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A comprehensive framework for thermoeconomic analysis of desalination systems

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ABSTRACT

Thermoeconomic analysis, a combined application of thermodynamic and economic analyses, has emerged as an important tool to optimize the performance of desalination systems. Contrary to conventional economic analysis, it offers flexibility to investigate and improve the performance of each component in the system, individually.

The current paper presents a comprehensive framework for conducting thermoeconomic analysis of desalination systems. In this regard, different energy calculation methods are discussed first. Then a detailed review of theoretical developments of thermoeconomic analyses is conducted to summarize the correlations/magnitude of important economic parameters. This is followed by a discussion on cost balance equations for important desalination components. Finally, a systematic thermoeconomic analysis model is developed for the mechanical vapor compression desalination system operating under different arrangements as an example. The monetary value of each stream calculated using appropriate fiscal parameters in the system is presented in the form of a cost flow diagram. The study can be used to conduct the thermoeconomic analysis of other commercial desalination systems.

1. Introduction

Energy consumption, environmental impacts, and water production cost are the decisive parameters frequently used to investigate the performance of desalination systems [1,2]. For the last few years, the research is particularly focused on minimizing the environmental impacts and cost through advancement in materials [3,4], retrofitting of conventional systems [5,6], and development of novel hybrid methods [7,8]. The conventional desalination processes involve multi-effect desalination (MED) with evaporation effects [9], multi-stage flash (MSF) with flashing chambers [10], mechanical/thermal vapor compression (MVC/TVC) [11], humidification-dehumidification (HDH) [12], and reverse osmosis (RO) systems [13]. These systems have been investigated extensively by researchers to discover improvement possibilities [14]. In this regard, the thermoeconomic analysis has emerged as an important tool to assess and improve the thermodynamic and economic performance of desalination systems, simultaneously [15–17].

The objective of this study is to present a detailed background for thermoeconomic analysis of desalination systems. The novelty of the study lies in its comprehensive presentation of energy calculation methods, summarizing the important capital cost correlations, fiscal parameter ranges, and up-to-date cost balance equations for major desalination system components. All the models introduced in this study can be employed to conduct rigorous thermoeconomic design and analysis of all thermal desalination systems.

The manuscript is structured systematically to make data useful for the readers. In this regard, the first part of the study is focused on the literature review on thermoeconomics of desalination systems, with particular emphasis on thermal desalination systems. The next section presents the input energy calculation methods, followed by discussion and review of important correlations and ranges of fiscal parameters like interest rate, electricity cost, chemical cost, thermal energy cost, and labor cost, etc. The second last section illustrates a comprehensive analysis procedure followed by solved examples for calculation of local and overall water production costs presented in the form of cost flow diagrams. In the end, a road map is suggested based on the findings of the present study and ongoing efforts in the form of different awards for sustainable developments in the future.

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Nomenclature

| Symbol | Description |
|--------|-------------|
| A      | area, $m^2$ |
| C      | stream cost, $$/h or $$/s |
| Cf     | fuel cost |
| Cf_w   | freshwater cost, $$/m$³ |
| C_index| cost index factor |
| E_eq   | equivalent electricity consumption, kWh/m$^3$ |
| ex     | specific exergy, kJ/kg |
| h_f_g  | latent heat of vaporization/condensation, kJ/kg |
| i      | interest/discount rate, % |
| k_s    | the salt permeability coefficient |
| m      | mass flow rate, kg/s |
| N      | number of effects |
| r_e    | nominal escalation rate |
| r_p    | pressure ratio |
| r_r    | recovery ratio |
| V      | volume flow rate, m$^3$/s |
| W      | work, kW |
| X      | flow exergy, kW |
| y      | years |
| y_Ø,RO| the salt mass fraction in the flow out from the RO membrane |
| y      | the average salinity through the membrane element |
| Z      | capital cost, $ |
| Z      | the annual rate of fixed cost, $$/y |
| i      | inlet |
| m      | mean |
| Mem    | membrane |
| misc   | miscellaneous |
| o      | outlet |
| S_t    | steam |
| t      | tube |
| v      | vapor |
| 0      | dead state |

Greek symbols

| Symbol | Description |
|--------|-------------|
| α      | amortization factor, 1/y |
| β      | brine management cost, $$/m$$^3$$ |
| Δ      | change in quantity |
| ε      | unit electricity cost, $$/kWh |
| η      | efficiency, % |
| λ      | stream cost, $$/s |
| ψ      | feed split ratio |
| Λ      | plant availability, % |
| ε      | plant life, years |
| ζ      | rate of fixed cost, $$/s |
| Γ      | maintenance cost, $ |
| Ω      | labor/manpower cost, $$/m$$^3$$ |
| Φ      | pretreatment/chemical cost, $$/m$$^3$$ |
| δ      | unit steam cost, $$/ton |
| γ      | rate of membrane replacement, 1/y |
| Π      | intake cost, $$/h |

Subscripts

| Symbol | Description |
|--------|-------------|
| B      | brine |
| BH     | brine heater |
| BP     | brine pump |
| che    | chemical |
| Comp   | compressor |
| D      | distillate |
| Deh    | dehumidifier |
| DH     | distillate heater |
| eff    | effective |
| Eq     | equivalent |
| EV     | evaporator |
| F      | feed |
| FC     | flash chamber |
| FP     | feed pump |
| gen    | generator |
| Ht     | heat transfer |
| Hum    | humidifier |

| Symbol | Description |
|--------|-------------|
| i      | inlet |
| m      | mean |
| Mem    | membrane |
| misc   | miscellaneous |
| o      | outlet |
| S_t    | steam |
| t      | tube |
| v      | vapor |
| 0      | dead state |

Superscripts

| Symbol | Description |
|--------|-------------|
| CI     | capital investment |
| y      | amortization years/economic life |

Abbreviations

| Symbol | Description |
|--------|-------------|
| AB     | absorption |
| ABHP   | absorption heat pump |
| BFP    | butane freezing process |
| BR     | brine recycle |
| CAPEX  | capital expenditure |
| CELF   | constant escalation levelization factor |
| CF     | conversion factor |
| CHP    | combined heat and power system |
| CRF    | capital recovery factor |
| DCFP   | direct contact freezing process |
| DP     | distillate pump |
| EAD    | evaporator as dehumidifier |
| EAH    | evaporator after humidifier |
| EGYP   | Egyptian |
| FC     | flash chamber |
| FO-LPRO| forward osmosis low-pressure reverse osmosis |
| GT     | gas turbine |
| HPP    | high pressure pump |
| HP     | heat pump |
| LC     | levelized cost |
| LCC    | levelized capital cost |
| LHV    | the lower heating value of fuel |
| LOC    | levelized operational cost |
| MBR-RO-AOP | membrane bioreactor reverse osmosis advanced oxidation process |
| MD     | membrane distillation |
| MEE/MED| multi-effect evaporation/desalination |
| MSF    | multistage flash |
| MVC    | mechanical vapor compression |
| M$    | million dollars |
| NF     | nanofiltration |
| O&M    | operation and maintenance |
| ORC    | organic Rankine cycle |
| PCM    | phase change material |
| PCF    | parallel cross feed |
| PR     | performance ratio, |
| PT     | Pelton turbine |
| PX     | pressure exchanger |
| RP     | recirculation pump |
| SD     | solar driven |
| SEC    | specific energy consumption, kWh/m$$^3$$ |
| SED    | single effect desalination |
| SOFC   | solid oxide fuel cell |
| SS     | stainless steel |
| SWRO   | seawater reverse osmosis |
| TVC    | thermal vapor compressor/ion |
| UPR    | universal performance ratio, |
| VDS    | visual design and simulation |
2. Literature review

The economic analysis of desalination systems has been widely adopted to estimate, compare, and improve the performance of desalination systems. For example, Madani [18] suggested the use of a solar still, solar-assisted MED, and RO at small scales and MVC, MED and MSF for medium and large scale applications based on economic analysis. Gaggioli et al. [19] applied the Second Law analysis to evaluate the improvement opportunities in the existing systems to minimize exergy destruction and fuel cost. Likewise, El-Mudir et al. [20] studied the performance of a small scale TVC system, and Nafey et al. [21] conducted the thermo-economics of different desalination systems, including MSF, MEE, and MVC system using a custom made visual design and simulation (VDS) software package.

