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A novel process of humidification-dehumidification with brine recirculation for desalination in remote areas of the world

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ABSTRACT

Among the thermal desalination process, the humidification-dehumidification (HDH) is quite energy demanding due to the presence of circulating air (as water carrier from the saline source to the condensation unit) that increases the required interface areas. Nevertheless, this process shows many unique attributes that are pushing a recent revival of R&D in the field of small-scale solar desalination. Within this framework, we believe in the importance of developing solutions for decentralized production in rural remote areas and suggest an innovative process scheme of HDH with brine recirculation. This process assumes the multiple extractions for the minimization of entropy generation and the internal recirculation of brine, in order to cut the cost of seawater pretreatment and the related environmental impacts. The multiple extraction HDH tower with partial rejection of heat and seawater introduces a new concept in the HDH panorama: the closed-air closed-water (CACW). The process scheme of the HDH with brine recirculation and the related mathematical model are presented in the paper. The degrees of freedom of the process and the way to fix the main parameters under conservative assumptions, are discussed in the following. Moreover, the preliminary results of a sensitivity analysis (such as the condenser surfaces and packing tower height) are shown. Although the paper presents the way to drastically improve the theoretical performances of HDH, a solution with acceptable operating costs and mild operating conditions is presented: the arrangement with three extractions and one stage of condensation (~10 m of humidification and dehumidification parallel towers) for an installed capacity of 50 m^3/day, has a GOR of 3 and a RR of 50% at a maximum process temperature of 65°C.

Keywords: HDH; Multi-stage; Brine recirculation; Variance; Process analysis, Solar desalination

1. Introduction

It is foreseen that, by 2025, over a half of the world's population will suffer from water shortage. Moreover, the climate changes have generated ever-increasing demands for freshwater and imbalance between supply and demand in several zones of the world [1,2]. The water issue, among the most concerning of the 21st century, markedly hits the rural areas (mostly in developing countries).

Centralized water supply is considered the optimal choice since it provides the cheapest service. However, only a half of global population receives the drinking water from a large-scale piped connection in the user's dwelling, plot or yard, with a large disparity between urban and rural communities [3]. Many of the areas, experiencing physical scarcity, are lacking in any kind of technology and facilities for the collection, production and distribution of water. Because they depend on the availability of energy sources, know-how, skilled personnel and suited infrastructures and facilities, a vicious circle inhibits the development. Rural

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and remote areas are particularly disadvantaged because often far away from centralized water supply systems and conventional water sources, and are often not connected to the electric power grid [2,3]. The sub-Saharan area, where around one third of the world population interested by this problem is located, is the most affected region [1].

So far economic interests have represented the main driving force leading the development of desalination technologies. Indeed, the most consolidated examples are large-scale intensive systems requiring a know-how and materials typical of industrialized countries. Drinking water supply in regions with physical water scarcity is generally ensured by means of reverse osmosis (RO) and, in minority, multiple effect desalination (MED) and multi-stage flash distillation (MSF). These methods are implemented in large-scale plants requiring big investments, highly skilled personnel and industrial infrastructures and facilities. For these reasons, these plants generally serve large urban areas, exploiting the plenty of facilities in the area. Although the R&D efforts, these desalination processes require a considerable amount of energy: taking into account the average energy demand of desalination processes (2–6 kWh/m^3), the global desalination capacity (>65 Mm^3/d) need approximately 200 GWh/d, equivalent to 73 TWh per year of electricity, whose only 1% from renewable energy sources [1–3].

