

Process intensification using energy-free highly enriched air: Application to seawater desalination plants

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1. Introduction

Fresh water is the very foundation of social and economic development [1]. Wherever fresh water is abundant, agriculture and industry easily proliferate with the consequent improvement in the wellness and healthcare for population [2]. Unfortunately, although largely abundant on the earth, only the 3% of the total is fresh water, whereas the rest is salt water. To tackle the potable water shortage of an increasing number of countries, the desalination processes and technologies have received a tremendous attention in last years with billions of m³ of seawater water desalinated per year. The attention is mainly focused on the possibility to make energy-sustainable the seawater desalination process since it is particularly energy-intensive. In fact significant power consumption or significant greenhouse emissions are the unavoidable result of the traditional technologies when fossil fuels are burnt for it. For this reason, many research activities are nowadays related to the integration of renewable energy [3] such as small solar plants for remote locations [4], innovative geothermal-powered systems [5], sustainable solar desalination [6], supply of bio-products [7], renewables for minimizing emissions and energy wastage [8], minimization of carbon footprint in biomass processing [9], optimization of renewable energy supply chains [10] and heat exchange

networks [11] and their layout/operational improvement [12], beet sugar processing [13] to quote a few. To do so, the overall desalination process layout changes in the energy supply system, apart from the technology adopted, whereas the target is still the separation of desalinated water from the brine. Whenever this separation is performed by evaporation, such as in multi-effect distillation (MED) technology, an additional target to intensify the plant can be introduced without additional operational costs and energy consumption and this new technology is the focus of the present research activity. The final aim is the production of a flow of enriched air, i.e. an oxygen/nitrogen mixture with an amount of O₂ in the range of 22–35 vol.%. Enriched air can be directly used in situ reducing the total flow of combustion air and/or the volume of the combustion facilities/burner units. Alternatively, enriched air can be stored in cylinders for diving or sold for industrial/medical uses.

The fundamental concepts which the proposed process intensification is based on are described in Section 2. The process simulation as feasibility study is described in Section 3. The experimental apparatus and results are given in Section 4. At last the industrial benefits are evaluated in Section 5 for the MED desalination process. Sustainability and economic appeal are summarized in Section 6.

2. Fundamental principles

The proposed intensification of MED technology allows to produce enriched air which largely overcomes the medical

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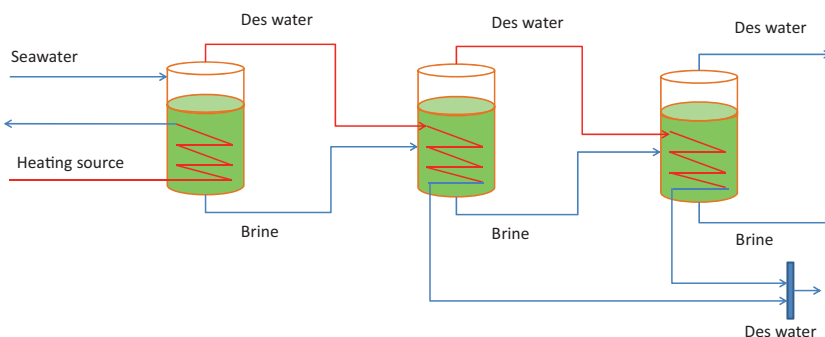


Fig. 1. Traditional MED seawater technology.

specification and the typical industrial uses, up to more than 30 vol.% of oxygen simply exploiting in a different way the energy supply already adopted to partially evaporate seawater. The basic principle which the idea is based on is the different Henry constant of oxygen and nitrogen in water, with the solution of about 6–20 NL of air in 1 m³ of seawater with 65 vol.% of nitrogen and 35 vol.% of oxygen [14]. When the operating conditions change (e.g. temperature increase and/or pressure reduction), the content of N₂, O₂, and other incondensable gases (argon and carbon dioxide) within the water decreases due to their consequent release into the atmosphere. Similarly, when the seawater is heated in an industrial process, the amount of incondensable gases released can be collected for several uses rather than venting them to the atmosphere.

Nevertheless, the heating and evaporation of large mass of water is an expensive operation due to the specific heat and evaporation enthalpy of water and for this reason the process is considered economically unappealing. This is not anymore true whenever the enriched air is obtained from processes that uses water as coolant in open-loop as well as from processes based on evaporating or boiling water such as the MED plants.

