

1 **Performance analysis of hybrid system of multi effect distillation and reverse osmosis for**
2 **seawater desalination via modeling and simulation**

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11 **Abstract**

12
13 The coupling of thermal (Multi Stage Flash, MSF) and membrane processes (Reverse Osmosis,
14 RO) in desalination systems has been widely presented in the literature to achieve an improvement
15 of performance compared to an individual process. However, very little study has been made to the
16 combined Multi Effect Distillation (MED) and Reverse Osmosis (RO) processes. Therefore, this
17 research investigates several design options of MED with thermal vapor compression (MED_TVC)
18 coupled with RO system. To achieve this aim, detailed mathematical models for the two processes
19 are developed, which are independently validated against the literature. Then, the integrated model
20 is used to investigate the performance of several configurations of the MED_TVC and RO
21 processes in the hybrid system. The performance indicators include the fresh water productivity,
22 energy consumption, fresh water purity, and recovery ratio. Basically, the sensitivity analysis for
23 each configuration is conducted with respect to seawater conditions and steam supply variation.
24 Most importantly, placing the RO membrane process upstream in the hybrid system generates the
25 overall best configuration in terms of the quantity and quality of fresh water produced. This is
26 attributed to acquiring the best recovery ratio and lower energy consumption over a wide range of
27 seawater salinity.

28
29 **Keywords:** Seawater desalination, MED_TVC+RO hybrid system, Mathematical modeling,
30 Sensitivity analysis.

1 **1. Introduction**

2 In the recent past, the demand for fresh water increased in many regions, especially in the
3 developing countries, which in turn pushed the researchers toward more energy-efficient ways for
4 seawater desalination. Coupling a power plant with a thermal desalination process allows to reach a
5 greater thermal efficiency. This is attributed to the thermal energy produced from the power plant
6 that would be used in the desalination process aside from wasting it. In this respect, the MSF was
7 considered as the preferred technology to couple with a power plant. However, the low-temperature
8 MED process proved to be more appropriate to couple with a power plant steam generator. This is
9 due to employing low temperature steam in the MED process ([Mahbub et al., 2009](#)).

10 Over the last decades, the use of RO process as a complementary option with MED process is
11 progressively increased. Interestingly, this technique acts in accordance with lower energy
12 consumption with attaining the regulated limits of potable water issued by the World Health
13 Organization ([WHO, 2011](#)). For instance, the Fujairah 2 desalination plant in the United Arab
14 Emirates is one of the biggest desalination facilities in the world, with a capacity of 591000 m³/day.
15 Quantitatively, this facility consists of a 2000 MW power plant coupled with a 450000 m³/day
16 MED plant and a 136000 m³/day RO plant ([Veolia Water, 2011](#)).

17 The desalination industry was growing very rapidly in the 2000s, and many researchers focused on
18 the development of more efficient desalination processes, including hybrid systems. The next
19 section illustrates several examples of the published research in the open literature regarding the
20 hybrid systems of MED, with or without the thermal vapor compression (TVC) section, coupled
21 with RO process.

22 [Hamed \(2005\)](#) reviewed the major features of commercially available hybrid desalination plants.
23 The study confirmed that Nanofiltration (NF) membranes can be the best technology to couple with
24 a thermal process, regarding fresh water productivity. Also, the full integration of membrane and
25 thermal desalination processes provided a higher thermal performance than the simple integration.
26 An economical evaluation of a small 2000 m³/day MED+RO system powered by natural gas and

1 includes heat recovery is carried out by [Cardona et al. \(2007\)](#). This in turn affirmed that the hybrid
2 process can be more economical, producing fresh water with a lower specific cost per cubic meter.
3 In the same context, [Rensonnet et al. \(2007\)](#) showed that the full hybridization of MED and RO is
4 the most economical option if the electricity cost is high, otherwise the standalone RO process can
5 be more convenient. [Mahbub et al. \(2009\)](#) proposed a detailed thermodynamic analysis of a
6 combined cycle power (CCP) plant with MSF, MED and RO (standalone), or with hybrid MSF+RO
7 and MED+RO. It is concluded that the specific energy consumption can be reduced by 17% with
8 the CCP+MED+RO system, compared to CCP+MSF+RO system. Furthermore, the lowest cost of
9 fresh water produced with the CCP+MED+RO option of about 1.09 \$/m³.

10 The techno-economic performance of an integrated system of concentrating solar plant (CSP) with
11 MED and Ultrafiltration (UF) is investigated by [Olwig et al. \(2012\)](#). The results showed the
12 necessity of the RO process to improve the economics of the integrated process compared to a
13 simple CSP+MED configuration. Specifically, a cost of fresh water of 1 \$/m³ was estimated based
14 on 0.24 \$/kWh as the electricity cost of CSP. [Manesh et al. \(2013\)](#) studied the optimal integration of
15 site utility and MED+RO desalination plant based on a simultaneous exergetic and economic
16 optimisation. Also, [Weiner et al. \(2015\)](#) modelled and optimised a hybrid MED+RO system. This
17 confirmed that the MED+RO hybrid system can be more energy efficient than a standalone MED
18 process and with a recovery ratio superior to a standalone RO process. Recently, a comprehensive
19 mathematical model is developed by [Sadri et al. \(2017\)](#) to describe the MED_TVC+RO integrated
20 system. Moreover, the performance of integrated process is maximised by using a Genetic
21 Algorithm (GA) technique.

22 The net outcome of the above literature review already showed that much attention been paid on the
23 integration of power and desalination technologies and consequent energetic and/or economic
24 assessment of the process. However, up to the authors' knowledge, the implementation of an
25 integrated hybrid system of MED_TVC process coupled with RO process, has not yet been fully
26 investigated. Also, it has been noticed that a parametric sensitivity analysis of several operating

1 conditions using a hybrid system of MED_TVC+RO processes has not yet been explored.
2 Therefore, the aim of this paper was to propose and evaluate different configurations in the context
3 of simple and full hybridization of MED_TVC+RO processes. Also, the integrated process
4 performance and sensitivity analysis to be explored via modeling and simulation. To systematically
5 conduct this aim, detailed mathematical models of both MED_TVC and RO processes are initially
6 developed. The mathematical models have been used to predict the performance of both
7 MED_TVC and RO processes with a minimum amount of assumptions and limitations, which is
8 rarely in other literature studies. This results in accurate models also the one developed for the
9 hybrid process. Occasionally, most studies neglected the TVC section, which can be important to
10 increase the performance ratio of the thermal process. The models developed of MED_TVC and
11 RO process are individually validated against the predictions of several previous models of MED
12 and the projected data collected from Toray Design System 2.0 (TDS2) for RO, respectively. Then,
13 five different configurations have been designed to explore the best one in terms of productivity,
14 fresh water quality, energy efficiency and recovery ratio of the whole hybrid process. A parametric
15 sensitivity analysis with respect to seawater conditions and steam available from the power plant
16 has been carried out in four of the proposed configurations. The output variables under
17 investigation are the fresh water productivity, fresh water purity, energy consumption, and recovery
18 ratio of the hybrid plant.

19

20 **2. Description of the process**

21 The description of both MED_TVC and RO processes is provided in [Sections S.F.1](#) and [S.F.2](#) in the
22 supplementary file, respectively. In this respect, the schematic diagrams of forward feed multiple
23 effect desalination process with thermal vapor compression and an industrial full-scale seawater RO
24 desalination plant are given in [Figs. S.F.1](#) and [S.F.2](#) in the supplementary file, respectively.

1 **Table 1** presents the technical specification and operating conditions of the MED and RO
 2 membrane processes. This also includes the permissible bounds of operating conditions of the
 3 membrane. The next section illustrates the description of the hybrid system of MED_TVC+RO.

4 **Table 1.** Specification and operating conditions of the MED and RO membrane processes

Operative parameter	Value	Unit
Number of effects	10	-
External steam flowrate	5.67	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
<hr/>		
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 width (W)	37.2	m
Module length (L)	1	m
$A_w(T_o)$ (m/ atm s) at 25 °C *	3.1591×10^{-7}	m/s atm
$B_s(T_o)$ NaCl (m/s) at 25 °C *	1.74934×10^{-8}	m/s
Spacer type	Naltex-129	-
Feed spacer thickness (t_f)	8.6×10^{-4} (34 mils)	m
Hydraulic diameter of the feed spacer channel d_h	8.126×10^{-4}	m
Length of spacer in the spacer mesh	2.77×10^{-3}	m
A' (dimensionless)	7.38	-
n (dimensionless)	0.34	-
ε (dimensionless)	0.9058	-

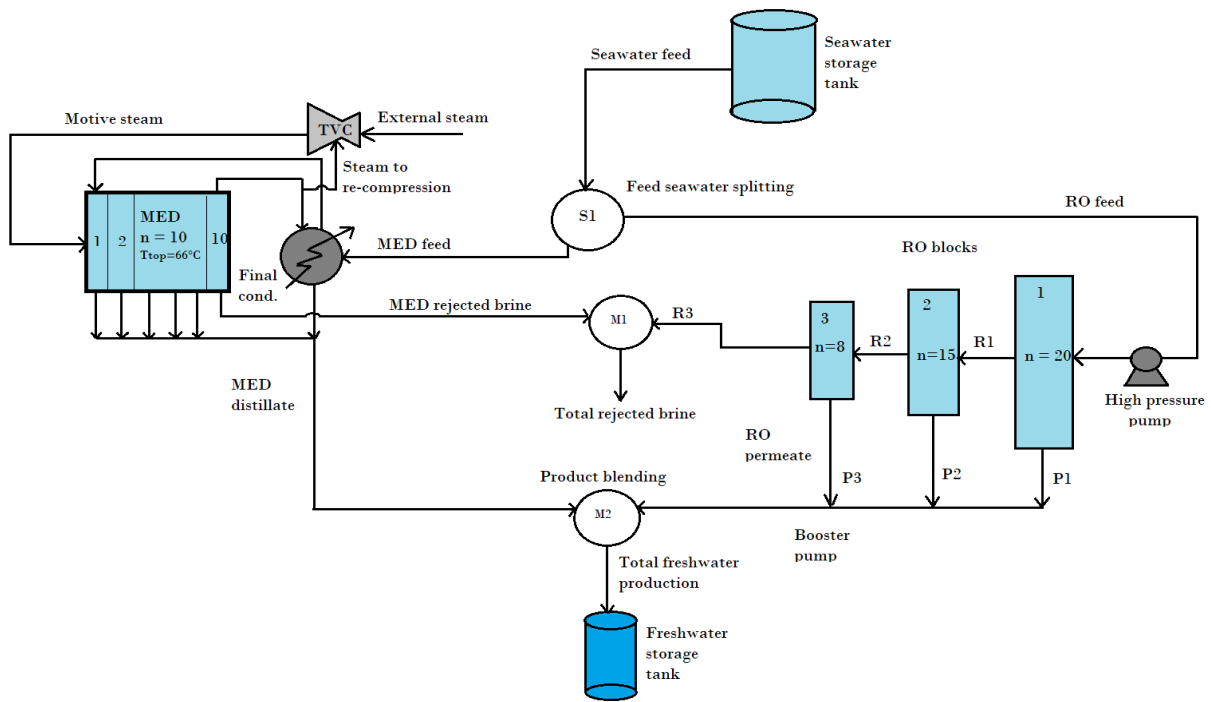
5 *: Estimated using parameter estimation in [Section 3.2.2](#)

