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Exergoeconomic optimization of a forward feed multi-effect desalination system with and without energy recovery

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ABSTRACT

The escalating freshwater demand is stimulating the researchers to optimize the performance of desalination technologies. The current study presents the exergoeconomic optimization of a forward feed multi-effect desalination (FF-MED) system under two configurations i.e., conventional MED and MED with energy recovery (MED-ER). A detailed numerical model concerning energy, exergy, and a component-based exergoeconomic analysis is employed to estimate the energy consumption, exergy destruction, and water production cost. Thereafter, the FF-MED-ER system is optimized using a Genetic Algorithm for four different objective functions i.e., maximum gain output ratio (GOR), and minimum specific energy consumption (SEC), exergy destruction, and water production cost. The constraint variables included steam temperature, brine salinity, and the last effect brine temperature. The analysis showed that the incorporation of an energy recovery section increased GOR by 17.9% and decreased SEC and water production cost by 14%, and 10.5%, respectively. Moreover, the optimization improved GOR by 9.26%, decreased SEC by 12.86%, exergy destruction by 12.59%, and the water production cost by 8.25% compared to the standard nonoptimal system.

Keywords: Multi-effect desalination, Genetic Algorithm, exergoeconomic analysis, cost flow method

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1. Introduction

The global water demand is continuously increasing and is expected to surge to 6900 Bm³ by 2030 from 4200 Bm³ in 2015 as shown in Figure 1 [1,2]. Currently, around 2.2 billion people don't have access to potable water and the situation is worsening with an increasing population [3,4], growing industrialization, and high living standards [5,6]. To cope with the escalating global water demand, desalination has emerged as one of the most feasible and practical options [7]. For this purpose, various desalination technologies are operating globally, producing about 95.37 million cubic meters/day of drinking water [8,9]. The share of different desalination technologies globally as well as in the Middle East is summarized in Figure 2 [10]. Among these desalination technologies, the thermal-based systems have attained significant importance in the regions with harsh feeds (e.g., Middle East) [11]. This is because of low pretreatment requirements, high system reliability, and flexibility to handle a variety of feeds [12,13]. However, these systems are energy and cost-intensive and have relatively high CO₂ emissions [14]. Of thermal systems, the MED has gained significant attention in the last few decades because of its high heat transfer capability, operational reliability, flexibility to operate on low-grade heat and integration with other systems (e.g., powerplants, vapor compression units, and adsorption systems etc.) [15,16].

Figure 1. World water demand comparison (Bm³) [2].

Figure 2. Desalination market shares [17,18]

Therefore, plenty of research has been directed on studying and improving the performance of MED systems. El-Dessouky et al. [19,20] studied the effect of feed flow arrangements and reported that the forward feed MED system is better than parallel feed and parallel crossfeed arrangements in terms of cost. Dastgerdi et al. [21,22] proposed the concept of a distributed boosted MED system which increased the waste heat performance ratio by 40% with a product cost between 0.9 and 1.3 \$/m³. Besides conventional standalone systems, the performance of MED integrated with vapor compression/other thermal systems has also been studied. For instance, Al-Mutaz et al. [23] optimized the location of the TVC suction point and revealed that it is more efficient to place it on the middle effects rather than the last effect. Elsayed et al. [24] reported that the exergy destruction in MED-TVC can be significantly reduced by controlling the motive steam pressure which reduces the unit product cost. Fathia et al. [25] combined the MED-TVC with a thermal power plant of Phosphoric Acid production unit and optimized for maximum exergetic efficiency and minimum product cost. Sayyaadi et al. [26] conducted the multi-objective optimization of a MED-TVC system and reported about 20.7% improvement in the water production cost. Ahmadi et al. [27] investigated the MED system operating on the heat rejected by a combined cycle of gas turbine and solid oxide fuel cell and highlighted several benefits from energy, economics, and environmental viewpoints. Likewise, Khalilzadeh et al. [28] modeled the MED system in cogeneration with a wind turbine power plant. They reported that the waste heat from 7.6 MW wind turbine power plant (i.e., 231 kW) can produce a freshwater of 45.069 m³/d. The economic optimization of MED integrated with an absorption heat pump was conducted by Esfahani et al. [29] which revealed that the optimization reduced avoidable cost rates by 33% and exergy destruction by 4.7%.

The literature review reveals that the thermodynamic and economic analysis of MED systems has been carried out extensively. However, an exergoeconomic based optimization of these systems has been seldomly addressed. The exergoeconomic method employs thermodynamic and economic analysis simultaneously to investigate how efficiently each component in the system works. Moreover, the optimization achieves the objective function by re-adjustment of operating conditions without any additional capital investment. Particularly, with the advancement in artificial intelligence, neural networks and numerical techniques (i.e., Sensitivity Analysis, Genetic Algorithm, Particle Swarm Optimization, etc.) have gained attention for the optimization of thermal systems [30–32]. These techniques optimize the system performance by minimizing/maximizing the objective functions through an intelligent selection of design variables from a specified range. It is also important to mention that the optimization is equally important for new designs as well as the existing systems because of the reduction in capital and operational costs.

Keeping in view the above discussion, the current study is focused on exergoeconomic optimization of conventional MED systems. The following research goals are achieved, a detailed performance comparison of FF-MED and FF-MED-ER to highlight the importance of energy recovery section, component level costing through exergoeconomic analysis to determine cost-intensive components of the systems, the effect of an increasing number of evaporators on thermal and economic performance, and system optimization for minimizing product cost, specific energy consumption, specific heat consumption, exergy destruction and maximum system gain output ratio using Genetic Algorithm (GA). It is important to emphasize that GA is employed for optimization because it is the most classic, probabilistic, robust, and best known evolutionary technique for optimizing the nonlinear, complex, and multi constrained models [33–36]. Moreover, its successful application on thermal systems like heat exchangers is well established [37,38] and relatively easier using simulation software even with nominal application skills, and no repeated trial and error process is required [39].