The typical MED technology consists of several effects where only the first one is supplied by thermal energy [15], whereas the following effects are operated by recovering the duty of the water vaporized in the upstream effect. It is worth underlining that several technologies are available for the effects from the oldest one adopting tube bundles to the newest one falling film evaporators are used with the seawater pulverized at the top. A qualitative and general scheme is given in Fig. 1. The inlet seawater is supplied to the first evaporator where an external heat source is also supplied. Such a heat can be supplied either from fossil fuel combustion or renewable source or hybrid source. The vapor stream of desalinated water and the liquid stream of brine are fed to the second evaporator. In any case, these two streams are kept segregated to ensure the effectiveness of seawater plant. The desalinated steam of the first evaporator achieves the second effect with slight superheated conditions of temperature for the ebullioscopic contribution. Conversely, the brine passes through a lamination valve. Thus, it achieves the second evaporator cooled and with lower pressure. This allows to use the latent heat of the condensing steam generated in the first evaporator to vaporize another portion of brine, which has the boiling point lower than the dew point of desalinated water. The outlet flows of the second evaporator are the condensed desalinated water that need to be further cooled, the novel steam evaporated within this effect, which is at lower pressure, and the brine with higher salt concentration. The steam is directly sent to the next evaporator, whereas the brine goes to the lamination valve, where temperature and pressure are lowered, and then achieves the next evaporator, to which the same concept above is extended. Only the last evaporator of the MED technology differs from others due to the need of venting the steam generated

inside, in order to ensure the partial vacuum conditions within the overall MED train (i.e. an ebullioscopic condenser is adopted).

As it is broached, the only thermal duty required to initialize the overall plant is the heat supplied to the first evaporator since the other ones are sustained by the upstream steam generations. Therefore, it is possible to split the energy supply for water evaporation in two steps (Fig. 2) to recover the enriched air before the boiling point of water. It is clear that no additional duty is required for this operation since the final water evaporation is the same with respect to the traditional process. In practice, seawater is fed to the first evaporator after a heat recovery train that heats it from the ambient temperature (15–30 °C for Mediterranean Sea conditions) to a temperature close to the operating conditions of the first evaporator. Nevertheless, a certain amount of heat is provided to further increase the temperature and evaporate a quantity of water in the first effect of MED desalination plants. Focusing on the first effect, the seawater partially evaporates and the desalinated water and brine are obtained (Fig. 2). If the duty is supplied in two steps, it is possible to collect the first vapor obtained at lower temperature, which is rich in incondensable gases, and produce the desalinated water at higher temperature in a second step. This process solution finds application in MED desalination plants, but also in steam generation processes, processes with exothermic reactions (e.g. methanol, combustion), and wherever the (fresh or sea) water boils/evaporates. Specifically, chemical/industrial processes that have an open-loop steam cycle (steam reforming, steam cracking, coal/biomass gasification...) could highly benefit from this intensification since they process a large amount of water.

As a matter of facts, the new process layout represents a kind of process-energy intensification of the existing plants since additional added value is obtained simply using differently the same amount of energy. Benefits of the intensification are tangible when the enriched air can be used directly within the plant that produces it, for instance in combustion chambers, partially diluted by air (the oxygen content is particularly high). It allows to reduce (i) the operational costs, (ii) the thermal inertia of air in combustion processes for steam generation, with the unavoidably reduction of fossil fuel consumption, and (iii) the volumes of unit operations.

3. Theoretical feasibility: enriched air process simulation

Several detailed process simulations have been developed as feasibility study and parametric sensitivity for the process to obtain enriched air. The simplest unit operations as well as the most conservative conditions have been selected not to overestimate oxygen concentration and flow of the enriched air stream; in particular, different simulations and thermodynamics have been tested following this purpose. Properties of the seawater of Mediterranean Sea and two different commercial packages well-established in oil refineries [16], petrochemical [17], and gas processing [18], such as PRO/II™ by Simulation Science Invensys and AspenHysys™ by

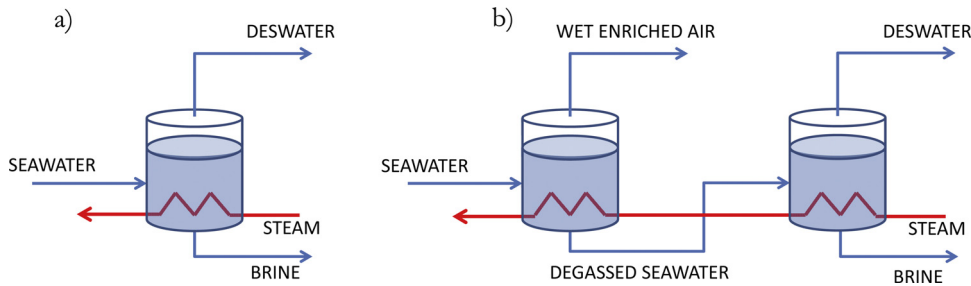


Fig. 2. Traditional (a) and two-stage supplied (b) first evaporator of MED plants.