To serve remote areas and small-scale water demand, most of the current desalination technologies are not suitable because of their high need for electricity supply and qualified maintenance. Centralized supply is often not technically or economically feasible. Even in the case of centralized water provision, decentralized supply can supplement other water uses in agriculture or at agglomerate of buildings. Moreover, especially in developing countries, high population growth in urban areas often leads to the establishment of informal (frequently illegal) settlements which remain disconnected from the urban supply lines. In this perspective, a low-cost technological solution requiring almost no pre-existing facilities and personnel with low technical skills would cope with the requirements of small communities far from urban areas and would reduce the water dependence of small communities (boosting rural areas development, cutting tensions due to water sharing issues). According to the BAT reported by [2], the projects in rural areas should have the goals of independence and autonomy of the local community and cannot ignore the socio-technical-institutional interdependence. The small scale Humidification-dehumidification (HDH) exploiting low-tech solar thermal energy could be a valid solution in this complex scenario. HDH mimics the natural water cycle of desalination: air as humidity carrier is used as the medium to convert seawater to freshwater in mild conditions (close to atmospheric temperature and pressure). Saline water is heated in the dehumidifier and sprayed down in a column in counter current to an upward draft of air. The humidified air is than conveyed to a condenser: the latent heat associated with the vapour condensation is used to preheat the saline feed water flowing inside the condenser tube [4,5]. This process is flexible (well complies with the dynamic of the solar source) and reliable, it can be realized with relatively cheap materials and simple arrangements and is able to exploit low-grade thermal energy. The higher input of thermal energy (with respect to the MED and MSF) and the difficulties in scaling up represent the main weak points [2].

HDH can be classified based on the cycle configuration: open air open water (OAOW), closed-water open- air (CWOA), closed-air open-water (CAOW). The other classification is based on the heated medium: air-heated or water-heated. First experimental and pilot units were constructed and operated in Iraq [6] and Jordan [7]. Ben Bacha reported many complete experimental and numerical works on solar HDH, evaluating the heat storage capacity, the dynamic performances and estimating the cost of desalted water [8–10]. Similar works, complete of experimental data and mass transfer modelling can be found in [11–14]. In recent years, many researchers suggested the use of multiple extraction for the reduction of local enthalpy pinch (thermal balance) with the minimization of entropy generation to optimize the process configuration [15–17]. A recent review of conventional and innovative HDH processes, is given by Narayan et al. [18]. The first operating plants introducing this concept with a similar approach have been designed by Muller Holst who developed the Multi-effect HDH system, (operated in Oman, Tunisia and Gran Canaria with a potentiality of 1 m^3/day) where air from the humidifier is extracted at various points and supplied to the dehumidifier at corresponding points for temperature stratification and natural circulation, resulting in a small temperature gap to keep the two complementary processes (humidification and condensation) [19,20]. Chafik et al. proposed a multi stage air heated CAOW to maximize the humidity ratio at the exit of the humidifier [21]. Other systems include variable pressure HDH in which water is separated using chambers operating at different pressures [22] and the mechanical compression driven HDH with the "shared-wall" solution claimed by the research group of MIT [23]. All these works have proved the use of HDH devices coupled with low-grade thermal energy (especially from renewable sources including solar, waste-heat recovery and geothermal) with the main advantages of reliability, low operative and maintenance costs, low environmental impacts as well as security and territorial independence (decentralized water production). These literature results showed the feasibility of small scale production (~1–10 m^3/d) with a thermal energy consumption in the range 150–300 kWh/m^3, an additional electrical energy consumption of 2–6 kWh/m^3 and a maximum overall GOR in the range 2–4 [5,16,18]. Although McGovern et al. reported higher performance limits of HDH (air-heated) [15], finite heat and mass transfer surfaces limits the GOR to 4.5 as discussed by He et al., that analysed the air-heated HDH desalination system with plate heat exchangers to recover the waste heat from the exhaust gas [24].

In this work we propose a solution for small-scale desalination with the following technological goals (beside the commonplace improvement of efficiency): i) to minimize the environmental impact by reducing the brine discharge and the treated seawater intake, ii) to use low-cost construction materials and low technology equipment, iii) to ensure long term durability of the installation and easiness of maintenance. A further improvement to HDH is presented in this paper: the HDH with multiple extraction and brine recirculation, with the aim to maximize the use of inlet water and the recovery of the internal energy. This solution, coming from the industrial experience in realizing MSF with brine recirculation [25,26] allows to reduce the water make-up and, consequently, the impacts of the required pretreatment as well as the brine discharge. In turn, the recirculation induces costs cut as well as the production of more distillate from a limited amount of saline water (strictly needed in case of limited availability of the source as groundwater and wastewater). The described process scheme can be named as "closed-air, closed-water" (CACW) because of the high percentage of brine recirculation. The CACW allows to reach much higher concentrations than the single crossing (or once-through) CAOW, resulting into a reduction of the treatment costs and a minimization of the discharge. In this paper, we present the fundamentals of the process analysis and the preliminary results obtained from the sensitivity analysis in order to fix the process fluid temperatures, flow rates, the number of extractions and the stages of condensation with seawater.