2. System description

In this study, a 50.5 kg/s water production capacity forward feed multi-effect desalination (FF-MED) system is considered in two configurations i.e., configuration-I: conventional MED and configuration-II: MED with energy recovery. The prime reason for the selection of forward feed configuration includes economic and operational advantages. Jamil et al. [12] compared the exergoeconomic performance of different feed flow arrangements (i.e., forward feed, parallel feed, and parallel crossfeed) and reported that the forward feed arrangement required the least investment cost and heat transfer area. Besides, it is also reported that the FF has smaller pumping power compared to other feed flow arrangements [40]. Moreover, FF has also shown superiority over others for the operating temperature limits with the highest top brine temperature thus allowing the maximum number of effects [41]. The process description and schematic diagrams for both configurations are given below.

2.1. Configuration-I: Conventional MED

The conventional forward feed MED system consists of evaporators, condensers, a cooling water pump, a feed pump, a distillate pump, and a brine pump as shown in Figure 3 (a). The evaporator feed is preheated to a set temperature in the condenser and sprayed on the tubes of the first evaporator. The brine rejected from the first evaporator is utilized as a feed for the next evaporator and the vapors of each effect serve the purpose of heating steam to the subsequent effects. The brine rejected in the last effect is discarded back to the sea. The distillate product is collected from all effects and is finally taken, stored, and bottled. The design and process parameters for this configuration are taken from a recent study by Elsayed et al. [40] for validation purposes.

2.2. Configuration-II: MED with energy recovery section

In this configuration, two additional plate heat exchangers are integrated with the system as shown in Figure 3 (b). These preheaters heat the intake seawater by recuperating heat from the distillate and brine streams which are wasted in the former system. Therefore, this configuration offers benefits from an energy and environmental sustainability viewpoint by minimizing energy losses and low-temperature brine disposals [42]. Steam from an external source (which can be a boiler, power plant exhaust, process plant, renewable energy source, etc.,) is used to produce vapors in the first effect. Low-temperature brine disposal lowers the entropy generation in bringing the discharge streams to the ambient temperature as depicted in Figure 4 [43]. The streams are considered in mechanical equilibrium with the environment but not in thermal equilibrium and are

127 brought to thermal equilibrium by irreversible heat transfer to the environment (heat reservoir). In
 128 Figure 4, $\dot{Q}_{ER}$ is the heat transferred to the feed water in the energy recovery section while $\dot{Q}_{drain}$
 129 is the heat retained by discharge streams. This means that $\dot{Q}_{ER}$ the portion is harnessed from the
 130 high-temperature brine and distillate streams and a smaller $\dot{Q}_{drain}$ remains the cause of entropy
 131 generation in the environment. In case of absence of the heat recovery section (refer 2.1), the
 132 rejected heat equals the sum of $\dot{Q}_{ER}$ and $\dot{Q}_{drain}$ which is the cause of incommodious entropy
 133 generation by the system.

Figure 3. Schematic diagram, (a) configuration-I and (b) configuration-II with preheaters.

Figure 4. Heat recovery and disposal for the thermal equilibrium of discharge streams [43].

3. Thermodynamic model

The above-described systems are the combination of heat exchangers and pumps. A detailed design for the critical analysis of these systems based on the laws of thermodynamics is presented in the following sections.

3.1. Heat exchangers design

Heat exchangers (HXs), are of vital importance in the design process of thermal desalination

systems. The plant's thermodynamic and exergoeconomic performance is significantly influenced by HXs design and they also affect the capital investment. So, keeping in view the significance, a comprehensive design of HXs used in the system of Figure 3 (b) is given.

Figure 5. Falling film evaporator.

The heat transfer rate $\dot{Q}$ of a HXs can be calculated as Eq. (1) where U is the overall heat transfer coefficient of evaporators, A is the heat transfer area, and $LMTD$ is the logarithmic mean temperature difference.

$$\dot{Q} = U \times A \times LMTD \quad (1)$$

The heat transfer within the evaporator faces three types of resistances i.e. cold side heat transfer coefficient, hot side heat transfer coefficient, and tube wall resistance as shown in Figure 5. These resistances are solved for the calculation of the overall heat transfer coefficient U using Eq. (3) [44].

$$\dot{Q}_{Ev,k} = \dot{m}_{s,k} h_{fg,s,k} = \dot{m}_{F,k} C_{pF} (T_{B,k} - T_{F,k}) + D_k h_{fg,k} \quad (2)$$

$$\frac{1}{UA} = \frac{1}{(hA)_c} + \frac{R_{f,c}}{A_c} + \frac{1}{(hA)_h} + \frac{R_{f,h}}{A_h} + R_w \quad (3)$$

Correlations for h_h , h_c and R_w for falling film evaporators and preheaters are given in

Appendix B. Fouling resistances of cold and hot side $R_{f,c}$ and $R_{f,h}$ for evaporators are 88×10^6 m² K/W and 1.7×10^6 m² K/W, for brine preheater 17×10^6 m² K/W and 52×10^6 m² K/W and distillate preheaters 52×10^6 m² K/W and 1.7×10^6 m² K/W respectively [45]. Other design parameters for preheaters are taken from [46].

The flow of cooling water is extracted from Eq. (4).

$$\dot{Q}_{Cond} = \dot{m}_{15} C_p (T_{16} - T_{15}) = \dot{m}_{13} C_p (T_{13} - T_{17}) + \dot{m}_{13} h_{fg} \quad (4)$$

For brine and distillate preheaters, energy balance is given as [47]:

$$\dot{Q}_{BH} = \dot{m}_3 C_p (T_5 - T_3) = \dot{m}_7 C_p (T_7 - T_9) \quad (5)$$

$$\dot{Q}_{DH} = \dot{m}_2 C_p (T_4 - T_2) = \dot{m}_8 C_p (T_8 - T_{11}) \quad (6)$$

The general formula to calculate *LMTD* for all heat exchangers is given by:

$$LMTD = \frac{(T_{h,o} - T_{c,i}) - (T_{h,i} - T_{c,o})}{\ln \left(\frac{(T_{h,o} - T_{c,i})}{(T_{h,i} - T_{c,o})} \right)} \quad (7)$$

Heat transfer areas of all the HXs (evaporators, condenser, brine preheater, and distillate preheater) are calculated separately from Eq. (1). Hence, the specific area of the whole configuration becomes:

$$A_{spc} = \frac{A_{EV} + A_{BH} + A_{DH} + A_{CD}}{\dot{m}_D} \quad (8)$$

The brine salinities of all effects are calculated from Eq. (9) and Eq. (10) for boiling point elevation (BPE) calculation as it is the function of effect temperature and brine salinity.

$$B_k \cdot S_{B,k} = F_k \cdot S_{F,k} \quad (9)$$

$$B_k = F_k - D_k \quad (10)$$

$$BPE = T_{B,k} - T_{V,k} \quad (11)$$

3.2. Pumps work

The power requirement of the pumps corresponding to the work done can be calculated by:

$$\Sigma \dot{W}_{pump} = \sum_{j=1}^n \frac{\dot{V}_j \Delta P_j}{\eta_{pump}} \quad (12)$$

Where $\dot{V}_j$ is the volume flow rate through j^{th} pump measured in m^3/s and ΔP_j is the pressure gradient of the respective pump in kPa.

