**Metrics Matter: Accurately Defining Energy Efficiency in Desalination**

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1 Introduction

Desalination is an inherently energy intensive process. The cost of energy can account for approximately 25–40% of the total cost of water [1]. Small improvements in plant energy efficiency can result in significant savings and a competitive edge for desalination businesses. Consequently, much research has focused on reducing the energy consumption of desalination technologies.

When researchers, plant designers, operators, and decision makers describe energy and energy efficiency in desalination, many different metrics and terms are used to characterize relevant aspects of plant performance. Common metrics in the literature include the specific energy consumption (SEC), the gained output ratio (GOR), the performance ratio (PR), the second law efficiency, the specific exergy consumption, and specific primary energy (PE) consumption. In addition, some researchers have proposed hybrid terms such as the universal performance ratio (UPR).

In communications among stakeholders, all parties need a common understanding of such terminology and tools to interpret these metrics within an engineering framework. Unfortunately, the plethora of available metrics [2–9] and a lack of understanding of their meaning, as well as the misuse of certain metrics, have impeded agreement about relative efficiency of different desalination technologies. In some cases, what is meant by a particular metric may not match what is understood by others reading that metric. The use of metrics inappropriate to a particular situation can also produce suboptimal outcomes.

In the first portion of this paper, we aim to clear up some of the confusion around desalination metrics. We examine a wide variety of metrics, and illustrate the appropriate usage of each metric with examples of desalination plants using reverse osmosis (RO), multiple effect distillation (MED), and multiple effect distillation with thermal vapor compression (MEDT).

In the course of this analysis, we show that large-scale RO systems outperform MED and MEDT in terms of energy consumption by a wide margin. These energetic advantages are increasingly borne out in practical economic decision making: 90% of new desalination capacity is expected to be in the form of RO in the near future [10]. The relative performance and cost effectiveness of these desalination systems may very well change in the future as novel technologies are discovered, developed, and implemented; but at present, RO is the most widely adopted technology.

In the second portion of this paper, we investigate the underlying reasons that RO outperforms MED from an energy consumption perspective. We do this by comparing the two different desalination systems at the plant level and at the point of separation of water from seawater. By investigating RO and MED entropy generation by transport processes at the membrane interface and in the evaporators, respectively, we gain a better understanding of the thermodynamic differences between chemical separation by membrane separation and by evaporation. By investigating the technologies at the plant level, we better understand the efficiency differences between these two processes in their broader operating context, taking into account the differences between the two technologies that exist in intakes, pumps, pretreatment technologies, and additional components such as feed heaters.

Finally, we show that practical RO systems outperform MED from an energetic perspective as a result of a combination of the costs of materials and the differences between the effective transport coefficients of membranes and heat exchangers.

2 Metrics of Energy Consumption for Desalination

Metrics for desalination energy consumption can broadly be broken down into two categories, based on the location at which energy consumption is measured. The first category is calculated
at the desalination plant inlet. These metrics measure the amount of heat, electricity, or both that enter the desalination plant.

The second category of metrics use a more complex calculation of energy consumption. These metrics take into account that the energy going into the desalination plant is derived from some primary energy source, whether that be a fossil fuel, heat from nuclear fission, or some form of renewable energy. In these cases, the power-generation system must be considered together with the desalination plant. Because some technologies require steam to drive desalination, power plants and desalination plants have often been colocated. The power plant provides steam and electricity to the desalination plant and sells electricity for the grid.

A coproduction system is shown schematically in Fig. 1. Several locations are highlighted. Metrics based on energy entering the desalination plant are evaluated at location 1 in Fig. 1. Metrics based on energy entering the power plant evaluate consumption at either point 2 or point 3. For the sake of simplicity, and so as not to involve the efficiency of combustors or solar absorbers, we will consider energy consumption at point 2. For additional discussion of this point, see Refs. [4], [11], and [12].

2.1 Measures of Energy at Desalination Plant Inlet. The most obvious way to measure energy consumption of desalination plants is by the amount of energy entering the desalination system itself, at location 1 in Fig. 1. This energy consumption is generally reported in one of three ways: PR, GOR, or specific energy consumption (SEC).

Performance ratio can be defined in several ways, all of which compare the actual heat input to some reference quantity, as in the following equation:

$$\text{PR} \equiv \frac{L_c}{q_{wp}}$$

(1)

In this equation, $q_{wp}$ is the heat input per unit mass of distillate and $L_c$ is some reference specific heat. The value of PR depends on how $L_c$ has been defined [2]:

- When $L_c$ is the latent heat of vaporization at 73 °C, which is 2326 kJ (1000 Btu), the performance ratio is the mass of distillate generated per this heating unit, in units of kg/2326 kJ.
- When $L_c$ is the enthalpy change of heating steam at its saturation temperature, the performance ratio becomes a mass ratio, describing the mass of product water per unit mass of heating steam. The units are kg/steam/kgsteam.
- When $L_c$ is the latent heat of vaporization at $T_0$ (the environment temperature), then the result is an energy ratio that is also called the GOR, with units of kJ/kJ. GOR can also be viewed as a measure of the number of times the latent heat of vaporization is recycled from condensing vapor to generate additional pure water vapor [13].

When the thermal energy input is provided by condensing steam, the performance ratio is related to GOR by Eq. (2) [13]

$$\text{GOR} = \frac{q_{wp}(T_0)}{h_g(T_0)}$$

(2)

These three formulations provide different values and units, but all are of the same order of magnitude, and the spread between them is usually unimportant. Further discussion of PR can be found in Mistry et al. [13] and El-Sayed and Silver [2].

Specific energy consumption is a common metric for work-driven systems, such as RO. SEC is a ratio of electrical energy or mechanical work consumption to either the mass or volume of product water. The formula for SEC is shown in the following equation:

$$\text{SEC} \equiv \frac{\rho W_{\text{desal}}}{m_p}$$

(3)

Specific energy consumption is commonly reported in units of kWh/m³, or by omitting $\rho$, in units of kWh/kg.

