of the thermal process and a lower specific cost (Fig. 7). Increasing the capacity of the MED process also aids in further reducing the salinity of blended fresh water and make easier to fulfil the constrain of fresh water purity. Regarding the RO membrane process, the optimisation methodology introduces a consistent increase of both the inlet feed flow rate and operating pressure of the RO compared to the one proposed by Filippini et al. (2018). The obtained value of 77.4 atm for the operating pressure is a very significant increment with respect to its previous value (+55%), but it is still lower than the maximum value

allowable for the TM820M-400 membrane of 81.91 atm. Seemingly, the optimisation constrains of inlet feed salinity of the MED process has a contribution in determining the optimal value of RO operating pressure. Apparently, the increase of operating pressure of the RO process is crucial to enhance the productivity of the RO and to mitigate the specific cost of fresh water. Also, the increase of feed flow rate of the RO process up to 0.107 m^3/s is important to increase the total product flow rate of the process that almost serves the reduction of fresh water cost.

Table 6. The non-optimised operating conditions from Filippini et al. (2018) and the optimised values.

Operating parameters	Non-optimised values	Optimised values
Steam temperature (°C)	70	68.1
Steam flow rate (kg/s)	8	9.71
Feed flow rate to RO (m^3/s)	0.058	0.107
Feed pressure to RO (atm)	50	77.9

Several considerations can be drawn from Table 7, where the non-optimised and optimised values of several operating condition of the hybrid MED+RO system including the estimated economic parameters are presented. The capital costs are increasing mainly due to the lower steam temperature, which means higher exchange area required for the effects of the thermal process, and because of the greater capacity of the RO process. Operative costs are increasing as well, mainly because of the substantial increase of pump work for the membrane process due to a greater flow rate must be compressed to a higher pressure. On the other hand, the productivity is significantly increased, especially for the membrane process, and this guarantees a lower specific cost of fresh water. From the energetic point of view, the optimisation had a negligible impact on the energy consumption of the single processes. However, the overall energy consumption is reduced by 19% due to a greater portion of fresh water is now produced with the least energetic

demanding process (RO). Based on the above results, the overall production cost decreases from 0.75 to 0.66 $/m^3, which is a reduction of almost 13%.

Moreover, Table 7 illustrates others relevant parameters. The fresh water salinity is decreased because of a significant increase occurred in the distillate of the MED process in addition to the production of lower salinity in the RO process. The rejected brine flow rate is also significantly increased, because of the increased inlet salinity into the thermal process and bigger capacity of the thermal process. For the same reason, the recovery ratio of the MED drops by 12%, while the recovery ratio of RO increases a lot. Consequently, the overall recovery is slightly reduced due to a significant impact of the thermal process. However, a reasonable value of 32.5% is achieved to consider the imposed constrain.

Table 7. Comparison between the non-optimised and optimised hybrid MED+RO systems.

Calculated parameter	**Non-optimised values**	**Optimised value**	**% Variation**
Total Cost ($/m^3)	**0.75**	**0.66**	**-12.76**
TCC (M$)	7.36	9.10	+23.64
AOC (M$/yr)	1.49	2.25	+50.44
Fresh water cost of MED ($/m^3)	0.80	0.77	-2.2
Fresh water cost of RO ($/m^3)	0.55	0.47	-14.1
Energy consumption (kWh/m^3)	**20.27**	**16.37**	**-19.27**
Energy consumption of MED (kWh/m^3)	25.66	25.89	+0.89
Energy consumption of RO (kWh/m^3)	4.29	4.36	+1.72
Total productivity (kg/s)	**88.81**	**143.5**	**+61.57**
MED distillate (kg/s)	66.41	80.01	+20.47
RO permeate (kg/s)	22.51	63.80	+183.46
Fresh water salinity (ppm)	144	122	-15.35
Rejected brine flowrate (kg/s)	168	240	+42.9
Overall Recovery Ratio (-)	**0.348**	**0.325**	**-6.4**
MED Recovery Ratio (-)	0.29	0.25	-12.3
RO Recovery Ratio (-)	0.37	0.57	+54.2

Fig. 10 shows the optimisation results for different values of seawater temperature and salinity against the optimal fresh water cost. The clarification of this issue is important to elucidate the possible optimisation fresh water cost against any proposed variation of both seawater temperature and salinity at fixed optimised parameters of Table 6. As expected, it is costlier to operate the hybrid plant with higher salinity seawater, as well as with lower temperature. This is belonging to a significant decrease in performance of both the thermal and the RO membrane processes at such conditions. Statistically, fresh water cost is reduced to 0.55 $/m^3 for a warm and low-saline seawater compared to a highest cost of 0.89 $/m^3 when a cold and highly saline seawater is fed to the hybrid plant.