Similarly, thermoeconomic performance analysis of hybrid desalination units coupled with power plants [22,23] and energy recovery devices (ERDs) have been conducted by researchers for low water costs. For example, Darwish et al. [24] studied the cost of three different cogeneration power desalting plants and reported the product cost as RO 0.36–0.51 $/m³, MVC 0.7–0.97 $/m³, and MSF 2.21–3.99 $/m³. Manesh et al. [25] performed a component-based exergy-economic optimization of an MSF system coupled with a nuclear reactor and proposed improvement in materials for minimum exergy destruction and cost. Najafi et al. [26] optimized an MSF system coupled with a solid oxide fuel cell (SOFC) driven gas turbine and they found exergetic efficiency, capital cost, and the payback period as 46.7%, $3.76/year and 9 years, respectively under optimum conditions. Alhazmy [27] studied the economic feasibility of a novel MSF desalination with brine-feed mixing and cooling. They reported the breakeven unit product cost from 0.5 to 0.9 $/m³. Wang and Lior [28] showed that the coupling of the absorption heat pump (ABHP) improves the thermoeconomic performance of conventional MED systems. Likewise, Esfahani et al. [29] reported the product cost of the MED-ABHP hybrid system could be as low as 0.13 $/m³. Sadri et al. [30] investigated the adsorption desalination system with silica gel and found the specific water production cost 0.57 $/m³.

Zhang et al. [31] studied the heat pump driven HDH system and reported the product cost 0.0412 $/kg. They also presented that optimization of the system can reduce the water production cost to 0.03846 $/kg [28]. Similarly, Ghofrani and Moosavi et al. [32] compared the exergo-economic performance of three brine recycle BR-HDH systems with, heat-driven (BR-HDH-H), heat pump driven with evaporator after the humidifier (BR-HDH-HP-EAH), heat pump driven with evaporator as dehumidifier (BR-HDH-HP-EAD). They reported the product cost of the mentioned 3 systems as 5.1, 6.2, 10.6 $/m³, respectively. Gasmí et al. [33] performed the techno-economic analysis of a large scale SWRO plant integrated with pressure exchanger (PX) and Pelton turbine (PT) as ERDs. They reported a profit of 120,000 (for PX) and 181,000 (for PT) Tunisian Dinars, annually. A similar effect of retrofitting RO plants with ERDs was presented by Penate and Rodríguez [34] and Jamil et al. [35]. The unit water production cost was reported from 0.54 to 0.66 €/m³ for different retrofits. Djebedjian et al. [36] investigated an actual MVC system at Abu Soma Bay and reported the water production and transportation costs 6.06 and 4.753 EGP-pounds/m³, respectively, based on energetic framework.

In addition to conventional and hybrid systems, some studies have also been focused on investigating the economic feasibility of renewable energy-based systems [37–39]. For instance, Fiorenza et al. [40] examined the economics of solar-driven (SD) desalination plants, including PV-RO and SD-MED, and reported high costs compared to conventional systems. Banat and Jwaied [41] studied a solar-driven membrane distillation (MD) system and reported the membrane and plant life to be the most influential economic parameters. The freshwater cost for small and large scale units was reported as 15 $/m³ and 18 $/m³, respectively. Karagiannis and Soldatos [42] reviewed the product cost of different desalination systems. They reported the range of product cost for conventional methods for seawater from 0.4 to 3 €/m³ and brackish water from 0.2 to 1.5 €/m³. They also found that the arrangements with renewable energy sources can produce water with a cost of reaching up to 15 €/m³. Some other recent studies in this regard...

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**Form of energy used**
(e.g. thermal, mechanical or electrical)

**Auxiliary energy input**
(e.g. for motors, pumps, compressors etc.)

**Supply source**
(e.g. from boiler, extract from cogeneration plant or from waste heat recovery unit)

**Equivalency of all input energy costs**
(e.g. electrical energy cost include the cost of heat source and generating equipment like generator, turbine etc.)

**Appropriate input energy calculation method**

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Fig. 1. Factors affecting input energy calculations [50].
include life cycle assessment [43], total revenue requirement analysis [44], and exer-go-environmental analysis [45].

It is important to emphasize that the water production cost in most of the above-mentioned studies and literature is calculated directly considering the whole plant as a single unit. The total expenses, including equipment, pretreatment, labor, operation, and maintenance costs invested at system boundaries, are divided by the production capacity of the plant to evaluate the unit production cost [9,46]. Though the method gives a quick preliminary estimation of the product cost, a component level investigation and optimization are not achievable with this approach. Therefore, thermo-economic analysis has emerged as an important tool for the detailed design and analysis of desalination systems at the component level.

The method employs thermodynamics and economic analyses simultaneously to assess how efficiently the input resources are utilized throughout the system by each component [47]. This joint application of exergetic analysis and economics can satisfactorily calculate product cost while offering flexibility for design improvement and malfunction diagnosis at the same time. Thus the effect of all endogenous (i.e., irreversibility) and exogenous factors (i.e., fiscal parameters) can be analyzed, individually [48,49]. Although this is an important and most accurate tool for desalination processes evaluation, there is no complete framework available in the literature as per author knowledge. Therefore, the novelty of this study is to present a comprehensive framework for thermo-economic analysis of commercial desalination systems.

3. Input energy calculation

In thermo-economic analysis, input energy cost is of paramount importance because of its major share in the plant economics, which governs the permeate cost. It depends upon the plant performance, unit energy price, and energy input mechanism. The appropriate calculation/allocation of input energy requires its quantity, quality (thermal, mechanical, or electrical) as well as supply source (i.e., from the grid, steam from a boiler, extracted from a cogeneration plant or waste heat recovery unit). Therefore, a comparison of energy consumption per unit permeate for different desalination systems operating with different types of energy inputs needs all facets to be considered, as shown in Fig. 1 [50].

To address the energy calculation issues, many approaches are presented in the literature such as exergetic analysis [51], equivalent electricity consumption (to append auxiliary energy inputs) [52], and universal performance ratio (for appropriate energy allocation in cogeneration and power systems) [53,54]. The application of these methodologies depends on specific parameters and can handle energy calculations satisfactorily across desalination plants. The most appropriate approaches are briefly discussed below.

3.1. Exergy

It is a valuable tool to assess the performance from thermodynamic viewpoint [55] and measures the maximum theoretical useful work that can be obtained from a system as it is brought into a complete thermodynamic equilibrium with the dead state [56]. Darwish et al. [50] used this approach to estimate the exergetic value of each stream in the desalination systems. The specific exergy of a fluid stream with negligible kinetic and potential energies is calculated.

$$\text{ex} = [(h - h_0) - T_0(s - s_0)]$$

where \( h \) and \( S \) represent the enthalpy and entropy of the stream and \( h_0, S_0 \) represents the enthalpy and entropy of the stream at dead state \( T_0, P_0 \).