2. Process description

The simplified process scheme is depicted in Fig. 1, the humidification tower is on the left side and dehumidification on the right. In this representation, there are three stages of humidification and dehumidification, equal to the number of extraction + 1. The stages between the air extractions are inserted as modules in a tower configuration and connected through a common wall with dampers and demisters. The humidification stages are arranged as packed units, while those for condensation are modular and extractable finned tube bundles. The air enters from the bottom of the tower, passes through the humidification unit flowing in counter current to the sprayed warm water, actually acting as the water carrier: the air drives humidity towards the right tower, releasing the distillate by condensation, and leaving the unit at the bottom. The brine and the freshwater are collected respectively at the bottom of the humidification unit and of the condensers. The 1st stage tube bundle is fed by seawater while the others are fed with the recirculated brine (mixed with the make-up flow). In this way the physical-chemical pre-treatment, needed to reach the maximum temperature (60–80 °C) of the cycle, is destined to the only make-up seawater. The recirculated brine, preheated in the condensers, is sent to the external heat source to reach the maximum temperature imposed by design. This configuration is perfectly compatible with low-tech solar thermal systems with high flexibility and low requirements of control and maintenance as the majority of the other examples of HDH devices available in the literature [7–18].

On these basis, the described process scheme can be named as "closed-air, closed-water" (CACW) because of the high percentage of brine recirculation. The CACW allows to reach much higher concentrations than the single crossing (or once-through) CAOW, resulting into a reduction of the treatment costs and a minimization of the discharge.

3. Modelling

The mathematical process scheme to perform the heat and material balances is represented in Fig. 2. B and G are the brine and air flow-rate; ΔG represent the extractions and Q_{IN} the required external thermal energy power. The seawater SW enters the bottom condenser stages and is partially mixed with the recirculating brine (make-up, SW') to adjust the salt concentration; the SW_R is rejected. The total number of theoretical stages (equal to the number of extraction + 1), is N. The brine recirculation occurs where the seawater is rejected. The total stages are divided for individuating the position of the brine recirculation in the condensation tower; the first ones below the brine recirculation are numbered

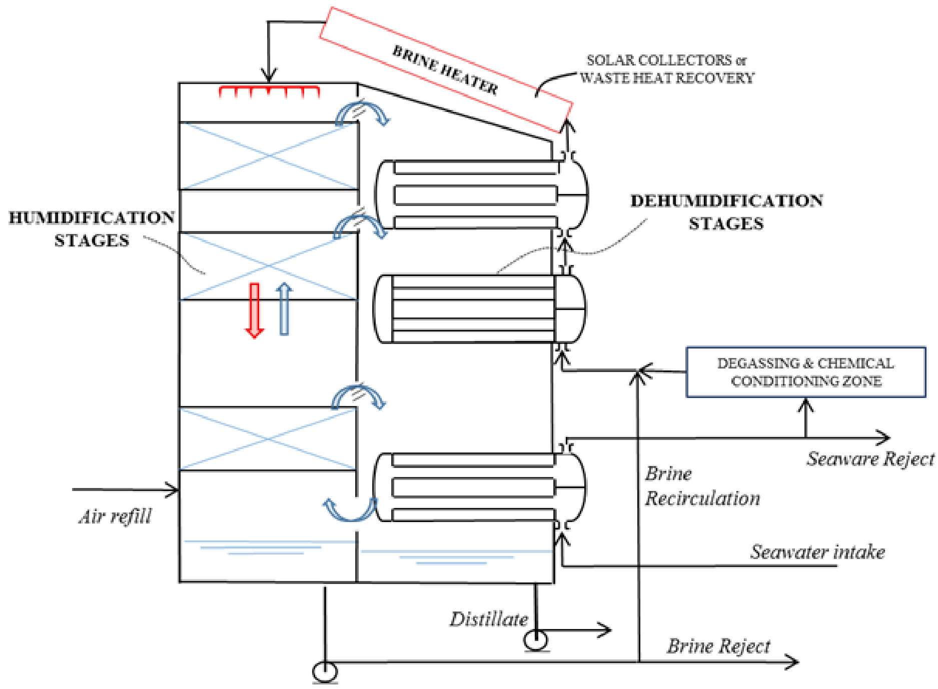

Fig. 1. Process scheme of HDH with brine recirculation.