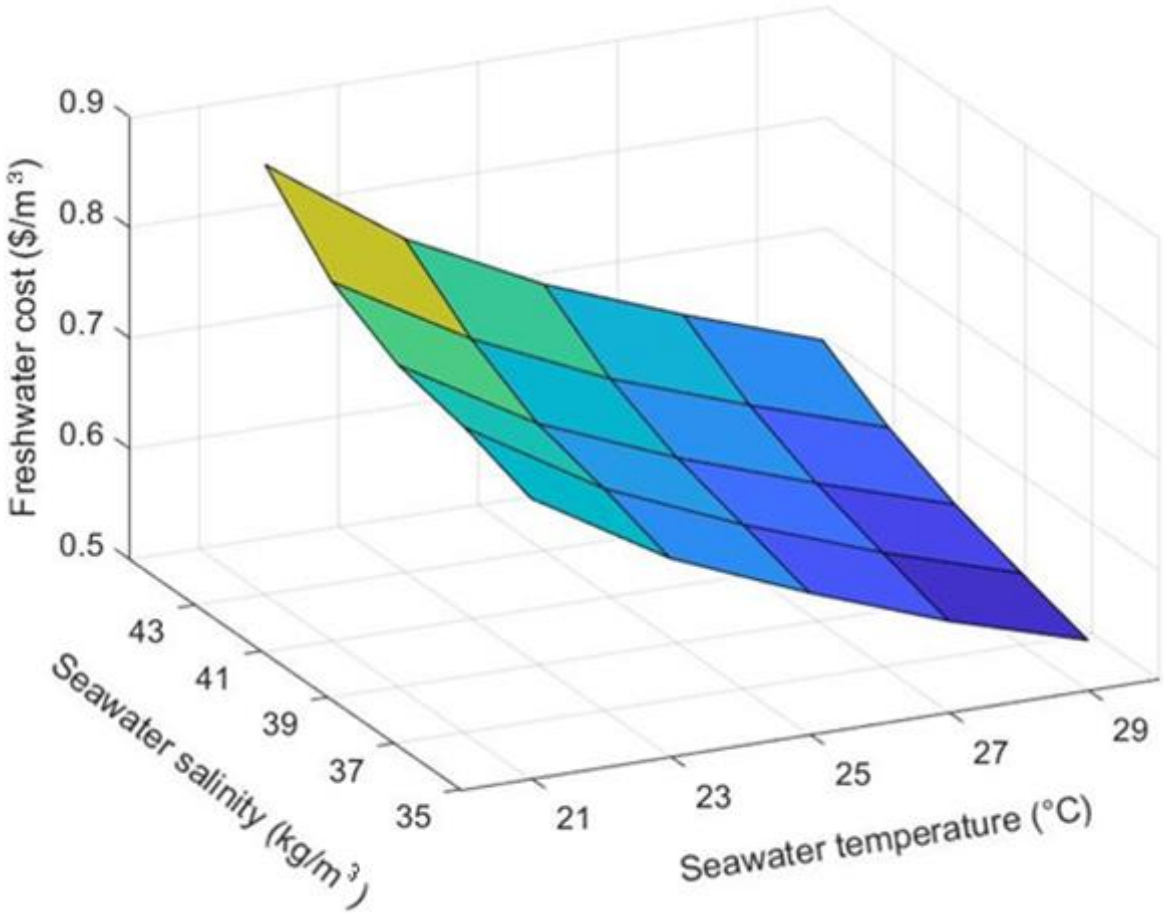

Fig. 10. Fresh water production cost after optimisation, considering different seawater properties. All parameters are fixed according to optimised value (Table 6), electricity cost is fixed at 0.09 $/kWh.

Furthermore, Fig. 11 shows the behavior of fresh water cost and annualised operating cost of the hybrid MED+RO system against the variation of electricity cost. The electricity cost is mainly dependent on the plant location. Therefore, a wide range of electricity cost is expected along different countries. In this respect, Fig. 11 shows a strong linear relationship between the overall optimised fresh water cost and the electricity price. This is attributed to the great variation of the annual operating costs; when electricity cost increases by 50%, fresh water production cost increases by about 22%. Basically, the electricity is already consumed in the high-pressure pump of the RO process and some pumps of the thermal MED process.