The metrics described above have several shortcomings. The first issue is that the methods above do not distinguish between various grades or qualities of energy. One kWh of thermal energy provided by steam at 100 °C can provide less work than one kWh of thermal energy from steam at 150 °C. Although the quantity of energy is the same, and the units are technically the same, the two quantities of energy do not have the same thermodynamic utility. To differentiate between the two, the exergy, or availability, should instead be considered. Exergy measures the maximum amount of useful work that can be extracted from a given stream or from a given heat transfer as the stream interacts with and reaches equilibrium with the environment or as heat transfer drives a reversible Carnot cycle rejecting heat to the environment.

If we were to compare a MED plant with a GOR of 10 to a multistage flash plant with a GOR of 10, we could correctly say that both plants are able to reuse the enthalpy of vaporization of condensing steam the same number of times, but we could not say that the thermal energy requirements were the same, because multistage flash plants require steam of a much higher temperature than MED plants. This concept is expanded upon in Sec. 2.2. To make a fair comparison, we would like the quantities of energy being compared to be able to produce the same amount of work.

The second issue is the cost of different grades of energy. The design of a desalination plant is an exercise in cost optimization, and so we should understand the cost implications of these energy metrics as well. One kWh of steam at 100 °C does not have the

![Fig. 1 Diagram of a generalized water and power coproduction system, where incoming heat is generated by either a solar collector and absorber or a fossil fuel combustor. The three highlighted numbers indicate three possible locations of measuring energy input.](image-url)
same dollar value as one kWh of steam at 150 °C, and neither has the same dollar value as one kWh of electricity. Therefore, comparing different desalination plants using GOR and PR is not an apples to apples comparison from either a thermodynamic or a financial perspective.

A third, essential issue is that these thermal metrics do not describe the entire energy consumption of desalination plants that use multiple sources and/or grades of energy. For example, a single MED plant with thermal vapor compression (MEDT) can have three distinct energy inputs: low temperature steam for normal operation of the MED plant, high temperature steam to operate the thermocompressor, and electricity to drive pumps throughout the plant. None of the metrics described above can combine all of these energy inputs, as all three energy inputs have a different thermodynamic (and economic) value. This issue also prevents us from analyzing more complex hybrid desalination plants incorporating multiple grades of energy.

Although these three factors preclude truly fair comparison of desalination plants in many cases, any of the above metrics can provide a fair comparison when the desalination plants being compared use energy of the same source and grade (e.g., steam at the same temperature, pressure, and quality). For example, we can make direct, fair comparisons between systems such as RO, mechanical vapor compression, and electrodialysis that are completely powered by electrical work. Thermal systems are much less amenable to fair comparison under these metrics.

Desalination metrics based on energy flows at only the desalination plant inlet do not take into account the quality of the energy, the source of the energy, or opportunity cost of turning the primary fuel into the derived energy being used for desalination. Translating these metrics back to the economic cost of energy can be very difficult when dealing with extracted steam (or similar energy streams), rather than electricity.

2.2 Measurement of Exergy at Desalination Plant Inlet.

The shortcomings and limitations of metrics like GOR, PR, and SEC might potentially be overcome by comparing energy inputs on the basis of exergy. Exergy, $\dot{E}$, also called availability, measures the maximum useful work that can be extracted when transferring heat from a high temperature stream at $T_H$ to an environment at $T_0$, through a reversible heat engine

$$\dot{E} = \dot{Q}(1 - \frac{T_0}{T_H})$$

Mechanical and electrical work can be interconverted without loss by a reversible system. Both can be equated to an equivalent magnitude of input exergy.

A desalination system’s exergy consumption may be normalized by dividing by the mass or volumetric flowrate of product water (specific exergy consumption). The exergetic or second law efficiency of a desalination plant is the ratio of least work (least exergy) of separation to actual exergy consumption

$$\eta_{\text{leq}} = \frac{\dot{W}_{\text{leq}}}{\dot{E}} = \frac{\dot{E}_{\text{leq}}}{\dot{E}}$$

Least work is the difference between the Gibbs free energy of the product and brine streams relative to the feed stream. The minimum least work is that for vanishingly small pure water recovery, as if separating a single drop of pure water from an ocean. For seawater at Arabian Gulf conditions (33 °C and 44 g/kg), which are the conditions we will focus on in this paper, the minimum least work of separation is 1.02 kJ/m³. In contrast, the least work of separation for a desalination plant operating at Arabian Gulf conditions and 50% recovery is 1.39 kJ/m³.

In this analysis, we calculate second law efficiency using the minimum least work of separation. Although different desalination plants operate at different recovery ratios, a fair comparison across plants should focus only on the useful end product: purified water. The value of that water is unrelated to the recovery ratio of the system that produced it. In other contexts, there are good reasons to consider the impact of recovery ratio on the thermodynamic minimum energy consumption, and therefore to define second law efficiency with respect to least work as opposed to minimum least work. For example, in brine concentration systems, high recovery ratios are required and the degree of recovery very strongly affects the energy utilization [14]. In this analysis, however, we assume that a nonzero recovery ratio serves only to minimize the levelized cost of water by reducing capital costs, without any thermodynamic benefit. Further discussion of the differences between least work and minimum least work can be found in the literature [11,13,14].

From a thermodynamic perspective, an exergetic comparison of desalination plants, at the point of entry (location 1 in Fig. 1), is a fair, apples-to-apples comparison. Regardless of the type or grade of energy going into the desalination plant, the energetic inputs are converted to a basis that has units of the same value—kW of exergy.

The exergetic comparison, however, does not resolve the need to compare energy on an economic basis, since different exergy inputs may have different costs: the cost of producing one kWh of steam exergy is not the same as the dollar cost of producing one kWh of electricity. Financial comparisons between thermal and membrane desalination technologies, in particular, are not resolved by exergy.

Furthermore, all exergy streams do not have the same thermodynamic cost, even though all exergy has the same thermodynamic value. We use power plants to convert energy—and therefore exergy—from one form to another, but the processes used are not ideal. Power plants convert primary energy, such as fossil fuels or sunlight, into derived energy, such as electricity or steam, at different efficiencies. To account for these varying efficiencies, we need to take a broader view of the system by considering primary energy.

2.3 Primary Energy.

Desalination plants are generally powered by derived energy sources such electricity or steam. For the latter, plants are frequently colocated with power plants that convert PE into the required secondary or derived energy, usually while producing additional electricity that is sold to the grid. Primary energy enters at location 3 in Fig. 1.

