Desalination and Water Treatment
www.deswater.com
doi: 10.1080/19443994.2014.940648

55 (2015) 3048–3055
September

Energy recovery using salinity differences in a multi-effect distillation system

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Received 31 March 2014; Accepted 16 June 2014

ABSTRACT

The use of Salinity Gradient methodologies to recover part of the osmotic energy in the brine of multi-effect distillation (MED) systems is explored here. Measurements from a membrane-based Pressure Retarded Osmosis laboratory system have been used to estimate the energy that would be recovered from this brine, when a source of low-salinity water is available locally (such as industrial or municipal wastewater). This methodology has been evaluated for a specific case study (72 m^3/d solar/gas MED system) at different temperatures.

Keywords: Energy recovery; Desalination; Pressure retarded osmosis; Multi-effect distillation

1. Introduction

Salinity Gradient (SG) energy is based on exploiting chemical potential differences between liquids with different concentrations of salts [1,2]. This renewable energy technique has been successfully developed during the last decades, with several pilot plants in operation. These pilot plants showed the feasibility of one of the proposed SG technologies: the Pressure Retarded Osmosis (PRO) technique [3,4]. Other test facilities, some based on alternative operating principles, are being developed throughout the world: it deserves special attention, the Reverse Electrodialysis, in development by REDStack, in Netherlands.

This paper is focused on the use of SG techniques to recover the osmotic energy from the brine of multi-effect distillation (MED) systems. The advantages of some low-temperature thermal desalination processes are their ability to be driven by low energy thermal sources, their reliability, easier operation and maintenance, and high purity of the product water. Among the thermal desalination processes, MED has the highest thermal efficiency and the lowest power consumption [1].

The brine obtained with MED processes has the particular characteristic of its relatively high and stable temperature (>35˚C), which improves the performance of the PRO membranes, as source of low-salinity water, municipal wastewaters have been frequently proposed and successfully evaluated in PRO processes [5–7],

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Presented at the Conference on Desalination for the Environment: Clean Water and Energy 11–15 May, 2014, Limassol, Cyprus

therefore they are assumed to be the source of low-salinity water for the SG Energy Recovery system assessed within this work.

In this paper, an evaluation of a PRO process coupled with a MED plant is presented. Some mathematical models are included to demonstrate the dependence of the recovered power with the brine and wastewater temperatures.

This paper is organized as follows. Sections 2 and 3 briefly describe MED and PRO processes. The proposed combination between these two processes is presented in Section 4. In Section 5 the case study is explained and evaluated. The results are discussed in Section 6.

2. MED process

There are many variations of MED plants, but in all of them, the distillation process is similar. The plant is divided in hermetic elements called effects, which are connected between them. During the normal operation of the plant a series of simultaneous evaporation/condensation processes in a decreasing sequence of pressures and temperatures. At the first effect, the one with the higher pressure, an external heat source produces the first boiling of seawater. The steam generated within the effect is used as the heat source of the next effect, so, while in one hand, the incoming steam is condensing, on the other, the seawater is boiling producing additional steam. This process is repeated in each effect.

A typical forward-feed vertically-stacked MED plant is schematized in Fig. 1. It consists of several effects vertically arranged, where the seawater goes from one effect to the other by gravity. Each effect has a preheater placed next to it with the aim of increasing the temperature of the feed seawater before it is introduced within the first effect. Here, seawater is sprayed over a horizontal tube falling film type evaporator powered by an external heat source (saturated steam or hot water). Part of the seawater evaporates within the effect and the rest, more concentrated, is sprayed over the second effect. The steam previously produced in the first effect flows through the preheater placed next to it, where only a small portion is partially condensed releasing his latent heat to the seawater that flows inside the tubes. Then, the steam and the distillate produced are taken inside the horizontal tube bundle of the second effect driving a new evaporation process at a lower temperature. The process is repeated from the third effect to the last one. Vapor condensed inside each horizontal tube bundle from the second to the last effects represents most of the distillate production of the MED plant. The vapor produced in this last effect is condensed in a final condenser cooled by seawater. The fraction of seawater that has not been evaporated reaches its highest concentration at the last effect where it is extracted by a pump. The total distillate production of the plant is also collected together at the final condenser and extracted by a pump.