3.3. Gain output ratio (GOR)

It is one of the most widely adopted performance parameters to describe how effectively the designed system works. The value of GOR is calculated as Eq. (13). Where D is the total distillate flow rate in kg/s and $\dot{m}_s$ is the mass flow rate of inlet steam in kg/s .

$$GOR = \frac{D}{\dot{m}_s} \quad (13)$$

3.4. Specific heat consumption (SHC)

Another important performance parameter is SHC which is defined as the amount of heat energy required to generate 1 kg of distillate. For a fixed temperature of steam, it is calculated by dividing vapor enthalpy (kJ/kg) at a saturation temperature of steam (normally around 50°C) by GOR [48] as given by Eq.(14).

$$SHC = \frac{2330}{GOR} \quad (14)$$

3.5. Specific energy consumption (SEC)

The concept of specific energy consumption combines input energies at different grades given to the desalination system and provides a single value that can be used as $\dot{W}_{in}$ for the analysis. Where,

$$\dot{W}_{in} = \Sigma \dot{W}_{pump} + \dot{W}_{st,eq} \quad (15)$$

$\dot{W}_{st,eq}$ is defined as the electricity generated in the steam turbine power plant by the same steam source that is injected into the desalination system if it were expanded in the steam turbine.

$$\dot{W}_{s,eq} = \dot{m}_s (h_{s,tur,in} - h_{s,tur,out}) \eta_G \quad (16)$$

Hence, $\dot{W}_{st,eq}$ it takes the enthalpy of steam at source temperature and at the turbine outlet

temperature which is normally 35°C for the Rankine cycle power plants. The isentropic efficiency of the steam turbine and electrical generator is considered to be 85% and 95% respectively [49,50]. Thus, the thermal energy is translated into the equivalent electrical energy which can now be simply added to the pumps' power $\Sigma \dot{W}_{pump}$, and finally the energy in kW is converted to kWh/m³ as under

$$SEC = \frac{\dot{W}_{in}}{3600 \dot{V}_p} \quad (17)$$

3.6. Exergy analysis and 2nd law efficiency

Second law analysis is based on exergy balance and it states that the difference of the summation of fuel and product exergy equals the sum of exergy destruction and exergy loss.

$$\Sigma \dot{X}_{product} = \Sigma \dot{X}_{fuel} - \dot{X}_D - \dot{X}_L \quad (18)$$

The exergy of fuel is the inlet stream exergy and that of product is only distillate exergy (useful product). The system exergy destruction refers to the lost available work due to associated irreversibility and exergy loss refers to lost available work due to rejected streams that carry exergy [43]. The second law efficiency becomes the ratio of exergy of useful product to the exergy input for the system that is only thermal and electrical input. The input seawater carries no exergy as stream exergy at dead state is zero. So, second law efficiency is given as:

$$\eta_{II} = \frac{X_{product}}{\dot{W}_{in}} \quad (19)$$

4. Exergoeconomic model

For each component, an exergoeconomic analysis is performed for both configurations as described above. The cost flow method is applied to the system as given by [51–55]. In this method, the cost of each flow stream of the system through all components is calculated by combining the fixed cost and stream cost. The fixed cost, also known as the current cost or capital cost, is calculated by correlations given in Appendix A [56,57]. The inputs are given to the exergoeconomic model like interest rate (i), amortization factor (AF), cost index factor (C_{index}), electricity cost (C_{elec}), and others are listed down in Table 1. The stream costs are based on the exergy of the respective stream, hence exergetic cost, the equations of which are listed down in Table 2.

231

Table 1. Input data for exergoeconomic analysis.

Parameter	Value	Ref.
Amortization period, t , years	30	[41]
Interest rate, i , %	5	[41]
Cost index factor, C_{index}	1.2	[58]
Electricity cost, C_{elec} , \$/kWh	0.06	[40]
Annual availability, %	90	[58]
Steam heating cost, \$/kJ	1.466	[41]
Chemical cost, $\dot{C}_{Chemical}$, \$/m ³	0.025	[40]

232

233

Table 2. Stream cost rates.

Component	Correlations	Eq. no.
Distillate preheater	$\dot{C}_4 = \dot{C}_2 + \dot{C}_8 - \dot{C}_{11} + \dot{Z}_{DH}, \frac{\dot{C}_8}{\dot{X}_8} = \frac{\dot{C}_{11}}{\dot{X}_{11}}$	(20), (21)
Brine preheater	$\dot{C}_5 = \dot{C}_3 + \dot{C}_7 - \dot{C}_9 + \dot{Z}_{BH}, \frac{\dot{C}_3}{\dot{X}_3} = \frac{\dot{C}_5}{\dot{X}_5}$	(22), (23)
Brine pump	$\dot{C}_{10} = \dot{C}_9 + \dot{W}_{BP} + \dot{Z}_{BH}$	(24)
Evaporator	$\dot{C}_v = \dot{C}_{F,EV,i} + \dot{C}_{S,EV,i} - \dot{C}_{cond,i} - \dot{C}_{B,i} + \dot{Z}_{EV}$	(25)
Distillate pump	$\dot{C}_{12} = \dot{C}_{18} + \dot{W}_{DP} \dot{C}_{electricity} + \dot{Z}_{DP}$	(26)
Condenser	$\dot{C}_{15} = \dot{C}_0 + \dot{W}_{CWP} \dot{C}_{elec} + \dot{Z}_{CWP}, \frac{\dot{C}_{13}}{\dot{X}_{13}} = \frac{\dot{C}_{17}}{\dot{X}_{17}}$ $\dot{C}_{16} = \dot{C}_{15} + \dot{C}_{13} - \dot{C}_{17} + \dot{Z}_{CD}$	(27), (28), (29)