A given source of PE provides energy at the same thermodynamic and economic cost and value regardless of the type of derived energy needed to power the desalination system being used. PE provides a fair basis for comparing any combination of power plant and desalination system in a coproduction scheme as long as they are powered by the same fuel source. We can also capture the entire energy footprint of a desalination plant, including intake pumping, pretreatment, desalination, post-treatment, and outfall pumping in a single PE metric. PE consumption is also easily converted into fuel cost.

In this paper, we consider PE at location 2 in Fig. 1, which is the point after solar absorption or fuel combustion. This approach simplifies calculations by separating the inefficiency of the solar absorber or the combustor. The result can easily be converted to solar radiation or fuel input at location 3 by dividing by the appropriate efficiency [4,11].

The literature contains several methods for computing PE, which has caused some confusion. We discuss three of these methods here.

2.3.1 Fixed Second-Law Efficiency Power-Loss Method. This approach is grounded on the important assumption that the power plant operates with a single second law efficiency, irrespective of the steam extraction location. The second law efficiency of a power plant is the ratio of actual power production to the maximum power that could be produced by a reversible heat engine.
Steam that is extracted for desalination is no longer available for conversion into power. The plant efficiency is then

\[
\eta_{pp}^{H} = \frac{\dot{W}}{\dot{Q}_{PE} \left( 1 - \frac{T_0}{T_H} \right) \eta_{pp}^{H}}
\]

(6)

The steam exergy and/or electricity extracted from the power plant to run the desalination plant results in a reduction of the amount of electricity available for sale to the grid—a power loss. To offset the power loss, additional primary energy (or exergy) is required. Thus, extracted electrical and steam energy must be evaluated in terms of primary exergy input. Under the assumption of a constant second law efficiency for the power plant, this conversion is described by the following equation:

\[
\dot{Q}_{PE,desal} \left( 1 - \frac{T_0}{T_H} \right) \eta_{pp}^{H} = \dot{W}_{desal} + \dot{Q}_{desal,steam} \left( 1 - \frac{T_0}{T_s} \right) \eta_{pp}^{H}
\]

(7)

Rearranging Eq. (7) leads to Eq. (8) and Eq. (9):

\[
\dot{Q}_{PE,desal} = \frac{\dot{W}_{desal}}{1 - T_0/T_H} \eta_{pp}^{H} + \dot{Q}_{desal,steam} \left( 1 - \frac{T_0}{T_s} \right) \eta_{pp}^{H}
\]

(8)

\[
\dot{Q}_{PE,desal} = \frac{\dot{W}_{desal}}{1 - T_0/T_H} \eta_{pp}^{H} + \dot{W}_{desal,steam} \eta_{pp}^{H}
\]

(9)

In Eqs. (6)–(9), \( \eta_{pp}^{H} \) is the second law efficiency of the power plant. \( T_0 \) is the dead state temperature, \( T_H \) is the postcombustion or postabsorption fuel temperature, \( T_s \) is the temperature of extracted steam, \( \dot{Q}_{desal,steam} \) is the rate of steam energy extraction from the power plant for desalination, \( \dot{W}_{desal} \) is the electrical power used for desalination, and \( \dot{Q}_{PE} \) is the rate of consumption of primary energy. These terms are also shown in Fig. 1.

The justification for assuming a fixed power plant second law efficiency is that the desalination plant generally represents a small portion of the total exergy consumed in a coproduction system. Shahzad et al. [7] calculated that an MED plant in a power–water coproduction scheme used about 3% of the exergy input to the coproduction system. Similarly, Altmann et al. [4] showed that the change in power-plant second law efficiency is less than 2% when desalination plants are added. Both analyses seem to imply that we can assume the efficiency remains constant without introducing large errors [11]. However, even a small change in \( \eta_{pp}^{H} \) can change the primary energy required for power production by an amount comparable to the primary energy needed for desalination, as is shown in Sec. 2.3.2. Thus, we cannot neglect changes in \( \eta_{pp}^{H} \) when calculating the primary energy consumption (or power loss) associated with the desalination process.

In cases of steam extraction, the turbine efficiency need not be the same before and after the location of extraction. At the beginning of the portion of the turbine, where the temperature and pressure of the steam are highest, power is generated at a different efficiency than the final portion of the turbine, where the steam temperature and pressure are much lower [15]. This variation is shown schematically by the hypothetical second law efficiency curve in Fig. 2.

The steam extracted, had it remained in the turbine, would generate power at the turbine’s average efficiency beyond the location of extraction, \( \eta_{post-extraction}^{H} \). On the other hand, steam added to make-up the power loss from extraction will generate power at the average efficiency for the turbine, \( \eta_{pp}^{H} \), which is different. As a result, the second law efficiency terms on either side of Eq. (7) are not likely to be equal, and the fixed-efficiency power-loss method will be inaccurate. More on this method and its shortcomings can be found in Altmann et al. [4].

2.3.2 Variable-Efficiency Method. This method compares a standalone power plant, generating a fixed amount of power for the grid, to a coproduction system producing the same amount of power for the grid and also a specified amount of water. The difference in PE between these two plants is then entirely attributable to the desalination system. This additional PE includes both the PE used to produce the steam and/or electricity for desalination and PE to compensate for any changes in power plant performance caused by the addition of the desalination plant. In this case, we assume, as above, that a system producing electricity and using RO for water production includes a power plant that is simply optimized for power production. This baseline plant has a fixed power cost and efficiency because the water and power systems are effectively decoupled when RO is used.

The approach of comparing a baseline plant without desalination to a coproduction system with desalination goes back at least as far as El-Sayed and Silver [2].