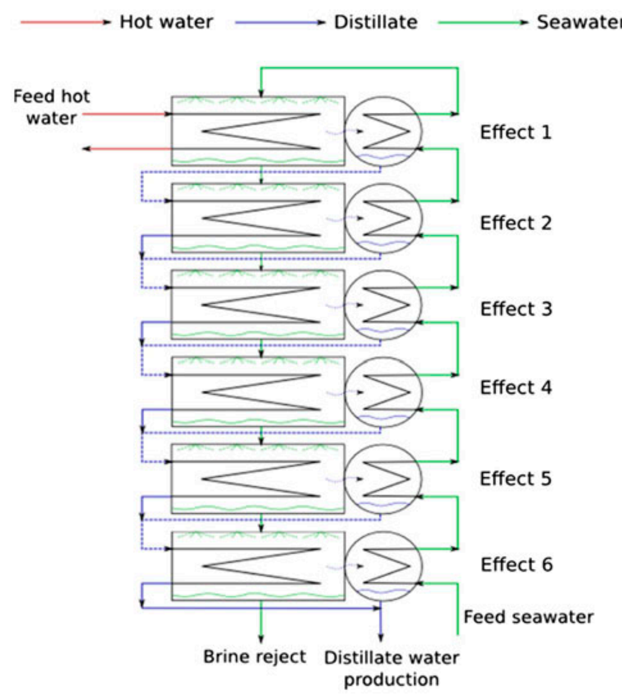

Fig. 1. Schematic process diagram of a forward feed vertically stacked MED plant.

3. PRO process

PRO has the ability to take advantage of the free energy of mixing when fresh river water flows into the sea for clean and renewable power generation. PRO utilizes the osmotic pressure difference that develops, when a semipermeable membrane separates two solutions of different concentration, to drive the permeation of water from the solution into the more concentrated solution. A hydraulic pressure less than the osmotic pressure difference is applied to the draw solution, thereby retarding water flux across the membrane, and a hydro-turbine extracts work from the expanding draw [8].

Water permeation flux (J_w) across an ideal semipermeable thin film that allows water passage but fully rejects all other solute molecules or ions can be related to the water permeability, A, the effective osmotic pressure difference, $\Delta\pi_m$, and the transmembrane hydraulic pressure difference, ΔP, as follows [9]:

$$J_w = A(\Delta\pi_m - \Delta P) \tag{1}$$

$$J_w = A(\pi_{D,m} - \pi_{F,m} - \Delta P) \tag{2}$$

where $\pi_{D,m}$ and $\pi_{F,m}$ are the osmotic pressure at the surface of the active layer at the support layer, respectively (see Fig. 2).

The salt flux J_s can be described as:

$$J_s = B\ \Delta C_m \tag{3}$$

$$\Delta C_m = C_{D,m} - C_{F,m} \tag{4}$$

where B is the salt permeability coefficient of the membrane active layer, and $C_{D,m}$ and $C_{F,m}$ are the solute concentrations at the interfaces of the active layer.

The external concentration polarization modulus is calculated using [10]:

$$\frac{\pi_{D,m}}{\pi_{D,b}} = \exp\left(-J_w/k\right) \tag{5}$$

where $k = \frac{\mathrm{Sh}D}{\delta}$ is the mass transfer coefficient in the draw solution, Sh is the Sherwood number, D is the diffusion coefficient of the brine water, and δ is the thickness of the boundary layer. Referring to [11], when the draw solution faces the active layer and the feed solution faces the support layer the salt of the feed water enters easily in the support layer by convection and therefore increases the concentration within the porous layer, and this is known as the internal concentration polarization (ICP) [11]. This phenomenon was described by Lee et al. using the following expression [12]:

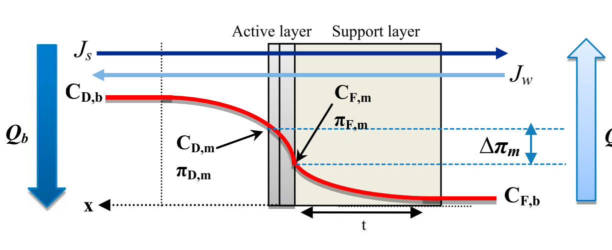

Fig. 2. Representation of flows at the membrane surface, generated by SGs. Internal and external concentration polarizations are also shown.

$$K = \left(\frac{1}{J_w}\right)\ln\frac{B + A\pi_{D,m} - J_w}{B + A\pi_{F,b}} \tag{6}$$

where $K = \frac{t\tau}{\varepsilon D} = \frac{s}{D}$ is the solute resistivity, t, τ, ε and s are the thickness, tortuosity, porosity, and structure parameter of the support layer, respectively.

For membranes, which reject salt to a high degree and operate at high flux, B is negligible compared to the other terms. Neglecting salt flux in the direction of water flux and any passage of salt from the permeate side, Eq. (6) is rearranged as follows:

$$J_w = A[\pi_{D,m} - \pi_{F,b}\exp\left(J_w K\right)] \tag{7}$$

The exponent factor is known as ICP modulus:

$$\frac{\pi_{F,m}}{\pi_{F,b}} = \exp\left(J_w K\right) \tag{8}$$

Combining Eq. (5) with Eq. (7), the water permeation is:

$$J_w = A\left[\pi_{D,b}\exp\left(\frac{J_w}{k}\right) - \pi_{F,b}\exp\left(J_w K\right) - \Delta P\right] \tag{9}$$

Then, assuming that the osmotic pressures depend on the temperature following $\pi = \beta CRT$ and substituting K and k with their expressions we conclude that:

$$J_w = A\beta R\left[T_{D,b}\exp\left(-\frac{J_w \cdot d_h}{\mathrm{Sh}\cdot D_D}\right) - T_{F,b}\exp\left(J_w\frac{S}{D_F}\right)\right] - A\Delta P \tag{10}$$

where β is the van't Hoff coefficient, D_F is the diffusion coefficient of the feed solution, and D_D is the diffusion coefficient of the solute in the draw solution.

The energy recovered by the PRO membranes can be estimated, assuming temporarily ideal conditions and a flat sheet PRO membrane:

$$W_e = Q_b \cdot \Delta P \cdot \Omega \tag{11}$$

with Q_b the flow of water that crosses the PRO membrane, ΔP the pressure difference between both sides of the PRO membrane, and Ω the efficiency of the turbine (that in practice is around 80–90%). In optimum conditions, it has been shown [3,13] that the maximum energy for the PRO process is obtained when the difference of pressures between both sides of the PRO membrane is half of the difference of osmotic pressures at the membrane (created by the difference of salinity between the draw and feed liquids):

$$\Delta P = \frac{\Delta\pi}{2} \tag{12}$$

The water flow through the membrane can be calculated from the applied pressure and the difference of osmotic pressures at both sides of the recovery membranes:

$$Q_b = S \cdot A\,(\Delta\pi - \Delta P) \tag{13}$$

where S is the active membrane surface and A is the intrinsic water permeability coefficient of the membrane (that depends on the membrane and the operating temperature). This gives that the total power that is recovered by the membrane is:

$$W_e = \Omega \cdot S \cdot A\,(\Delta\pi - \Delta P) \cdot \Delta P \tag{14}$$

4. Proposed recovery of osmotic energy in MED

The current work explores the possibility of using SG to recover osmotic energy from MED systems. The process would operate as follows: the osmotic energy of the outlet brine in MED plants would be transformed into hydraulic pressure by using wastewater as low-salinity water source; this pressure is then transformed into electricity in a turbine (it could also be used to assist pumping within the MED process, by exchanging pressure using standard hydraulic energy recovery devices, but this possibility is outside the scope of this work). From the available SG approaches [1–4] PRO has been selected because the high temperature of the brine improves the performance of PRO processes [14]. In fact, brines are frequently cooled down in MED plants using seawater and a heat exchanger. As an alternative, the wastewater that would be used as feed water in the PRO process can be used in the heat exchanger, so that the PRO process operates at the best operating point for each specific membrane (feed temperature has the biggest effect in the process). It must be pointed out that the proposal is tested in the laboratory, as the membranes specific for PRO, that are needed to get good performance, are not yet commercially available [9,10,15,16], but are expected to reach the market soon (see [17] for some recent developments).