6

7 **2. Description of the Hybrid MED_TVC+RO process**

8 **Figs. 1** to **4** show the proposed configurations of the hybrid MED_TVC+RO process under
 9 investigation. In each configuration, the permeate of the RO membrane process is blended with the
 10 product of the thermal process, which is a distillate with a salinity close to zero. However, a value
 11 of 10 ppm is assumed for the salinity of the distillate to account a few seawater droplets that can be
 12 entrained in the vapor phase beyond the demisters. According to the World Health Organization

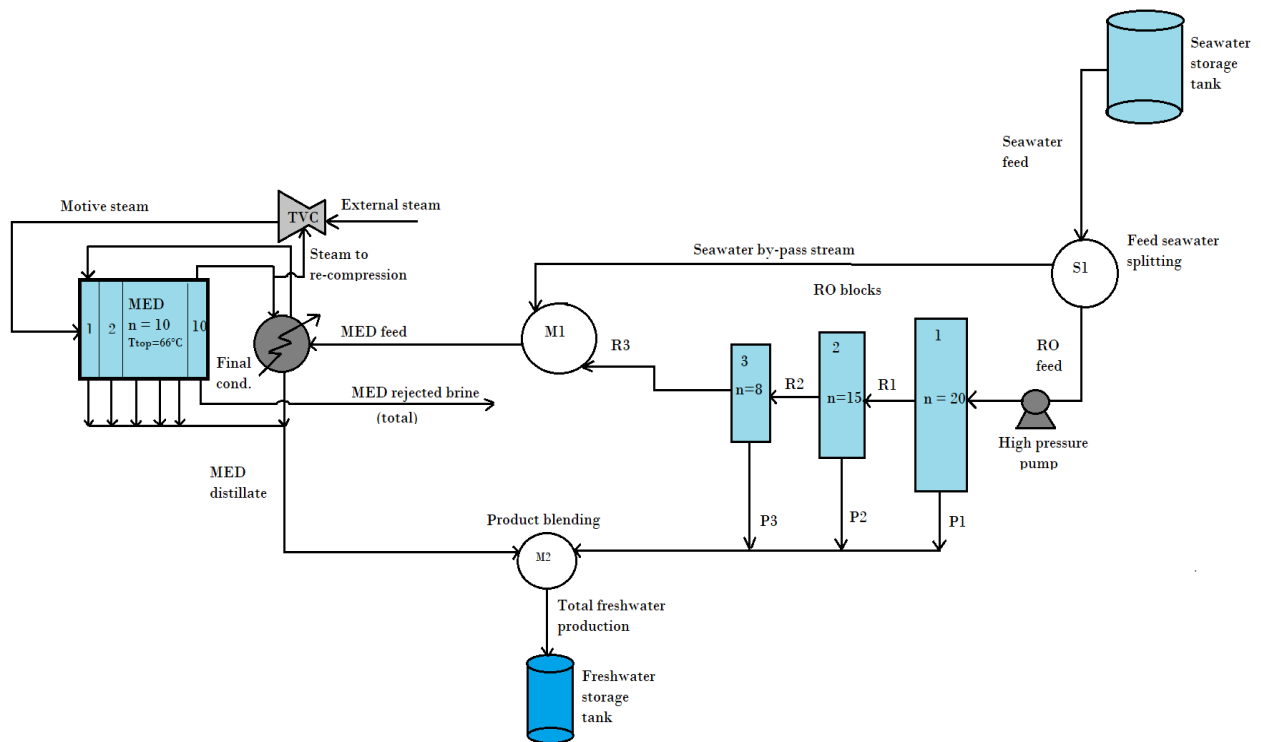
1 (WHO), the salinity of a good quality drinking water should be below 300 ppm, and precisely
 2 below 200 ppm for the most tap water (WHO, 2011). Therefore, the MED_TVC process has been
 3 designed to have a capacity approximately 4 times bigger than the RO process to produce enough
 4 distillate for the blending and commensurate with a salinity of the final product below 200 ppm. .
 5 Fig. 1 shows the so-called simple hybridization of the thermal and pressure driven desalination
 6 processes. The seawater feed is split between the two processes, which operates unconnectedly. In
 7 other words, the operating conditions of one process have no effect on the other one, since the
 8 connection is only at the level of final products (fresh water) and rejected brine streams.



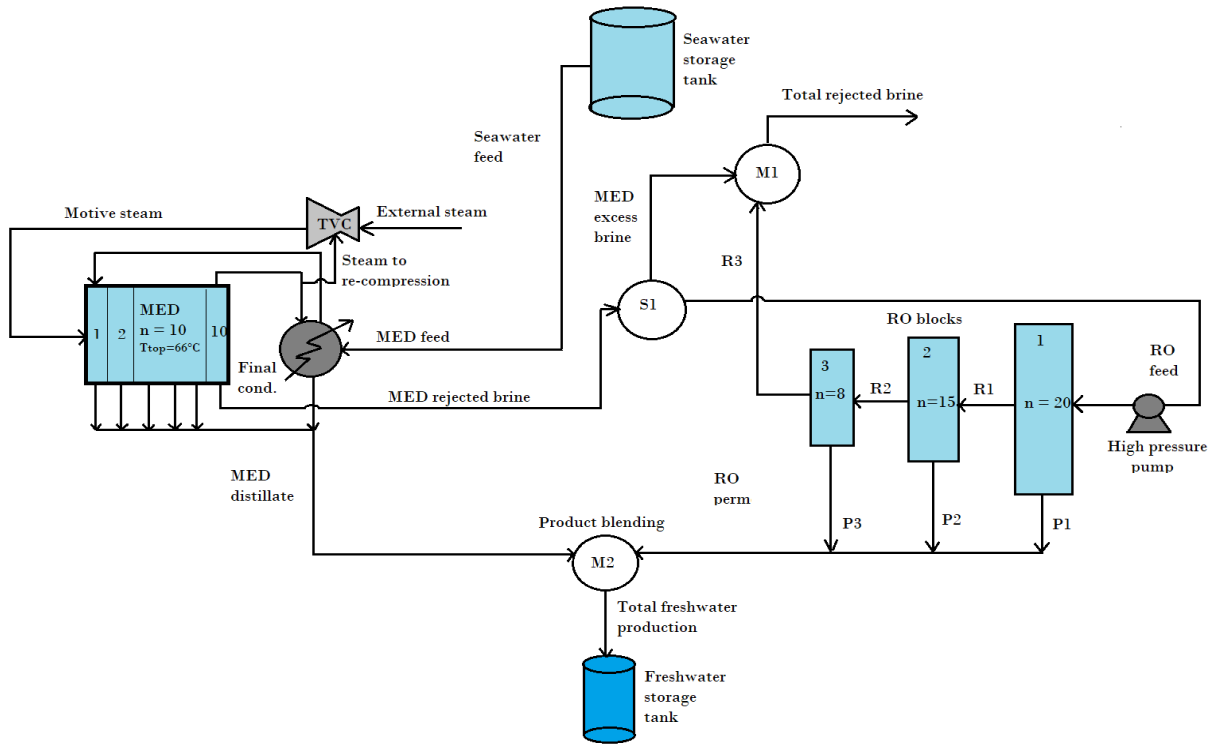
9
 10 **Fig. 1.** Schematic diagram of the simple MED_TVC+RO hybridization.
 11

12 Fig. 2 shows the full hybridization when the membrane process is placed upstream. This design has
 13 considered that the seawater feed is partially fed to the RO process and the rest is mixed with the
 14 retentate to form the inlet stream of the MED_TVC process. The option of blending a by-pass
 15 stream with the retentate is to accommodate the operating flow rate of MED_TVC process, which
 16 works at a greater capacity. Moreover, this option would reduce the feed salinity of the MED_TVC

1 process. In this configuration, the rejected brine is made up only of the brine from the thermal
 2 process.

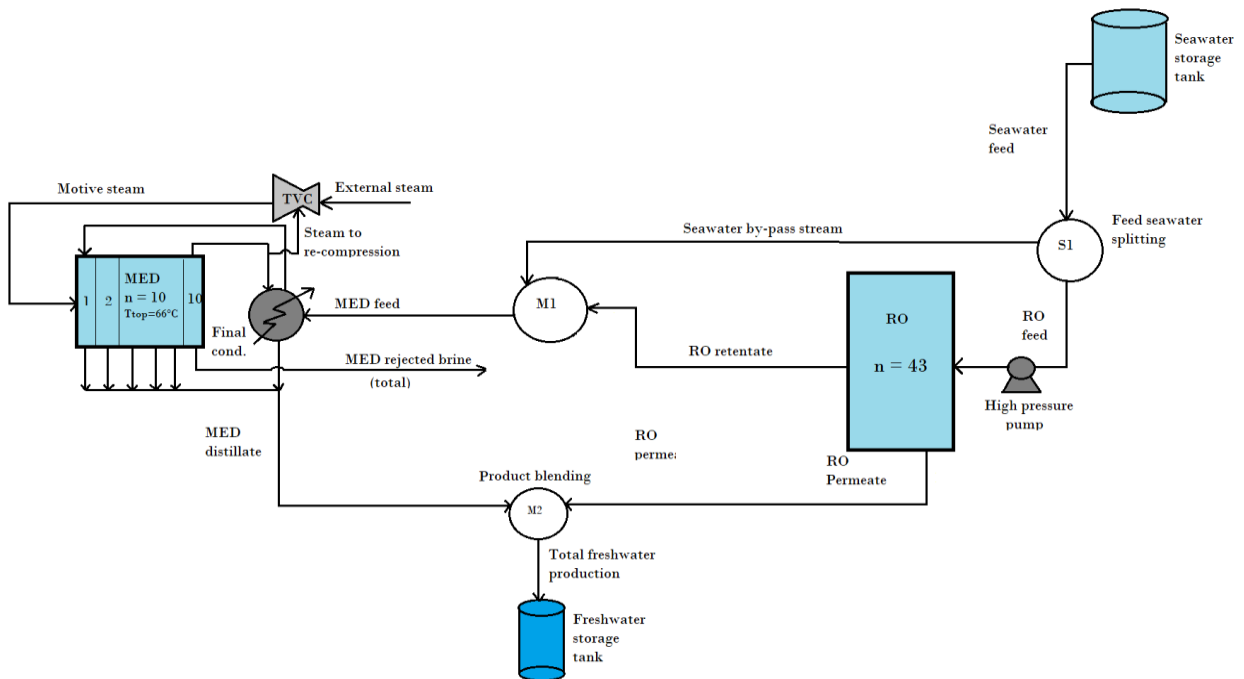


3
 4 **Fig. 2.** Schematic diagram of full MED_TVC+RO system. RO process is upstream with respect to the MED process.
 5
 6 Another option for a full hybridization with the thermal MED_TVC process placed upstream is
 7 given in Fig. 3. In this configuration, the membrane process is fed with the rejected brine of the
 8 thermal process, which has a temperature of 40 °C and a salinity limited to 50 kg/m³ to avoid a very
 9 fast membrane deterioration. As a result, the MED_TVC process is forced to operate in a small
 10 salinity window. The remaining brine of MED is blended with the RO retentate and rejected.
 11 Finally, Fig. 4 shows the coupling of the MED process with a simple RO process of a single block
 12 comprises a total number of 43 of pressure vessels, where each pressure vessel includes 8 elements
 13 in series. Note that the total number of 43 pressure vessels has been already considered for other
 14 configurations, as well as the total seawater flow rate entering the membrane process.