To derive PE consumption using this method, we first consider a power plant with no desalination system, optimized for power production. Using plant simulation tools (such as Thermoflow) or data, we can determine the plant’s first law efficiency

\[
\eta_{f}^{H} = \frac{\dot{W}_{grid}}{\dot{Q}_{PE,grid}}
\]

(10)

When no steam is extracted, we have the baseline first law efficiency \( \eta_{f}^{H} \). Next, we can simulate the combined power/water coproduction plant, which will produce a different first law efficiency

\[
\eta_{f}^{H} = \frac{\dot{W}_{grid} + \dot{W}_{desal}}{\dot{Q}_{PE,co-prod}}
\]

(11)

Both plants produce the same amount of power for the grid, \( \dot{W}_{grid} \). We attribute the difference between \( \dot{Q}_{PE,grid} \) from Eq. (10) and \( \dot{Q}_{PE,co-prod} \) to the desalination plant, as shown in Eq. (11). These two equations are solved for the PE consumption of the desalination system:
\[ \hat{Q}_{\text{PE, desal}} = W_{\text{grid}} \left( \frac{1}{\eta_{\text{PP,b}}} - \frac{1}{\eta_{\text{PP,e}}} \right) + W_{\text{desal}} \frac{1}{\eta_{\text{PP,e}}} \]  

The benefit of this method is that it makes no assumptions about power plant efficiency. The power plant efficiency is allowed to change depending on the amount and grade of steam being extracted from the plant. One exception is that the power plant efficiency is assumed not to change when only additional electricity is required, and the plant is simply scaled up to produce more power.

To account for changes in power production efficiency, we can perform a similar analysis that considers average baseline and extraction efficiencies of the power plant, but uses second law efficiencies

\[ W_{\text{grid}} = \hat{Q}_{\text{PE, grid}} \left( 1 - \frac{T_0}{T_H} \right) \eta_{\text{PP,b}}^{\text{II}} \]

\[ W_{\text{grid}} + W_{\text{desal}} = \left( \hat{Q}_{\text{PE, grid}} + \hat{Q}_{\text{PE, desal}} \right) \left( 1 - \frac{T_0}{T_H} \right) \eta_{\text{PP,e}}^{\text{II}} - \hat{Q}_{\text{desal,steam}} \left( 1 - \frac{T_0}{T_e} \right) \eta_{\text{PP,e}}^{\text{II}} + W_{\text{desal}} \frac{1}{\eta_{\text{PP,e}}^{\text{II}}} \]

Combining and rearranging these equations

\[ \hat{Q}_{\text{PE, desal}} \left( 1 - \frac{T_0}{T_H} \right) = W_{\text{grid}} \left( \frac{1}{\eta_{\text{PP,e}}} - \frac{1}{\eta_{\text{PP,b}}^{\text{II}}} \right) + \hat{Q}_{\text{desal,steam}} \left( 1 - \frac{T_0}{T_e} \right) + W_{\text{desal}} \frac{1}{\eta_{\text{PP,e}}^{\text{II}}} \]  

Equation (15) will produce the same results as Eq. (12) for a given coproduction system. Equation (15) also clearly shows why, if \( W_{\text{grid}} \) is large compared to the amount of exergy used in the desalination plant, a small change in \( \eta_{\text{PP,e}}^{\text{II}} \) can lead to large changes in the amount of primary energy used by the desalination plant.

The drawback of this method is that it requires accurate models for power plant operation and knowledge of how steam turbine operation changes with variable steam extraction. Altmann et al. [4] provide a large dataset using real power plant data and industry desalination plant data. Models describing steam extraction turbines are also available [15-18].

The results of the variable-efficiency method are almost always more favorable to thermal desalination technologies than the fixed-efficiency method. The low temperature steam that is used to power desalination plants is extracted from the sections of the turbine that have lower efficiency. This has the effect of increasing the overall average second law efficiency of the power-producing sections of the turbine, which in turn means that the actual power loss for steam-powered desalination systems is less than would be predicted by the fixed-efficiency method of Sec. 2.3.1.

2.3.3 Alternative Methods. Another method for finding primary energy consumption, proposed by Shahzad and coworkers, uses conversion factors to convert derived energy (steam and electricity) back into primary energy [6,7]. The conversion factor they use for electricity is the second law efficiency of the power plant, which is consistent with Eq. (9). To determine the primary energy used to generate steam for the desalination system, the authors considered the exergy utilization within the major components of a representative combined-cycle gas turbine (CCGT) and thermal desalination coproduction system. Their analysis includes the gas turbine, heat recovery steam generator, steam turbines, and desalination system. They take the percentage of exergy embodied in the desalination steam, \( \Xi_{\text{desal,steam}} \), as a function of the total primary exergy input into the cogeneration system, \( \Xi_{\text{PE}} \), to determine a conversion factor

In theory, one could arrive at the same results as Eq. (9) (apart from differences between the power plants considered) without even requiring this conversion factor; however, these authors appear to have multiplied thermal energy terms by this conversion factor, which is an exergetic factor. In addition, they treated the primary exergy used to produce electricity as primary energy. These errors artificially shift the outcome of the analysis to be more favorable to thermal technologies, as claimed in the cited papers. The authors also propose a single conversion factor for use with multiple grades of steam; we would argue that each grade of steam needs to have its own conversion factor to account for the difference in grade.

2.3.4 Metrics Derived From Primary Energy. Primary energy consumption may be normalized or reported in several ways to create a more useful and understandable metric. Most obviously, PE consumption can be divided by desalination plant production to produce a specific primary energy consumption with units of (kWhPE/m³).

Primary energy can also be converted into primary exergy using a temperature factor, as in Eq. (4). Then, a primary second law efficiency can be found, where the least work of separation is divided by the primary exergy consumption, as in Eq. (5). This describes how efficiently the entire system converts fuel (or in this case the thermal exergy from burned fuel or absorbed solar) into desalinated water relative to thermodynamic limits.

When examining large coproduction systems (like those in Sec. 3), we can determine a system-wide combined second law efficiency

\[ \eta_{\text{PP,e, system}} = \frac{W_{\text{min, least}} + W_{\text{grid}}}{\hat{Q}_{\text{PE, desal}}} \]  

This metric compares all the useful work done by the system, both in producing electrical work for the grid and producing useful chemical separation work, to the total primary exergy input to the system. However, the difference between systems will be greatly affected by the relative sizes of the power and desalination plants. In particular, if the power plant produces much more useful work as grid electricity than in the form of desalinated water, this system-wide second law efficiency will closely follow the second law efficiency of the power plant, and vice versa.