In some cases, the chemical potential of feed and brine also contributes as chemical exergy, and, in those cases, the total specific exergy of a stream is calculated as.

$$\text{ex} = [(h - h_0) - T_0(s - s_0)] + \text{ex \_che}$$

This \( \text{ex \_che} \) represents the specific chemical exergy calculated about a reasonably selected dead state (usually intake seawater conditions). For calculation of specific exergy of seawater streams, different models have been developed over the years, as discussed by Fitzsimons et al. [57]. Among these models, the seawater functions developed by Sharqawy et al. [58,59] are reported to be user-friendly and accurate for a wide range of salinity and temperature. These functions were later updated with pressure dependency by Nayyar et al. [60] and are extensively used in the latest studies.

Once the specific amount is known, the total exergy \( X \ (kW) \) of each stream is calculated using the respective flow rate \( m(\text{kg/s}) \) as.

$$X = m\text{ex}$$

Finally, the specific energy consumption (SEC) in kWh/m³ of the system is calculated as,

$$\text{SEC} = \frac{X + W}{m_0}$$

where \( W \) is the work input to the plant in kW and \( m_0 \) the mass flow rate of the product in kg/s.

3.2. Equivalent electricity consumption

The equivalent electricity consumption \( (E_{eq}) \) measures the amount of electricity that could have been produced using thermal energy, which was bled to a desalination system [52]. It is used to rationally compare the performance of thermal and electricity-driven desalination systems. Also, the \( (E_{eq}) \) is used to reasonably append the auxiliary energy in the total energy consumption of the system i.e., the electricity consumption of pumps in steam-driven MED and MSF systems. Likewise, in MVC systems, the makeup steam used can also be converted into equivalent electricity consumption to estimate the SEC. The electrical work is calculated, assuming that the steam used in the desalination plant was instead expanded in a steam turbine, as discussed by Naryan et al. [61].

$$W_e = m_{steam}(h_i - h_0) \quad (\text{KW})$$

where \( h_{steam} \) is the efficiency of the electrical generator, and it is assumed 95% [52,61]. The exit steam temperature is taken to be 35 °C, and the isentropic efficiency of 85% is assumed for the steam turbine [52,61]. The equivalent electricity consumption is calculated as.

$$E_{eq} = \frac{W_e}{3.6 \sum m_0} \quad (\text{kWh/m}^3)$$

where \( W_e \) represents the turbine work produced in kW, \( m_0 \) is the distillate flow rate in kg/s and 3.6 is the conversion factor from kW/ kg to kWh/m³ taking water density 1000 kg/m³.

3.3. Universal performance ratio

Shahzad et al. [62] proposed the universal performance ratio (UPR) method to evaluate the desalination processes on a common platform based on primary energy consumption. The method is particularly useful for cogeneration plants that face unfair primary fuel cost apportionment to electricity and desalination. Using the UPR method the authors investigated that in a cogeneration combined cycle power and desalination plant, the gas turbine was undercharged by 40%, the steam turbine was overcharged by 71% and desalination was overcharged by 350% by conventional energetic apportionment methods. They suggested that the use of exergetic analysis for primary fuel percentage apportionment to all components in the cycle according to the quality of working fluid utilized on a large scale exterminate this issue.

In contrast to conventional performance ratio (PR) formula (given in Eq. (7)) which is based on derived energy and does not accommodate
the quality (work potential) of different derived energies, a new UPR was proposed. In the proposed UPR (given in Eq. (8)), the derived energies are multiplied with respective conversion factors (CF) to convert into primary energies; those can be then added to calculate total input to desalination process. The conversion factors are calculated based on the exergy destruction across the components corresponding to primary fuel exergy.

\[
PR = \frac{\text{evaporative energy}}{\text{derived energy}} = \frac{h_{fg}}{\left(\frac{\text{kWh}}{\text{m}^3}\right)_{\text{electrical}} + \left(\frac{\text{kWh}}{\text{m}^3}\right)_{\text{thermal}}} 
\]

(7)

\[
UPR = \frac{\text{evaporative energy}}{\text{primary energy input}} = \frac{h_{fg}}{3.6 \left(\frac{\text{kWh}}{\text{m}^3}\right)_{\text{electrical}} + \text{CF}_1 \left(\frac{\text{kWh}}{\text{m}^3}\right)_{\text{thermal}} + \text{CF}_2 \left(\frac{\text{kWh}}{\text{m}^3}\right)_{\text{renewable}}} 
\]

(8)

The above discussion asserts that the exergy (available energy) gives the real picture of plant input energy. Therefore, the framework presented in the later sections of this paper is based on the calculation of exergetic cost of each fluid stream as it enters or leaves any component in the desalination plant.

4. Parameters of thermoeconomic analysis

The product cost of desalination systems depends upon a wide range of economic parameters as shown in Fig. 2 [63,64]. However, in thermoeconomic analysis, the main targeted economic parameters are capital expenditure (CAPEX), energy cost, fiscal parameters, operation, and maintenance cost; a detailed explanation of these parameters is presented in the subsequent sections.

4.1. Capital cost

The thermoeconomic analysis starts with the provision/calculation of capital cost (Z) of each component of the system. It generally reflects the component’s purchasing cost; however, in some cases, the commissioning and outset operation and maintenance costs are also included [65]. The most reliable option to obtain Z is through the market survey or supplier’s quotations as it fluctuates with location, the distance of the delivery site, local taxes, and availability of alternatives.
However, the cost obtained by this method is only suitable for specified conditions and does not offer design flexibility.

Therefore, to accommodate these variations in the equipment cost, different correlations have been developed that can satisfactorily approximate the equipment cost [66]. These correlations are modeled as a function of design parameters such as mass flow rate, heat transfer rate, duty, efficiency, energy consumption, pressure, and temperature differentials [67–69]. The correlations for the capital cost of common desalination system equipment are presented in Table A.1 of the appendix.

4.2. Cost index factor

The cost index factor ($C_{\text{index}}$) is used to escalate the capital cost of equipment from the original year to the current year in correlation development [70]. For instance, the capital cost of equipment calculated in 2019 using a correlation proposed 10–20 years back cannot reliably predict the required expenses [71]. Therefore, it is recommended to escalate the capital cost to the current year using cost indices that accommodate the inflation and variations in market scenarios that happened over the years [72]. For this purpose, a systematic methodology is used to calculate the $C_{\text{index}}$ using the Chemical Engineering Plant Cost Index (CEPCI) of the reference year and the current year, as given below [73].

\[
C_{\text{index}} = \frac{\text{CEPCI}_{\text{current}}}{\text{CEPCI}_{\text{reference}}} \tag{9}
\]

Then the current capital cost of the equipment is given as [74],

\[
Z^I = C_{\text{index}} \times Z^I_{\text{reference}} \tag{10}
\]

For understanding, $C_{\text{index}}$ is 1.7 for a correlation developed in 1990 as CEPCI$_{1990}$ is 390 [75], and CEPCI$_{2020}$ is 650 [76]. However, for rigorous design and analysis purposes, the effect of $C_{\text{index}}$ is studied for a wide range of values, as presented in the study [15].