Aspen Technology, have been adopted to characterize the proposed system. The use of two packages also allows to cross-check the numerical results. The flowsheet developed using the AspenHysys suite, identical to those developed with PRO/II, is reported in Fig. 3. Detailed simulation data are provided in Appendix A. The process simulation consists of two separators, V-101 and V-102, and two heat exchangers, E-100 and E-101. The first unit, V-100, is necessary to simulate the seawater equilibrium with the atmosphere. A blower on the enriched air of stream 8 is also included. According to the conservative approach to the process simulation, in this general simulation, flash drum separators are adopted. The seawater (stream 2) enters the unit E-100 where it is heated to a specific temperature, which is in any case lower than the boiling point. Then, it enters the unit V-102 for the degassing operation to separate the wet flow of incondensable gases (stream 11) from the degassed seawater (stream 5), which is sent to the MED desalination process. The stream 11 is cooled in the unit E-101 to condensate the most of the water within the stream. By doing so, a negligible portion of the incondensable gases are re-absorbed in water (stream 4) since the water amount is significantly smaller than the original one. The condensed phase (stream 6) is separated from the enriched air (stream 7). For instance, at the conditions given in Section 5, the total amount of enriched air is estimated in the order of 1.4 kg/h for 1 m³/h of seawater with over the 32 vol.% of oxygen, which is air highly enriched. Considering the large volumes processed in a typical desalination plant, a relevant flow of enriched air can be recovered with the proposed intensification. It is worth remarking that the energy supply required to operate the intensified process is exactly the same of the traditional desalination process. At last, differences between the two simulators adopted are negligible and results obtained from one of them only are proposed in this paper.

4. Theory validation: experimental activity

An experimental plant has been predisposed to validate the theory. The plant is continuous and it is designed to process a flow till 20L/h of fresh water. It is well-known that fresh water dilutes a different amount of air with different N₂/O₂ composition, but the

differences with respect to seawater are negligible [14]. Thus, fresh water has been used to better characterize properties, operating conditions and parameters for the experimental apparatus. In this plant the water is drawn from a tank of about 25 L, kept open and in contact with the air in order to ensure the dissolution equilibrium at ambient temperature (20 °C) of oxygen and nitrogen. The flow of water (4–20L/h) is sent by a pump to a heater exchange to raise its temperature at a fixed value of 80 °C. At this temperature the high content of oxygen and nitrogen dissolved in the water does not correspond to an equilibrium condition; nevertheless we experimentally verified that it is not easy to strip these gas from water. In others words the simple temperature action is not sufficient to obtain the desired flow of enriched air from the water. Consequently, in order to solve this problem, the hot water flow enters a stripping column for the degassing operation. This column, height 48 cm and with an internal diameter of 4.5 cm, is filled with Sulzer Mellapak CX type structured packing and maintained at 80 °C by an external heating jacket. The flow of water is fed at the top of the column and the stripping gas, helium, at the bottom. Helium is selected to favor the oxygen composition measurement in the enriched air flow. The stripping gas is useful to favor the separation of enriched air and helium has been selected not to spoil the gas chromatograph measurements. The flow outgoing from the column then enters in a cold trap for the separation of the evaporated water and it is finally sent to a micro-gas chromatograph for the on-line quantification of oxygen and nitrogen. These analyses are performed using an Agilent mod. 3000A micro-GC with molecular sieves column, operating with helium as carrier and an isothermal analysis at 80 °C. A picture of the experimental plant is reported in Fig. 4. The experimental results are reported in Fig. 5 and Fig. 6. Different flows of helium as stripping carrier are used and specifically 4 NL/h, 10 NL/h and 20 NL/h, in order to better characterize the behavior of the experimental system. In particular, the progressive enrichment of the oxygen content in the gas outgoing from the column it is reported in Fig. 5, with the following operative parameters: water flow = 10 L/h, He flow = 10 NL/h, water temperature = 80 °C. It is evident that the system reaches a constant oxygen concentration after few minutes. In Fig. 6 the

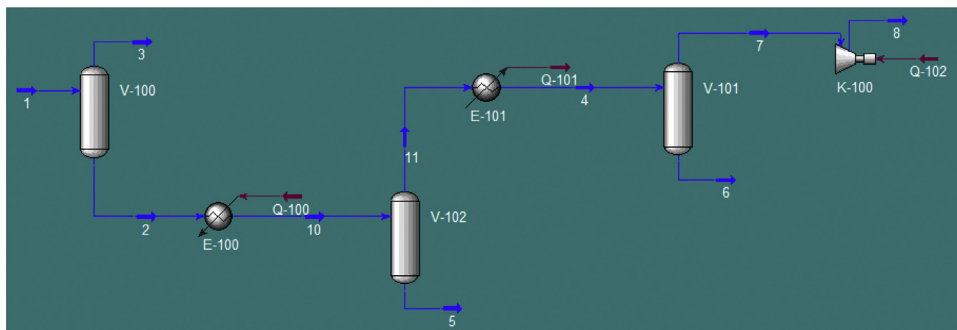


Fig. 3. Process layout for the simulation of enriched air separation and production.