1
2 **Fig. 3.** Schematic diagram of full MED_TVC+RO system. MED process is upstream with respect to the RO process.

3



4
5 **Fig. 4.** Schematic diagram of full MED_TVC+RO system. RO process is upstream and made of a single block.

6

7

8 **3. Model of the hybrid MED-RO process**

1 The development of an accurate and detailed mathematical model is an importance target to express
2 the essential phenomena of any industrial process, which enables to generate accurate results via
3 simulation. In the next sections, the description of the models developed for the MED_TVC and
4 RO processes is represented and followed by a validation study for each individual process.

5

6 **3.1 MED_TVC process**

7 The following model of MED process is adapted from [Darwish et al. \(2006\)](#). Interestingly, some
8 modifications are made with respect to the original model. Specifically, detailed thermodynamic
9 correlations are used to evaluate all the relevant thermodynamic properties of the system as a
10 function of temperature, salinity, and fouling Note that all these characteristics were assumed as
11 constants in the original work by [Darwish et al. \(2006\)](#). To accommodate the industrial reality, the
12 equal exchange area of all the effects is imposed by means of a procedure for de-linearizing the
13 temperature profiles. This new technique devised by the authors shows a very fast convergence,
14 being able to approximately equalize area in a single iteration. The model for the thermal vapor
15 compression section (TVC) is adapted from [Dessouky et al. \(2002\)](#) and given in [Table A.1](#) in
16 [Appendix A](#).

17

18 **3.1.1 Assumptions**

- 19 1. Steady state process.
- 20 2. The vapour phase is salt free.
- 21 3. Energy loss to the surroundings is negligible.
- 22 4. Equal transfer area in all the effects.
- 23 5. Non-equilibrium allowance (NEA) and pressure drops are neglected.
- 24 6. Boiling point elevation and specific heat are considered as a function of temperature and
25 salinity.

7. Latent heat of evaporation and overall exchange coefficient are considered as a function of temperature. For the heat exchange, experimental correlations that consider fouling are implemented.

8. Steam from the external utility is provided saturated and leaves as saturated liquid.

3.1.2 Model equations

The model is made of a series of material and energy balances together with the thermodynamics correlations, which are provided in the [Appendix A](#). Steam flow rate M_s (kg/s) and steam temperature T_s (°C) are assumed to be known, since generated from an upstream process (i.e. a co-generation power plant or a renewable energy facility), while fresh water production is evaluated. The feed flow rate M_f , the total distillate flow rate M_d , and the rejected brine flow rate M_b are evaluated according to simple overall material balances. Moreover, the salinity of the feed x_f (kg/m³) and of the rejected brine x_b are known.

$$M_d = M_f \frac{x_b - x_f}{x_b}$$

(1)

$$M_b = M_f - M_d$$

(2)

The sensible power $Q_{sensible}$ (kW) is used to heat the feed from the feed temperature after pre-heating tI , up to the boiling temperature in the first effect $T1$. The latent power Q_{latent} is used for vaporizing a quantity of distillate equal to DI , where $\lambda(T_s)$ is the latent heat of vaporization (kJ/kg) at steam temperature T_s .

$$M_f = \frac{M_s \lambda(T_s)}{Q_{sensible} + Q_{latent}}$$

(3)

$$Q_{sensible} = M_f \int_{t1}^{T1} c_p(T1, x1) dT$$

(4)

$$1 \quad Q_{latent} = D1 \lambda(Tv1)$$

$$2 \quad (5)$$

3 Linear temperature profiles can be defined as a *first attempt* by imposing an equal temperature drop
 4 (ΔT) among the effects and an equal temperature increase (Δt) among the feed pre-heaters, where
 5 Tb is the temperature of the rejected brine, equal to the temperature in the last effect. n is the
 6 number of effects.

$$7 \quad \Delta T = \frac{T1-Tb}{n-1} \quad \text{or} \quad \Delta T = \frac{Ts-Tb}{n} \quad (6)$$

$$8 \quad \Delta T = \Delta t$$

$$9 \quad (7)$$

10 The feed temperature in the first effect ($t1$), after $n-1$ pre-heaters, can be evaluated starting from the
 11 temperature tn at the exit of the final condenser, which is assumed to be 11°C higher than seawater
 12 temperature. The temperature of the vapor phase Tv is lower than the brine temperature by the
 13 Boiling Point Elevation (BPE).

$$14 \quad t1 = tn + (n - 1) \Delta t$$

$$15 \quad (8)$$

$$16 \quad Tv = T - BPE(T, x)$$

$$17 \quad (9)$$

18 A small fraction of brine rejected by each effect ($D_{flash,i}$) is flashed to a pre-heater for heating the
 19 feed stream. α is defined as the fraction of brine rejected by effect $i-1$ (B_{i-1}) that is flashed in the
 20 associated pre-heater, evaluated at mean temperature and salinity of the plant.

$$21 \quad D_{flash,i} = \alpha B_{i-1}$$

$$22 \quad (10)$$

$$23 \quad \alpha = \frac{cp(T_{mean}, x_{mean}) \Delta T}{\lambda(T_{mean})}$$

$$24 \quad (11)$$

1 Where $T_{mean} = \frac{T1+Tb}{2}$, $x_{mean} = \frac{xf+xb}{2}$ (12,

2 13)

3 The fraction of the total distillate produced by evaporation in each effect will be denoted as β . This
 4 value can be evaluated as a function of known parameters (number of stages, initial salinity, final
 5 salinity, α) by rearranging the material balances as follows;

$$D1 = D_{flash1} + D_{boil1} = \alpha Mf + \beta Md$$

$$6 \quad B1 = Mf - D1 = (1 - \alpha)Mf - \beta Md$$

$$D2 = D_{flash2} + D_{boil2} = \alpha B1 + \beta Md$$

$$B2 = B1 - D2 = (1 - \alpha)B1 - \beta Md$$

$$7 \quad B2 = (1 - \alpha)[Mf(1 - \alpha) - \beta Md] - \beta Md$$

$$B2 = (1 - \alpha)^2 Mf - \frac{\beta Md}{\alpha} [1 - (1 - \alpha)^2]$$

8 Similarly, the brine rejected stream of the last effect can be evaluated with the following equation:

$$9 \quad Bn = Mb = (1 - \alpha)^n Mf - \frac{\beta Md}{\alpha} [1 - (1 - \alpha)^n]$$

10 (14)

11 Substituting Eq. (1) and (2) in Eq. (14), yields:

$$12 \quad \frac{xb - xf}{xb} - 1 = \frac{xb - xf}{xb} (1 - \alpha)^n - \frac{\beta}{\alpha} [1 - (1 - \alpha)^n]$$

13 (15)

14 Eq. (15) can be re-arranged to explicit the parameter β :

$$15 \quad \beta = \frac{\alpha [xb(1 - \alpha)^n - xf]}{(xb - xf)[1 - (1 - \alpha)^n]}$$

16 (16)

17 Accordingly, the amount of distillate boiled in each effect $D_{boiled,i}$, the total distillate (Md), and the
 18 brine flow rates B_i can now be evaluated, as well as the salinity profile.

$$19 \quad D_{boiled,i} = \beta Md \quad (17)$$

$$20 \quad D_i = D_{boiled,i} + D_{flash,i} \quad (18)$$

$$1 \quad B_i = B_{i-1} - D_i$$

$$2 \quad (19)$$

$$3 \quad x_i = \frac{x_{i-1}B_{i-1}}{B_i}$$

$$4 \quad (20)$$

5 The thermal loads in every effect Q_i (kW) and exchange areas of evaporators $A_{ev,i}$ (m²) and pre-
6 heaters $A_{ph,i}$ can be estimated using a simple energy balance, where U_{ev} is the overall heat exchange
7 coefficient.

$$8 \quad Q_i = U_{ev,i}A_{ev,i}\Delta T_{ev,i}$$

$$9 \quad (21)$$

$$10 \quad Q_i = D_{boiled,i-1} \lambda(T_{v,i-1})$$

$$11 \quad (22)$$

$$12 \quad \Delta T_{ev,i} = T_{v,i-1} - T_i = T_{i-1} - BPE_{i-1} - T_i = \Delta T - BPE_{i-1}$$

$$13 \quad (23)$$

14 In the first effect, the thermal load Q_s is directly provided by the external steam;

$$15 \quad Q_s = Ms \cdot \lambda(T_s) = A_{ev,1}U_{ev,1}(T_s - T_1)$$

$$16 \quad (24)$$

17 In the feed pre-heaters, heat exchange is between the flashed distillate at temperature T_{v_i} and
18 liquid feed stream at a temperature t_i .

$$19 \quad Mf \cdot \int_{t_{i+1}}^{t_i} cp(t, xf) dt = U_{ph,i}A_{ph,i}\Delta t_{\log,i}$$

$$20 \quad (25) \quad \Delta t_{\log,i} = \frac{\Delta t}{\log\left(\frac{T_{v_i} - t_{i+1}}{T_{v_i} - t_i}\right)}$$

21 (26) Since the exchange areas are evaluated using linear temperature profiles presented
22 in Eq. (6), it is impossible to guarantee the fulfilment of Assumption 5, where equal area in all the

1 effects is assumed ([Assumption 4](#)). Therefore, temperature profiles can be de-linearized according
 2 to the following procedure devised by the authors to achieve a fast equalization of exchange areas.
 3 First, mean area of evaporators is evaluated using [Eq. \(27\)](#). [Eq. \(28\)](#) is solved by modifying the
 4 value of the vectors $\Delta T_{ex,i}$. Finally, [Eq. \(29\)](#) is solved to evaluate the vector ΔT_i which can be used
 5 to calculate the new non-linear temperature profiles.

$$6 \quad A_{ev,mean} = \frac{\sum_{i=1}^n A_{ev,i}}{n} \quad (27)$$

$$7 \quad A_{ev,mean} - \frac{Q_i}{U_{ev,i} \Delta T_{ex,i}} = 0 \quad (28)$$

$$8 \quad \Delta T_i = \Delta T_{ex,i} - BPE_i \quad (29)$$

$$9 \quad T_i = T_{i-1} - \Delta T_i \quad (30)$$

$$10 \quad T v_i = T_i - BPE(T_i, X_i) \quad (31)$$

11 All the process variables are then re-evaluated considering the new temperature profiles. The
 12 equality of areas is checked according to [Eq. \(32\)](#).

$$13 \quad \Delta A_{ev} \% = \frac{\max(A_{ev}(2:10)) - \min(A_{ev}(2:10))}{A_{ev,mean}} \cdot 100\%$$

14 (32)

15 This procedure has been proved as an effective method to quickly equalizing the areas. In this
 16 respect, [Table 2](#) shows the percentage error drops from 13.28 % to 0.76 % for the evaporator areas
 17 after a single iteration. However, the first effect is exempted, since it receives a different thermal
 18 load being the temperature difference between steam and brine in the first effect fixed at 4°C.