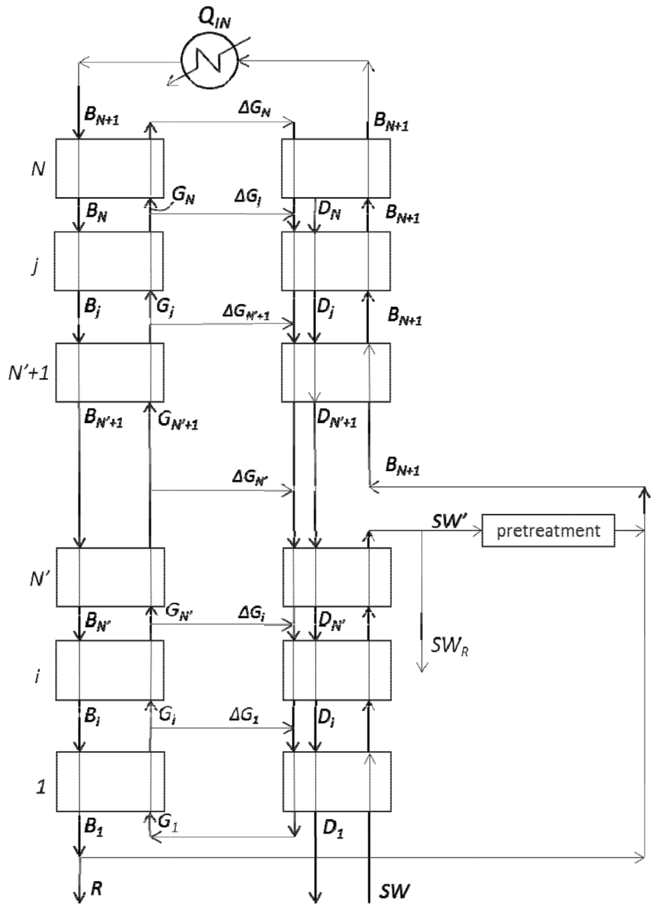

Fig. 2. Schematic for the process analysis of HDH with Brine Recirculation.

from 0 to N'. The process scheme allows for the evaluation of system variance and determination of the governing equations. The variance analysis is introduced to choose the independent variables in the optimization and to define the mass and energy balances The external parameters are fixed depending on the environmental conditions. Once the process variables are fixed, the overall performances (as the GOR in Eq. (1)) are determined.

$$GOR = \frac{Q_{in}}{\lambda(T_{N+1})D_1} \tag{1}$$

where D_1 is the total distillate produced (sum of the distillate obtained from each stage of condensation). The external heat balance allows to calculate the required thermal power Q_{in} by Eq. (2). Enthalpy of the humid air and saline water are calculated according to [27,28].

$$Q_{in} = R \cdot h_R(T_1) + SW_R\, h(T_{N'}) - SW\, h(T_0) + D_1 h(T_1) \tag{2}$$

The independent variables related to the process analysis (Fig. 2) are listed in Table 1. The air flow rate in the humidification stage is equal to the one in the respective dehumidification stage. The temperatures have been indicated with letters t, T and θ respectively for the gas, the recirculating brine and the seawater. Since the air is saturated, it is possible to calculate the enthalpy from the temperature at the top and bottom of each stage (as well as to calculate the temperature form the saturated air enthalpy). The values of the dependent ones can be simply obtained by fixing the external parameters and some of the process and equipment variables to perform the heat and material balances to each stage and to the overall process (Eqs. (3)–(7)).

Table 1
Independent variables of the process HDH with brine recirculation

Equipment Variables	Number of stages (number of extraction +1), N	1
Operation Variables	Position of the seawater condenser (and position of the brine recirculation)	1
	Brine flow-rate profile, B_i	N–1
	Profile of the liquid/gas ratio, B_i/G_i	N
	Air temperature profile, t_i	N
	Brine discharge temperature, T_1	1
	Concentration ratio between R and SW, r	1
	Maximum brine temperature, T_{N+1}	1
External Parameters	Distillate Production, D_1	1
	Temperature of the seawater, θ_1	1
	Salinity of seawater, s	1
Total		$3N$ +7

$$SW' = R + D_1 \tag{3}$$

$$SW' = R \cdot r \tag{4}$$

$$D_1 = \sum_{1}^{N} D_i = \sum_{i}^{N} \Delta G_i \left(Y_{i+1} - Y_i\right) \tag{5}$$

$$G_{i+1} = G_i + \Delta G_i \tag{6}$$

$$B_{i+1} \cdot h(T_{i+1}) + G_i \cdot H(t_i, Y_i) = B_i \cdot h(T_i) + G_{i+1} \cdot H(T_{i+1}, Y_{i+1}) \tag{7}$$