Fig. 4. Experimental plant: stripping column (working temperature = 80 °C) filled with Sulzer Mellapak CX type structured packing for the stripping of the wet flow of oxygen and nitrogen from the degassed water.

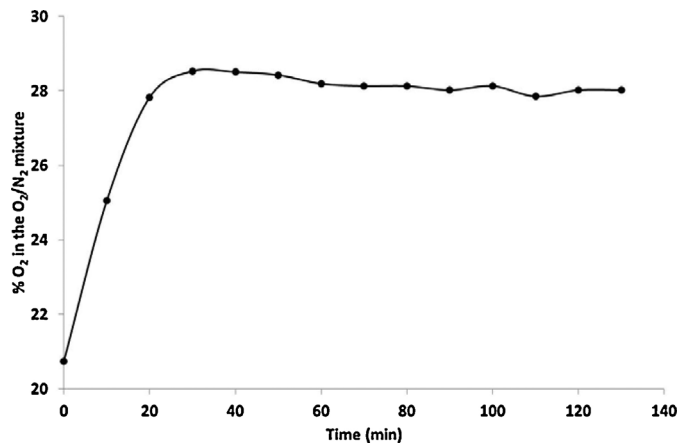


Fig. 5. Oxygen volume percentages of the recovered enriched air vs. time (min). Experimental conditions: water flow = 10 L/h; He flow = 10 NL/h; water temperature = 80 °C.

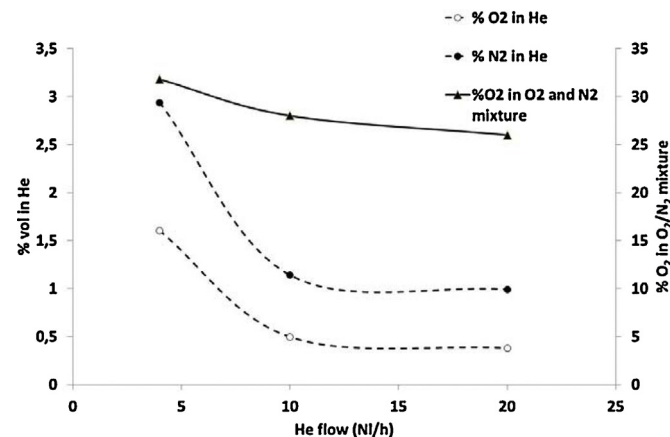


Fig. 6. Oxygen and nitrogen vol.% in helium (primary axis) and Oxygen vol.% in enriched air (secondary axis) vs. Helium flow. $T = 80\text{ }^{\circ}\text{C}$; water flow = 10 L/h.

results obtained using different He flows are compared. Obviously, increasing the carrier flow the oxygen and nitrogen volume percentages decrease, due to a dilution effect (at the same temperature in fact their molar flow outgoing from the hot water is the same). In the same figure the oxygen concentration in the enriched air (O₂/N₂ mixture) is reported too. This concentration should be constant, being not dependent from the helium flow; nevertheless it changes in the range between 25 and 30 vol.%. This phenomenon is explained considering the effect of the cold trap after the stripping column: the water condensed in this point in fact can re-adsorb part of the enriched air and its action depends from the He flow because by increasing this parameter a greater flow of water is condensed in the trap and this water excess decreases the oxygen concentration.

The absolute flow of enriched air obtained in these runs is in the range of 0.05–0.08 NL/h, calculated from the area of chromatographic peaks of nitrogen and oxygen, demonstrating the effectiveness of the proposed intensification.

5. Industrial-scale feasibility

The process to produce highly enriched air can be used to intensify all the industrial processes that have water boiling/evaporation. The proposed process can be used to intensify also the processes that uses large amount of water as coolant, without achieving the boiling point of water. The enriched air can be appropriately stored or used locally. A possible application is the MED desalination plant, especially the plants using fossil fuels in the combustion chamber to generate the steam and supply the first evaporator. Actually, such plants can directly use the enriched air to operate the combustion chamber. In practice, the enriched air must be diluted with air so as not to have more than 28–29 vol.% of oxygen, which leads to the technological limit of operating temperature within the combustion chambers. Whenever higher concentrations of oxygen are obtained, the enriched air can be opportunely diluted with air.

5.1. Seawater MED desalination plant at Pantelleria Island

The typical layout of desalination plants consists of several evaporators where the steam generated in the n th evaporator is used as duty for the $(n + 1)$ th evaporator. The simplified seawater desalination plant designed for civil use of fresh water in Pantelleria Island already described elsewhere [15] is adopted as industrial case study and it is reported in Fig. 7 for the sake of completeness.