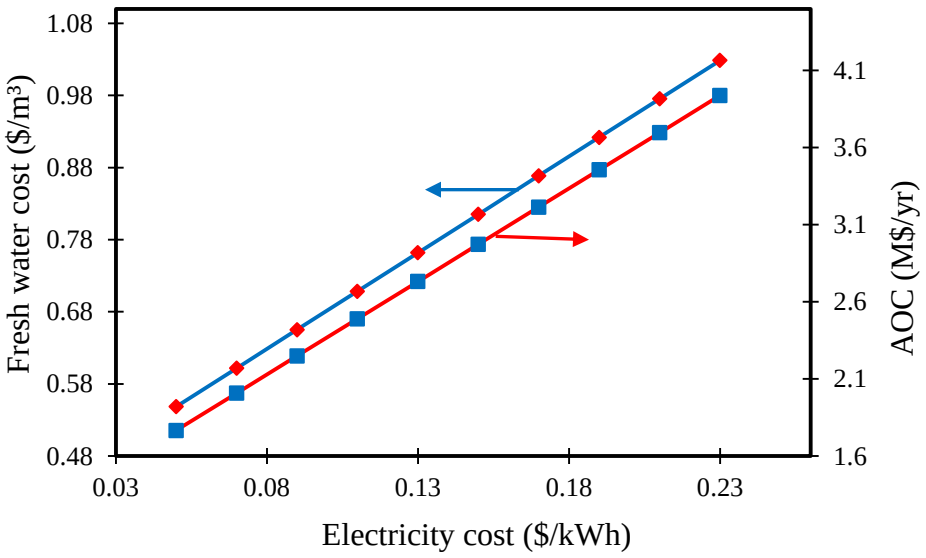

Fig. 11. Fresh water cost and AOC of the hybrid MED+RO system against electricity cost. All parameters are fixed according to optimised value (Table 6), seawater salinity fixed as 39.000 ppm and temperature at 25°C.

Finally, Fig. 12 shows the dependence of fresh water production cost and AOC on steam cost per kg. A noticeable increase of around 7% in the cost of fresh water is noticed as a result to

increasing the cost of steam by 50%. However, this increase in the fresh water cost is less relevant than the dependence on electricity price. This confirms that installing the proposed plant in a region where electricity is cheap is of paramount importance to maintain a low cost of produced fresh water.

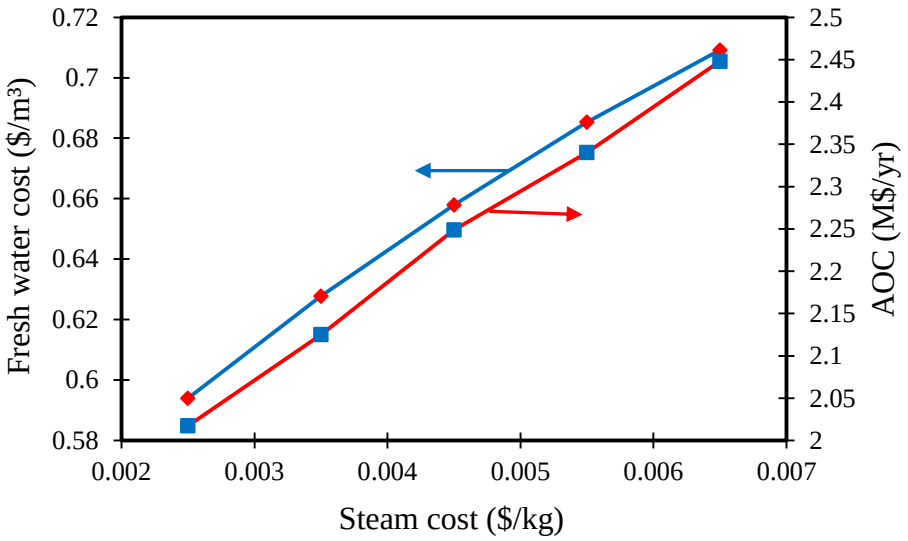

Fig. 12. Fresh water cost and AOC of the hybrid MED+RO system against steam cost. All parameters are fixed according to optimized value (Table 5), seawater salinity fixed as 39.000 ppm and temperature at 25°C.

5.3. Practical Implications of the current research

Several practical implications can be drawn from the current research including;

- It can be used to investigate the performance indicators and the fresh water production cost of any size of MED+RO hybrid system due to the availability of a robust mathematical and cost models. Also, it can be efficiently used to estimate the performance indicators of different configurations of RO process in the hybrid system.