234 After the fixed cost is calculated, all costs are converted into the rate of fixed cost by taking the
 235 product of fixed cost and amortization factor (AF). A separate value for each component is
 236 calculated and finally merged into the exergetic costs of streams of the respective component as
 237 given in Table 2.

$$238 \quad Z_{Annual} = Z_{comp} AF \quad (30)$$

239 Where,

$$240 \quad AF = \frac{i(1+i)^t}{(1+i)^t - 1} \quad (31)$$

241 Here, t represents the plant life in years and i is the interest rate. Now we have the components
 242 cost in \$/year (annual fixed cost). As the present MED system uses cost in \$/s, so the actual cost is
 243 given as below [59]:

$$244 \quad \dot{Z}_{actual} = \frac{Z_{Annual}}{365 \times 24 \times 3600 \times Availability} \quad (32)$$

245 The stream cost rate is calculated by applying the fuel, product principle to all components of
 246 the system. This principle says that the cost rate of the product equals the summation of fuel cost
 247 rate and fixed cost rate. Here, the term fuel refers to inlets to the component and product refers to
 248 the outlets from the same component. The general formula is shown by Eq. (23) which is equal to
 249 the sum of the fixed cost of components and the cost of fuel [58].

$$250 \quad \sum \dot{C}_p = \sum \dot{C}_f + \dot{Z}_{Comp} \quad (33)$$

251 Based on the exergy of the stream entering and leaving the components, auxiliary equations
 252 are developed. Combining these equations, stream costs are calculated as given in Table 2 refers
 253 to the developed costing model.

254 Similarly, the live components that use electricity input also accommodate electricity usage in the
 255 cost stream. For example, the calculation of feed pump stream cost which passes through it is equal
 256 to the fixed cost of the feed pump, electricity cost, and intake stream cost.

$$257 \quad \dot{C}_1 = \dot{C}_0 + C_{elec} \dot{W}_{FP} + \dot{Z}_{FP} \quad (34)$$

258 Finally, the product cost in \$/s is calculated by summing up the exergoeconomic cost of
 259 distillate $\dot{C}_{12}$, brine $\dot{C}_{10}$, cooling water $\dot{C}_{16}$, chemical cost, input steam cost $\dot{C}_{14}$, operational &
 260 maintenance cost OM, and labor cost LB as given by Eq. (25) Equation based on Figure 3(b).

$$261 \quad \dot{C}_D = \dot{C}_{12} + \dot{C}_{10} + \dot{C}_{16} + \dot{V}_{12} C_{chem} + \dot{C}_{14} + OM + LB \quad (35)$$

262 Final Product Cost in \$/m³ is then calculated using Eq. (26).

$$263 \quad \dot{C}_P = \frac{\dot{C}_D}{\dot{V}_D} \quad (\$/m^3) \quad (36)$$

5. Numerical solution and assumptions

Steady-state models are developed in Engineering Equation Solver (EES) for both configurations of the MED system. The function of these models is described by a model flow map in Figure 6. Mass balance is applied to the system to provide intermediate brine salinities and their flow. While energy balance provides the flow rates of cooling water, steam, and distillate. The input values given to the system are given in Table 3.

The generated steady-state models are based on the following assumptions. (a) the intake seawater pressure for each configuration is $P_o = 101.325$ kPa, salinity $S_o = 35$ g/kg and temperature of intake is $T_o = 22.5$ °C, (b) the system operates on a steady-state condition, (c) heat transfer per unit area for all the evaporators is the same, (d) thermophysical properties of vapor are temperature and pressure-dependent and that of seawater is salinity and temperature-dependent, (e) BPE effect is non-negligible, (f) product salinity is assumed as zero, and (g) pressure losses in pumps and heat losses in heat exchangers are negligible [60,61].

Table 3. Input data for thermodynamic analysis.

Parameter	Value
Number of effects, N	4
Intake seawater temperature, $T_{sw}, ^\circ C$	22.5
Intake seawater pressure, P_{sw}, kPa	101.325
Intake seawater salinity, $S_o, g/kg$	35
Brine salinity, $S_7, g/kg$	53
Plant capacity, $\dot{m}_D, kg/s$	50.5
Feedwater temperature for FF-MED, $T_F, ^\circ C$	31.5
Inlet steam temperature, $T_s, ^\circ C$	56
Last effect evaporation temperature, $T_B, ^\circ C$	45.4
Brine to sea temperature for FF-MED-ER, $T_v, ^\circ C$	25
Distillate temperature for FF-MED-ER, $T_D, ^\circ C$	25

Figure 6. Flow map of the steady-state model including optimization.

6. Genetic Algorithm

The genetic algorithm is an evolutionary technique used for parametric optimization of the systems since 1975 [62]. The method is one of the best techniques applied to optimize the nonlinear, complex, and multiple constrained problems. The technique is based on the Darwinian Theory of evolution i.e. survival of the fittest in which the population is naturally selected for bringing up the new generations based on their fitness [63]. This means that individuals with higher fitness values are favored for cross-over than those with lower fitness. The mutation is caused by arbitrarily in the genome of an individual to cause a real change in the upcoming generation [33]. The mechanism of survival selection is similar to the parent selection mechanism i.e. fitness level [34]. The working

sequence of this algorithm is concisely shown in Figure 7. Every design variable of the system is a gene (genotype) with its length being the chromosome. Each stage contains a set of values of the design variables in a population and gives rise to a new generation. The new generation becomes a population for the subsequent crossover and so forth. The process continues till the required optimized population is met which is fittest of all [64].