19 **Table 2.** Exchange areas in evaporators. Subscript *old* means before the equalizing procedure.

20 Parameters for simulation are set according to [Table 1](#).

Effect number	$A_{ev,old}$ [m ²]	A_{ev} [m ²]
Effect 1	1893.9834	1893.9834
Effect 2	2229.3867	2302.6501
Effect 3	2256.7415	2301.7188
Effect 4	2285.8877	2300.9355
Effect 5	2317.0193	2300.3308
Effect 6	2350.3570	2299.9404
Effect 7	2386.1506	2299.8086

Effect 8	2424.6880	2299.9875
Effect 9	2466.3015	2300.5405
Effect 10	2511.3794	2301.5420
Error %	13.28 %	0.76%

1

2 After the equalizing procedure, it is possible to proceed with the thermal vapor compression (TVC)
3 section modeling. All the equations are summarized in [Table A.1](#) in the [Appendix A](#).

4 The last part of the process to be modelled is the final condenser, which receives a vapor flow rate
5 (M_{COND}) to be condensed equal to the distillate from the last effect (D_n) minus the vapor fraction
6 entrained in the TVC section (M_{TVC}).

$$7 \quad M_{TVC} = Ms - Mm$$

8 (40)

$$9 \quad M_{COND} = D_n - M_{TVC} \quad (41)$$

10 In the final condenser the seawater flow rate is heated up to a fixed temperature, exchanging the
11 latent heat Q_{COND} provided by the condensation of steam. The unit can be modelled like a bigger
12 pre-heater. [Eq. \(42\)](#) and [\(43\)](#) are used to evaluate the area of the final condenser A_{COND} and the total
13 seawater flow rate M_w at temperature T_w , which is required by the MED. Indeed, the required flow
14 rate is important to know, especially when it is provided by the RO process placed upstream, to
15 design properly the by-pass stream.

$$16 \quad Q_{COND} = U_{COND} A_{COND} \Delta T_{log,COND} \quad (42)$$

$$17 \quad Q_{COND} = M_{COND} \lambda(Tv_n)$$

18 (43)

$$19 \quad Q_{COND} = M_w \int_{T_w}^{tn} cp(T, xf) dT$$

20 (45)

$$21 \quad \Delta T_{log,COND} = \frac{tn - T_w}{\log\left(\frac{T_{V_n} - T_w}{T_{V_n} - tn}\right)}$$

22 (44)

23

1 **3.1.3 Validation of MED process**

2 The accuracy of any developed model should be tested before implementing the model in any
 3 parametric sensitivity analysis. Thus, the model developed in [Section 3.1](#) of MED_TVC must be
 4 first computed and validated with the results of those from the literature. Specifically, the model
 5 validation has been carried out in terms of Gained Output Ratio (GOR) by comparing the prediction
 6 of the model developed against the prediction of other consolidated literature models, namely
 7 [Dessouky et al. detailed \(1998\)](#), [El-Sayed et al. \(2001\)](#), [Dessouky et al. simplified \(2002\)](#), [Darwish
 8 et al. \(2006\)](#), and [Mistry et al. \(2012\)](#). The GOR is defined as the quantity of distilled fresh water
 9 (Md) produced by the process over the quantity of steam utilized (Ms) as an external utility in the
 10 first effect. More importantly, the model validation is carried out in the feasible range of 60 – 80 °C
 11 of steam temperature. The reason behind this is that running the MED process at low temperatures
 12 would require high exchange area, while a significant drop in the process performance is occurred
 13 at elevated steam temperatures ([Dessouky et al., 2002](#)). [Table 3](#) shows that the prediction of the
 14 current model is closer to the one of an adaptive model of [Mistry et al. \(2012\)](#). Having said this, an
 15 acceptable convergence is noticed after comparing the recent model against [El- Sayed et al. \(2001\)](#)
 16 and [Darwish et al. \(2006\)](#) models. However, significant discrepancies are revealed after comparing
 17 the recent model against the models of [Dessouky et al. detailed \(1998\)](#), and [Dessouky et al.
 18 simplified \(2002\)](#). This can be ascribed to severe thermodynamics assumptions were made to
 19 develop the latterly models. Consequently, it is fair to say that the recent model developed is
 20 accurately able to predict the performance of MED due to low deviations of only 1.13-1.85%
 21 compared to the latest literature model. However, it is important to mention that this comparison
 22 has referred to the MED process without TVC. Therefore, the TVC section has been deactivated.

23

24 **Table 3.** Comparison of the present model with respect to literature model regarding GOR, for different steam
 25 temperatures, in the range of feasible values for low-temperature MED process.
 26 Parameters for simulation: $n=8$, $T_n=40^\circ\text{C}$, $T_w=25^\circ\text{C}$, $x_f=42000$ ppm, $x_n=70000$ ppm.

Gained Output Ratio (GOR)										
	Present	Dessouky et		El-Sayed		Dessouky et		Darwish		Mistry

ure (°C)

	model	al. (1998) detailed	% error	et al. (2001)	% error	al. (2002) simple	% error	et al. (2006)	% error	et al. (2012)	% error
60	7.06	6.33	10.34	6.72	4.82	7.90	-11.90	7.44	-5.38	6.98	1.13
62	7.001	6.21	11.33	6.68	4.53	7.88	-12.56	7.33	-4.73	6.917	1.20
64	6.942	6.09	12.33	6.65	4.25	7.86	-13.22	7.22	-4.07	6.854	1.27
66	6.883	5.96	13.35	6.61	3.95	7.84	-13.90	7.12	-3.39	6.791	1.34
68	6.824	5.84	14.39	6.57	3.66	7.82	-14.60	7.01	-2.71	6.728	1.41
70	6.765	5.72	15.45	6.54	3.36	7.80	-15.30	6.90	-2.01	6.665	1.48
72	6.706	5.60	16.52	6.50	3.05	7.78	-16.02	6.79	-1.30	6.602	1.55
74	6.647	5.48	17.62	6.47	2.74	7.76	-16.74	6.69	-0.58	6.539	1.62
76	6.588	5.35	18.73	6.43	2.42	7.74	-17.49	6.58	0.16	6.476	1.70
78	6.529	5.23	19.87	6.39	2.09	7.72	-18.24	6.47	0.91	6.413	1.78
80	6.47	5.11	21.02	6.35	1.82	7.70	-19.01	6.36	1.64	6.35	1.85

1

2 3.2 RO process

3 The model developed in this paper for an individual spiral wound RO process is based on the model
4 of [Abbas \(2005\)](#) that originally based on the principles of the solution diffusion model suggested by
5 [Lonsdale et al. \(1965\)](#) to express the transport phenomena of water and solute through the
6 membrane. The model developed is formerly considered the following assumptions:

- 7 1. The membrane characteristics and the channel geometries are assumed constant.
- 8 2. The film theory model is used to express the concentration polarisation.
- 9 3. Constant pressure of 1 atm at the permeate channel.
- 10 4. Isothermal process.

- 11 1. The correlation of [Da Costa et al. \(1994\)](#) is used to elucidate the pressure drop in the
12 membrane feed channel.

13 Interestingly, several modifications are made on the model of [Abbas \(2005\)](#) as follows:

- 14 • Considering the impact of operating temperature on the membrane transport parameters
15 using the proposed correlations of Toray membrane;
- 16 • The permeate concentration is estimated based on [Al-Obaidi et al. \(2017b\)](#), which is
17 developed to consider solute transport parameter;
- 18 • The variation of physical properties against feed concentration and temperature is
19 considered based on the developed correlations of [Koroneos \(2007\)](#) compared to constant
20 physical properties assumed by [Abbas \(2005\)](#).

21

1 3.2.1 Model equations

2 The water Q_p (m³/s) and solute Q_s (kg/m² s) fluxes through the membrane are calculated as

$$3 \quad Q_p = A_{w(T)} \left(P_f - \frac{\Delta P_{drop,E}}{2} - P_p - \pi_w - \pi_p \right) A_m$$

4 (46)

$$5 \quad Q_s = B_{s(T)} (C_w - C_p)$$

6 (47)

7 $A_{w(T)}$, P_f , $\Delta P_{drop,E}$, P_p , π_w , π_p , A_m , $B_{s(T)}$, C_w , C_p are water permeability constant at operating
8 temperature (m/s atm), feed pressure (atm), pressure drop along the membrane element (atm),
9 permeate pressure (atm), osmotic pressure at the membrane surface and permeate channel (atm),
10 effective membrane area (m²), solute transport parameter at operating temperature (m/s), membrane
11 wall concentration (kg/m³), and permeate concentrations (kg/m³), respectively. The osmotic
12 pressure is calculated as ([Abbas, 2005](#))

$$13 \quad \pi_w = 0.76881 C_w$$

14 (48)

$$15 \quad \pi_p = 0.7994 C_p$$

16 (49)

17 The impact of temperature T (°C) on transport parameters is calculated based on the transport
18 parameters of water and solute at the reference temperature (Toray membrane)

$$19 \quad A_{w(T)} = A_{w(25\text{ }^\circ\text{C})} \exp[0.0343 (T - 25)] \quad < 25\text{ }^\circ\text{C}$$

20 (50)

$$21 \quad A_{w(T)} = A_{w(25\text{ }^\circ\text{C})} \exp[0.0307 (T - 25)] \quad > 25\text{ }^\circ\text{C}$$

22 (51)

$$23 \quad B_{s(T)} = B_{s(25\text{ }^\circ\text{C})} (1 + 0.08 (T - 25)) \quad < 25\text{ }^\circ\text{C}$$

24 (52)

$$1 \quad B_{s(T)} = B_{s(25\text{ }^\circ\text{C})} (1 + 0.05 (T - 25)) \quad > 25\text{ }^\circ\text{C}$$

$$2 \quad (53)$$

3 The pressure drop $\Delta P_{drop,E}$ (atm) per element is calculated as proposed by [Da Costa et al. \(1994\)](#),

$$4 \quad \Delta P_{drop,E} = \frac{9.8692 \times 10^{-6} A' \rho_b Q_b^2 L}{2 d_h Re_b^n (W t_f \epsilon)^2}$$

$$5 \quad (54)$$

6 A' , ρ_b , Q_b , L , d_h , Re_b , n , W , t_f and ϵ are the feed spacer characteristic (-), bulk density (kg/m³),
7 bulk flow rate (m³/s), membrane length (m), the hydraulic diameter of the feed spacer channel (m),
8 Reynolds number (-), feed spacer characteristic (-), membrane width (m), feed channel height (m),
9 and the membrane porosity (-), respectively.