- The sensitivity analysis covers most the required operating conditions of both MED and RO processes. This in turn would help the managers to investigate the proper one that need to be implemented in such hybrid system and take correct decisions.
- It is a perfect tool to investigate the advantages of coupling the proposed hybrid system of MED+RO with an alternative source of energy such as solar power energy.

6. Conclusions

In this paper, an earlier authors' model for the hybrid MED_TVC+RO system has been augmented with detailed cost models for the individual processes of MED_TVC and RO gathered from the literature to estimate the fresh water cost. The low-temperature MED process has been identified as more cost competitive with respect to the MED_TVC process. Therefore, the TVC section has been deactivated for the optimisation. In this respect, a sensitivity analysis was carried out for the proposed hybrid MED+RO process with respect to steam temperature, and steam flow rate of the MED process, and operating pressure and feed flow rate of the RO process. The impact of the considered parameters in terms of fresh water cost was investigated by varying a single parameter at a time, and then by considering the joint variation of MED parameters and RO parameters at the same time. This in turn aids to understand the interaction between the process performance include the fresh water cost and operating conditions. The analysis highlighted a higher cost dependence on the operating conditions of the RO compared to those of the MED.

An optimisation study was conducted to minimise the fresh water cost of the hybrid MED+RO system by manipulating the operating conditions of RO process, as well as the feed flow rate and steam flow rate of the MED process. An optimal point, corresponding to a fresh water cost of

0.66 $/m^3, was identified. This value was obtained considering average values of seawater salinity (39 kg/m^3) and temperature (25 °C). However, the optimisation methodology has demonstrated the insignificant impact of seawater salinity and temperature on the fresh water cost where different optimal point were found for different seawater conditions. Finally, the dependence of fresh water cost on electricity cost was investigated, showing how desalination cost can substantially higher in countries where electricity is costlier. The main limitation of the present study is that the proposed plant is a theoretical one, entirely model-based. In this respect, the results in terms of fresh water cost refers to the performance of this hypothetical plant. However, those results and the sensitivity analysis could certainly be an instrument for managers and engineers of a real similar plant when deciding the best design and operative conditions. As a further development of this work, a renewable energy plant, i.e. a photovoltaic or concentrating solar power farm, could be coupled with the presented hybrid desalination system, since a great portion of the operative costs has been found to be linked with the electricity cost. The aim of the proposed future study could be the evaluation of the fresh water cost when the necessary energy to run the desalination unit is provided by an alternative source.

Nomenclature

A^* : Feed spacer characteristic (-)

A_m : Effective membrane area (m²)

$A_{w(T)}$: Water permeability constant at operating temperature (m/s atm)

$A_{ev,i}$: Exchange area of i-th evaporator (m^2)

$A_{ph,i}$: Exchange area of i-th pre-heater (m^2)

A_{cond}: Exchange area of final condenser (m^2)

$A_{ev,mean}$: Mean exchange area of evaporators (m^2)

$A_{ph,mean}$: Mean exchange area of pre-heaters (m^2)

B_i : Brine rejected by the i-th effect (kg/s)

$B_{s(T)}$: Solute transport parameter at operating temperature (m/s)

C_b : Bulk concentration of a single membrane (kg/m³)

C_f : Feed concentration of a single membrane (kg/m³)

$C_{f(plant)}$: Plant feed concentration (kg/m³)

C_p : Permeate concentration at the permeate channel of a single membrane (kg/m³)

C_r : Retentate concentration of a single membrane (kg/m³)

C_w : Membrane surface concentration of a single membrane (kg/m³)

CR: Compression ratio in the steam ejector (-)

D_i : Total distillate produced in i-th effect (kg/s)

D_b : Diffusivity parameter (m²/s)

d_h : Hydraulic diameter of the feed spacer channel (m)

$D_{boil,i}$: Distillate produced by boiling in i-th evaporator (kg/s)

$D_{flash,i}$: Distillate produced by flashing in i-th flashing box (kg/s)

E_s : Specific energy consumption (kJ/kg)

ERD : Energy recovery device (-)

J_w : Water flux through a single membrane (m/s)

k : Mass transfer coefficient (m/s)

k_{dc} : Constant (-)

L : Membrane length (m)

L_f : Length of filament in the spacer mesh (m)

m_f : Coefficient

Mb: Rejected brine flowrate (kg/s)

$M_{COND:}$ Flowrate of steam in the final condenser (kg/s)

Md: Distillate from MED process (kg/s)

Mf: Water intake in the first effect (kg/s)