As it is depicted in Fig. 3, two flows are in contact through the PRO membranes:

(1) A brine flow (Q_d) which in this case corresponds to the MED brine at the last effect. This flow is assumed to have high osmotic pressure, and would be regulated to have the optimal pressure for the PRO process.
(2) A feed water flow (Q_f); in this case wastewaters that cannot be used for recovery of water for human consumption. This flow is assumed to have low osmotic pressure, and would be regulated to have low pressure.

5. Case study: the AQUASOL MED plant

The AQUASOL plant is a solar thermal desalination system, located at the Plataforma Solar de Almería (southern Spain), which is going to be used as a case study to test the current proposal of sustainable desalination process. Currently, the experimental plant operates as an hybrid solar-gas plant that combines a MED process and a low temperature solar field with a Double Effect Absorption Heat Pump coupled with a gas boiler [18] (Fig. 4). The MED plant is a 14-effect forward feed unit with a vertical arrangement. The nominal operating parameters are presented in Table 1. A static compound parabolic concentrator solar field provides the thermal energy required for the MED process during sunshine hours. This thermal energy is stored in two water tanks. The gas boiler used by the DEAPH can provide heat for the MED process at variable loads from 30 to 100% when no solar energy is available. A three-way regulating valve (V2) is used to reach the nominal first-effect inlet temperature by mixing water

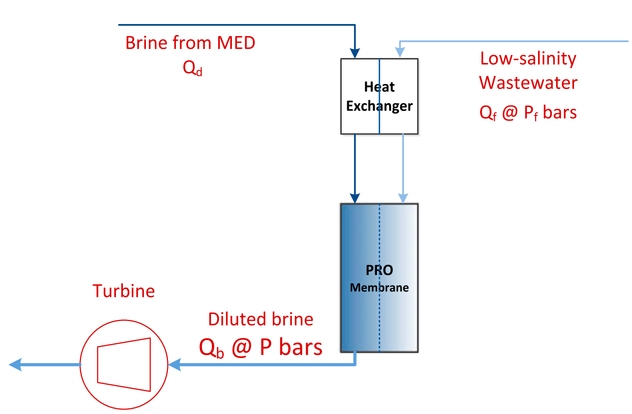

Fig. 3. Basic concept of the pressure-retarded osmosis process for osmotic energy recovery of MED brines.