Decision variables

The length of decision variables is selected according to the flexibility of the system in which it can be operated. In the present study, the assorted decision variables are the operating parameters which include inlet steam temperature, the last effect brine temperature, and brine salinity. The upper and lower constraints of these parameters are given as under:

$$\text{Inlet Steam Temperature, } T_s \text{ }^{\circ}\text{C [65]} \quad 55 \leq T_s \leq 70$$

$$\text{Last Effect Salinity, } S_{B,N} \text{, ppm [40,45]} \quad 50 \leq S_{B,N} \leq 70$$

$$\text{Last Effect Brine Temperature, } T_{B,N} \text{ }^{\circ}\text{C [26]} \quad 42 \leq T_{B,N} \leq 48$$

Using this optimization criterion within the prescribed ranges, the four effect FF-MED-ER system is optimized to maximize GOR, minimize SEC, minimize exergy destruction, and to minimize product cost. The system was run for 100 generations, 3334 iterations with 0.025 mutation rate. The results converged after 35 generations for GOR, 65 for SEC, 50 for exergy destruction, and 19 for the product cost.

Figure 7. Genetic Algorithm working sequence.

6. Results and discussion

7.1. Model validation

The FF-MED system without energy recovery is first validated with a system presented by Elsayed et al. [40]. The simulation was conducted under the same operating conditions for thermodynamic analysis. The system is based on four number of effects with forward feed configuration. The proposed model has good agreement with results presented by Elsayed et al. within $\pm 0.187\%$ deviation as shown in Table 4. It shows that model assumptions and parameters are reasonable, and it can be used for further investigation.

Table 4. FF-MED model validation

Parameter	Literature [40]	Current Model	% Error
Distillate production (D), kg/s	50.49	50.5	± 0.019
Gain output ratio ($GOR=D/\dot{m}_s$)	3.5	3.507	± 0.002
Brine flow (B), kg/s	98.006	98.19	± 0.187
Feed flow (F), kg/s	148.5	148.7	± 0.001
Cooling seawater flow rate, kg/s	824.726	823.2	± 0.185
Specific heat transfer area (S_A) $m^2/kg/s$	143.86	143.9	± 0.027
Specific heat consumption (SHC), kJ/kg	665.71	664.5	± 0.181

7.2. Performance comparison of configuration I & II

The significance of the heat recovery section is justified by comparing the performance and economic benefits obtained by its integration with the conventional MED systems. Both systems are operated under the same operating conditions as given in Table 3.

The performance improvement starts from a 35% increment in feed temperature for FF-MED-ER by recuperating the waste heat. The high-temperature feed in the first effect reduces the required flow rate of steam because of less sensible heating requirements before evaporation for fixed steam temperature. This resulted in a 17.9% increase in GOR and a 15.1% decrease in SHC of FF-MED-ER as given in Table 5. The comparison of SEC which is a function of electrical and thermal input shows a 13.8% decrement for FF-MED-ER. As far as individual values of electrical and thermal inputs are concerned, the electrical input slightly increases for FF-MED-ER however, the sheer reduction in thermal input dominates the SEC minimization which results in an overall

improvement of 13.8%. The specific heat transfer area is reduced by 0.42% despite the inclusion of the pre-heaters area. This reduction in area successively saves the evaporators fixed cost which is the main issue of hybrid systems [66].

Table 5. Improvements by using energy recovery system: a comparison

Parameter	FF-MED	FF-MED-ER	% Improvement
Feed temperature, T_f °C	31.5	42.5	34.92
Steam flow rate, m_s kg/s	16.56	14.05	9.11
Feed flow, m_f kg/s	148.7	148.7	-
Brine out temperature, T_b °C	45.4	25	44.93
Distillate out temperature, T_d °C	44.97	25	44.4
Brine flow, m_b kg/s	98.19	98.19	-
Gain output ratio, GOR	3.049	3.595	17.9
Specific heat consumption, SHC kJ/kg	764.1	648.1	15.1
Specific energy consumption, SEC kWh/m ³	11.86	10.22	13.82
Specific heat transfer area, S_A , m ² /kg/s	590.1	587.6	0.42
Product cost, \$/m ³	3.154	2.823	10.49
System exergy destruction, kW	3016	2859	5.2

The unit product cost, on one hand, is reduced from 3.154 \$/m³ to 2.823 \$/m³ (10.5% reduction) with the addition of preheaters due to the lowering of fixed cost. On the other hand, the lower exergetic cost of product and discharge streams and lower steam cost consolidate with this 10.5% depreciation. Besides the performance and economic benefits, FF-MED-ER also claims environmental sustainability benefits. The recuperation of brine and distillate streams energies in the energy recovery section lowers the stream temperatures by 50% and 45% respectively.

7.3. Exergy analysis

A component-based exergy analysis is performed to analyze both proposed configurations from the exergy destruction viewpoint. The proportion of exergy destruction covered by each component is shown in Figure 8. The total system exergy destruction of FF-MED and FF-MED-ER as presented in Table 5 are 3016 kW and 2859 kW respectively, out of which condensers show the highest value followed by the evaporators collectively for both of the configurations. For FF-MED, the exergy destruction values are 1727 kW, 1278 kW, and 10.96 kW for condenser, evaporators,

and pumps respectively. Similarly, for FF-MED-ER, these values are 1733 kW for the condenser, 979.8 kW for evaporators, 69.44 kW for brine preheater, 64.79 kW for distillate preheater, and 11.5038 kW for pumps. The substantial difference in the component's exergy destruction value is justified by the irreversibilities associated with phase change which is occurred in evaporators and condensers only [45]. The peak values for condenser gives the idea of using hybrid technologies downside of the MED like MED-AD as proposed by Shahzad et al. [16].

Figure 8. Component-based exergy destruction, (a) configuration I and (b) configuration II.

The exergy of the product (distillate stream) coming out of the system can be calculated from Eq. (18). The value for FF-MED comes out to be 307.4 kW which is lowered to 138 kW with energy recovery in FF-MED-ER. The corresponding values for $\dot{W}_{in}$ also decrease from 2171 kW to 1864 kW sequentially however, the second law efficiency is seen more sensitive to product stream exergy reduction in the present study. Hence the system shows the degradation of 47.7% in second law efficiency.