It consists of several sections: (I) the thermal energy supply (from fossil fuel or renewable source, e.g. geothermal); (II) the seawater heat integration; (III) the multiple effects (evaporators); (IV) the collection of desalinated water; and (V) the pump section for desalinated and brine waters. The thermal energy supply is constituted by a make-up/purge of water (GEO_LOOP), a mixer (M1) to collect the water within the thermal supply loop and the fresh make-up, a pump (P2) to regulate the flow rate circulating inside it, a heat exchanger (HX.GEO) that characterizes the heat exchange with the fluid within the thermal energy loop, and the heat exchanger (HX.EFF1) included as tube bundle in the first evaporator. The only point of contact between the thermal energy loop and the desalination process is the first evaporator that receives preheated seawater. Actually, the fresh seawater is pumped (P1) into the geothermal desalination plant. It encounters a train of economizers (HX.ECON1, HX.ECON2, HX.ECON3) designed for pre-heating the seawater as well as for cooling the desalinated steam and the brine water before sending these to the tank storage and back to the sea, respectively. The preheated seawater enters the first evaporator (HX.EFF1 and F1) and exchanges the heat with the geothermal steam. The seawater reaches boiling point and a portion of it is boiled and exits from the top of the evaporator and is

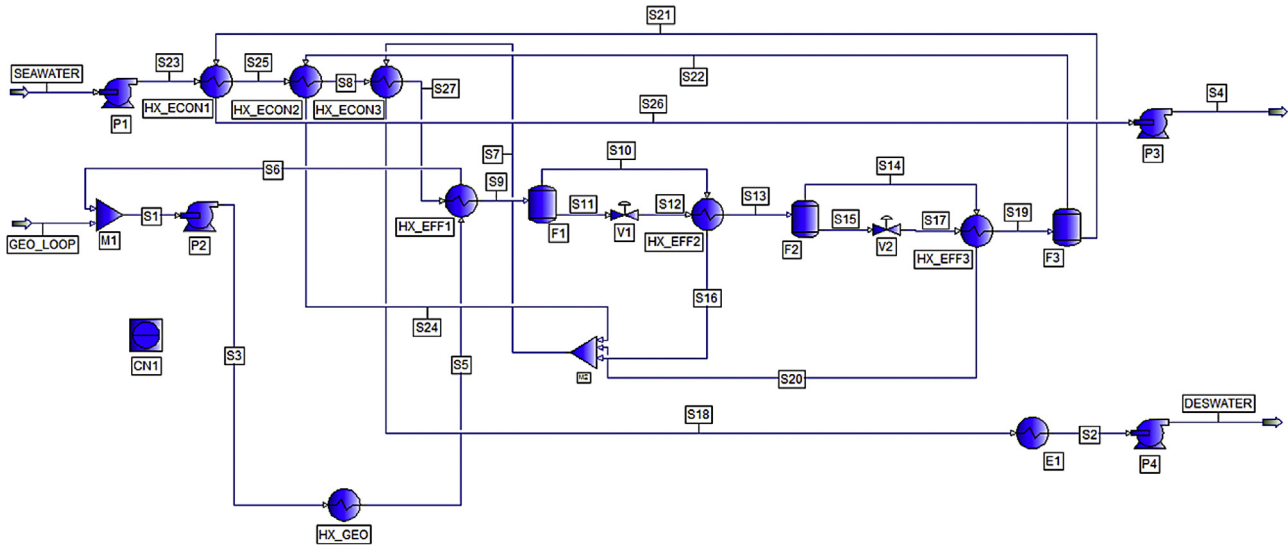


Fig. 7. Simplified process layout of the seawater desalination plant designed for Pantelleria Island [15].

sent to the second evaporator. The vapor is desalinated and slightly superheated for the ebullioscopic effect of boiling seawater. The brine water exits from the bottom of the flash drum separator of the first evaporator, passes through a lamination valve, and enters the second evaporator (HX.EFF2 and F2). The brine water loses pressure within the valve and its temperature decreases slightly, allowing for effective heat exchange with the superheated steam coming from the same effect. The steam is condensed and is the first production of desalinated water that is sent to the collector M2, whereas a portion of the brine water boils further. Similarly to the previous evaporator, the brine water is laminated and sent to the third evaporator (HX.EFF3 and F3), where it exchanges heat with the steam generated in the previous evaporator. The steam is condensed and collected into M2. The procedure is iterated for the evaporators installed downstream. For the sake of simplicity, we considered only three co-current evaporators, knowing that the simulation results can be rearranged easily for the economic evaluation of desalination plants with more effects [19]. The brine water outflowing from the third evaporator preheats the fresh seawater in the unit HX.ECON1. Once it is cooled, it is sent back to the sea by way of a dedicated pump (P3). Conversely, the steam outflowing from the third evaporator is sent to the heat integration train before being collected into M2 together with the previous desalinated water lines. It is worth remarking that in industrial plants, the steam generated in the last evaporator is usually lost in terms of desalination efficiency and water production since the steam is sent to a vacuum chamber for condensation and especially to keep the vacuum in the entire line of evaporators. Since the final steam flow rate is negligible with respect to the total amount of steam generation, and since we have selected a simplified layout (three evaporators) for our study, it is more realistic to collect the final steam generation as desalinated water, thus it is collected in M2. All the streams of desalinated water are sent to the heat integration train to preheat (with the unit HX.ECON3) the fresh seawater fed to the geothermal desalination plant before being pumped (P4) to a dedicated set of water tanks. It is worth considering that additional exchangers could be useful to condense the desalinated water (HX.E1).