Eqs. (6) and (7) have been written for the N dehumidification stages. These can be applied to the dehumidification as well, by taking into account the effect of salinity on the sensible heat of brine and the portion of heat loss in the distillate. The system variance can be further reduced through additional simplifying assumptions. The extraction temperature is equal to that in the condensation stages (N–1 variables t_i consequently fixed). Moreover, the brine flow rate profile in the humidifier is calculated from the difference of the humidity in the air above and below each stage. The portion of feed seawater is imposed by the concentration ratio [Eq. (4)]. The B/G ratio (liquid/gas in the humidification tower) in each stage has been fixed as proportional to the local derivate of the humid air enthalpy at the saturation conditions [Eq. (8)] with $c_i = 0.9 \div 1.1$ as suggested by a previous thermodynamic analysis [15]. This equation allows to know the operating enthalpy of the air-vapour mixture at any stage.

$$\frac{B_i}{G_i} = c_i \cdot \left.\frac{dH}{dt}\right|_{\bar{t}} \quad (N\,\text{equations}) \tag{8}$$

The independent variables assumed to perform the following sensitivity analysis, are reported in Table 2. The described heat and material balances of each stage can be solved by assuming the external independent variables such as the distillate D_1, the seawater temperature θ_1 and salinity s according to the specific case study. The temperature of the seawater reject $\theta_{N'+1}$ controls the flow rate of sea water SW and can be fixed in function of the fouling in the bottom condenser stages (N'). In this work, the temperature of the SW′ and SWR is around 35°; the temperature of the saturated gas exiting the 1st stage of condensation is calculated by assuming a to limit the surface of the condenser. On the other hand, the brine discharge temperature is fixed from the difference between the saturation and operating curve $\Delta T = T_1 - t_1$ at the bottom of the humidification tower [27]. Fig. 3 shows the graphical representation of this concept in the typical design diagram of HDH. The temperature of the saturated gas in the

Table 2
Values or range of the independent variables to perform the sensitivity analysis

Independent variables	
Θ_1 [°C]	25
T_{N+1} [°C]	65
s [ppm]	35000
r	2
D [m³/d]	50
Position of the recirculation	1st stage
$\Delta T = T_1 - t_1$ [°C]	2–5
N	1–11

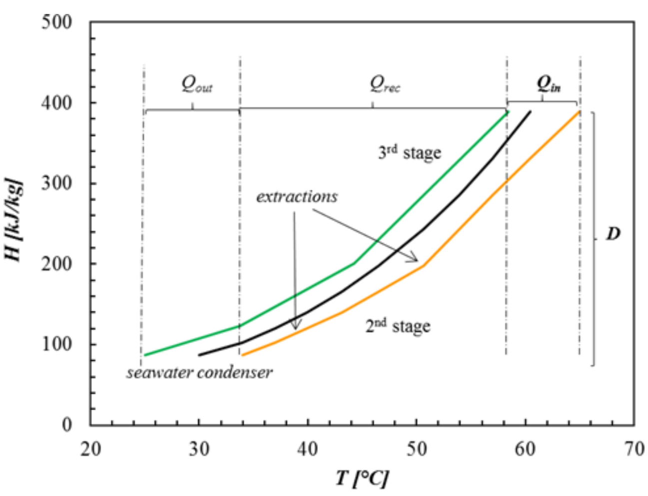

Fig. 3. Sketch of the operating work plane H,T and highlight of the variables to optimize.

dehumidification zone depends consequently on both the number of extractions (N–1) and the pinch between the operating and saturation curves, ΔT. By reducing the temperature differences as well as by increasing the maximum temperature and the number of extraction, it is possible to better approximate the slope of the saturation curve and to obtain a better distribution of the driving forces and the minimization of entropy generation. In this way, the distillate D can be maximized or, equivalently, the required Q_{in} (proportional to the temperature difference between the liquid at the top of the two columns) can be minimized. Some further consideration can be made in the stage of preliminary optimization of the basic design. For instance, the position of the seawater reject gives the number of bottom stages N' and is crucial for the GOR optimization. By moving the 1st stage upward (increasing N'), the condensation surfaces of the dehumidification zone can be reduced but, at the same time, the required Q_{IN} is higher (thus increasing the external energy need and reducing the GOR of the process).