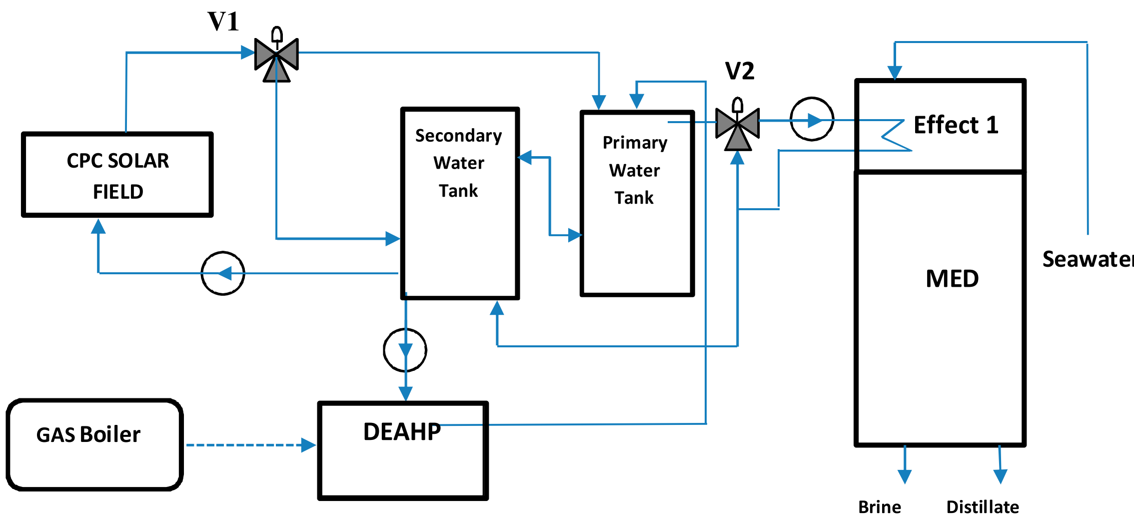

Fig. 4. Schematic diagram of the AQUASOL plant.

from the primary tank with the return flow coming back from the first effect.

The salt water for the AQUASOL MED plant is obtained by brackish water wells of Tabernas Dessert whose salinity is 3.3 g/l. The experimental brine salinity obtained by the plant is too low so extrapolations for usual seawater salinity were required. The simulations have been performed over a non-lineal first-principles model designed according to the experience with real data from the plant [19]. Table 2 shows the results of the simulations for different inlet seawater conditions.

To validate the proposed energy recovery technique, experiments were carried out at the Fraunhofer Institute for Interfacial Engineering and Biotechnology. Self-developed cellulose acetate membranes with an optimized internal structure [15] were studied under realistic operating conditions for a scaled-down membrane surface. More precisely, process parameters such as salt concentrations, pressures and flow rates were varied in order to validate the models and get information on the most adequate operating conditions and the expected energy production (for details see [14]).

Assuming the MED brine temperature and concentrations presented in Table 2 and a low- salinity water of 0.5 g/l, the PRO operating pressure, flows and the energy that could be recovered are presented in Table 3, where the results are obtained by scaling up the laboratory results obtained for different temperatures of feed water. It can be seen that a significant portion of the Osmotic Pressure can be recovered before discharge, especially operating at higher temperatures: up to 22 kW if it were possible to operate

Table 1
Nominal operating parameters of the Aquasol MED system

Number of effects	14
Feed seawater flow rate	8 m^3/h
Brine flow rate from the last effect	5 m^3/h
Hot water flow rate	12.0 l/s
Total distillate output	3 m^3/h
Cooling seawater flow rate at 25˚C	20 m^3/h
Vapor production in the last effect at 35˚C	159 kg/h
Heat source energy consumption	200 kW
Performance ratio	>9
Vacuum system	Hydro-ejectors (seawater at 3 bar)
Inlet/outlet hot water temperature	75.0/71.0˚C
Brine temperature (on the first cell)	68˚C
Feed and cooling sea water temperature at the outlet of the condenser	33˚C

Table 2
Concentrations and temperatures of the brine in the PSA AQUASOL plant for different inlet seawater conditions

Case study	SW concentration (g/l)	SW temperature (°C)	Brine concentration (g/l)	Brine temperature (°C)
#1	40	29	57	40
#2	38	20	56	31
#3	36	30	51	41

Table 3
Expected energy recovered using the proposed system for different seawater and feed water conditions

Case Study	Operating pressure (bars)	Feed water flow (m^3/h)	Feed water temperature (°C)	Discharge concentration (g/l)	Power recovered (kW)
#1	23.5	5	20	39.8	14.2
			30	35.5	19.8
			40	33.7	22.6
#2	22.5	5	20	39.9	13.5
			30	36.1	17.5
			40	33.7	21.0
#3	21.2	5	20	35.7	12.8
			30	32.5	17.1
			40	31.1	19.2

with draw water at 40°C (this temperature could be achieved by heat exchange in the first stages of the MED process). As a secondary positive effect, the salinity of the discharge is reduced from 57 g/l to 32–35 g/l (as the brine is mixed with low-salinity water not suitable to produce drinking water); this concentration is near the seawater concentration, facilitating discharge to the sea. The membrane area needed to reproduce the results should be around 3 m^2, which is acceptable for this process.