$$10 \quad Re_b = \frac{\rho_b d_h Q_b}{t_f W \mu_b}$$

$$11 \quad (55)$$

$$12 \quad Q_b = \frac{Q_f + Q_r}{2} \quad (56)$$

13 μ_b , Q_f , Q_r are kinematic viscosity (kg/m s), feed and retentate flow rates (m³/s), respectively. The
14 bulk concentration C_b (kg/m³) is the average of feed C_f (kg/m³) and retentate C_r (kg/m³)
15 concentrations as can be shown in [Eq. \(57\)](#)

$$16 \quad C_b = \frac{C_f + C_r}{2}$$

$$17 \quad (57)$$

18 The membrane surface concentration C_w (kg/m³) is expressed by the film theory model developed
19 by [Michaels, 1968](#) which is corresponding to the mass transfer coefficient k (m/s) ([Da Costa et al.,](#)
20 [1994](#))

$$21 \quad \frac{(C_w - C_p)}{(C_b - C_p)} = \exp\left(\frac{Q_p/A_m}{k}\right)$$

$$22 \quad (58)$$

$$k = 0.664 k_{dc} Re_b^{0.5} Sc^{0.33} \left(\frac{D_b}{d_h}\right) \left(\frac{2d_h}{L_f}\right)^{0.5}$$

(59)

$$Sc = \frac{\mu_b}{\rho_b D_b}$$

(60)

k_{dc} , Sc , D_b , L_f are constant (-), Schmidt number (-), diffusivity parameter (m²/s), and length of filament in the spacer mesh (m), respectively. The physical properties of seawater are calculated based on [Koroneos \(2007\)](#).

$$\rho_b = 498.4 m_f + \sqrt{[248400 m_f^2 + 752.4 m_f C_b]}$$

(61)

$$m_f = 1.0069 - 2.757 \times 10^{-4} T \quad (62)$$

$$D_b = 6.725 \times 10^{-6} \exp\left\{0.1546 \times 10^{-3} C_b - \frac{2513}{T+273.15}\right\} \quad (63)$$

$$\mu_b = 1.234 \times 10^{-6} \exp\left\{0.0212 C_b + \frac{1965}{T+273.15}\right\} \quad (64)$$

The total mass and solute balance of the whole unit gives

$$Q_f = Q_r + Q_p \quad (65)$$

$$Q_f C_f - Q_r C_r = Q_p C_p \quad (66)$$

The permeate concentration C_p (kg/m³) is estimated by the correlation of [Al-Obaidi et al. \(2017b\)](#)

$$C_p = \frac{B_s C_f e^{\frac{J_w}{k}}}{J_w + B_s e^{\frac{J_w}{k}}} \quad (67)$$

J_w denotes the water flux through the membrane (m/s). The overall solute rejection and recovery rate are

$$Rej = \frac{C_f - C_p}{C_f} \quad (68)$$

$$Rec = \frac{Q_p}{Q_f} \quad (69)$$

1 The above completed simulation model of an individual spiral wound RO process is used to build
2 the full modelling package of the proposed configurations of multi-stage RO process including
3 retentate reprocessing design of [Fig. S.F.2](#) (given in the supplementary file). [Table A.2](#) show the
4 simulation model of the proposed configurations of multi-stage RO process, including the overall
5 plant performance of solute rejection and total recovery and the interconnected streams of three
6 blocks for retentate reprocessing design. Moreover, the model encompasses the calculation of
7 product concentration, retentate concentration, and overall energy consumption. Finally, the model
8 code is written and solved using gPROMS model builder software ([general Process Modelling
9 System by Process System Enterprise Ltd., 2001](#)). The gPROMS environment can be used as a
10 modelling platform for the steady state and dynamic simulation, optimisation, experiment design
11 and parameter estimation.

12

13 **3.2.2 Estimation of unknown model parameters**

14 The RO model developed in [Section 3.2.1](#) contains two unknown transport parameters of water and
15 NaCl permeability constants at 25 °C ($A_{w(25\text{ }^{\circ}\text{C})}$, $B_{s(25\text{ }^{\circ}\text{C})}$) that will be used with the known
16 parameters to solve the model equations. The gEST parameter estimation tool of gPROMS is used
17 to investigate these parameters based on the projected data from the Toray Design System 2.0
18 (TDS2) that is a commercial projection software provided by the membrane manufacturer, i.e.,
19 Toray. Therefore, a set of projected data is gathered from TDS2 for a single pressure vessel holds
20 eight membranes type TM820M-400/ SWRO (Toray) connected in series at several operating
21 conditions. The estimated transport parameters are given in [Table 1](#).

22

23 **3.2.3 Validation of RO process**

24 [Table 4](#) shows the consistency between the model predictions of several operating parameters
25 against the projected data of TDS2 at relatively small errors in the most parameters. Upon
26 investigation of the validity of RO process model, it is fair to say that this model is valid enough to

- 1 be augmented with the model of MED_TVC to represent the modelling of the hybrid process of
- 2 MED_TVC+RO.
- 3

Table 4. RO model validation against TDS2 data

No.	Parameter	EXP	Model	Error%	No.	Parameter	EXP	Model	Error%	No.	Parameter	EXP	Model	Error%
1	$Q_f/0.0197$	$C_f/35$	$T/25$	$P_f/55.91$	5	$Q_f/0.0099$	$C_f/35$	$T/25$	$P_f/50.35$	9	$Q_f/0.0066$	$C_f/35$	$T/25$	$P_f/51.1$
	Q_r	0.0181	0.018	-0.80		Q_r	0.0083	0.0083	0.00		Q_r	0.005	0.005	-0.25
	C_r	38.03	37.779	0.66		C_r	41.65	41.412	41.65		C_r	46.02	46.041	-0.05
	C_p	0.1186	0.132	-11.38		C_p	0.1079	0.1314	0.10		C_p	0.1173	0.135	-15.13
	Q_p	0.0016	0.001	9.09		Q_p	0.0016	0.0015	0.00		Q_p	0.0016	0.001	0.78
	P_r	38.2	32.896	13.88		P_r	43.95	43.434	43.95		P_r	47.76	47.819	-0.12
	Rec	8	7.382	7.71		Rec	16	15.533	16		Rec	24	24.052	-0.21
	Rej	99.661	99.622	0.03		Rej	99.691	99.624	0.06		Rej	99.664	99.614	0.05
2	$Q_f/0.0158$	$C_f/35$	$T/25$	$P_f/53.91$	6	$Q_f/0.0088$	$C_f/35$	$T/25$	$P_f/50.34$	10	$Q_f/0.0061$	$C_f/35$	$T/25$	$P_f/51.53$
	Q_r	0.0142	0.014	-0.25		Q_r	0.0072	0.007	-0.62		Q_r	0.0045	0.004	-0.10
	C_r	38.875	38.829	0.12		C_r	42.66	42.482	0.41		C_r	47.26	47.347	-0.19
	C_p	0.1202	0.124	-3.87		C_p	0.11	0.131	-19.82		C_p	0.1202	0.136	-13.59
	Q_p	0.0016	0.001	2.30		Q_p	0.0016	0.001	2.828		Q_p	0.0016	0.001	0.28
	P_r	40.79	38.231	6.27		P_r	45.02	44.750	0.59		P_r	48.61	48.706	-0.19
	Rec	10	9.893	1.06		Rec	18	17.667	1.84		Rec	26	26.153	-0.59
	Rej	99.656	99.643	0.01		Rej	99.685	99.623	0.06		Rej	99.656	99.609	0.04
3	$Q_f/0.0131$	$C_f/35$	$T/25$	$P_f/52.87$	7	$Q_f/0.0079$	$C_f/35$	$T/25$	$P_f/50.48$	11	$Q_f/0.0056$	$C_f/35$	$T/25$	$P_f/52.05$
	Q_r	0.0116	0.011	1.07		Q_r	0.0063	0.0063	-0.51		Q_r	0.0041	0.004	2.41
	C_r	39.76	39.937	-0.45		C_r	43.72	43.63	0.21		C_r	48.56	48.933	-0.76
	C_p	0.122	0.122	-0.04		C_p	0.1123	0.1326	-18.15		C_p	0.1234	0.138	-12.51
	Q_p	0.0016	0.001	-1.53		Q_p	0.0016	0.0015	2.03		Q_p	0.0016	0.001	0.05
	P_r	42.62	41.592	2.40		P_r	45.98	45.891	0.19		P_r	49.47	49.655	-0.37
	Rec	12	12.401	-3.34		Rec	20	19.840	0.79		Rec	28	28.554	-1.98
	Rej	99.651	99.651	0.00		Rej	99.679	99.620	0.05		Rej	99.647	99.603	0.04
4	$Q_f/0.0113$	$C_f/35$	$T/25$	$P_f/52.32$	8	$Q_f/0.0072$	$C_f/35$	$T/25$	$P_f/50.74$	12	$Q_f/0.0053$	$C_f/35$	$T/25$	$P_f/52.64$
	Q_r	0.0097	0.009	0.53		Q_r	0.0056	0.005	-0.39		Q_r	0.0037	0.003	0.32
	C_r	40.68	40.972	-0.72		C_r	44.84	44.786	0.11		C_r	49.95	50.236	-0.58
	C_p	0.124	0.121	1.80		C_p	0.1147	0.133	-16.60		C_p	0.1268	0.139	-10.30
	Q_p	0.0016	0.001	-3.25		Q_p	0.0016	0.001	1.37		Q_p	0.0016	0.001	-0.74
	P_r	44.03	43.645	0.87		P_r	46.89	46.876	0.02		P_r	50.35	50.494	-0.28
	Rec	14	14.620	-4.43		Rec	22	21.917	0.37		Rec	30	30.414	-1.38
	Rej	99.645	99.652	-0.00		Rej	99.672	99.617	0.05		Rej	99.637	99.600	0.03

1 **4. Modelling of the hybrid MED_TVC+RO processes**

2 The earliest sections provided the validation of the models developed for the thermal and
3 membrane processes. Therefore, it is possible to connect them in several ways to
4 accommodate the proposed configurations, as illustrated in [Section 2](#).