On the other hand, by reducing the pinch, the HDH tower would be higher, so the system would be more complex (i.e. the air flow pressure drop and extraction regulation could be crucial) as well as the plant would be bigger and more expensive. Therefore, for each process configuration being simulated, this work reports the basic design of the heat and mass transfer surfaces: the column height z and the overall heat exchanger surface A (Eq. (9),(10)).

$$z = \frac{G}{SK_y a}\int_1^N \frac{dH}{H_s - H} = H_{tOG} \cdot N_{tOG} \tag{9}$$

$$Q_{in} = U_D \cdot A \cdot LMTD \tag{10}$$

where the packing height can be calculated as the product of the height and the number of the gas-enthalpy transfer units, respectively H_{tOG} and N_{tOG} [27] once the process variance has been determined. The driving force is the difference between enthalpy of the gas at the saturation H_s and at the operating condition at each point of the columns, S is the column section, K_y is the overall mass transfer coefficient and a is the related specific surface. The terms, out of the integral, dependent on the flow mass velocities of the two fluids, varies little (= 0.4–0.7 m) with a specific pressure drop ranging in 120–270 N/m² in the process conditions [27]. Hence, the column height depends on the enthalpy pinch between the saturation and the operating curve. For the evaluation of the required heat transfer surface in the dehumidifier [Eq. (10)], the estimated heat transfer coefficient for the tube finned bundles (21 mm ID) is calculated through for each temperature profile and is in the range 18–30 W/m² K. This coefficient is limited by the transfer of sensible heat (gas cooling), a non-negligible part of the total duty of the dehumidifier stages; temperature differences to calculate the LMTD obtained from any solution in the process analysis. The velocities of the outside gaseous phases are calculated knowing the mass flow rate and the cross sections (2–4 m/s for the gas and around 1 m/s for the water inside the tubes). The coefficient of condensation in presence of vapour and incondensable gases have been estimated from [29] in the case of tube exchangers with extended surfaces (annular fins).

4. Results

Fig. 4 there shows the operating curves of different simulations with increasing number of extractions and increasing maximum temperature of the cycle. The maximum cycle temperature improves the transfer of water into the air and allows for a higher transportation of water into the condensing stages (by increasing the saturation humidity) and, therefore, it reduces the required flowrate of air and seawater to produce the same amount of distillate. The GOR varies from 1 to 3 increasing the number of stages up to $N = 4$. On the other hand, to work at higher temperatures, requires a constant heat source above this limit, in other words a large volume of the thermal storage tank in case of Solar HDH. Moreover, the higher the temperature, the higher the corrosion and scaling phenomena. Durability and functionality versus costs is the balance to be kept. Therefore, in the following sensitivity analysis, a maximum temperature of 60°C has been fixed. Other values of independent variables are previously reported in Table 2. The effect of the number of extraction and of the temperature difference at the bottom has been analysed. A summary of the simulation results is given in Fig. 5.

The lower the ΔT, the higher the GOR of the HDH. By increasing the heat recovery in the condensing/preheating unit, it is possible to reduce the required energy input. The mass flow rate of the carrier-gas mixture is varied by extracting or injecting the carrier-gas mixture from at least one intermediate location in the fluid flow path to reduce the average local enthalpy pinch in the device. The GOR range obtained in this sensitivity analysis is in the upper range of the literature results cited in the Introduction Section, that are limited below GOR = 4.5 [8–10,19–21,23]. The recent researches of the Massachusetts Institute of Technologies [15–18,23] show how it is possible to reach very high energy efficiency by implementing this optimization approaches (GOR→10) with very low pinch point temperature difference. On the other hand, by approaching the saturation curve, the area required for the heat and mass