Mm: Motive steam flowrate (kg/s)

Ms: Total steam flowrate (kg/s)

Mw: Intake water flowrate (kg/s)

M_{TVC}: Vapor flowrate entrained in TVC section (kg/s)

n: Number of effects of MED process (-)

PFC: Pressure Correction Factor (-)

P_v: Pressure of saturated steam at temperature T_v (kPa)

P_s: Pressure of saturated steam at temperature T_s (kPa)

P_m: Pressure of saturated steam at temperature T_m (kPa)

P_{ev}: Pressure of saturated entrained vapor (kPa)

P_{crit}: Critical pressure of water (kPa)

P_f : Operating feed pressure of a single membrane (atm)

$P_{f(plant)}$: Plant feed pressure (atm)

P_p : Permeate pressure at the permeate channel (atm)

P_r : Retenate pressure of a single membrane (atm)

$P_{r(plant)}$: Plant retenate pressure (atm)

Q_b : Bulk flowrate of a single membrane (m^3/s)

Q_f : Feed flowrate of a single membrane (m^3/s)

$Q_{f(plant)}$: Plant feed flow rate (m^3/s)

Q_p : Total permeate flow rate of a single membrane (m^3/s)

$Q_{p(plant)}$: Plant permeate flow rate (m^3/s)

$Q_{p(PV)}$: Permeate flow rate of single pressure vessel (m^3/s)

Q_r : Retentate flowrate of a single membrane (m^3/s)

$Q_{r(plant)}$: Plant retentate flowrate (m^3/s)

Q_s : Total solute flux through the membrane ($kg/m^2\ s$)

Q_{COND}: Thermal load in final condenser (kW)

$Q_{sensible}$: Sensible heat used in first effect (kJ/kg)

Q_{latent}: Latent heat used in first effect (kJ/kg)

Q_i: Thermal load at i-th evaporator (kW)

Q_s: Thermal load of steam (kW)

Ra: Entrainment ratio (-)

Re_b : Reynolds number (-)

Rec : Total recovery rate of a single membrane (-)

$Rec_{(plant)}$: Plant recovery rate (-)

Rej : Total solute rejection (-)

$Rej_{(plant)}$: Plant solute rejection (-)

Sc : Schmidt number (-)

t_i : Feed temperature after i-th pre-heater (°C)

t_f : Height of feed channel of the membrane (m)

tn: Feed temperature after final condenser (°C)

$T1$: Top brine temperature (Ttop) (°C)

Tb: Temperature of rejected brine (°C)

Ts: Steam temperature (°C)

Tv_i: Temperature of the vapor phase in i-th effect (°C)

Tw: Temperature of the cooling water (°C)

T_{mean}: Mean temperature in the plant (°C)

T_{crit}: Critical temperature of water (°C)

TCF: Temperature Correction Factor (-)

$U_{ev,i}$: Global heat exchange coefficient in i-th evaporator (kW/m^2 °C)

$U_{ph,i}$: Global heat exchange coefficient in i-th pre-heater (kW/m^2 °C)

U_{cond}: Global heat exchange coefficient in final condenser (kW/m^2 °C)

U_b : Cross flow velocity of a single membrane (m/s)

W : Membrane width (m)

x_i: Salinity in i-th evaporator (ppm or w/w%)

xb: Salinity in rejected brine (ppm or w/w%)

xf: Salinity in the feed (ppm or w/w%)

x_{mean}: Mean salinity in the plant (ppm or w/w%)

Greek

α: Fraction of rejected brine from previous effect flashed in the associated pre-heater (-)

β: Fraction of total distillate boiled in each evaporator (-)

$\Delta A_{ev}\%$: Percentage error on evaporators' areas (%)

$\Delta A_{ph}\%$: Percentage error on pre-heaters areas (%)

$\Delta T_{ex,i}$: Driving force for heat exchange in i-th evaporator (°C)

$\Delta t_{\log,i}$: Driving force for heat exchange in i-th pre-heater (°C)

$\Delta T_{\log,cond}$: Driving force for heat exchange in final condenser (°C)

ΔT_i : Temperature drop between two evaporators (°C)

Δt_i : Temperature increase between two pre-heaters (°C)

$\Delta P_{drop,E}$: Total pressure drop along the membrane element (atm)

λ: Latent heat of evaporation (kJ/kg)

π_p : Total osmotic pressure at the permeate channel (atm)

π_w : Total osmotic pressure at the membrane surface (atm)

ρ_b : Density parameter (kg/m³)