6 **4.1 Simple hybridization**

7 Referring to [Fig. 1](#), simple material balances on mixers M1 and M2 are used to describe the
8 blending of the rejected brine and fresh water. Md_{MED} is the distillate produced by the
9 thermal process with a salinity xd_{MED} , Mp_{RO} is the permeate produced by RO with a salinity
10 xp_{RO} , and $M_{freshwater}$ is the total productivity of the plant, with a salinity equal to
11 $x_{freshwater}$. Note that the salinity of the distillate from MED is always assumed equal to 10
12 ppm.

$$13 \quad Md_{MED} + Mp_{RO} = M_{freshwater}$$

14 (70)

$$15 \quad Md_{MED}xd_{MED} + Mp_{RO}xp_{RO} = M_{freshwater}x_{freshwater}$$

16 (71)

17 It is also important to evaluate the flow rate of rejected brine M_{reject} as the sum of the
18 rejected brine of the two processes, as well as its salinity x_{reject} . Note that the salinity of the
19 rejected brine from MED is fixed at 60 kg/m^3 .

$$20 \quad Mb_{MED} + Mr_{RO} = M_{reject}$$

21 (71)

$$22 \quad Mr_{MED}xr_{MED} + Mp_{RO}xr_{RO} = M_{reject}x_{reject}$$

23 (72)

24

25 **4.2 Full Hybridization, RO upstream**

1 Referring to Fig. 2, Eqs. (70) and (71) are used to evaluate the flow rate of fresh water
 2 produced by the plant and its purity, while the rejected brine is entirely produced by the MED
 3 process. For this configuration, it is necessary to quantify the by-pass flow rate M_{bypass} to
 4 provide the proper feed $M_{W_{MED}}$ in the thermal process, with a salinity equal to $x_{f_{MED}}$.

$$5 \quad M_{W_{MED}} = M_{r_{RO}} + M_{bypass}$$

$$6 \quad (73)$$

$$7 \quad M_{W_{MED}}x_{f_{MED}} = M_{r_{RO}}x_{r_{RO}} + M_{bypass}x_{seawater}$$

$$8 \quad (74)$$

9

10 **4.3 Full Hybridization, MED upstream**

11 Referring to Fig. 3, Eqs. (70) and (71) are used to evaluate the flow rate of fresh water
 12 produced by the plant and its purity. The rejected brine is evaluated accordingly to Eqs. (75)
 13 and (76), which model the blending of RO retentate and excess MED brine, where $M_{r_{RO}}$ is
 14 the retentate from the membrane process and $x_{r_{RO}}$ its salinity. Note that the MED is now
 15 forced to produce a brine with a salinity of 50 kg/m³, to obtain a suitable inlet condition for
 16 the RO process.

$$17 \quad M_{b_{MED}} + M_{r_{RO}} - M_{f_{RO}} = M_{reject}$$

$$18 \quad (75)$$

$$19 \quad (M_{r_{MED}} - M_{f_{RO}})x_{r_{MED}} + M_{p_{RO}}x_{r_{RO}} = M_{reject}x_{reject}$$

$$20 \quad (76)$$

21

22 **4.4 Parameters for comparison**

23 The comparison between the different proposed configurations is essentially based on the
 24 following chosen quantities: the productivity of the hybrid plant ($M_{freshwater}$), purity of the
 25 product ($x_{freshwater}$), specific energy consumption (Es), and recovery ratio (RR). A

1 sensitivity analysis of those parameters has been performed. Also, the quantity of rejected
2 brine (M_{reject}) and its salinity (x_{reject}) have been evaluated for every configuration, where
3 this parameter is important for environmental reasons. The total energy is evaluated using Eq.
4 (77) by considering the energy requirement of both processes. In this respect, the energy
5 consumed by the thermal process is calculated by Eq. (78) and linked with the steam enthalpy
6 that converted into kWh/m³, where only a small fraction ($E_{el} = 2$ kWh/m³) is considered as an
7 electrical energy consumed by pumps (Gude et al., 2010). However, the electrical energy
8 required by the membrane process is given by Eq. (79), which represents the required
9 pumping energy to compress the feed up to 50 atm. Eq. (80) is used to estimate the total
10 recovery ratio.

$$11 \quad E_s = \frac{E_{s,MED}M_{d,MED} + E_{s,RO}M_{p,RO}}{M_{freshwater}}$$

12 (77)

$$13 \quad E_{s,MED} = \frac{M_{steam} \lambda(T_s)}{M_{d,MED}} + E_{el}$$

14 (78)

15

$$16 \quad E_{s,RO} = \frac{P_{RO}M_{f,RO}}{\eta_{pump}M_{p,RO}}$$

17 (79)

$$18 \quad RR = \frac{M_{freshwater}}{M_{seawater}}$$

19 (80)

20

21 **5. Results and discussion**

22 In this section, a sensitivity analysis is performed to simultaneously compare the four
23 proposed configurations and to investigate the variation of the parameters when external
24 inputs such as seawater conditions and steam supply change for each considered
25 configuration. A variation of $\pm 12\%$ for steam consumption and seawater salinity and $\pm 8\%$ of

1 seawater temperature has been considered with respect to the initial values reported in [Table](#)
2 [1](#), where also the operating conditions of MED and RO processes are reported. It is
3 noteworthy to mention that a by-pass stream is necessary to satisfy the feed requirement of
4 the thermal process, being the latter more productive, when the RO process is placed
5 upstream. The by-pass ratio, defined as the quantity of seawater fed to the MED process over
6 the quantity of seawater fed to the RO process, which is already calculated as a function of
7 the operating conditions. Specifically, its value is around 3; this means that the by-pass
8 stream is larger than the feed stream to the RO process by around three folds. Then, the
9 simulation results are compared against the performance of other proposed configurations in
10 terms of productivity, energy consumption, the purity of the product, and recovery ratio to
11 identify the best one.

12

13

14

15 **5.1 Sensitivity analysis**

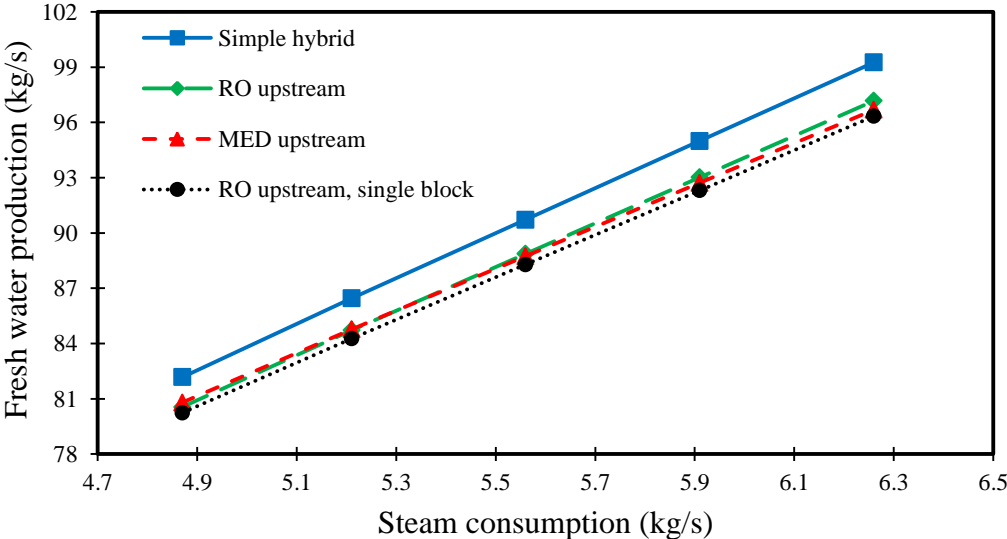
16 Performing a sensitivity analysis is important for the design and operation perspectives of any
17 industrial process. This in turn would offer the feasible operating parameters that serve the
18 process performance. Results obtained from the simulation of different configurations of
19 MED_TVC+RO hybrid processes are shown in [Figs. 5 – 10](#). These figures show the value of
20 the performance indicators of the hybrid plant, in relation to the variation of the most
21 important operating parameters. The selected performance indicators are at the same level of
22 importance and commonly used in the literature.

23 [Figs. 5](#) and [6](#) show the effect of steam supply variation of the MED process on the key
24 performance indicators of hybrid system, i.e., the overall productivity and fresh water
25 salinity.

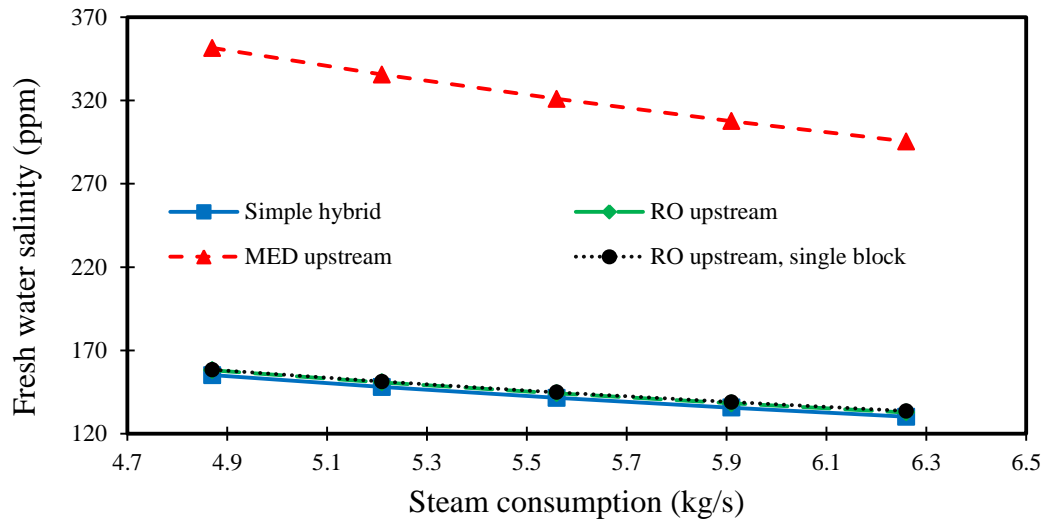
1 This in turn confirmed that the production of fresh water linearly increases as well as its
 2 purity as a result to increasing the steam fed to the thermal process. Apparently, the hybrid
 3 plant productivity is, for every configuration, strongly dependent on the quantity of steam
 4 used. This is because the MED process accounts for approximately $\frac{3}{4}$ of the total fresh water
 5 production (Fig. 5).

6 The comparison of four proposed configurations based on the product salinity is investigated
 7 based on the steam consumption in Fig. 6. This in turn shows that the configuration with
 8 MED upstream generates a product with a salinity always above 300 ppm. Specifically, this
 9 is quite comparable to all the other proposed configurations, which produce fresh water with
 10 salinity under 200 ppm for every operating condition. To systematically resolve this problem,
 11 a more productive MED plant should be designed to dilute even more the high-salinity RO
 12 permeate, or a different RO process structure must be implemented to generate a purer
 13 permeate.