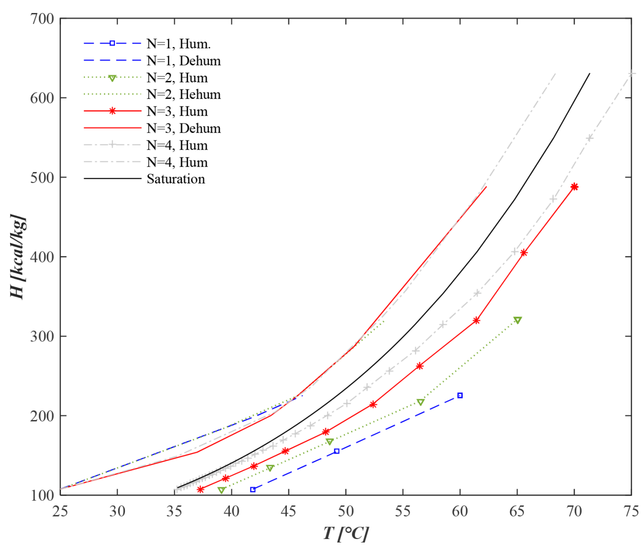

Fig. 4. Operating curves at different number of extractions and maximum temperature of the preheated brine in the work plan H,T.

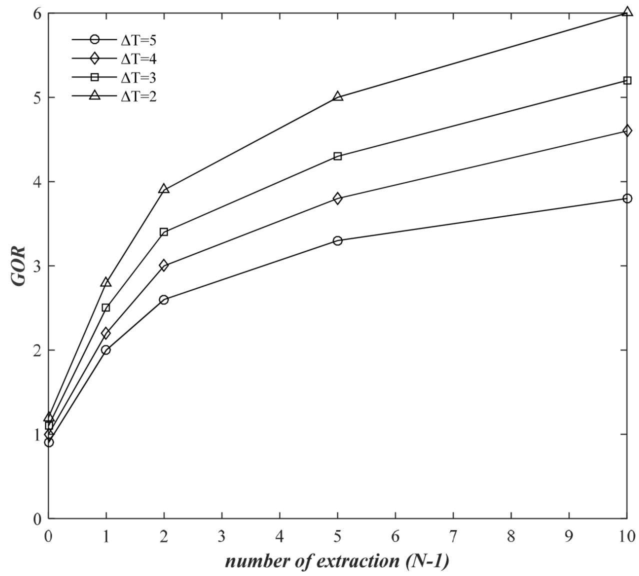

Fig. 5. Summary of the simulation results: independent variables are listed in Table 2.

transfer becomes infinite. In analogy with the humidification pinch analysis, these considerations can be applied for the temperature difference in the condensation zone. The estimation of the tower height (number of theoretical transfer units) and the required heat transfer area (for the results of Fig. 5) are described by the curves of Fig. 6.

In the light of the preliminary results of the sensitivity analysis, a possible plant configuration corresponding to GOR = 3 is presented in the following. In the case of decentralized desalination, the balance between investment and operational costs is one of the main aspects for key decisions regarding two optional strategies to minimize lifetime water costs: low investment or low operational costs [2,18–21]. In these preliminary simulations has been chosen in order to obtain a reliable, economic and simple mechanical configuration. Table 3 summarizes the operating parameters and

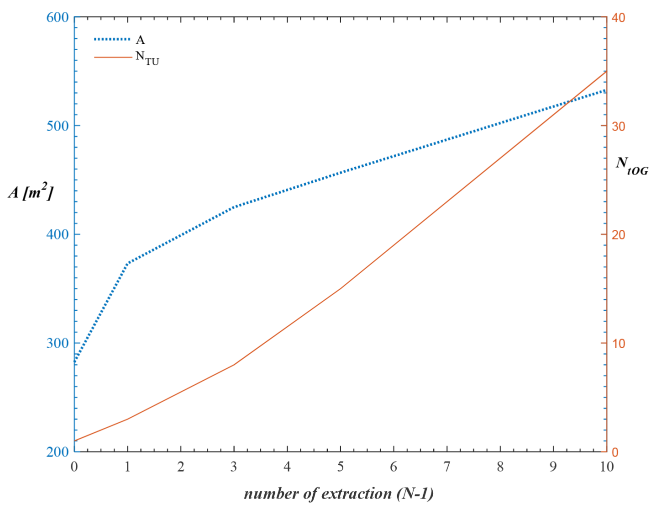

Fig. 6. Number of transfer unit N_{tOG} (representative of the column height) and heat transfer area versus the number of extraction.