The simulation has been solved using PRO/II commercial package [20] with the same stringent assumptions as above. The operating conditions are: seawater at 22 °C with average Mediterranean Sea salinity content (0.35%), inlet flow of seawater 25 kg/s, pressure drops at the second and third evaporators is 0.35 bar,

the final temperature of desalinated water is 51 °C. The simulation provides the following results: 13.09 kg/s of desalinated water, 11.91 kg/s of brine with about double salt concentration according to local restrictions, and 0.034 kg/s of highly enriched air (32.2 vol.% of O₂). The power requirement to supply the first evaporator is estimated in the order of 23.176 MW. Simulation results for traditional MED plant can be found in Manenti et al. [15].

5.2. Sustainable economic considerations

The process above can be improved as reported in the modified layout of Fig. 8 that merges the proposed process to enrich air and the desalination process discussed above. It is assumed that the steam generation to supply the first evaporator of seawater desalination plant is obtained using methane with stoichiometric air in a combustion chamber. The following calculations deal with the performance improvement of the overall plant in terms of reduced flow of combustion air, which is unavoidably related to lower operational costs. Nevertheless, it is worth considering that other important benefits, such as the reduced volumes of the unit operations, are not considered in the present work. The calorific value of methane is assumed to be 55,255 kJ/kg. To generate the power of about 23 MW required to the first evaporator to initialize the overall desalination plant, 0.4194 kg/s of methane are necessary, under the assumption that flue gases have a negligible effect on the general balance (usually, a convection section allows to recover this heat). Also, assuming that methane is completely burnt with stoichiometric air within the combustion chamber according to the general oxidation path:



the total amount of combustion air required is calculated as follows:

- Molar flow of methane:

$$0.4194 \text{ (kg/s)} / 16 \text{ (kmol/kg)} \times 1000 \text{ (mol/kmol)} = 26.2 \text{ (mol/s)} \quad (2)$$

- Molar flow of oxygen:

$$26.2 \text{ (mol/s)} \times 2 \text{ (kmol/kmol)} = 52.4 \text{ (mol/s)} \quad (3)$$

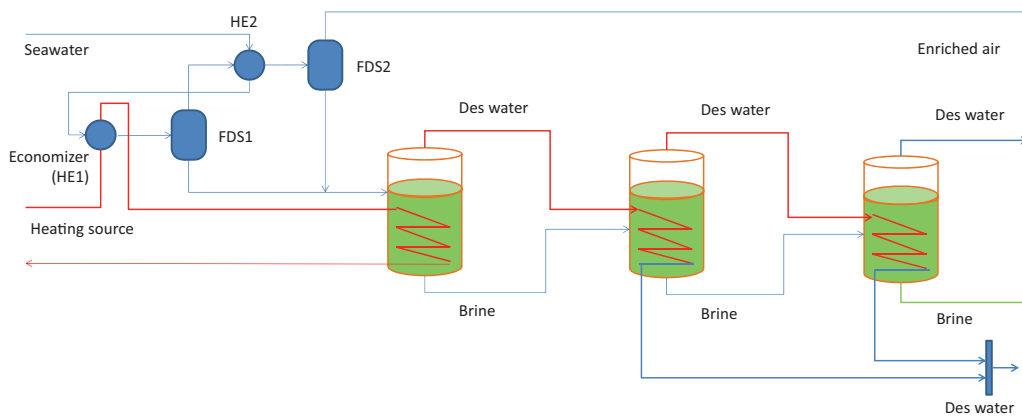


Fig. 8. Intensification IP1 of the seawater desalination plant designed for Pantelleria Island; FDS1 and FDS2 are the selected separation units (in this case flash drums); HE1 and HE2 represents heat exchanger network.