Shahzad and coworkers have proposed the UPR [6], which they also call standard universal performance ratio (SUPR) [7]. This metric compares the primary energy used for desalination, \( \hat{Q}_{\text{PE, desal}} \), to the latent heat of vaporization of water, 2326 kJ/kg, as in the following equation:

\[ \text{UPR} = \frac{2326 \text{ (kJ/kg)}}{3.6 \times \hat{Q}_{\text{PE, desal}} \text{ (kJ/kg)}} \]  

The resulting UPR can range from zero to the limit imposed by the laws of thermodynamics, which was computed in Ref. [6] by setting \( \hat{Q}_{\text{PE, desal}} \) equal to the minimum least work (as \( T_{\text{II}} \to \infty \)) in Eq. (18). When the minimum least work is set to 0.78 kWh/m³, the resulting maximum UPR is approximately 830 [6]. However, we argue that this limit cannot be reached even by a system consisting of ideal, thermodynamically reversible power and desalination systems, unless we consider primary exergy in the denominator instead of primary energy, as even a reversible Carnot engine cannot convert a given amount of fuel energy directly into the same amount of power. The maximum achievable UPR varies depending on the seawater conditions; and it will be higher for lower salinity feed streams. The actual significance of the unitless number produced by the UPR or SUPR equation is similar to
PR, in that we can think of it as the number of times that the enthalpy of vaporization is recycled within a system or as the number of times more water we can produce when the alternative is boiling off water vapor using the specified type of primary energy. When UPR is computed for systems analyzed by Altmann et al. [4], UPR for CCGT-powered systems ranged from 31 for MEDT to 104 for RO, with a maximum possible UPR of 632.

While UPR could be used to compare any desalination systems that are powered with the same primary energy source, it lacks true universality in that it cannot compare systems that are powered by different primary energy sources, although no metrics described in this article can. While we include values for UPR in our analysis in Sec. 3, the primary-exergy second-law efficiency is a more direct way to compare actual performance of a coproduction system to the second-law limit, and it does not involve an opaque conversion factor.

2.3.5 Limitations. The variable-efficiency method of calculating PE consumption can combine all energy input to a plant into a single metric that has a consistent value in terms of both thermodynamics and cost, and which easily translates into fuel costs. We consider the variable-efficiency method of calculating PE consumption to be the surest comparison of energy consumption over a wide variety of desalination plants.

Although comparisons of PE are robust, several potential limitations of this metric should be mentioned, as previously noted by Altmann et al. [4].

- Comparing by PE consumption only allows for plants with one fuel source to be compared. If comparison over various different fuel sources is desired, a full technoeconomic analysis is required.
- Complex systems that cascade energy through a series of different subsystems can make PE consumption difficult to calculate in practice. While the overall framework and concepts laid out here still apply to complex systems, formulas to calculate efficiency, such as Eq. (11), may need to be adapted for particularly innovative applications, which is outside the scope of this paper.
- Comparing by PE consumption does not constitute a comparison of total operating expense. Costs of labor, chemicals, membrane replacement, downtime, cleanings, etc. are not represented.
- Comparing by PE consumption requires knowledge of how power plants operate both with and without steam extraction in order to make accurate estimates for thermal desalination plants. Simulations or complicated models may be needed.

Care must be taken in determining what operating conditions or parameters are held constant when comparing systems in order to fairly evaluate differences. For example, comparing desalination plants with different recovery ratios or different feed salinities is not necessarily fair from a thermodynamic point of view because a plant operating at higher recovery ratio or with more saline feed is performing a greater amount of separation work than a plant operating at low recovery ratio or lower salinity.

In selecting desalination plants, the determining factor is usually not GOR, SEC, exergy consumption, primary energy consumption, or even fuel cost, but instead the levelized cost of water. Producing water at the lowest cost while meeting any performance, environmental, legal, location, and other constraints will win bids. These constraints will continue to shift as our scientific understanding and societal values regarding water and the environment continue to change and develop. For example, we do not include any analysis of the climate change impacts of the technologies considered in this paper, although emissions reduction is an increasingly important driver of technology selection, such as in the NEOM project in Saudi Arabia [19]. Additionally, the recent steep decline in solar energy costs, especially in the Middle East [20], could lead to paradigm shifts, such as smaller, decentralized desalination plants located closer to end users to reduce conveyance costs. In the selection of technologies and plant designs, it is vital that this changing landscape of values and constraints is considered, while still minimizing levelized cost of water.

3 Sample Comparison of Desalination Technologies

Many of the points made in Sec. 2 are best illustrated with examples. Using selected data from Altmann et al. [4], we can compute all of the relevant performance metrics (Table 1). Here, we consider RO, MED, and MEDT, which are shown in Figs. 3–5. We also look at plants with nanofiltration (NF) pretreatment, which enables thermal plants to achieve higher recovery ratios than MED to match RO in terms of primary energy consumption. These desalination plants and their operating conditions are described in depth by Altmann et al. [4], but we note here that the selected data reflect current state-of-the-art technology for desalination plants in the Gulf Cooperation Council (GCC) region, including the use of energy recovery devices. In an effort to make a true “apples to apples” comparison, the data account for energy consumed for the entire treatment process (including intake, pretreatment, outfall, etc.). Each system is constructed according to real-world best practices to minimize levelized cost of water [21]. In the coproduction scheme, all systems produce the same amount of power and water. Thus, the plants considered do not have the same recovery ratio and are not optimized to minimize energy consumption, but provide a more accurate representation of the energy consumption of these technologies under real world operating conditions.

The data illustrate several of the points made in Sec. 2. If the only information available to us were GOR, PR, and SEC, then MEDT would seemingly outperform MED and NF-MEDT would look energetically favorable to NF-MED. However, when we consider exergy instead of energy at the desalination inlets, we see that the MED and NF-MED use less exergy than MEDT and NF-MEDT, respectively, changing the outcome of the comparison. As was noted in Sec. 2.2, comparing on the basis of exergy instead of energy helps to take into account the relative value of steam at varying grades and gives a better idea of the thermodynamic cost of desalination.

If we take this comparison all the way back to primary energy consumption, which is directly correlated to fuel consumption, we see that MED outperforms NF-MED, as well as the other thermal systems, again shifting the relative order of performance. Additionally, while MED uses approximately three times the exergy of an RO system at the desalination system inlet (location 1 in Fig. 1), MED uses less than twice the primary energy (at location 2 in Fig. 1) of an RO system. The gap between RO and thermal technologies is reduced when we consider primary energy because of the larger thermodynamic penalty paid to convert primary energy into electricity rather than to convert primary energy into steam.