14

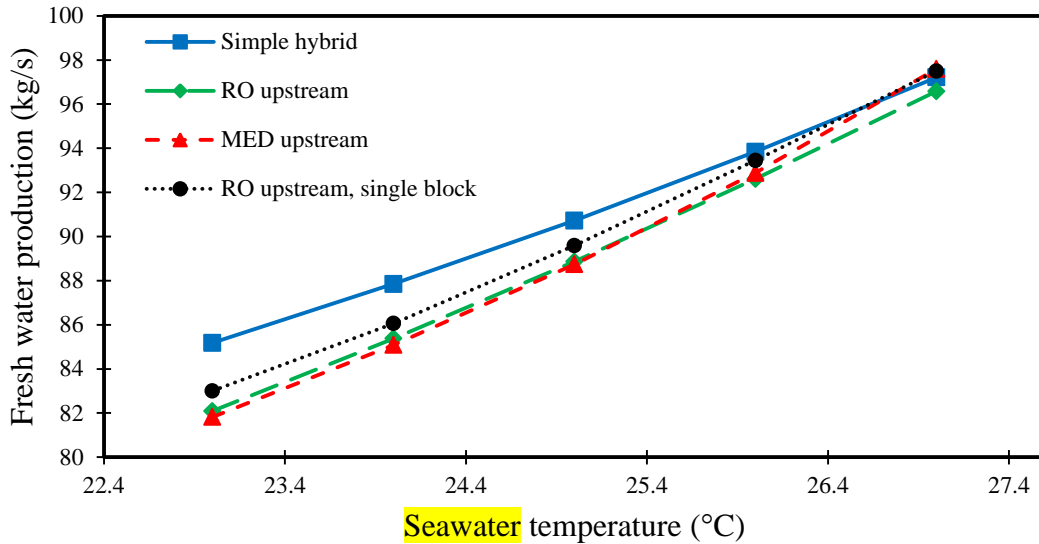


15
 16 **Fig. 5.** Fresh water production versus steam consumption in the thermal process for different configurations of
 17 the hybrid process.

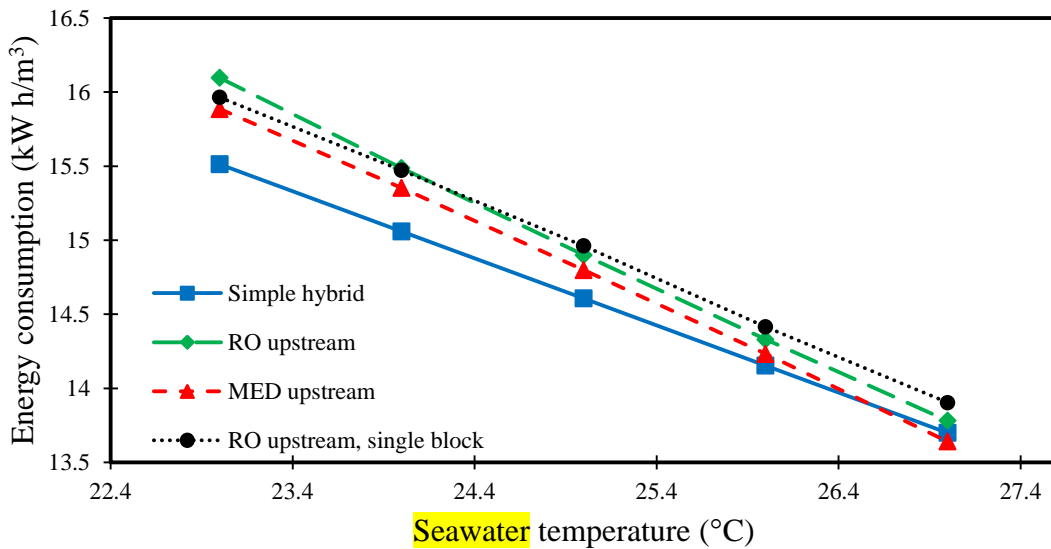


1
2 **Fig. 6.** Fresh water salinity versus steam consumption in the thermal process for different plant configurations of
3 the hybrid process.
4

5 The effect of seawater temperature variation on the fresh water productivity and energy
6 consumption of the hybrid system is plotted in Figs. 7 and 8, respectively. The simple hybrid
7 configuration presented in Fig. 1 is the least sensitive to variation of external seawater
8 temperature, due to its simplicity and straightforward operation, while the configuration with
9 MED upstream is the more sensible configuration. Specifically, Figs. 7 and 8 confirm that the
10 simple hybrid configuration performs a slightly higher productivity and a little lower energy
11 consumption. However, those advantages tend to invalidate at high seawater temperatures,
12 which is the most realistic scenario when considering hot and arid regions as possible sites to
13 install the proposed plant.
14



1
2 **Fig. 7.** Fresh water production versus inlet seawater temperature for different plant configurations of the hybrid
3 process.
4

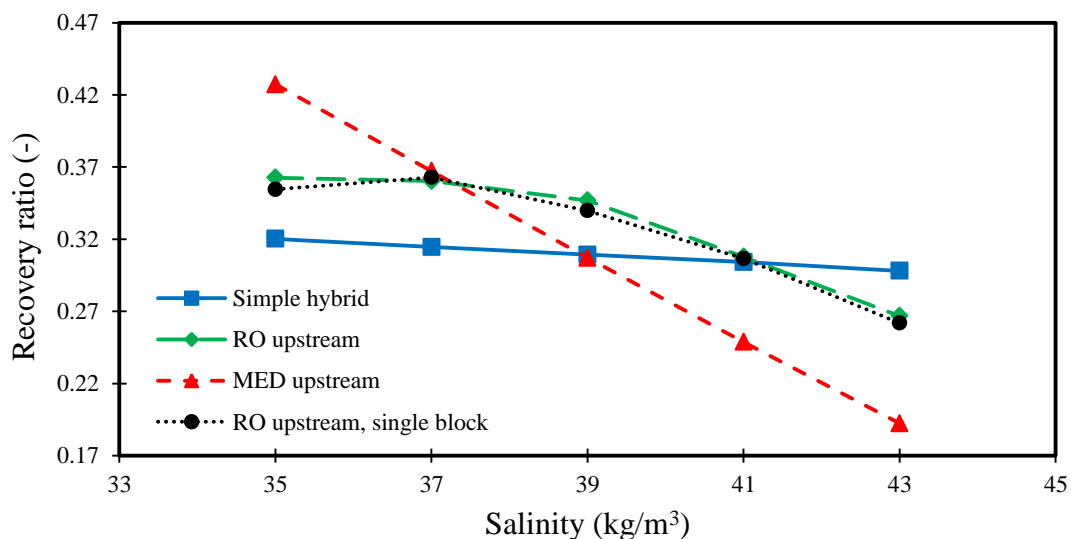


5
6 **Fig. 8.** Specific energy consumption versus inlet seawater temperature for different plant configurations of the
7 hybrid process.
8

9 **Figs. 9 and 10** show the effect of seawater salinity variation on the energy consumption and
10 the overall recovery ratio of the hybrid system. The full hybrid configuration with RO
11 upstream shows a relatively higher performance under every aspect for both the three blocks
12 and single block configurations represented in **Figs. 2 and 4**, respectively. Specifically, **Fig. 9**
13 confirms the superiority of this configuration regarding the recovery ratio, except for very

1 high seawater salinity (over 41 kg/m^3). The reason for this behavior is that all the rejected
 2 brine of the membrane process is re-utilized as feed for the thermal process, which reduces
 3 the need for an external seawater feed. In contrast, the full hybrid configuration with MED
 4 upstream has some issues related to the fact that the thermal process is forced to produce a
 5 lower salinity brine to feed the membrane process. Accordingly, this limits the salinity
 6 windows in which it can operate and thus reducing the MED upstream performance. This is
 7 especially true when seawater salinity is high (for instance, $40 - 43 \text{ kg/m}^3$). Moreover, the
 8 MED process operates very poorly at a noticeable increase of energy consumption (Fig. 9)
 9 and a significant reduction of the recovery ratio (Fig. 10). However, the MED upstream
 10 design allows to reach a considerable recovery ratio that commensurate with the lowest
 11 energy consumption if the seawater salinity is low (for instance 35 kg/m^3). Energy
 12 consumption is generally moderately dependent on seawater conditions, except for the MED
 13 upstream configuration, which shows a strong dependence (Figs. 8 and 10). Recovery ratio is
 14 linear dependent on seawater salinity for the simple hybridization (weakly) and MED
 15 upstream hybridization (strongly), while there is a moderate non-linear dependence for the
 16 RO upstream configurations (Fig. 9).

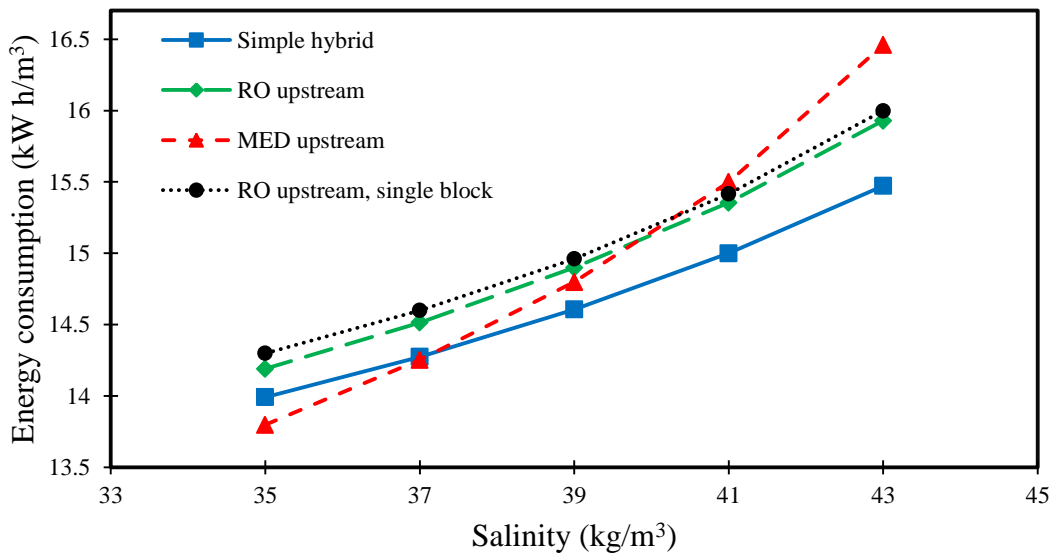
17



18

1 **Fig. 9.** Recovery ratio versus inlet seawater temperature for different plant configurations of the hybrid process.

2



3

4 **Fig. 10.** Specific energy consumption versus inlet seawater temperature for different plant configurations of the hybrid process.

5

6
7 **Table 5** presents the simulation results of all the proposed configurations with a fixed
8 seawater salinity of 37 kg/m³. All other parameters are set according to **Table 1**. Moreover,
9 the evaluation of the flowrate and salinity of the rejected brine is included.

10

11

12 **Table 5.** Performance comparison of the proposed configurations. Simulations performed with seawater salinity
13 of 37 kg/m³, other parameters set accordingly with **Table 1**

Configuration type	Productivity (kg/s)	Product salinity (ppm)	Rejected flow (kg/s)	Rejected salinity (kg/m ³)	Energy Consumption (kW h/m ³)	Recovery Ratio (-)
Simple hybrid	93.36	136	162.84	60.72	14.27	0.3146
RO upstream	91.80	138	165.73	60.00	14.51	0.3603
RO upstream, 1 block	93.25	135	154.75	60.00	14.93	0.3521
MED upstream	92.42	306	198.17	53.08	14.25	0.3673

14

15 **Table 5** shows how the configuration with MED upstream produces the highest brine flow
16 rate despite attaining the lowest rejected salinity compared to other configurations. This is
17 due to considering of 50 kg/m³ as the rejected brine concentration of the thermal process

1 instead of 60 kg/m^3 when placed upstream. Another relevant observation is that using the RO
2 process in a single block can lead to a reduction of about 7% of the rejected flow rate
3 compared to RO upstream configuration. Up to the authors' knowledge, this reduction is not
4 enough to justify the feasibility of this configuration compared to the conventional 3 RO
5 blocks, which presents higher recovery ratio and slightly lower energy consumption.