4.3. Capital recovery factor

The capital recovery factor (CRF) defines the annuities in terms of money transacted out of a present value after a defined period throughout the economic life of the equipment [21]. It is used to estimate the annual rate of capital investment $Z$ ($$/yr$) of equipment or a plant using interest rate (i) and amortization years/economic life of the plant (y) as given by Eesen et al. [77]:

\[
\text{CRF} = \frac{i \times (1 + i)^y}{(1 + i)^y - 1} = \frac{i}{r} \tag{11}
\]

\[
Z = \text{CRF} \times X \tag{12}
\]

Finally, the rate of fixed cost $\zeta$ (in $$/yr) is calculated based on the plant availability (A) as [78],

\[
\zeta = \frac{Z}{365 \times 24 \times 3600 \times A} \tag{13}
\]

4.4. Cost balance equations

In thermoeconomic analysis, each component of the plant is investigated as a system with its local inputs, outputs, and fixed cost. For this purpose, a cost balance equation relating to the rate of expenditures (inputs) to the products (output) is applied to each component as given below [13].

\[
C_o = \Sigma C_i + \zeta \tag{14}
\]

where $C_i$ represents the cost of the local output stream, $C_o$ the cost of the input stream, and $\zeta$ the rate of capital cost.

For the systems with inclusive operation and maintenance costs, the above equation is written as [79].

\[
C_o = \Sigma C_i + \zeta + C_{\text{M}} \tag{15}
\]

It is important to mention that for the components with a single outlet stream (i.e., pumps, compressors, blowers, etc.), the cost balance equation can be solved by providing the inlet costs. While for the components with multiple outlet streams (i.e., heat exchangers, evaporators, flash chambers and membrane modules, etc.), additional auxiliary equations are required. For a system with “k” outlet streams, “k – 1” number of auxiliary equations are required to solve the system [80]. These equations are based on the equality of the average cost of inlet and outlet streams and are of the form as given in Eq. (16) [51].

\[
\frac{C_1}{X_1} = \frac{C_0}{X_0} \tag{16}
\]

The cost balance equations for commonly used desalination system components and the corresponding auxiliary equations are presented in the following sections.

4.4.1. Pump

Pumps are used to maintain pressure and flow rate of different water streams, including feed, permeate, and brine. In membrane-based systems (particularly RO) pumps have a significant share in the total plant cost because of high pumping pressures (≥6000 kPa) that require robust size and high energy input. Contrarily, in thermal-based systems, pumps maintain pressure merely to overcome the pressure drops in pipes, heat exchangers and evaporators. As the pressure in these systems is slightly above atmospheric pressure (≤500 kPa), their energy consumption and cost do not contribute significantly. Resultantly, in most of the thermally driven desalination processes, the energy consumption of pumps is either ignored or calculated under auxiliary energy inputs, as mentioned above (refer 2.2). However, for a detailed and accurate analysis, a complete plant layout is to be considered (including pumps). Fig. 3 demonstrates the cost flow diagram of pumps, and the corresponding cost balance equation is given in Eq. (17).

\[
C_o = C_i + \zeta W_{\text{pump}} + \zeta_{\text{pump}} \tag{17}
\]

4.4.2. Compressor

A compressor is an integral part of mechanical vapor compression based thermal desalination systems. It is used to increase the temperature and pressure of water vapors produced during the desalination process. These compressed vapors are used to desalinate the subsequent feed water streams. The process offers two major benefits including flexibility to operate on electrical energy (suitable for remote areas) and a significant reduction in intake steam flow rate which abates the demand of a robust steam generation facility. Therefore, in such systems compressor is a major energy-consuming (~95%) device and thus have a major share in plant economics. Fig. 4 illustrates the cost-flow diagram of a vapor compressor and the corresponding cost balance is given in Eq. (18).

\[
C_o = C_i + \zeta W_{\text{comp}} + \zeta_{\text{comp}} \tag{18}
\]
in a condenser as a distillate, but in multi-effect systems, these are used as steam in the subsequent effects. The cost balance equation for the evaporator is given in Eq. (23) and the auxiliary equations are given in Eqs. (24) and (25).

\[ C_{V,o} = C_{F,i} + C_{SW,i} - C_{D,o} + \xi_{SV} \]  
(23)

\[ \frac{C_{S,1,i}}{X_{S,1,i}} - \frac{C_{D,o}}{X_{D,o}} = 0 \]  
(24)

\[ \frac{C_{F,i}}{X_{F,i}} - \frac{C_{R,o}}{X_{R,o}} = 0 \]  
(25)

4.4.6. Flash chamber

In flashing-based thermal-desalination systems, e.g., MSF, the flash chambers are used to produce vapors. First, the intake seawater is heated to the feed temperature by passing through the condenser tubes. It is then fed to the brine pool at the bottom of the chamber, where it vaporizes partly and the remaining leaves as brine, which serves as a feed for subsequent effects. The flashed vapors pass through a demister and are collected as condensate in the distillate tray after exchanging heat with the intake seawater. The distillate stream (from previous effects) is mixed with condensate and is collected at the outlet, as shown in Fig. 8. The cost balance equation for the flash chamber is given as [21].

\[ C_{D,o} = C_{F,i} + C_{SW,i} + C_{D,1} - C_{R,o} + \xi_{FC} \]  
(26)

The two auxiliary equations required to solve this cost balance are given in Eq. (27) and (28).

\[ \frac{C_{F,i} - C_{R,o}}{X_{F,i} - X_{R,o}} = 0 \]  
(27)

\[ \frac{C_{F,1} - C_{D,o}}{X_{F,1} - X_{D,o}} = 0 \]  
(28)

4.4.7. Humidifier and dehumidifier

In HDH systems, the feed water is sprayed in a humidifier where it humidifies the air using the evaporative potential of air. The moist air from the humidifier is directed to the dehumidifier, where it condenses by exchanging heat with the intake seawater, as shown in Figs. 9 and 10. The cost balance equations and auxiliary equations for a humidifier are given in Eqs. (29) and (30) and for a dehumidifier in Eqs. (31) and (32) given below [83].

\[ C_{\text{moist air},o} = C_{\text{dry air},i} + C_{F,i} - C_{R,o} + \xi_{\text{Hum}} \]  
(29)

\[ C_{F,i} = C_{R,o} \]  
(30)

\[ C_{D,o} = C_{\text{moist air},i} + C_{SW,i} - C_{SW,o} - C_{\text{dry air},o} + \xi_{\text{Deh}} \]  
(31)

\[ \frac{C_{\text{dry air},i} - C_{\text{moist air},o}}{X_{\text{dry air},i} - X_{\text{moist air},o}} = 0 \]  
(32)

It is important to mention that in addition to the above-discussed parameters, some other indicators like payback period, Levelized cost, etc. have also been used by some researchers to illustrate the economics.
of desalination systems. These indicators are not an intrinsic part of thermoeconomic analysis and are rarely used. However, they provide valuable information from a monetary viewpoint. A brief overview of these economic indicators is presented in the appendix. Besides, a detailed review of existing studies on thermoeconomic analysis of desalination technologies is conducted to summarize the range of important fiscal parameters as presented in the appendix. Table A.2 shows that water production cost varies remarkably with location, technology, and time. The prime reasons for fluctuations in the prices are the variations in system performance and fiscal parameters. The technological advancements and development of better materials have considerably reduced the energy consumption of desalination systems, thus providing the potential for cutting the operational expenses. However, the associated research and development expenses, inflation in material costs, hikes in energy prices, and strict environmental legislations have imposed additional monetary overheads.