Clearly, RO outperforms the thermal technologies for all cases shown in this paper. This finding is by no means novel; however, there remain the questions of why, from a fundamental thermodynamic standpoint, RO outperforms evaporative technologies and of what would need to change for an evaporative technology to achieve greater efficiency than RO.

4 Fundamental Thermodynamic Comparison of Reverse Osmosis and Multiple Effect Distillation

The goal of this section is to uncover the fundamental reasons, from a thermodynamics and heat and mass transfer perspective, for RO’s lower primary energy consumption compared to MED, and to understand what advancements would be required in order for MED to match RO in terms of primary energy consumption. We consider two approaches to answer this question.

4.1 Plant-Level Comparison. First, we examine the RO and MED systems from Altmann et al. [4] to determine what would
need to change within the MED system in order to achieve the same primary energy and primary exergy consumption as RO. Selected plant data are presented in Table 2, and additional data regarding the operating conditions and environmental parameters of desalination systems considered here can be found in the literature [4].

According to Altmann et al., an RO plant operating at Arabian Gulf conditions has a total primary exergy consumption of 4.96 kWh/m³ [4]. Our hypothetical improved MED plant should consume the same primary exergy, while having a design similar to the MED plant described by Altmann et al. Primary exergy will be converted in the power plant into three different grades of derived exergy to be consumed by the MED system: (1) electricity to power pumps and other needs within the plant; (2) medium pressure (MP) steam for the ejectors that provide vacuum; and (3) low pressure (LP) steam to provide the thermal energy that drives evaporation in the effects.

The aim of this analysis is to determine the maximum amount of LP steam exergy can be used by the MED plant without the system exceeding the allowed total primary exergy consumption. Our focus is on improving heat recovery in the effects so that overall steam consumption will be reduced. We assume that the electrical consumption and the MP steam consumption will be essentially the same for the improved MED plant as for the original MED plant, and we hold the recovery ratio and environmental conditions fixed.

To determine how primary exergy is split between the three exergy needs, we use simple conversion factors derived from the values in Table 2. The derivation of these factors is explained in Appendix A. With these conversion factors, we find that the

| Performance metric | RO     | MED    | NF-MED | MEDT   | NF-MEDT |
|-------------------|--------|--------|--------|--------|---------|
| Specific thermal energy input kWh/m³ | 0.0    | 72.9   | 52.2   | 59.0   | 41.4    |
| SEC kWh/m³ | 3.7    | 1.9    | 2.2    | 1.6    | 2.1     |
| PR kg Brayton cycle/2326 kJ | —     | 9.4    | 13.5   | 11.8   | 17.5    |
| GOR | —     | 9.2    | 13.3   | 12.5   | 18.1    |
| Specific exergy input kWh/m³ | 3.7    | 11.3   | 10.7   | 15.9   | 12.3    |
| Desal second law efficiency at desal plant inlet % | 27.8   | 9.1    | 9.5    | 6.4    | 8.3     |
| Power plant first law efficiency % | 59.1   | 57.7   | 57.7   | 56.0   | 57.0    |
| Power plant second law efficiency % | 74.0   | 74.8   | 74.5   | 73.9   | 74.1    |
| Specific PE consumption, variable-efficiency method kWh/m³ | 6.2    | 11.3   | 11.9   | 20.8   | 15.9    |
| Specific PE consumption, fixed-efficiency method kWh/m³ | 6.2    | 14.9   | 14.4   | 20.7   | 16.3    |
| Desal second law efficiency, from PE % | 22.7   | 12.4   | 11.8   | 6.8    | 8.8     |
| UPR/SUPR | —     | 104    | 57     | 54     | 31      |
| Co-production overall second law efficiency % | 73.01  | 71.92  | 71.79  | 69.88  | 70.93   |

Table 1  A variety of performance metrics for several simulated desalination plants. All plants produce 100,000 m³/day, operate in Arabian gulf conditions, and are powered by a CCGT plant [4]
maximum allowed LP exergy for the new MED plant is 1.89 kWh/m³, which is far less than the LP steam exergy consumption of 8.2 kWh/m³ for the plant described by Altmann et al. Because the LP steam in both the new and old cases has the same temperature, quality, and pressure, we see that the PR must improve by approximately a factor 4, approaching PR = 40.

To determine the operating conditions that would be necessary to achieve such a high PR, we simulate MED plant performance using the equations provided in Appendix B. We assume the same top brine temperature, dead state temperature, and recovery ratio as the previous MED plant. The baseline heat transfer efficiencies in the MED effects and feed heaters is taken from a design correlation by El-Dessouky and Ettouney [22]. In order for the MED system to be able to achieve a PR of 40, the terminal temperature difference in the feed heaters must be 0.22 °C. The resulting plant requires 52 effects, has a flux of 0.03 L/m² h, and operates with a thermal driving force of only 0.01 °C! These numbers are unachievable with current technology.

This plant would be unrealistically large, as the original MED plant in Ref. [4] had a flux of 6.6 L/m² h. Further, the temperature differences mentioned would be very difficult to sustain. If we could increase the heat transfer coefficient within MED systems without increasing the cost per unit area, we could keep the same footprint as the previous MED plant, approximately holding capital costs constant. In order to do this, while still matching RO’s primary exergy consumption—and thus approximately matching the operating costs of RO—the plant would need to operate at a much higher flux. The overall heat transfer coefficients of the effects and the feed heaters (U_f, U_h) would have to go from 2250 to approximately 51,600 W/m² K, an increase of a factor of 23. This is the target overall heat transfer coefficient in the effect (including condensation and evaporation). This overall heat transfer would require substantial reductions in the resistance of the liquid film and a highly efficient convective condensation process within the tubes.

### 4.2 Comparison at Separation Interface

The approach described above requires assumptions about conversion factors and what constitutes “similar” plants. Depending on how the problem is framed, the required increase in performance by the MED system could vary. In this section, we take a different approach. We will compare the separation phenomena occurring at the membrane for an RO system to that in the evaporators of MED systems.