7.4. Exergoeconomic analysis

A comprehensive component-wise exergoeconomic analysis is carried out to check how much every component contributes to the final product cost. The monetary comparison of the two configurations is well explained in Figure 9 (a) and (b) in which all the component's costs and their respective stream costs are presented in \$/h. The fixed cost of HXs is a function of a specific heat transfer area, so evaporators take the highest value and condenser takes the second highest in both configurations. Evaporators cost in Figure 9 corresponds to the sum of costs of all evaporators being

operated. The fixed costs of pumps used in FF-MED are given as 0.6896 \$/h, 0.848 \$/h, 0.1768 \$/h, 0.2972 \$/h and that of FF-MED-ER are 0.6843 \$/h, 0.8986\$/h, 0.1768 \$/h, and 0.2972 \$/h for cooling water pump, feed pump, brine pump, and distillate pump, respectively. The product cost comes out to be 3.154 \$/m³ and 2.823\$/m³ for FF-MED and FF-MED-ER respectively. The prime reasons for the reduction in cost include exergetic cost recovery in preheaters, a decrease in sensible heating requirements, and lower fixed costs. The only rise is shown by the cooling water which can be accommodated well by hybridization as explained earlier. But to avoid the huge investments in the hybrid system, mere adjustment of operating parameters can prove more suitable for further cost reduction as given under.

Figure 9. Cost Flow of MED (a) configuration I and (b) configuration II.

7.5. Effect of number of evaporators

The analysis done so far is based on four number of effects, but as in literature, the higher number of effects bring about a progressive influence on thermodynamic performance as well as unit product cost [45,67]. Hence, the discussed systems have also been critically analyzed for the number of effects up to eight. Figure 10 (a) through (c) presents the comparison of FF-MED and FF-MED-ER first law performance i.e. GOR, SHC, and SEC. Similarly, Figure 11 (a) through (c) presents the comparison of second law analysis in terms of second law efficiency, system exergy destruction, and product cost for varying number of effects. Figure 10 (a) shows a gradual increase in GOR with the number of effects increasing from 3 to 8. This is because of the smaller thermal input requirement for fixed capacity plant. An increasing number of effects will provide better utilization of heat input into more recoveries hence a smaller heat source is required by the system. This reduction of required steam leads to the gradual lowering of SHC Figure 10 (b), and SEC Figure 10 (c) with a higher number of effects.

(a)

(b)

Figure 10. Comparison of configuration I & II performance versus the number of effects (a) GOR, (b) SHC, and (c) SEC.

387 Figure 11 (a) shows an increasing trend of second law efficiency with higher number of effects
 388 due to the gradually decreasing input exergy requirement. The system total exergy destruction is
 389 lowered as shown in Figure 11 (b) with a greater number of effects. The component-based analysis
 390 reveals that the major reduction of exergy destruction is seen in condensers as shown in Figure 12
 391 (a) and (b). For FF-MED the exergy destruction value is highest for the condenser up to 6 number
 392 of effects however, for seven or more effects, the value dominates for evaporators (Figure 12 (a)).
 393 In FF-MED-ER, the whole dominance is shown by the condenser throughout the variation from 3
 394 to 8 effects (Figure 12 (b)).

Figure 11. Comparison of configuration I & II performance versus the number of effects (a) second law efficiency, (b) system exergy destruction (c) product cost.

This shows that configuration II is more suitable for the downstream hybridization than configuration I with 7 or more number of effects. As far as the trend is concerned, the value for condenser reduces from 2312 kW for 3 effects to 866 kW for 8 effects system. Other components also show a decrease however, the decrement is small. The economic advantage from 3.608 $\text{\$/m}^3$ to 2.752 $\text{\$/m}^3$ for FF-MED and from 3.318 $\text{\$/m}^3$ to 2.342 $\text{\$/m}^3$ for FF-MED-ER as shown in Figure 11 (c) is partially due to evenly falling steam cost and also cooling water cost. The specific heat transfer area and corresponding fixed cost increases but these increments are less significant than exergetic stream costs reduction hence, an overall drop in the product cost is seen.

Figure 12. Exergy destruction trend with increasing effect numbers (a) configuration I and (b) configuration II.

7.6. Optimization to maximize GOR

Comparison of the two configurations above from thermodynamic as well as economic aspects demonstrates the supremacy of FF-MED-ER over FF-MED. Hence, optimization is done only for FF-MED-ER for assorted operating conditions.

As mentioned earlier, a typical desalination system performance is evaluated by GOR. The system is optimized for a maximum value of GOR by finding out a combination of operating parameters without any design changes. The optimization results are given in Table 6 which shows that GOR is increased by 9.3%. This is because a higher temperature feed as compared to the reference case enters the first evaporator that reduces steam flow. This reduction in steam

progressively improves the SHC and SEC by 8.47% and 11.82% respectively. The second law efficiency is improved by 13.45% due to reduced exergy destruction by about 12.06% compared to the reference case. This optimization shows considerable thermodynamic benefits however, economic benefits are less significant as product cost improves by just 0.14%.

Table 6. Optimization to maximize GOR

Parameter	Reference case	Opt. (maxi. GOR)	Improvement %
Inlet steam temperature, T_s °C	56	55.23	1.37
Last effect brine temperature, T_{bN} °C	45.4	47.95	5.61
Feed temperature, T_f °C	42.5	45.66	7.43
Steam flow rate, m_s kg/s	14.05	12.86	8.46
Outlet brine salinity, S_b g/kg	53	65.22	23.05
Feed flow, m_f kg/s	148.7	109	26.69
Brine flow, m_b kg/s	98.19	58.49	40.43
Gain output ratio, GOR	3.595	3.928	9.26
Specific heat consumption, SHC kJ/kg	648.1	593.2	8.47
Specific energy consumption, SEC kWh/m ³	10.22	9.011	11.82
Second law efficiency, %	7.405	8.401	13.45
Product cost, \$/m ³	2.823	2.819	0.14
System exergy destruction, kW	2859	2514	12.06

7.7. Optimization to minimize SEC

The system is also optimized for the lowest SEC value. The results in Table 7, show that the SEC is lowered by 12.9% which occurred due to the lowering of required steam input and a minor portion of this decrement is due to less feed being pumped into the system. This is due to concentrated brine coming out of the system that means a small feed amount is separating the required distillate which ultimately reduces the pumps' work. The subsequent improvements are seen in GOR by 9.4% and SHC by 8.7%. Second law efficiency is improved by 14.8% as less thermal and electrical input is required for the same product exergy. The last effect brine temperature rise and relatively lower steam temperature distribute a small temperature drop per effect which increases the specific area by 44%. This causes product costs to rise by 0.8%.