μ_b : Kinematic viscosity (kg/m s)

ϵ : Membrane porosity (-)

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Appendix A

Table A.1. Specification of the MED and RO processes with the operating conditions (Adapted from Filippini et al. (2018))

Operative parameter	**Value**	**Unit**
Number of effects	10	-
Steam flow rate	8	kg/s
Steam temperature	70	°C
Rejected brine temperature	40	°C
Rejected brine salinity	60	kg/m^3
Seawater temperature	25	°C
Seawater salinity	39	kg/m^3
External steam pressure	1300	kPa
Effective operating pressure in RO	50	atm

Membrane properties	**Value**	**Unit**
Membrane:	TM820M-400/ SWRO	-
Supplier	Toray membrane	-
Membrane material and module configuration	Polyamide thin-film composite Spiral wound element	-
Maximum operating pressure	81.91	atm
Maximum operating feed flow rate	0.00536	m^3/s
Minimum operating feed flow rate	0.001	m^3/s
Maximum pressure drop per element	0.987	atm
Maximum operating temperature	45	°C
Effective membrane area (A_m)	37.2	m^2
Module length (L) and width (W)	1 and 37.2	m
$B_{s(T_o)}$ NaCl and $A_{w\,(T_o)}$ at 25 °C	1.74934×10^{-8} and 3.1591×10^{-7}	(m/s) and (m/s atm)
Feed spacer type	Naltex-129	-
Feed spacer thickness (t_f) and hydraulic diameter (d_h)	8.6×10^{-4} (34 mils) and 8.126×10^{-4}	m
Length of filament in the spacer mesh	2.77×10^{-3}	m
Spacer characteristics (A')	7.38	-
Spacer characteristics (n)	0.34	-
Voidage (ε)	0.9058	-

Table A.2. Modelling of individual Reverse Osmosis process (Adapted from Filippini et al. (2018))

Expression	Notes	Unit
$Q_p = A_{w(T)} \left(P_f - \frac{\Delta P_{drop,E}}{2} - P_p - \pi_w - \pi_p\right) A_m$	Water flux through the membrane	m³/s
$Q_s = B_{s(T)}(C_w - C_p)$	Solute flux through the membrane	kg/m² s
$\pi_w = 0.76881\ C_w \qquad \pi_p = 0.7994\ C_p$	The osmotic pressure in feed and permeate channels	atm
$A_{w(T)} = A_{w(25\,C)} \exp[0.0343\ (T-25)] < 25\ °C$ $A_{w(T)} = A_{w(25\,C)} \exp[0.0307\ (T-25)] > 25\ °C$	The impact of temperature on water transport parameter	m/s atm
$B_{s(T)} = B_{s(25\,C)}\ (1 + 0.08\ (T-25)) \quad < 25\ °C$ $B_{s(T)} = B_{s(25\,C)}\ (1 + 0.05\ (T-25)) \quad > 25\ °C$	The impact of temperature on solute transport parameter	m/s
$\Delta P_{drop,E} = \frac{9.8692x10^{-6}\ A^* \rho_b\ {Q_b}^2\ L}{2 d_h\ Re_b^n\ (W\ t_f\ \epsilon)^2}$ $Re_b = \frac{\rho_b\ d_h\ Q_b}{t_f\ W\ \mu_b}$	The pressure drop per element and Reynolds number	atm, -
$Q_b = \frac{Q_f + Q_r}{2} \qquad C_b = \frac{C_f + C_r}{2}$	The bulk flowrate and concentration	m³/s, kg/m³
$\frac{(C_w - C_p)}{(C_b - C_p)} = exp\left(\frac{Q_p/A_m}{k}\right)$	The membrane surface concentration	-
$k = 0.664\ k_{dc}\ Re_b^{0.5}\ Sc^{0.33}\ \left(\frac{D_b}{d_h}\right)\left(\frac{2d_h}{L_f}\right)^{0.5}$ $Sc = \frac{\mu_b}{\rho_b\ D_b}$	The mass transfer coefficient and Schmidt number	m/s, -
$\rho_b = 498.4\ m_f + \sqrt{\left[248400\ m_f^2 + 752.4\ m_f\ C_b\right]}$ $m_f = 1.0069 - 2.757x10^{-4}\ T$	Density parameter	kg/m³
$D_b = 6.72510^{-6}\ exp\left\{0.154610^{-3}\ C_b - \frac{2513}{T+273.15}\right\}$	Diffusivity parameter	m²/s
$\mu_b = 1.234x10^{-6}\ exp\left\{0.0212\ C_b + \frac{1965}{T+273.15}\right\}$	Viscosity parameter	kg/m s
$Q_f = Q_r + Q_p \qquad Q_f\ C_f - Q_r\ C_r = Q_p\ C_p$	The total mass and solute balance of the whole unit	m³/s
$C_p = \frac{B_s\ C_f\ e^{\frac{J_w}{k}}}{J_w + B_s\ e^{\frac{J_w}{k}}}$	The permeate concentration	kg/m³