It is important to mention that the calculation of water production cost (presented above) is a multifaceted procedure and necessitates a systematic approach for the calculation of required parameters. Keeping in view, a component-based comprehensive theoretical framework calculating local and overall water production cost, is presented in the subsequent section. For this purpose, a technically sophisticated desalination system, i.e., Mechanical Vapor Compression Desalination System working under different operating scenarios, is selected as an example. The system demonstrates all major thermal, hydraulic, and thermodynamic processes involved in the conventional desalination systems. Therefore, the developed model can be applied to other systems with minor relevant modifications.

5. A sample thermoeconomic model

Based on the earlier presented capital cost correlations, cost balance equations, and fiscal parameters, a comprehensive thermoeconomic model is developed for MVC systems, as an example. Different operating arrangements are considered for a complete plant layout. Firstly, the exergy analysis is conducted to calculate the flow exergy (kW) of all the streams, including feed, distillate, brine, and vapors. At the second stage, exergy destruction in each component is calculated to assess the thermodynamic performance. Lastly, the cost balance equations are
employed to calculate the monetary cost of all the streams using exergetic cost-based auxiliary equations.

5.1. Single effect MVC system

A single effect MVC system consists of an evaporator, a compressor, pumps, and two preheaters. The governing equations for thermo-economic analysis of SEE-MVC are summarized in Eqs. (33)-(47) [15].

\[
C_1 = C_0 + \varepsilon W_{FP} + \zeta_{FP}
\]  
(33)

\[
C_2 = \psi C_1
\]  
(34)

\[
C_1 = (1 - \psi) C_1
\]  
(35)

\[
C_4 = C_2 + C_0 - C_{14} + \zeta_{SW}
\]  
(36)

\[
\frac{C_0}{X_0} = \frac{C_{14}}{X_{14}}
\]  
(37)

\[
C_3 = C_3 + C_{11} - C_{12} + \zeta_{SW}
\]  
(38)

\[
\frac{C_{11}}{X_{11}} = \frac{C_{12}}{X_{12}}
\]  
(39)

\[
C_6 = C_4 + C_5
\]  
(40)

\[
C_{16} = C_3 + C_{17} - C_8 - C_9 + \zeta_{SW}
\]  
(41)

\[
\frac{C_7}{X_7} = \frac{C_8}{X_8}
\]  
(42)

\[
\frac{C_3}{X_3} = \frac{C_{17}}{X_{17}}
\]  
(43)

\[
C_{17} = C_{26} + \varepsilon W_{Comp} + \zeta_{Comp}
\]  
(44)

\[
C_{15} = C_{14} + \varepsilon W_{FP} + \zeta_{SW}
\]  
(45)

where, \( C_{\text{misc}} \) is the miscellaneous cost such as blowdown, cooling, condensate, chemical, post-treatment, etc.

The thermoeconomic analysis for the SEE-MVC system is conducted using the above-presented model. The important thermodynamic and economic parameters used in the analysis are summarized in Table 1. The cost-flow diagram presenting the cost rate (in $/h) at each component’s outlet is presented in Fig. 11 [15,84].

5.2. Multi-effect MVC systems

The multi-effect MVC systems use many evaporation effects (normally 2 to 10) to desalinate feed water. The vapor produced in each evaporator is used as a heat source for subsequent effects. Finally, the vapors from the last effect are directed to a vapor compressor where they are compressed and superheated to serve as a heat source for the first effect. However, in some high capacity systems, makeup steam is added with the compressed vapor to reduce the compressor size, energy consumption, and cost. It is worth mentioning that the multi-effect systems can operate in different feed flow arrangements (see Fig. 12), which have different operating conditions, energy requirements, and product costs for the same capacity. The thermoeconomic models for three commonly used flow configurations are presented below:

5.2.1. Forward feed (FF)

In this arrangement, the total feed is sprayed in the first evaporator, and the brine of each evaporator serves as a feed for the next effects. Finally, the brine from the last effect is directed to the brine preheater to preheat the intake seawater. The cost balance equation is generalized in terms of the number of evaporators which can vary as \( j = 2 \) to \( N \). The governing equations in the general form are given as [82]:

\[
C_V[j] = C_{\psi j} + C_{\psi j - 1} - C_0[j] + \zeta_{SW}
\]  
(48)

For the first effect \( j = 1 \), the vapor and feed stream costs are calculated as:

\[
C_V[1] = C_{\psi 1} + C_{\psi 0} + \zeta_{SW}
\]  
(49)

\[
C_{30}[1] = C_{30}[N] + \varepsilon W_{Comp} + \zeta_{Comp} + C_{\text{makeup steam}}
\]  
(50)

For \( j = 2 \) to \( N \)

\[
C_V[j] = C_{\psi j} + C_{\psi j - 1}
\]  
(51)

\[
C_{31}[j] = C_{\psi j - 1}
\]  
(52)
5.2.2. Parallel feed (PF)

In this configuration, the feed is equally disbursed in all the evaporators at feed temperature. The brine from all effects is collected and sent to the brine preheater and the remaining operation same as that of forward feed. The general form of equations for calculation of different costs are given below [85].

\[
C_{r}[j] = C_{F}[j] + C_{D}[j] - C_{B}[j] + \xi_{EV} 
\]  

(53)

\[
C_{r}[j] = \frac{C_{F, DH} + C_{F, BH}}{N} 
\]  

(54)

\[
C_{S,1}[1] = C_{F}[N] + \xi W_{Comp} + \xi_{Comp} + C_{makeup, steam} 
\]  

(55)

For \([j = 2\) to \(N] \)

\[
C_{S,1}[j] = C_{r}[j - 1] 
\]  

(56)

5.2.3. Parallel crossfeed (PCF)

In this arrangement, the feed is equally distributed in all the effects like PF; however, brine from each effect is introduced in the next effect at the bottom. The addition of brine in the subsequent effects results in addition to vapor production with flashing because of the difference in temperature and pressure. The governing cost equations are given below [85].

\[
C_{r}[j] = C_{F}[j] + C_{S}[j] - C_{D}[j] + \xi_{EV} + C_{B}[j - 1] 
\]  

(57)

\[
\hat{C}_{r}[j] = \frac{C_{F, DH} + C_{F, BH}}{N} 
\]  

(58)

\[
C_{S,1}[1] = C_{F}[N] + \xi W_{Comp} + \xi_{Comp} + C_{makeup, steam} 
\]  

(59)

For \([j = 2\) to \(N] \)

\[
C_{S,1}[j] = C_{r}[j - 1] 
\]  

(60)

It is important to emphasize that the feed flow arrangements discussed above have their advantages and limitations from thermodynamic and economic viewpoints, which have been investigated in detail by Jamil et al. [84,85]. The cost flow diagram for 4-effects MVC systems operating under the above-discussed arrangements is presented in Fig. 13.

6. Discussion of results

The thermoeconomic analysis conducted above presented the cost rate (in $/h) and water production cost (in $/m \textsuperscript{3}) for different MVC system configurations. For example, in a single effect system, the feed is assigned a cost rate of 4.6 $/h at a preheater inlet because of pre-treatment. After being heated to feed temperature (at evaporator inlet) from the intake seawater temperature, the cost of stream surged to 795 $/h because of the capital cost of preheaters and the hot water stream costs coming from the evaporator. The cost of vapors produced in the evaporator is calculated as 5216 $/h, which increased to 5274 $/h at the compressor outlet because of the addition of compressor fixed cost as well as input energy cost. Finally, the distillate and brine stream costs after releasing its heat in the preheater are calculated as 94 $/h and 33 $/h, respectively. The water production cost for this system with a distillate capacity of 13 kg/s is calculated as 1.7 $/m \textsuperscript{3}.