Table 7. Optimization to minimize SEC

Parameter	Ref. case	Opt. (mini. SEC)	Improvement%
Inlet steam temperature, T_s °C	56	55.01	1.76
Last effect brine temperature, T_{bN} °C	45.4	47.98	5.68
Feed temperature, T_f °C	42.5	45.64	7.38
Steam flow rate, m_s kg/s	14.05	12.83	8.68
Outlet brine salinity, S_b g/kg	53	65.46	23.50
Feed flow, m_f kg/s	148.7	108.5	27.03
Brine flow, m_b kg/s	98.19	58.03	40.90
Gain output ratio, GOR	3.595	3.936	9.48
Specific heat consumption, SHC kJ/kg	648.1	591.9	8.67
Specific energy consumption, SEC kWh/m ³	10.22	8.905	12.86
Second law efficiency, %	7.405	8.5	14.78
System exergy destruction, kW	2859	2490	12.90

431 7.8. Optimization to minimize system exergy destruction

432 The system exergy destruction is minimized up to 12.6% as given in Table 8 due to lower operating
 433 temperatures than the reference case. Besides temperature, an 8.6% lesser steam flow is also
 434 providing lesser exergy to the evaporators. The salinity of brine coming out of the system is
 435 increased by 23.54%. Concentrated brine corresponds to smaller feed demand to harness the
 436 distillate and correspondingly smaller vapors flow to the condenser which destructs exergy the
 437 most. Due to the major contribution of the condenser in exergy destruction, a significant reduction
 438 in overall system exergy destruction is seen. This minimized exergy destruction improves second
 439 law efficiency by 14.27%. Other improvements are recorded as 9.4% in GOR, 8.5% in SHC, and
 440 12.5% in the SEC. The specific heat transfer area increases by 42% due to the smaller temperature
 441 drop per effect which increases product cost by 0.53%.

Table 8. Optimization to minimize system exergy

Parameter	Ref. case	Opt. (mini. system exergy)	Improvement %
Inlet steam temperature, T_s °C	56	55.1	1.60
Last effect brine temperature, T_{bN} °C	45.4	47.9	5.50
Feed temperature, T_f °C	42.5	45.62	7.34
Steam flow rate, m_s kg/s	14.05	12.84	8.61
Outlet brine salinity, S_b g/kg	53	65.48	23.54
Feed flow, m_f kg/s	148.7	108.5	27.03
Brine flow, m_b kg/s	98.19	57.99	40.94
Gain output ratio, GOR	3.595	3.933	9.40
Specific heat consumption, SHC kJ/kg	648.1	592.4	8.59
Specific energy consumption, SEC kWh/m ³	10.22	8.946	12.46
Second law efficiency, %	7.405	8.462	14.27
System exergy destruction, kW	2859	2499	12.59

444 7.9. Optimization to minimize product cost

445 After that the thermodynamic optimization is elaborated in detail, the economic side is also
 446 highlighted by cost minimization. This cost minimization is solely based on varying the operating
 447 parameters rather than the conventional way of varying the economic factors like interest rate, cost
 448 index factor, or cost of electricity, chemicals, labor, and maintenance. The reference case and
 449 optimization results are compared in Table 9. This table shows that the product cost is improved by
 450 8.25% due to a 38% reduction in the area which is the only cost controlling variable in this case.
 451 This is because the reduction in product cost has occurred at the expense of SEC, system exergy
 452 destruction, and second law efficiency which declines by 41%, 30%, and 29% respectively.
 453 However, SHC and GOR slightly improve by 2.2% and 2.25% respectively as compared to the
 454 reference case due to high-temperature source in a smaller amount. The required optimized feed
 455 temperature increases that also justifies the requirement of preheaters as designed in the current
 456 study.

457 The whole optimization results show that three out of four scenarios deal with thermodynamic
 458 improvements (Table 6, Table 7, and Table 8) and the fourth one deals with the economics of the
 459 desalination system in Table 9. The two (thermodynamic and economic) paradigms, however,

operate at contrasting operating conditions and either one can be opted according to the site requirement.

Table 9. Optimization to minimize product cost

Parameter	Ref. case	Opt. (mini. product cost)	Improvement %
Inlet steam temperature, T_s °C	56	67.14	19.89
Last effect brine temperature, T_{bN} °C	45.4	47.99	5.70
Feed temperature, T_f °C	42.5	48.73	14.65
Steam flow rate, m_s kg/s	14.05	13.74	2.20
Outlet brine salinity, S_b g/kg	53	63.88	20.52
Feed flow, m_f kg/s	148.7	111.7	24.88
Brine flow, m_b kg/s	98.19	61.2	37.67
Gain output ratio, GOR	3.595	3.676	2.25
Cooling seawater flow, m_{cw} kg/s	849.3	846	0.38
Specific heat consumption, SHC kJ/kg	648.1	633.8	2.20
Specific heat transfer area, $m^2/kg/s$	587.6	363.5	38.13
Product cost, $\$/m^3$	2.823	2.59	8.25

8. Conclusion

The current study presents a exergoeconomic optimization of a forward feed MED system operating with and without energy recovery. The effect of various operating parameters on the system performance is investigated. Genetic Algorithm is employed to achieve optimal design from energy, exergy, and economic viewpoints. The major findings of the study are as follows.

- The integration of the energy recovery section with MED significantly increased the system performance. About 17.9% increase in GOR, 15.1% decrease in SHC, 13.8% in SEC, 0.42% in the heat transfer area, and a 10.5% reduction in $\dot{C}_P$ is observed due to energy recovery. Moreover, the distillate and brine stream temperatures decreased by 45% and 50%, respectively which is an additional benefit from an environmental perspective.
- The actual values of SHC, SEC and $\dot{C}_P$ are observed to be varying from 933 to 511 kW, 14.4 kW to 8 kW and 3.6 to 2.75 $\$/m^3$ for FF-MED and 829 to 386 kW, 13 to 6.2 kW, and 3.31 to

2.37 \$/m³ for FF-MED-ER, respectively when the number of effects changes from 3 to 8.