$Rej = \frac{C_f - C_p}{C_f}$ $Rec = \frac{Q_p}{Q_f}$	Rejection and recovery rate	-

Table A.3. Modelling of MED process (Adapted from Filippini et al. (2018))

Description	Equation	Unit
Feed flowrate	$Mf = \frac{Ms\ \lambda(Ts)}{Q_{sensible} + Q_{latent}}$	kg/s
Sensible heat in first effect	$Q_{sensible} = Mf \int_{t1}^{T1} cp(T1, x1)dT$	kJ/s
Latent heat in first effect	$Q_{latent} = D1\ \lambda(Tv1)$	kJ/s
Temperature drop among effects (first attempt)	$\Delta T = \frac{Ts - Tb}{n}$	°C
Temperature drop among pre-heaters (first attempt)	$\Delta T = \Delta t$	°C
Feed temperature in first effect	$t1 = tn + (n-1)\ \Delta t$	°C
Temperature of vapor phase	$Tv = T - BPE(T, x)$	°C
Flowrate of flashed distillate	$D_{flash,i} = \alpha B_{i-1}$	kg/s
Fraction of distillate by flashing	$\alpha = \frac{cp(T_{mean}, x_{mean})\Delta T}{\lambda(T_{mean})}$	-
Mean temperature	$T_{mean} = \frac{T1 + Tb}{2}$	°C
Mean salinity	$x_{mean} = \frac{xf + xb}{2}$	ppm
Fraction of distillate by evaporation	$\beta = \frac{\alpha[xb(1-\alpha)^n - xf]}{(xb - xf)[1-(1-\alpha)^n]}$	-
Flowrate of evaporated distil.	$D_{i,boiled} = \beta M_D$	kg/s
Total distillate	$D_i = D_{i,boiled} + D_{i,flashed}$	kg/s
Rejected brine flowrate	$B_i = B_{i-1} - D_i$	kg/s
Salinity profile in the effects	$x_i = \frac{x_{i-1}B_{i-1}}{B_i}$	ppm
Area of i-th effect	$\frac{Q_i}{U_{ev,i}\Delta T_{ev,i}} = A_{ev,i}$	m^2
Heat load in i-th effect	$Q_i = D_{boiled,i-1}\ \lambda(T_{v,i-1})$	kJ/s
Temperature drop in heat exchangers	$\Delta T_{ev,i} = \Delta T - BPE_{i-1}$	°C
Area of i-th pre-heater	$Mf \cdot \int_{t_{i+1}}^{t_i} cp(t, xf)dt = U_{ph,i} A_{ph,i} \Delta t_{\log,i}$	m^2
Logarithmic temperature difference in pre-heaters	$\Delta t_{\log,i} = \frac{\Delta t}{\log(\frac{Tv_i - t_{i+1}}{Tv_i - t_i})}$	°C
Area of final condenser	$Q_{COND} = U_{COND} A_{COND} \Delta T_{log,COND}$	m^2
Heat load in final condenser	$Q_{COND} = D_n \lambda(Tv_n)$	kJ/s
Logarithmic temperature difference in final condenser	$\Delta T_{\log,COND} = \frac{tn - Tw}{\log(\frac{Tv_n - Tw}{Tv_n - tn})}$	°C

Table A.4. Modelling of hybrid MED+RO system (Adapted from Filippini et al. (2018))

Description	Equation	Unit
Seawater feed to MED process	$Mw_{MED} = Mr_{RO} + M_{bypass}$	kg/s
BM for inlet salinity to MED process	$Mw_{MED}xf_{MED} = Mr_{RO}xr_{RO} + M_{bypass}x_{seawater}$	ppm
Total freshwater production	$Md_{MED} + Mp_{RO} = M_{freshwater}$	kg/s
BM for salinity of freshwater	$Md_{MED}xd_{MEd} + Mp_{RO}xp_{RO} = M_{freshwater}x_{freshwater}$	ppm
Total rejected brine	$Mb_{MED} = M_{reject}$	kg/s
BM for rejected brine salinity	$xr_{MED} = x_{reject}$	ppm