For any chemical separation, including desalination systems, energy efficiency can be improved by reducing entropy generation or exergy destruction within the system. The exergy consumption in RO and MED systems can be broken into five main terms, as shown in the following equation:

\[
\tilde{\varepsilon} = \tilde{\varepsilon}_{\text{feed}}^{\text{min}} + \tilde{\varepsilon}_{\text{RR} \leq 0} + \tilde{\varepsilon}_{\text{flux}} + \tilde{\varepsilon}_{\text{non-equ}} + \tilde{\varepsilon}_{\text{par}}
\]  

(19)

These terms represent the minimum least work of separation (required to produce a miniscule quantity of pure water from the feed solution), the additional exergy required to achieve the desired nonzero recovery ratio, the excess exergy required to achieve a nonzero flux, the additional exergy lost due to nonequivalencing of entropy generation, and all other sources of exergy destruction, respectively [23–26]. The first four right-hand-side terms are directly related to the separation of pure water from the feed stream. The last term, \(\tilde{\varepsilon}_{\text{par}}\), includes all parasitic destruction of exergy that occurs outside the separation process itself, such as pump or turbine inefficiencies, chemical mixing, heat loss to the environment, and viscous dissipation in pipes, membranes, and heat exchangers. For this portion of the analysis, we ignore parasitic exergy destruction and focus on the separation interface. We note that Mistry et al. found that over half the entropy generation of typical RO and MED systems occurred in the RO module and MED effects [13, 27].

For desalination plants that operate at the same recovery ratio and with the same feed solution, the first two terms of Eq. (19) are always equal. The difference in exergy consumption at the point of separation will come down to two factors: the exergy required to achieve the desired nonzero flux, and the exergetic cost of system imbalance. Generally, large-scale MED systems, with their high number of effects, are very well equipartitioned compared to standard continuous RO systems that use one or two stages. In this analysis, we will consider perfectly equipartitioned MED and RO systems.

#### 4.2.1 Reverse Osmosis

First, we consider an RO system. The mass flux of water through the membrane from feed water to product is given by the solution–diffusion equation

\[
j = \rho A (\Delta P - \Delta \pi)
\]

(20)

where \(j\) is flux in kg/m² s, \(\rho\) is liquid density, \(A\) is membrane permeability in m³/m² s Pa (or, more commonly, in L/m² h bar), and \((\Delta P - \Delta \pi)\) is the hydraulic pressure difference across the membrane minus the osmotic pressure difference across the membrane.

The instantaneous local entropy generation per unit volume in any system where mass, heat, or momentum is transferred from one region to another can be determined from the product of the appropriate flux and its associated thermodynamic driving force [12, 23, 26]. In the case of RO, a detailed analysis of the driving force [12] shows that the entropy generation per unit membrane area is

\[
S_{\text{gen,RO}}^\sigma = \frac{j (\Delta P - \Delta \pi)}{\rho T_0}
\]

(21)

where \(T_0\) is the RO system’s dead state temperature. Substituting Eq. (20) and multiplying by membrane area, \(A_{RO}\), we find

\[
S_{\text{gen,RO}}^\sigma = A_{RO} \frac{\Delta g_{RO} (\Delta P - \Delta \pi)^2}{T_0}
\]

(22)

Representative operating conditions for seawater RO systems are an average volumetric flux of approximately 15 L/m² h, with membrane permeabilities of about 1 L/m² h bar [30]. For a perfectly equipartitioned system, this flux would be constant at every point in the system. In real systems, the flux varies from approximately 25 L/m² h in the first elements of a membrane train, where the osmotic pressure difference across the membrane is low, to

---

3The original design of RO and MED systems presumably aimed to minimize the levelized cost of water, which entails optimization between capex and opex. Here, we have effectively fixed the capex and reduced opex, so we are no longer operating at the cost-optimal point.

4Recently introduced batch RO and closed-circuit RO systems improve the equipartitioning of entropy generation both in time and space [28, 29].

5In Eq. (31) of Ref. [12], membrane permeability has units of mol/m² s Pa.----

---

#### Table 2 Plant design data for three desalination plants [4]

| Quantity           | RO   | MED | MD   |
|--------------------|------|-----|------|
| Top brine temperature °C | 66.0 | 66.0 |
| PR                 | 9.4  | 9.0 |
| LP steam temperature °C | 74.6 | 74.6 |
| LP steam exergy utilized kWh/m³ | 8.2 | 8.5 |
| MP steam temperature °C | 265.3 |
| MP steam exergy utilized kWh/m³ | —    |
| SEC                | 3.67 | 1.85 | 1.95 |
| Recovery ratio     | 40   | 30  | 60  |
| Primary exergy—thermal kWh/m³ | 6.48 | 5.55 |
| Primary exergy—electrical kWh/m³ | 4.96 | 2.56 | 2.69 |
| Primary exergy—total kWh/m³ | 4.96 | 9.04 | 8.24 |
| Second law efficiency | 20.6 | 11.3 | 12.4 |
4.2.2 Multi-Effect Distillation. For an MED system, the heat flux across the evaporator is determined by Eq. (23), and the condensation mass flux by Eq. (24).

\[ q = U (\Delta T - \text{BPE}) \]  
\[ j = \frac{U}{h_{fg}} (\Delta T - \text{BPE}) \]

where \( q \) is the heat flux per unit area, \( U \) is the average overall heat transfer coefficient of the heat exchanger, and \( \Delta T \) is the temperature difference between subsequent effects, and BPE is the boiling point elevation caused by salts [31].

The entropy generation per unit area in an MED process due to transport of pure water from the hot salty feed stream to the cooler vapor can be expressed as the heat flux times the thermodynamic driving force for a heat transfer system is \( \dot{S}_{\text{gen}} \).

Upon the linearizing of the reciprocal absolute temperature difference

\[ \dot{S}''_{\text{gen,MED}} \approx q \times \frac{1}{T_{\text{eff}}} (\Delta T - \text{BPE}) \approx \frac{j}{h_{fg}} \frac{1}{T_{\text{eff}}} (\Delta T - \text{BPE}) \]  

The exergy destruction, or extra work, required to achieve a nonzero flux is

\[ \dot{X}_{\text{dest}} = \dot{S}''_{\text{gen,MED}} T_0 \]  

Combining Eqs. (26) and (24), and multiplying by the heat transfer area

\[ \dot{S}_{\text{gen,MED}} = UT_{\text{eff}}^{\text{MED}} (\Delta T - \text{BPE})^2 \]

A standard MED plant may operate with a temperature difference in the evaporators of 2–3 °C, a top brine temperature of about 70 °C, and an average overall heat transfer coefficient of 2400 W/m² K.