Similarly, for multi-effect systems, the pretreated intake seawater cost is taken the same as 4.61 $/h for all the feed flow arrangements. However, the preheaters are different in size for all three systems because of different inlet and outlet temperatures; therefore, the feed water cost is calculated as 338 $/h, 122 $/h, 116 $/h for FF, PF and PCF thus showing the largest investment in the case of FF. This is because of the reason that in FF, the total feed is sprayed in the first evaporator, which has significantly higher evaporation temperature.
Fig. 12. Multi-effect MVC system operating under (a) forward feed, (b) parallel feed, and (c) parallel crossfeed [84,85].
requirements (i.e., 47 °C) than PF and PCF (i.e., 36 °C). The distillate cost at the outlet of evaporator and compressor section is calculated as 316 $/h, 156 $/h, and 183 $/h, and the corresponding brine stream costs are 117 $/h, 62 $/h, and 32 $/h. Finally, the water production cost is calculated as the highest for FF with 0.867 $/m³, followed by PCF with 0.865 $/m³ and PF with 0.842 $/m³. Therefore, the analysis concluded that the single effect system has the highest water production cost, followed by FF, PCF, and PF, respectively.

7. Closing remarks and future road map

The current study presents a comprehensive theoretical framework for a component-based thermoeconomic analysis of commercial-scale desalination systems. The key economic parameters that govern the analysis and water product cost are discussed in detail. A comprehensive review of existing studies is conducted to cluster the magnitude of critical fiscal parameters. The cost balance equations are developed for
### Table A1

Correlations for capital cost.

| Component | Correlation | Limitations | Ref. |
|-----------|-------------|-------------|------|
| Pump      | $Z_{\text{Pump}} = 13.92 m_{\text{th}} \Delta P_{\text{pump}}^4 0.05$ | $2 \leq m \leq 32100 \leq \Delta P \leq 6200$ $1.8 \leq c \leq 9$ | [94] |
|           | $Z_{\text{Pump}} = 705.48 W_{\text{pump}}^{0.71} (1 + 0.2/1 - \eta)$ | — | [95] |
|           | $Z_{\text{Pump}} = 2300(W/10)^{0.25} (1 - \eta/\eta_{\text{pump}}^3)$ | — | [71] |
|           | $Z_{\text{Pump}} = 1785(W/10)^{0.25} (1 + 0.2/1 - \eta)$ | — | [72] |
|           | $Z_{\text{Pump}} = 996 V_{\text{pump}}^{0.8}$ | — | [13,96] |
|           | $Z_{\text{Pump}} = 3500 (W/10)^{0.77}$ | — | [97] |
| Preheater (liquid-liquid plate heat exchanger) | $Z_{\text{Preheater}} = 1000(12.86 + A_{\text{th}}^{0.5})$ | $40 \leq Q \leq 185$, $2.5 \leq \Delta T \leq 80.7$ $\leq \Delta T \leq 50$, $0.007 \leq \Delta T \leq 9$ | [20] |
|           | $Z_{\text{Preheater}} = 157.81 \Delta T_{\text{in}}^{0.7}, \Delta T_{\text{out}}^{0.04}$ | $A \geq 18.6m^2$ $\leq 8$ $\leq 50\times 0.007 \leq 50$ | [94] |
|           | $Z_{\text{Preheater}} = 1839 \Delta F_{\text{pump}}^{0.631}$ | $P \leq 10$ bar, $T \leq 160^\circ$ | [98–100] |
|           | $Z_{\text{Preheater}} = 781 F_{\text{pump}}^{0.714}$ | $A \geq 18.6m^2$ $\leq 8$ $\leq 50\times 0.007 \leq 50$ | [94] |
|           | $Z_{\text{Preheater}} = 1281 \Delta F_{\text{pump}}^{0.887}$ | $A \geq 18.6m^2$ $\leq 8$ $\leq 50\times 0.007 \leq 50$ | [98–100] |
|           | $Z_{\text{Preheater}} = 702 F_{\text{pump}}^{0.6907}$ | $A \geq 18.6m^2$ $\leq 8$ $\leq 50\times 0.007 \leq 50$ | [94] |
|           | $Z_{\text{Preheater}} = 1391 A_{\text{th}}^{0.778}$ | SS | [101–103] |
|           | $Z_{\text{Preheater}} = 635.14 A_{\text{th}}^{0.778}$ | SS | [101–103] |
|           | $Z_{\text{Preheater}} = 201.67 Q \Delta T_{\text{in}}^{1.0} \Delta P_{\text{in}}^{0.15}$ | — | [104,105] |
| Evaporator | $Z_{\text{Evaporator}} = 250 Q \Delta T_{\text{in}}^{2.0} \Delta P_{\text{in}}^{0.1}$ | $150 \leq Q \leq 800$, $2 \leq \Delta T_{\text{in}} \leq 220.06 \leq \Delta P_{\text{in}} \leq 0.3503 \leq \Delta P_{\text{in}} \leq 0.06$ | [94] |
| Mechanical Vapor Compressor | $Z_{\text{MVC}} = 73644 m_{\text{th}} \eta_{\text{r}} \phi_{\text{r}}$ | $10 \leq m_{\text{th}} \leq 455$, $1.1 \leq \eta_{\text{r}} \leq 2$ $2.3 \leq \phi_{\text{r}} \leq 11.5$ | [94] |
|           | $Z_{\text{MVC}} = 704.68 W_{\text{comp}}^{0.66}$ | — | [106,107] |
| High-pressure pump | $\log (W) = 3.3892 + 0.05361 \times \log (Q) + 0.1538(\log (W))^{2}$ | $W_{\text{HP}}(kW)$ | [13] |
|           | $Z_{\text{High-pressure pump}} = 52(B_{\text{p}} V)$ | $V \leq 200m^3/h$ | [97,108] |
|           | $Z_{\text{High-pressure pump}} = 81(B_{\text{p}} V)$ | $V \leq 450m^3/h$ | [97,108] |
|           | $Z_{\text{High-pressure pump}} = 393000 + 10710 \times P_{\text{p}}$ | $P_{\text{p}}(\text{bar})$, $V'(\text{m}^3/\text{h})$ | [97,108] |
| Pelton turbine | $\log (Z_{\text{Turbine}}) = 2.2476 + 1.4956 \times \log (W_{\text{Turbine}})$ | $W_{\text{Turbine}}(kW)$ | [13] |
| Membrane module (RO) | $Z_{\text{Membrane}} = 10 \times \frac{1.5}{(1.5 + \frac{1.5}{\frac{1.5}{\frac{1.5}{1.5}}})}$ | $z_{\text{RO}} = 2.03 \times 10^{-5}(-m^2/m^2 \text{kPa})$ | [13,109–111] |
|           | $\tau = \frac{(\text{m} - m_{\text{th}}) \times 10^{-5} \times m_{\text{th}}}{m_{\text{th}} \times 10^{-5} + (\text{m} - m_{\text{th}})}$ | | |
|           | $V_{\text{th}} = 1.5(-m^3/h)$ | | |
|           | $\phi_{\text{r}} = 0.55$ | | |
| Thermal vapor compressor (TVC) | $Z_{\text{TVC}} = 15962.74(R/10)^{0.87} P_{\text{th}}^{0.75}$ | — | [47] |
|           | $Z_{\text{TVC}} = 31.05(6340.13 V_{\text{T}}/10^{0.007}) \times 10^6$ | — | [112] |
| Humidification Dehumidification (HDH) | $Z_{\text{Humidification}} = 746.75 m_{\text{th}}^{0.71} (T_{\text{in}}^{0.15} - T_{\text{out}}^{0.57})$ | — | [71,83,113,114] |
|           | $Z_{\text{Humidification}} = 31.05(6340.13 V_{\text{T}}/10^{0.007}) \times 10^6$ | — | [71,83,113,114] |
| Water heater | $Z_{\text{Water heater}} = 130(A_{\text{TH}}/0.093)^{0.78}$ | — | [72,83] |
| Condenser | $Z_{\text{Condenser}} = 800(A_{\text{Condenser}}/10)^{0.8}$ | $A (m^2)$ | [72,83] |
|           | $Z_{\text{Condenser}} = 150(A_{\text{Condenser}}/0.8)$ | | [97] |
|           | $Z_{\text{Condenser}} = \frac{120.74 \times Q}{T_{\text{cond}}/T_{\text{in}}}$ + 746 $\times +$ $70.5 \times +$ $0.6936 \cdot \ln (\Delta T) + 2.1898$ | | [47] |
| MED/MSF complete plant fixed cost | $Z_{\text{MED/MSF}} = 3054(\text{TH})^{0.9731}$ | Including supply of equipment, engineering, commissioning, erection, instrumentation, and control has been derived as a function of daily production up to 10,000 (m$^3$/day) | [115,116] |