6. Discussion on the effect of the temperature

The experimental results obtained so far in the laboratory have shown that SG techniques depend significantly on temperature. We believe that the main effect is due to the change of fluid parameters with the temperature. For example, a preliminary evaluation of some physical parameters is presented in Table 4 for different temperature and concentration values in MED systems. In this case, the temperatures of the streams were kept equal ($T_{D,b} = T_{F,b} = T$). Sherwood (Sh), Reynolds (Re) and Schmidt (Sc) numbers corresponding to the feed stream were also illustrated. It can be seen in Fig. 5 that increasing the temperature leads to a better performance of the process. Increasing the temperature will lead to the change of physiochemical properties of the membrane and solution which can directly influence the osmotic membrane performance. The result can be justified by the fact that the changing of the physical parameters of the two streams caused by the rise of the temperature improve the water flux crossing the membrane. In fact, the rise of the temperature reduces the viscosity of the water at the surface of the membrane and increases the diffusivity of the water, thus, this the ICP that can occur at the surface of the membrane support layer will be reduced. Also, the osmotic

Table 4
Variation of the viscosity, the diffusivity, the difference of the osmotic pressure, Sherwood, Reynolds and Schmidt numbers with the operating temperature

T (°C)	η_F (Pa.s)	η_D (Pa.s)	D_F (m^2/s)	D_D (m^2/s)	Sh (–)	Re (–)	Sc (–)	$\Delta\pi$ (bar)
20	1.00×10^{-3}	1.07×10^{-3}	3.80×10^{-9}	3.56×10^{-9}	25.1	99.5	265	45.0
30	7.98×10^{-4}	8.23×10^{-4}	4.93×10^{-9}	4.78×10^{-9}	26.8	125	162.4	46.6
40	6.53×10^{-4}	6.87×10^{-4}	6.23×10^{-9}	5.92×10^{-9}	28.3	150	106.7	48.1

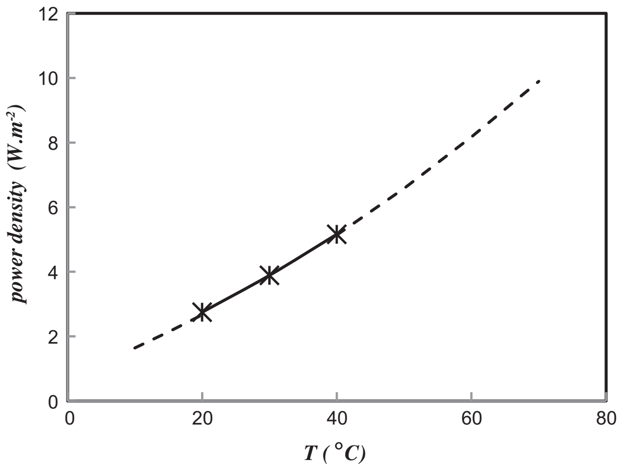

Fig. 5. Variation of the power density with the temperature.

pressure difference also increases at higher temperature which improves the driving force of the process. Consequently, this result leads to higher values of J_w. Rising the temperature from 20 to 60°C is followed by an increase of two times the amount of energy produced. This result is not only caused by to the change of physiochemical properties of the solution. In fact, membrane transport parameters are also affected by the operating temperature.

Fig. 6 shows the percentage of energy that can be produced at different operating temperatures. It is clear that is better to operate with inlet solutions with high temperatures to cause enhanced PRO performance