- In the first optimization case (maximum GOR), the value of GOR increased by 9.3% compared to the reference system and SEC decreased by 11.82% and $\dot{C}_P$ by 0.14%.
- In the second optimization case (i.e., minimum SEC), the value of SEC decreased by 12.9% which shows an advantage of GOR by 9.4%, however, $\dot{C}_P$ increased by 0.8%.
- In the third optimization case (minimum exergy destruction), the exergy destruction decreased by 12.6%, increased GOR by 9.4%, thus decreasing SEC and $\dot{C}_P$ by 12.5% and 0.53%, respectively.
- Lastly, the optimization (minimum water production cost) decreased $\dot{C}_P$ by 8.25% and increased GOR by 2.25%.

Overall, it is concluded that the energy recovery section not only proves well for a sustainable environment but also increases the thermal and economic performance of the system. A higher number of evaporators are economically more feasible for the regions where equipment purchasing is less expensive even for higher electricity rates due to smaller SEC. The optimized operating parameters can control the performance as well as product cost without any system design changes.

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498 **Nomenclature**

A	area (m^2)
B	brine
$\dot{C}$	the rate of monetary cost (\$/s)
$\dot{C}_p$	final product cost (\$/m ³)
C_p	specific heat ($kJ/kg.K$)
C_{index}	cost index factor
h	heat transfer coefficient
i	interest rate
k	thermal conductivity
L	tube length
$\dot{m}$	mass flow rate (kg/s)
Nu	nusselt number
P	pressure (kPa)
Pr	prandtl number
$\dot{Q}$	heat transfer rate, kW
Re	Reynold's number
R_f	fouling resistance
R_w	wall resistance
S	salinity (g/kg)
T	temperature ($^{\circ}C$)
U	overall heat transfer coefficient (kW/m^2-K)
$\dot{V}$	volume flow rate (m^3/s)
W	work
$\dot{X}$	exergy (kW)
Z	fixed cost (\$)
$\dot{Z}$	rate of fixed cost (\$/s)
<i>Greek letters</i>	
Δ	change in quantity
ρ	density
Υ	feed mass flow per unit length of each side of tube
η	efficiency
<i>Subscripts</i>	
B	brine
C	cold side
BH	brine heater
CD	condenser
$Comp$	component
CWP	cooling water pump
$chem$	chemical

D	distillate
DH	distillate heater
EV	evaporator
elec	electricity
eq.	equivalent
F	feed
FP	feed pump
G	gas turbine
<i>f</i>	fuel
H	hot side
in	input
<i>i</i>	inlet condition
k	effect number
L	lost
N	no. of effects
<i>O</i>	outlet condition
P	product
SW	seawater
<i>v</i>	vapor
Spc	specific
<i>s</i>	steam
tur.	turbine
<i>W</i>	water
II	second law
<i>Superscripts</i>	
<i>t</i>	plant life in years
<i>Abbreviations</i>	
AF	amortization factor
FF-MED	forward feed multi-effect desalination
FF-MED-ER	forward feed multi-effect desalination energy recovery
BH	brine heater
BP	brine pump
CWP	cooling water pump
CD	condenser
DH	distillate heater
DP	distillate pump
FP	feed pump
GOR	gain output ratio
LMTD	log mean temperature difference
BPE	boiling point elevation
SEC	specific energy consumption
SHC	specific heat consumption
SEE	single effect evaporation

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Appendix

Appendix A

Table 10. Correlations for the fixed cost of components.

Component	Equation	Ref.	Comments
Pump	$Z_{Pump} = 32 \times 0.435 \times \dot{m}_{water}^{0.55} \Delta P^{0.55} \left(\frac{\eta}{\eta - 1} \right)^{1.05}$	[56]	$\Delta P (kPa)$ $\dot{m} (kg/s)$
MED effect	$Z_{EV} = 201.67 \times UA_{Evap.} \Delta P_t^{0.15} \Delta P_s^{-0.15}$	[56]	$U (kW/m^2.k)$
Preheater	$Z_{PH} = 1000(12.86 + A_{HT}^{0.8})$	[57]	plate & frame heat exchanger
1.1 Condenser	$Z_{Condenser} = 430 \times 0.582 UA_{Condenser} \Delta P_t^{-0.01} \Delta P_s^{-0.01}$	[56]	$A (m^2)$
Steam cost	$C_{Steam, Annual} = \frac{(1.466 \times h_{fg, steam} \times Availability \times 365 \times \dot{V}_{destillate})}{1000 \times GOR}$	[41]	steam heating cost(\$1.466/MJ)

691 **Table 11.** Overall heat transfer coefficient correlations.

Parameter	Equation	Ref.
Preheaters wall resistance	$R_{w,PH} = \frac{t_{plate}}{Ak}$	[68]
Preheaters heat transfer coefficient	$Nu_{PH} = C_h Re^n Pr^{\frac{1}{3}} \left(\frac{\mu}{\mu_w} \right)^{0.17}$ $C_h \text{ and } n \text{ are constants}$	[46]
Evaporator wall resistance	$R_w = \frac{1}{2\pi kL} \ln \left(\frac{d_o}{d_i} \right)$	[68]
Evaporator cold side heat transfer coefficient	$Nu_{ev} = \frac{h_{c,ev} \left(\frac{\mu_l^2}{\rho_l^2 g} \right)^{\frac{1}{3}}}{k_l} = 0.009 Pr^{0.65} (q)^{0.15} Re^{0.2}$	[69]
	$Re = \frac{4\dot{\Upsilon}}{\mu}$	
	$h_{c,ev} = \Phi \int_{\Upsilon_o}^{\Upsilon_l} Re^{0.2} d\Upsilon$	
	$\Phi = 0.009 k_{min.} \left(\frac{\mu_l^2}{\rho_l^2 g} \right)^{\frac{1}{3}} Pr^{0.65} (q)^{0.15}$	
Evaporator hot side heat transfer coefficient	$h_h = \frac{0.023 Re_l^{0.8} Pr_l^{0.4} k}{d_i} \left(0.55 + \frac{2.09}{p^{0.38}} \right) \left(\frac{\mu_l}{14\mu_g} \right)^{(0.0058+0.557P)}$	