Table 3
Preliminary results of process optimization

Dependent variables	
G dry air	27000 kg/h@30°C
SW	45000 kg/h @65°C
Extractions	2 (4500 kg/h @40°C, 10800 kg/h @46°C)
N_{tOG}	9
Q_{IN}	320000 kcal/h
EC	<3 kWh/m^3
GOR	3
RR	50%

the geometrical and process variables of the proposed configuration. Beside the decreasing of the energy dispersion (GOR increasing) by approaching a closed water-closed air system (CWCA), this solution brings the consistent advantage of RR increasing with respect to HDH the state of the art. An RR up to 50% generates a consequent reduction of the chemical treatments and the minimization of the operative costs (lower volume of seawater to be treated) and environmental impacts related to the discharge. In these conditions, the electrical need (EC) has been estimated, including the ventilation and pump power (internal need) and is 2.7 kWh/m^3. With very low maintenance and chemicals/pre-treatments, the plant O&M costs are in the lower range of renewable desalination technologies and can be considered very feasible at this scale. The solution also requires the use of standardized components for cost effective production and the incorporation of transport casing and device containment. A standard container ensuring low freight rates during international transportation can be implemented to carry the main components (condensers and evaporators). This is an essential step towards cost reduction and is made possible by operating at mild conditions (T_{max} = 65°C). By referring to the proposed solution of Tables 2 and 3, in case of an overall installation cost (conservative estimation) of 900 k€ in area of the Northern Africa, a capacity factor of 0.60 and 20 years of depreciation, the installation cost would be around 4 €/m^3, generating a total cost of produced water of 5 €/m^3.

5. Conclusions

In this paper, the process analysis of the HDH with brine recirculation is presented in terms of analysis of freedom degrees and definition of the strategy of process optimization. A preliminary intermediate solution, compromise between operative and capital costs (<3 kWh/m^3, GOR = 3) with the reference capacity of 50 m^3/day can be coupled with a low grade thermal energy source, operating at maximum temperatures between 60°C and 70°C at the atmospheric pressure. Thermal energy can be supplied by low-cost solar collectors, simplifying the O&M and reducing the operating complexity. The low operative temperature allows to minimize the feed pretreatment and to use of low-tech renewable sources or waste energy to supply the process heat duty. The brine recirculation allows to increase the recovery ratio up to 50%. Therefore, the main advantages of the presented decentralized installation are the low maintenance requirements and the minimization of the chemical pretreatment. Furthermore, the supply of low temperature heat as main energy source allows the application of waste heat from generators, where available, or allows the supply of relatively inexpensive solar heat, where sufficient and stable energy supply from the grid is not available. The modular design and the mechanical solutions can bring high flexibility to the process and the potentiality of series production and commercialization (for the capacity range 25–200 m^3/day). For small-middle scale installations, the water cost is relatively low compared to the state of the art (~5 €/m^3).

Symbols

A — Heat transfer surface, m^2
a — Specific interfacial surface based on volume of packed section, m^2/m^3
B_i — Brine flowrate at the stage i, kg/h
D — Flowrate of the distilled water, kg/h
EC — Electrical energy consumption, kWh/m^3
G_i — Flowrate of the gaseous carrier of water (air), kg/h
GOR — Gained output ration, kg/kg
H — Enthalpy of the humid air (dry basis), kJ/kg
H_s — Enthalpy at saturation of the humid air (dry basis), kJ/kg
H_{tOG} — Overall height of the gas-enthalpy transfer unit, m
h — Enthalpy of the saline water, kJ/kg
K_y — Overall gas-phase mass transfer coefficient, kg/(m^{2}s)
N — Number of stages (number of extraction +1)
N' — Number of the bottom stages (seawater condenser)
N_{tO} — Number of overall gas transfer units
t_i — Air temperature at the i-th stage, °C
T_i — Brine temperature at the i-th stage, °C
r — Concentration ratio
RR — Recovery ratio
S — Cross section of the humidification column, m^2
s_i — Water salinity, kg/kg
Θ_i — Temperature of the saline cooling water, °C
Y — Umidity of the humid air (dry basis), kg/kg

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