- Corresponding molar flow of nitrogen:

$$\begin{aligned} & 52.4(\text{mol/s})/20.8(\text{mol fract O}_2) \times 79.2(\text{mol fract N}_2) \\ & = 199.6(\text{mol/s}) \end{aligned} \quad (4)$$

- Combustion air molar flow:

$$52.4(\text{mol/s}) + 199.6(\text{mol/s}) = 252(\text{mol/s}) \quad (5)$$

With 0.034 kg/s of enriched air, the reduction of operational costs is calculated as follows:

- Average molecular weight of enriched air:

$$\begin{aligned} & 32(\text{kg/kmol}) \times 0.33(\text{mol fract O}_2) + 28(\text{kg/kmol}) \\ & \times 0.33(\text{mol fract N}_2) = 29.32(\text{kg/kmol}) \end{aligned} \quad (6)$$

- Molar flow of the enriched air stream:

$$0.034(\text{kg/s})/29.32(\text{kg/kmol}) \times 1000(\text{mol/kmol}) = 1.16(\text{mol/s}) \quad (7)$$

- Flow of nitrogen saved with enriched air:

• Nitrogen content of enriched air:

$$1.16(\text{mol/s}) \times 0.67 = 0.777(\text{mol/s}) \quad (8)$$

• Oxygen content of enriched air:

$$1.16(\text{mol/s}) \times 0.33 = 0.383(\text{mol/s}) \quad (9)$$

• Nitrogen content in the corresponding combustion air:

$$0.383(\text{mol/s})/20.8 \times 79.2 = 1.457(\text{mol/s}) \quad (10)$$

• Nitrogen saved:

$$1.457(\text{mol/s}) - 0.777(\text{mol/s}) = 0.68(\text{mol/s}) \quad (11)$$

It means that the performance for the operational costs is apparently low:

$$\frac{252(\text{mol/s}) - 0.68(\text{mol/s})}{252(\text{mol/s})} = 0.27\%$$

making the intensification not economically appealing. Nevertheless, if we consider that the energy supply with fossil fuel is progressively replaced by renewable energy sources, which is the modern trend to make sustainable the desalination processes (i.e. with geothermal or solar source), the contribution of enriched air is much more relevant. For instance, we can suppose that the heat recovery combined with renewable energy sources covers the 90% of the energy needs. It is worth considering that fossil fuel is always in spare to bridge possible gaps due to the discontinuous nature of some renewable energy sources and it is kept operating for prompt actions. In these cases the benefits are in the order of several percent. The overall economic benefit is in the order of 1 million m³/year of air enriched over the 30% in oxygen.

6. Conclusions

The paper demonstrated the possibility to further intensify several processes simply modifying their current energy supply section. Specifically, it is possible to separate enriched air with a high content of oxygen (up to about one third in volume) in all the industrial plants that process a certain amount of water as coolant or for steam generation. The theoretical aspects are discussed and, therefore, applied in specific process simulations and corroborated by experimental data. Finally, the case of a large-scale seawater desalination plant is adopted as case study.

The industrial feasibility and effectiveness has been demonstrated by applying the intensification to the seawater desalination plant adopting the multi-effect distillation technology. Specifically, the case of desalinated water production for the fresh water of municipal use in the Pantelleria Island has been selected and the benefits have been quantified in terms of operational costs.

Acknowledgement

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Appendix A. Appendix A

Simulation data for the process layout of Section 3 and Fig. 3 is provided hereinafter. Specifically, flow rates are given in Table A.1, compositions in Table A.2, and operating conditions of process units in Table A.3.

Table A.1
Process flow rates.