(continued on next page)
Table A1 (continued)

| Component        | Correlation                                                                 | Limitations                                                                 | Ref.  |
|------------------|-----------------------------------------------------------------------------|------------------------------------------------------------------------------|-------|
| HDH complete unit| \( Z_{H D H} = 529.71 \left( B_1 + B_2 P_{m} F_p \right) C^0 \) \[117\] | \( B_1 \) and \( B_2 \) are constants, \( P_{m} \) the material factor, \( F_p \) the pressure factor, \( A \) represents the capacity or size \( K_2, K_3, K_4 \) and \( C_1, C_2, C_3 \) are constants |       |
| Boiler           | \( Z_{s o l a r} = A' \times m_A + B' \) \[104,105\]                         | —                                                                            |       |
|                  | \( A' = 0.249 P_{s o l a r} + 47.19, B' = 3.29 P_{s o l a r} + 624.6 \)      | —                                                                            |       |
|                  | \( Z_{s o l a r} = 3.28 \times 10^3 (m_A / 20000)^{0.41} \)                | —                                                                            |       |
| Vapor generator  | \( Z_{v a p e r \_g e n e r a t o r} = 10393.6 U A \Delta P^{0.13} \Delta p^{1.26} \) \[105\] | —                                                                            |       |
| Solar field      | \( Z_{s o l a r \_f i e l d} = 150 (A)^{0.95} \) \[97\]                      | —                                                                            |       |

where, all values are in SI units i.e. \( Q(kW), A(m^2), T(K), \Delta T(\degree C), P(kPa), m(i(kg/s)), W(kW) \), \( r_p = P_{o} / P_{r}, \epsilon = \eta / (1 - \eta) \), IF: Installation factor for PHX ranging 1.5–2.0 \[98\]

commonly used desalination system components. Finally, a detailed mathematical procedure is developed for thermo-economic analysis of desalination systems using one of the most panoptic and sophisticated systems, i.e., MVC. The proposed procedure can be satisfactorily used to investigate the thermo-economic performance of other desalination systems as well.

In conclusion, an extensive review and assessment of literature reveals that despite lots of improvement efforts, the leveling-off water production cost for existing desalination systems is hovering around 0.9 ± 0.3 $/m^3 (refer Fig. 14). Most of the developing countries are unable to pay such a high cost, where the average salary is close to $1–2/day \[86\]. Hence, for a quantum decrease in water production cost, a technological breakthrough is essential that can address the key limitations like operational limits (~0.78 kWh/m^3), material properties, brine management, monetary expenses, and energy consumption, etc., simultaneously. For affordable water supply to all developing countries, the estimated operational cost must be closer to 0.3 ± 0.05 $/m^3 \[87\]. In this regard, some noticeable efforts are underway such as Global Water Award (SUQIA UAE) \[88\], the Water Challenge (Water Aid Australia) \[89\], Global Water Awards \[90\], the IWA Global Water Award \[91\] and the IDA Water Awards \[92\] to motivate water experts for innovative solutions to reduce the water cost.

The overall target is to achieve low cost and sustainable water supply zone, as shown in Fig. 14. Based on recent COVID-19 global pandemic experience when US oil prices dropped below zero, water has emerged as the new oil, and its importance is endorsed in every sector of life. To cope this kind of uncertain situation in the future, water security, and low-cost supply are the most important. So, it is very timely to investigate out-of-box solutions for uninterrupted and low-cost water supply to save lives.

Declaration of Competing Interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

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Appendix

A-1. Additional economic indicators

The following additional economic indicators are also used in the literature to present the economic performance of desalination systems.

Payback period

It refers to the period in which total capital investment is recovered by selling the product \[79\]. The equations for calculation of payback period is given as \[26\].

\[
E_{C,p} = \left( \sum_{k} Z_k \right) (1 + i)^p + 3600 \sum_{m=1}^{p} C_f N (1 + i)^{p-m}
\]

(A1)

where, \( "E_{C,p}" \) represents the worth of the plant in “\( p^{th}\)" year of operation with “\( Z\)” as capital cost, “\( i\)” as interest rate, “\( C_f\)” as fuel cost rate, “\( p\)” as the payback period, and “\( N\)” as operational hours per year.

Levelized cost

It measures and compares the cost of different alternative systems giving the same output. It gives the real price for long term projects in net
Table A2
Ranges of economic parameters used by different authors.