Table A.5. Equations describing the TVC section modelling. (Dessouky et al., 2002)

Description	Equation	Unit
Pressure Correction Factor	$PCF = 3e\text{-}7 \cdot Pm^2 - 0.0009 \cdot Pm + 1.6101$	-
Temperature Correction Factor	$TCF = 2e\text{-}8 \cdot Tv_n^{\,2} - 0.0006 \cdot Tv_n + 1.0047$	-
Pressure at vapor temperature	$Pv = P_{crit} e^{(\frac{T_{crit}}{Tv_n}+273.15) - 1} \cdot \sum_{j=1}^{8} f_j$	bar
Pressure at steam temperature	$Ps = P_{crit} e^{(\frac{T_{crit}}{Ts}+273.15) - 1} \cdot \sum_{j=1}^{8} f_j$	bar
Calculate Compression Ratio	$CR = \frac{Pv}{Ps}$	-
Calculate Entrainment Ratio	$Ra = 0.296 \frac{Ps^{1.19}}{Pev^{1.04}} \frac{Pm^{0.015}}{Pev^{0.015}} \frac{PCF}{TCF}$	-
Calculate motive steam flowrate	$Mm = Ms \frac{Ra}{1 + Ra}$	kg/s

Coefficient	$f1$	$f2$	$f3$	$f4$	$f5$	$f6$	$f7$	$f8$
Value	-7.4192	0.29721	-0.1155	0.00868	0.00109	-0.0043	0.00252	-0.00052

Appendix B

Correlations for MED process

Collected from : El-Dessouky HT, Ettouney H.M., 2002. *Fundamentals of salt water desalination*. Elsevier.

Boiling Point Elevation

Correlation valid in the range: 1% < w < 16%, 10°C < T < 180°C

$$w = x \cdot 10^{-5} \quad [w/w\%]$$

$$BPEa = 8.325 \cdot 10^{-2} + 1.883 \cdot 10^{-4} \cdot T + 4.02 \cdot 10^{-6} \cdot T^2$$

$$BPEb = -7.625 \cdot 10^{-4} + 9.02 \cdot 10^{-5} \cdot T - 5.2 \cdot 10^{-7} \cdot T^2$$

$$BPEc = 1.522 \cdot 10^{-4} - 3 \cdot 10^{-6} \cdot T - 3 \cdot 10^{-8} \cdot T^2$$

$$BPE = BPEa \cdot w + BPEb \cdot w^2 + BPEc \cdot w^3 \quad [°C]$$

Specific heat at constant pressure

Correlation valid in the range: 20000 ppm < x < 160000 ppm, 20°C < T < 180°C

$$s = x \cdot 10^{-3} \quad [gm/kg]$$

$$cpa = 4206.8 - 6.6197 \cdot s + 1.2288 \cdot 10^{-2} \cdot s^2$$

$$cpb = -1.1262 + 5.4178 \cdot 10^{-2} \cdot s - 2.2719 \cdot 10^{-4} \cdot s^2$$

$$cpc = 1.2026 \cdot 10^{-2} - 5.3566 \cdot 10^{-4} \cdot s + 1.8906 \cdot 10^{-6} \cdot s^2$$

$$cpd = 6.8777 \cdot 10^{-7} + 1.517 \cdot 10^{-6} \cdot s - 4.4268 \cdot 10^{-9} \cdot s^2$$

$$cp = \frac{cpa + cpb \cdot T + cpc \cdot T^2 + cpd \cdot T^3}{1000} \quad [\frac{kJ}{kg \cdot °C}]$$

Latent heat of evaporation

$$\lambda = 2501.89715 - 2.40706 \cdot T + 1.19221 \cdot 10^{-3} \cdot T^2 - 1.5863 \cdot 10^{-5} \cdot T^3 \quad [\frac{kJ}{kg}]$$

Global heat exchange coefficients

$$U_{ev} = 1.9695 + 1.2057 \cdot 10^{-2} \cdot T - 8.5989 \cdot 10^{-5} \cdot T^2 + 2.5651 \cdot 10^{-7} \cdot T^3 \quad [\frac{kW}{m^2 \cdot °C}]$$

$$U_{cond} = U_{ph} = 1.7194 + 3.2063 \cdot 10^{-3} \cdot T + 1.597 \cdot 10^{-5} \cdot T^2 - 1.9918 \cdot 10^{-7} \cdot T^3 \quad [\frac{kW}{m^2 \cdot °C}]$$