Stream name		S ₁	S ₂	S ₃	S ₄	S ₅	S ₆	S ₇	S ₈
Description phase		Liquid-vapor	Liquid	Vapor	Vapor	Liquid	Vapor	Liquid	Vapor
Total stream properties									
Rate	kmol/h	2655.937	2632.382	23.555	7.216	2625.166	2.495	4.721	2.495
	kg/h	50,383.535	49,709.699	673.835	156.827	49,552.867	71.701	85.125	71.701
Std. liquid rate	m ³ /h	50.000	49.224	0.776	0.167	49.057	0.082	0.085	0.082
Temperature	°C	20.000	20.000	20.000	90.000	90.000	20.000	20.000	86.690
Pressure	atm	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.600
Molecular weight		18.970	18.884	28.607	21.735	18.876	28.739	18.032	28.739
Enthalpy	MJ/h	4.777	4.740	0.037	0.240	18.530	0.004	0.007	0.008
	kJ/kg	94.807	95.347	54.939	1533.241	373.942	57.060	83.975	106.171
Total liquid fraction		0.9911	1.0000	0.0000	0.0000	1.0000	0.0000	1.0000	0.0000
Reduced temperature		0.4385	0.4355	2.0461	0.7676	0.5390	2.0208	0.4535	2.4805
Pressure		0.0047	0.0046	0.0245	0.0064	0.0046	0.0240	0.0046	0.0384
Acentric factor		0.3357	0.3384	0.0413	0.2396	0.3386	0.0413	0.3445	0.0413
Watson K (UOPK)		8.653	8.688	6.070	7.500	8.691	6.013	8.752	6.013
Standard liquid density	kg/m ³	1007.671	1009.877	867.799	939.722	1010.116	878.153	998.700	878.153
Specific gravity		1.0087	1.0109	0.8687	0.9406	1.0111	0.8790	0.9997	0.8790
API gravity		8.784	8.478	31.395	18.928	8.445	29.475	10.045	29.475
Vapor phase properties									
Rate	kmol/h	23.555	n/a	23.555	7.216	n/a	2.495	n/a	2.495
	kg/h	673.835	n/a	673.835	156.827	n/a	71.701	n/a	71.701
	m ³ /h	566.625	n/a	566.625	215.017	n/a	60.015	n/a	46.055
Std. vapor rate	m ³ /h	527.968	n/a	527.967	161.729	n/a	55.920	n/a	55.920
Specific gravity (Air = 1)		0.988	n/a	0.988	0.750	n/a	0.992	n/a	0.992
Molecular weight		28.607	n/a	28.607	21.735	n/a	28.739	n/a	28.739
Enthalpy	kJ/kg	54.939	n/a	54.939	1533.241	n/a	57.060	n/a	106.171
c _p	kJ/kg °C	1.972	n/a	1.972	1.818	n/a	1.956	n/a	1.026
Density	kg/m ³	1.189	n/a	1.189	0.729	n/a	1.195	n/a	1.557
Thermal conductivity	W/m K	0.02520	n/a	0.02520	0.02616	n/a	0.02521	n/a	0.03002
Viscosity	Pa s	0.00002	n/a	0.00002	0.00002	n/a	0.00002	n/a	0.00002
Liquid phase properties									
Rate	kmol/h	2632.382	2632.382	n/a	n/a	2625.166	n/a	4.721	n/a
	kg/h	49,709.699	49,709.699	n/a	n/a	49,552.867	n/a	85.125	n/a
	m ³ /h	49.540	49.540	n/a	n/a	52.611	n/a	0.086	n/a
Std. liquid rate	m ³ /h	49.224	49.224	n/a	n/a	49.057	n/a	0.085	n/a
Specific gravity (H ₂ O @ 60F)		1.0109	1.0109	n/a	1.0111	n/a	0.9997	n/a	n/a
Molecular weight		18.884	18.884	n/a	n/a	18.876	n/a	18.032	n/a
Enthalpy	kJ/kg	95.347	95.347	n/a	n/a	373.942	n/a	83.975	n/a
c _p	kJ/kg °C	3.980	3.980	n/a	n/a	3.998	n/a	4.184	n/a
Density	kg/m ³	1003.423	1003.423	n/a	n/a	941.870	n/a	993.443	n/a
Surface tension	N/m	0.0735	0.0735	n/a	n/a	0.0602	n/a	0.0737	n/a
Thermal conductivity	W/m K	0.48694	0.48694	n/a	n/a	0.46225	n/a	0.55071	n/a
Viscosity	Pa s	0.00120	0.00120	n/a	n/a	0.00036	n/a	0.00102	n/a

Table A.2
Flow compositions.

Stream name Description phase	Component rates							
	S ₁ Mixed	S ₂ Liquid	S ₃ Vapor	S ₄ Vapor	S ₅ Liquid	S ₆ Vapor	S ₇ Liquid	S ₈ Vapor
Temperature (°C)	20.000	20.000	20.000	90.000	90.000	20.000	20.000	86.690
Pressure (atm)	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.600
Molecular weight	18.970	18.884	28.607	21.735	18.876	28.739	18.032	28.739
Component molar rates (kmol/h)								
H ₂ O	2545.409	2544.897	0.511	4.772	2540.126	0.058	4.714	0.058
Oxygen	5.539	0.762	4.777	0.593	0.170	0.591	0.002	0.591
Nitrogen	20.574	2.309	18.265	1.848	0.461	1.844	0.004	1.844
CO ₂	0.008	0.005	0.003	0.002	0.003	0.002	0.000	0.002
Total (kmol/h)	2655.937	2632.382	23.555	7.216	2625.166	2.495	4.721	2.495
Component mole fractions								
H ₂ O	0.9584	0.9668	0.0217	0.6613	0.9676	0.0231	0.9986	0.0231
Oxygen	0.0021	0.0003	0.2028	0.0821	0.0001	0.2370	0.0003	0.2370
Nitrogen	0.0077	0.0009	0.7754	0.2561	0.0002	0.7392	0.0008	0.7392
CO ₂	0.0000	0.0000	0.0001	0.0003	0.0000	0.0008	0.0000	0.0008

Table A.3
Unit operations.

Unit name Type	F1 Flash	F2 Flash	F3 Flash
Unit data			
Temperature (°C)	20.000	90.000	20.000
Pressure (atm)	1.000	1.000	1.000
Pressure drop (atm)	0.000	0.000	0.000
Duty (MJ/h)	0.000	14.031	-0.229
Molar flow (kmol/h)	2655.937	2632.382	7.216
Liquid volume flow (m ³ /h)	49.540	52.611	0.086
Gas volume flow (m ³ /h)	566.625	215.017	60.015

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