| Reference | Technologies (Product cost) | Economic parameters |
|-----------|-----------------------------|---------------------|
| Ettouney et al. [46] 2002 | MSF (1.17 $/m^3)  
MEE (1.01 $/m^3)  
MVC (0.501 $/m^3)  
RO (0.669$/m^3) | $i = 5\%, \xi = 30y, \Lambda = 90\%, \phi = 0.025 $/m^3, \Omega = 0.05 to 0.1 $/m^3, \varepsilon = 0.05 $/kWh, \delta = 1.466 $/MJ, \gamma = 5\%$ (for brackish water with good pretreatment) to 20% (for seawater) |
| Fiorenza et al. [40] 2003 | PV-RO (2 $/m^3)  
Solar driven MEE (2 $/m^3) | $i = 8\%, \xi = 25y, \Gamma = 2\% of PEC, \Omega = 0.1 $/m^3, \phi = 0.035 $/m^3  
i = 8\%, \xi = 25y, \Gamma = 2\% of PEC, \Omega = 0.1 $/m^3, \phi = 0.025 $/m^3 |
| Terno et al. [111] 2005 | RO (76.7 c€/m^3) | $\xi = 20y, \Gamma = 2 to 4\% of PEC, \phi = 5.5 c€/$, \varepsilon = 6c/kWh, |
| Marcovecchio et al. [110] 2005 | RO (0.8 – 1.018 $/m^3) | $i = 20\%, \xi = 25y, \gamma = 20\% /y, \Lambda = 90\% |
| Mudir et al. [20] 2004 | TVC (3.79 $/m^3)  
RO-PWT (0.685 $/m^3)  
RO-PX (0.572 $/m^3) | $i = 5\%, \xi = 20y, \Gamma = 15–25\% of PEC, \Lambda = 90\%, \varepsilon = 0.06 $/kWh |
| Drioli et al. [118] 2006 | MF-NF-RO (0.51 $/m^3) | $i = 5\%, \xi = 30y, \Gamma = 0.033 $/m^3, \Omega = 0.03 $/m^3, \phi = 0.025 $/m^3, \Lambda = 90\%, \gamma = 10\%, \varepsilon = 0.09 $/kWh, \delta = 0.005 $/m^3, \beta = 0.0053 $/m |
| Nafey et al. [21] 2006 | MSF (2.63 $/m^3) | $i = 5\%, \phi = 0.07 $/m^3, \xi = 20y, \Lambda = 90\%, \gamma = 0.1 $/m^3, \Delta = 13.4 $/ton, \Pi = 3.15 $/h |
| Nafey et al. [9] 2006 | MEE-FF (1.87 $/m^3)  
MEE-MSF (1.70 $/m^3)  
MEE-PCF (2.58 $/m^3) | $i = 5\%, \xi = 20y, \phi = 3.7 $/kg of chemicals, \Lambda = 90\%, \Delta = 4.5 $/m^3, \varepsilon = 0.098 $/kWh, \Pi = 0.072 $/m |
| Nafey et al. [11] 2007 | MEE-MVC (1.7 $/m^3)  
MEE-MVC (2.2 $/m^3) | $i = 7\%, \phi = 5.9 $/h, \xi = 20y, \Lambda = 90\%, \epsilon = 0.09 $/kWh, \Delta = 15.6 $/ton, \Pi = 4.6 $/h |
| Nafey et al. [97] 2010 | RO (0.898 $/m^3)  
RO-PWT (0.685 $/m^3)  
RO-PX (0.572 $/m^3) | $i = 5\%, \xi = 20y, \Gamma = 15–25\% of PEC, \Lambda = 90\%, \varepsilon = 0.06 $/kWh |
| Mabroud et al. [78] 2007 | MSF (2.63 $/m^3)  
MEE-TVC (3.4 $/m^3)  
MEE-MVC (1.7 $/m^3)  
SWRO (1.34 $/m^3) | $i = 5\%, \phi = 3.7 $/kg of chemicals, \xi = 10y, \Lambda = 90\%, \epsilon = 0.098 $/kWh, \Delta = 4.5 $/m^3, \Pi = 0.072 $/m |
| Ophir et al. [119] 2007 | MED-TVC (0.69 $/m^3)  
MED-MVC (0.60 $/m^3) | $i = 6\%, \Gamma = 1\% of PEC, \Omega = 0.031 $/m^3, \phi = 0.05 $/m^3, \Lambda = 95\%, \epsilon = 20y, \epsilon = 0.05 $/kWh |
| Banat and Jwaideh [41] 2008 | SD-MD (15–18 $/m^3) | $i = 5\%, \Gamma + \Omega = 20\% of PEC, \phi = 0, \Lambda = 90\%, \varepsilon = 20y, \gamma = 20\% |
| Lukic et al. [106] 2010 | MVC (N/A) | $i = 4\%, \Omega = 0.1 €/m^3, \phi = 0.02 €/m^3, \xi = 30y, \varepsilon = 0.09 €/kWh, \lambda = 0.05783 |
| Sayyad and Saffari [44] 2010 | MED-TVC (0.962 $/m^3) | $\xi = 30y, \Gamma = 1\% of PEC, \Omega = 0.031 $/m^3, \phi = 0.05 $/m^3, \Lambda = 95\%, \varepsilon = 0.03 $/kWh, \Delta = 0.2 $/m |
| Sharaf et al. [120] 2011 | SD-MED-PF-TVC (1.323 $/m^3)  
SD-MED-PF-MVC (0.94 $/m^3) | $\Gamma = 2\% of PEC, \phi = 0.025 $/m^3, \Omega = 0.025 $/m^3, \varepsilon = 0.06 $/kWh, \delta = 1.466 $/MJ |
| Periante and Rodríguez [34] 2011 | SWRO -ERD (0.5–0.6 $/m^3) | $\Gamma = 0.140 €/m, \xi = 15y, \epsilon = 0.12 €/kWh, \phi = 8y, \Lambda = 95\% |
| Eman and Dincer [13] 2014 | SWRO-ERD (2.451 $/m^3) | $i = 8\%, \phi = 0.018 $/m^3, \text{Cartridge} = 0.01 $m^3, \xi = 20y, \gamma = 10\%, \Lambda = 8760 h, |
| Hanafizadeh et al. [121] 2016 | MED (0.96 $/m^3) | $i = 6\%, \xi = 20y, \varepsilon = 0.06 $/kWh, \phi = 2.63 $/GJ (based on LHV) |
| Linares et al. [43] 2016 | FO-LPRO (0.636 $/m^3)  
MBR-RO-AOP (0.637 $/m^3)  
SWRO (0.737 $/m^3)  
MED (3.13 $/m^3) | $i = 6\%, \xi = 20y, \varepsilon = 0.08 $/kWh, \gamma = 8y, \Lambda = 95\%, \gamma = 10\%, \phi = 5y (RO and FO membrane) |
| Piacentini [122] 2015 | MED (3.13 €/m) | $i = 5\%, \xi = 20y, \phi = 0.0852 €/m, \Omega = 0.27 €/Nm³ (NG), \Lambda = 2.838 \times 10^7 s, \delta = 5 \times 10^{-5} €/kJ |
| Mohktari et al. [123] 2016 | GT-MED (1.79 $/m^3)  
GT-MED-RO (2.3–2.8 $/m^3) | \Lambda = 0\%, \xi = 20y, \phi = 0.08 $/kWh |
| Delgado et al. [124] 2016 | Solar thermal power plant (0.76 to 1.26 €/m³)  
MED (0.76 to 1.26 €/m³)  
RO (0.76 to 1.26 €/m³) | $\xi = 20y, \phi = 5.7 h/d, \phi = 1500 €/kWh, \phi = 300 €/m³ |
| Ahmadi et al. [125] 2017 | GT-MED-TVC (1.96 $/m^3)  
SOFC-GT-MED-TVC (2.88 $/m^3) | $i = 1.3\%, \xi = 10y, \phi = 0.03 $/m^3, \Omega = 1000 $/month/person, \Lambda = 4000 h/y, |
present value in terms of Levelized Capital Cost (LCC) and Levelized Operational Cost (LOC). The term Levelized asserts that these costs are presented in the form of a series of equivalent payments

\[
(\text{A2})
\]

where \( LCC \) is given by

\[
(\text{A3})
\]

Here, "\( K_t \)" is the capital expense ($) accruing in the year "\( t \)", "\( V_t \)" is distillate amount (m\(^3\)) in "\( t \)", "\( r \)" is the required rate of return, and "\( y \)" is the amortization period.

Likewise, the LOC is calculated in terms of operational cost "\( O_t \) ($)" accruing in year "\( t \)" as.

\[
(\text{A4})
\]

In addition to LCC and LOC, another parameter called Constant Escalation Levelization Factor (CELF) is also calculated. It gives a reliable estimate of the present value of a system, and is given as [13,79].

\[
(\text{A5})
\]

where \( CRF \) is the capital recovery factor, "\( \varepsilon \)" is economic life, and "\( \kappa \)" is constant and calculated as.

\[
(\text{A6})
\]

In the above equation "\( r_n \)" is the nominal escalation rate of energy and O&M costs with time and "\( i_{\text{eff}} \)" is the effective discount rate.

### A-2. Summary of various cost elements

The capital cost of different desalination system components is calculated using the correlations that are summarized in Table A1, whereas the
ranges of important fiscal parameters and the product cost are described in Table A2.
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