6 To investigate the robustness of the proposed hybrid system, the earlier simulation results are
7 compared with the findings of a detailed study of MSF+RO hybrid plant carried out by [Helal
8 et al. \(2004\)](#). Specifically, [Helal et al. \(2004\)](#) have investigated all the possible alternatives
9 for integrating the thermal MSF process and the membrane RO process in a hybrid system.
10 The authors also conducted an economic analysis to estimate the cost of fresh water for every
11 proposed configuration. The output of this study has affirmed that the best configuration was
12 the one where the RO and the MSF plants were partially integrated. In other words, a fraction
13 of the heated feed from the intake was fed to the single-stage RO plant, then the RO permeate
14 was mixed with the MSF distillate and the reject stream was combined with the MSF
15 blowdown. This configuration was able to generate fresh water at around 500 ppm, with an
16 overall recovery ratio of 32.4 %. Interestingly, the investigated MED_TVC+RO system in the
17 current study is in turn able to generate fresh water with a salinity lower than 200 ppm, with
18 an overall recovery ratio up to 37 %. However, the performance of this system is quite
19 sensible to the variation of seawater properties. Most importantly, the current study explored
20 the impact of possible variations of seawater properties (i.e. seasonal changes) on the hybrid
21 process performance. According to our results, the best overall configuration appears to be
22 the MED_TVC+RO full hybrid with RO process placed upstream This is due to the best
23 recovery ratio over a wide range of seawater salinity, moderately dependence of other
24 parameters on changing seawater conditions, and low salinity of the produced freshwater. No
25 great differences are highlighted between the triple block RO upstream configuration and the

1 single block configuration. However, the use of three separate blocks allows a slightly lower
2 energy consumption. Finally, it can be said that the simple hybrid could be a more feasible
3 option in case of operating at cooler and very low salinity seawater. However, this will not be
4 the case because this kind of plants is usually installed in hot regions with fairly high
5 seawater salinity (i.e. Gulf regions).

6

7 **6. Conclusions**

8 In this paper, the interest is on the MED_TVC+RO hybrid desalination systems, that are less
9 well studied in the literature compared to other more popular hybrid configurations, such as
10 MSF+RO. Detailed mathematical models for both the thermal and the membrane processes
11 have been developed and validated against literature and projected data of TDS2,
12 respectively, providing a good agreement. Four different possibilities to connect the
13 processes have been investigated. Moreover, a performance sensitivity analysis of the
14 proposed configurations was performed by running the simulations with variable seawater
15 properties and steam supply. The productivity of the various configurations, the purity of the
16 fresh water, recovery ratio, and energy consumption, were considered as the performance
17 indicators. The results confirmed that placing the MED process upstream results unfeasible
18 for a high seawater salinity due to bad operation of the thermal process, bounded in a narrow
19 salinity window. In other words, the MED process upstream hybrid system is significantly
20 sensible with respect to seawater salinity. Additionally, the generated fresh water salinity
21 appears to be too high. On the other hand, placing the RO process upstream in a full
22 hybridized configuration provides an enhanced recovery ratio for seawater salinity under 41
23 kg/m³. This configuration proved to be competitive also from the point of view of
24 productivity and energy consumption. Therefore, this configuration was identified as the best
25 one overall among the four proposed configurations.

1

2 **Appendix A**

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

4 Elsevier.

5 **Boiling Point Elevation**

6 Correlation valid in the range: $1\% < w < 16\%$, $10^\circ\text{C} < T < 180^\circ\text{C}$

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

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

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

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

$$BPE = BPEa \cdot w + BPEb \cdot w^2 + BPEc \cdot w^3 \quad [^\circ\text{C}]$$

8

9 **Specific heat at constant pressure**

10 Correlation valid in the range: $20000 \text{ ppm} < x < 160000 \text{ ppm}$, $20^\circ\text{C} < T < 180^\circ\text{C}$

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

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

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

11 $cpc = 1.2026 \cdot 10^{-2} - 5.3566 \cdot 10^{-4} \cdot s + 1.8906 \cdot 10^{-6} \cdot s^2$

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

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

12 **Latent heat of evaporation**

13 $\lambda = 2501.89715 - 2.40706 \cdot T + 1.19221 \cdot 10^{-3} \cdot T^2 - 1.5863 \cdot 10^{-5} \cdot T^3 \quad \left[\frac{kJ}{kg} \right]$

14 **Global heat exchange coefficients**

15 $U_{ev} = 1.9695 + 1.2057 \cdot 10^{-2} \cdot T - 8.5989 \cdot 10^{-5} \cdot T^2 + 2.5651 \cdot 10^{-7} \cdot T^3 \quad \left[\frac{kW}{m^2 \cdot ^\circ\text{C}} \right]$

$$1 \quad U_{cond} = U_{ph} = 1.7194 + 3.2063 \cdot 10^{-3} \cdot T + 1.597 \cdot 10^{-5} \cdot T^2 - 1.9918 \cdot 10^{-7} \cdot T^3 \quad \left[\frac{kW}{m^2 \cdot ^\circ C} \right]$$

2

3

Table A.1. Equations describing the TVC section modelling. Reference: [Dessouky et al. \(2002\)](#)

No.	Title	The Mathematical Expression
1	Pressure Correction Factor	$PCF = 3e^{-7} \cdot Pm^2 - 0.0009 \cdot Pm + 1.6101$
2	Temperature Correction Factor	$TCF = 2e^{-8} \cdot Tv_n^2 - 0.0006 \cdot Tv_n + 1.0047$
3	Pressure at vapor temperature	$Pv = P_{crit} e^{\left(\frac{T_{crit}}{Tv_n} + 273.15\right) - 1} \cdot \sum_{j=1}^8 f_j$
4	Pressure at steam temperature	$Ps = P_{crit} e^{\left(\frac{T_{crit}}{Ts} + 273.15\right) - 1} \cdot \sum_{j=1}^8 f_j$
5	Calculate Compression Ratio	$CR = \frac{Pv}{Ps}$
6	Calculate Entrainment Ratio	$Ra = 0.296 \frac{Ps^{1.19} Pm^{0.015} PCF}{Pev^{1.04} Pev^{0.015} TCF}$
7	Calculate motive steam flowrate	$Mm = Ms \frac{Ra}{1 + Ra}$

Effect number	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

Table A.2. The mathematical modelling of retentate reprocessing RO desalination plant ([Fig. S.F.2 in the supplementary file](#))

Model Equations	Specifications	Eq. no
$Q_{f(plant)} = Q_{r(plant)} + Q_{p(plant)}$	Plant feed flow rate	1
$Q_{f(plant)} C_{f(plant)} = Q_{r(plant)} C_{r(plant)} + Q_{p(plant)} C_{p(plant)}$	Plant feed concentration	2
$Q_{r(plant)} = Q_{r(Block\ 3)}$	Plant retentate flow rate	3
$C_{r(plant)} = C_{r(Block\ 3)}$	Plant retentate concentration	4
$C_{p(Plant)} = \frac{C_{p(Block\ 1)} Q_{p(Block\ 1)} + C_{p(Block\ 2)} Q_{p(Block\ 2)} + C_{p(Block\ 3)} Q_{p(Block\ 3)}}{Q_{p(plant)}}$	Plant product concentration	5
$Q_{p(Plant)} = Q_{p(Block\ 1)} + Q_{p(Block\ 2)} + Q_{p(Block\ 3)}$	Plant permeate flow rate	6
$T_{f(plant)} = T_{r(plant)}$	Plant constant temperature	7
$P_{f(plant)} = P_{f(Block\ 1)}$	Plant feed pressure	8
$P_{r(plant)} = P_{r(Block\ 3)}$	Plant retentate pressure	9
$Rec_{(plant)} = \frac{Q_{p(plant)}}{Q_{f(plant)}} \times 100$	Total plant permeate recovery	10
$Rej_{(plant)} = \frac{C_{f(plant)} - C_{p(plant)}}{C_{f(plant)}} \times 100$	Total plant rejection	11

$C_{f(Block\ 1)} = C_{f(plant)}$	Feed concentration of 1 st block	12
$Q_{f(Block\ 1)} = Q_{f(plant)}$	Feed flow rate of 1 st block	13
$Q_{p(Block\ 1)} = \sum_{PV=1}^{20} Q_{p(PV)}$	Permeate flow rate of 1 st block	14
$C_{p(Block\ 1)} = \frac{\sum_{PV=1}^{20} C_{p(PV)} Q_{p(PV)}}{Q_{p(Block\ 1)}}$	Permeate concentration of 1 st block	15
$Rej_{(Block\ 1)} = \frac{C_{f(Block\ 1)} - C_{p(Block\ 1)}}{C_{f(Block\ 1)}} \times 100$	Total solute rejection of 1 st block	16
$Rec_{(Block\ 1)} = \frac{Q_{p(Block\ 1)}}{Q_{f(Block\ 1)}} \times 100$	Total permeate recovery of 1 st block	17

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 (m²)

$A_{ph,i}$: Exchange area of i-th pre-heater (m²)

A_{cond} : Exchange area of final condenser (m²)

$A_{ev,mean}$: Mean exchange area of evaporators (m²)

$A_{ph,mean}$: Mean exchange area of pre-heaters (m²)

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 in Eq. (59) in (-)

L : Membrane length (m)

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

m_f : Coefficient in Eq. (62)

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 (-) and the spacer characteristics in RO 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 (m³/s)

Q_f : Feed flowrate of a single membrane (m³/s)

$Q_{f(plant)}$: Plant feed flow rate (m³/s)

Q_p : Total permeate flow rate of a single membrane (m³/s)

$Q_{p(plant)}$: Plant permeate flow rate (m³/s)

$Q_{p(PV)}$: Permeate flow rate of single pressure vessel (m³/s)

Q_r : Retentate flowrate of a single membrane (m³/s)

$Q_{r(plant)}$: Plant retentate flowrate (m³/s)

Q_s : Total solute flux through the membrane (kg/m² 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 ($^{\circ}C$)

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

t_n : Feed temperature after final condenser ($^{\circ}C$)

Tl : Top brine temperature (T_{top}) ($^{\circ}C$)

Tb : Temperature of rejected brine ($^{\circ}C$)

Ts : Steam temperature ($^{\circ}C$)

Tv_i : Temperature of the vapor phase in i-th effect ($^{\circ}C$)

Tw : Temperature of the cooling water ($^{\circ}C$)

T_{mean} : Mean temperature in the plant ($^{\circ}C$)

T_{crit} : Critical temperature of water ($^{\circ}C$)

TCF : Temperature Correction Factor (-)

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

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

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

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%)

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

x_f : 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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