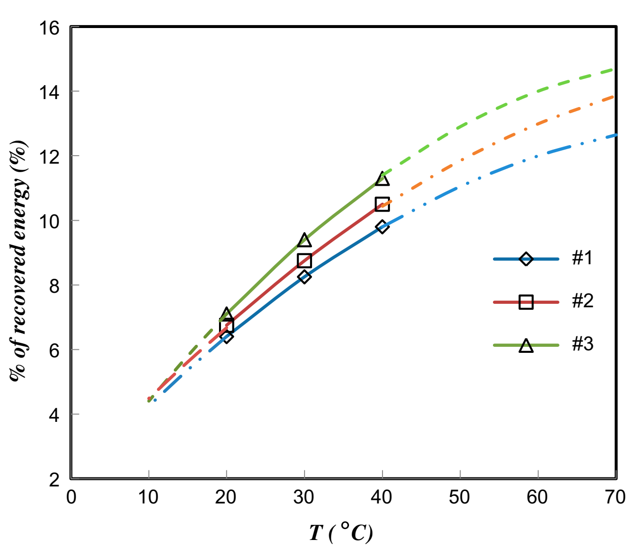

Fig. 6. Percentage of estimated energy for MED unit at different operating temperatures.

(around 10% of the energy can be recovered at 40°C, compared with 7% at 20°C). Dashed lines present the extrapolation of the result for non-studied values of temperature. Extrapolations show that using membranes that can stand with high temperature (no swelling, no collapse) can be very benefitial in terms of energy production using PRO.

7. Conclusions

Preliminary results of the capability of recovering osmotic energy from the brine of MED processes have been presented. Using PRO membranes it is shown that it is possible to recover as hydraulic pressure a significant part of the osmotic energy of the MED brine, and simultaneously reduce the salinity of the discharge. The effect of the temperature was also investigated.

Acknowledgements

This work has been funded by MiCInn POWER project (DPI2010-21589-C05). The authors would like to thank partners of the INES network and Luis Yebra for many helpful discussions.

List of symbols

A	—	water permeability coefficient (m/s/Pa))
B	—	salt permeability coefficient (m/s)
C_D	—	salt concentration of the feed stream (g/l)
$C_{F,m}$	—	salt concentration on the membrane surface at the side of the feed (g/l)
$C_{D,b}$	—	salt concentration of the feed stream (g/l)
$C_{F,b}$	—	salt concentration on the membrane surface at the side of the feed (g/l)
ΔC_m	—	concentration difference on the membrane surface (g/l)
d_h	—	hydraulic diameter of the flow channel (m)
D_D	—	diffusion coefficient of the draw solution (m^2/s)
D_F	—	diffusion coefficient of the feed solution (m^2/s)
J_w	—	water flux that crosses the membrane (m/h)
J_s	—	salt flux that crosses the membrane (g/m^2 s)
K	—	mass transfer coefficient (m/s)
K	—	solute resistivity (s/m)
ΔP	—	transmembrane Pressure (Pa)
$\Delta\pi$	—	difference of osmotic pressure between the draw water and the feed water (Pa)
Q_b	—	the flow of water that crosses the PRO membrane (m^3/s)
Q_d	—	draw water flow with high-salinity (m^3/s)
Q_f	—	feed water flow with low-salinity (m^3/s)
R	—	gas constant (J/mol/K)
Re	—	Reynolds number (−)
S	—	effective Surface of the membrane (m^2)
s	—	structure parameter of the support layer (m)

Sc	—	Schmidt number (−)
Sh	—	Sherwood number (−)
$T_{D,b}$	—	temperature of the draw water bulk (°C)
$T_{F,b}$	—	temperature of the feed water bulk (°C)
W	—	power density (W/m)
W_e	—	power recovery (W)
η_D	—	dynamic viscosity of the draw solution (Pa.s)
η_F	—	dynamic viscosity of the feed solution (Pa.s)
$\pi_{D,m}$	—	osmotic pressure at the surface of the active layer (Pa)
$\pi_{F,m}$	—	osmotic pressure at the surface of the support layer (Pa)
$\pi_{D,b}$	—	osmotic pressure at the draw bulk (Pa)
$\pi_{F,b}$	—	osmotic pressure at the feed bulk (Pa)
t	—	length of the support layer (m)
τ	—	tortuosity of the membrane (−)
ε	—	porosity of the membrane (−)
β	—	van't Hoff coefficient (−)
Ω	—	turbine efficiency (%)

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