4.2.3 Comparison of Reverse Osmosis and Multiple Effect Distillation Driving Force and Transfer Coefficient. The excess specific exergy to achieve nonzero flux in RO and MED increases as the square of \( \Delta P - \Delta \pi \) and \( \Delta T - \text{BPE} \), respectively. However, these two terms cannot directly be compared since they do not have the same dimensions. In order to compare the entropy generation of RO and MED using terms with driving differences in similar units, we can combine Eqs. (20) and (21), and Eqs. (24) and (26)

\[ \dot{S}''_{\text{gen,RO}} = j^2 \left( \frac{1}{Ap^2 T_0} \right) \]  
\[ \dot{S}''_{\text{gen,MED}} = \frac{j}{h_{fg}} \left( \frac{h_{fg}}{UT_{\text{eff}}} \right) \]

We may think of these equations as analogous to the electrical power loss equation

\[ P = I^2 R \]

where the resistance \( R \) is equal to \( 1/G \), with \( G \) being the electrical conductance. The analogous conductance of the membrane or heat transfer material is \( Ap^2 \) or \( UT_{\text{eff}}/h_{fg} \), respectively, with units of (kg K/J) × (kg/m² s).

From Eqs. (29) and (30), we can infer that for two desalination processes operating with the same area and recovery ratio (and therefore the same flux), the process with the higher conductance will produce the least entropy and be the most efficient.

For an RO plant with permeability of 1 L/m² h, water density of 1000 kg/m³, and dead state temperature of 33 °C [4], the conductance is \( 8.5 \times 10^{-4} \) (kg K/J) × (kg/m² s). For MED, with an average \( h_{fg} \) is \( 2.39 \times 10^6 \) J/kg, a heat transfer coefficient of 2400 W/m² K, and an average effect temperature of 50 °C, the conductance is \( 4.38 \times 10^{-5} \) (kg K/J) × (kg/m² s), nearly 20 × less than for RO. Thus, from a completely different perspective, we have shown in this section that the overall heat transfer coefficient in the MED system has to be increased by a factor of around 20 × to be competitive with current RO technology, which aligns with results from Sec. 4.1. This does not take into account the effects of nonequipartitioning in the RO system or any parasitic losses.

4.3 Discussion. In all methods that we compared RO and MED from a thermodynamic perspective, RO outperformed MED. Economics play an important role in this difference. Desalination plants are designed to minimize the levelized cost of water. Several factors that go into these cost optimizations include the cost of energy, the cost of membrane area or heat exchanger area, and the material properties of the heat exchanger or membrane, especially membrane permeability, and the conditions that set the overall heat transfer coefficient. The cost of fuel should be the same for any two plants in the same coproduction system, and we have already shown that more plant-level exergy can be extracted from a given amount of fuel when we are performing a thermal separation as opposed to a separation that requires electrical work. Thus, the difference between the two technologies may be seen as the different between the cost of polymer membranes versus the cost of metal heat exchangers and the relatively higher effective conductance of the membranes in comparison heat exchangers. The cost optimization of these factors results in MED systems that are designed to be more compact than they would have to be to operate at very high energy efficiency.

The key design needs for cost-optimal MED to perform as well as cost-optimal RO are a significant decrease in the price of heat exchangers (at fixed overall heat transfer coefficient) and/or a significant (≈25 times) increase in the overall heat transfer coefficient (without changing the price per unit area of heat exchangers).

5 Conclusions

When choosing between desalination technologies, and when optimizing the operating parameters of desalination plants, the ultimate goal is often to minimize the cost of water. To offer the lowest bid and win contracts for desalination plants, the cost of
fuel must be carefully understood, and this cost is directly tied to
the energy efficiency of the desalination plant. Additional impor-
tant design considerations include environmental impact, resil-
liency of water supply, and local regulatory requirements. In this
paper, we have shown that:

- A variety of metrics that can be used to measure the energy
efficiency of a desalination plant, each with different pur-
poses and conveying different information.
- Primary energy consumption is the metric most closely tied
to fuel consumption and ultimately to operating costs. This
metric is the most fair way to compare the energy consump-
tion of desalination technologies.
- Under any accurately defined metric in use today, RO outper-
forms MED on the basis of primary energy consumption.
- RO’s efficiency advantage results from the cost and
“conductivity” advantages of membranes over heat
exchangers.
- Significant improvements in heat exchanger costs or heat
transfer coefficients are needed to improve thermal desalina-
tion technologies such as MED.

Nomenclature

Roman Symbols

- \( A \) = membrane permeability, \( \text{L/m}^2 \text{h bar} \)
- \( A \) = area, \( \text{m}^2 \)
- \( C_p \) = specific heat, \( \text{J/kg K} \)
- \( F \) = derived to primary exergy conversion factor
- \( h_{lg} \) = latent heat of vaporization, \( \text{kJ/kg} \)
- \( h_{lg,s} \) = mean latent heat of vaporization, \( \text{kJ/kg} \)
- \( J \) = flux, \( \text{kg/m}^2 \text{s} \)
- \( L_r \) = reference specific heat
- \( m_p \) = mass flow rate of pure water, \( \text{kg/s} \)
- \( N \) = number of effects
- \( P \) = hydraulic pressure, \( \text{bar} \)
- \( q \) = heat flux, \( \text{W/m}^2 \)
- \( Q \) = heat transfer rate, \( \text{W} \)
- \( q_{sp} \) = specific heat consumption, \( \text{kJ/kg} \)
- \( S \) = entropy, \( \text{kJ/kg K} \)
- \( S_{gen} \) = entropy generation rate, \( \text{W/K} \)
- \( S_{gen}^0 \) = entropy generation per unit area, \( \text{W/m}^2 \text{K} \)
- \( T \) = temperature, \( \text{K} \) or \( \text{°C} \)
- \( U \) = average heat transfer coefficient, \( \text{kW/m}^2 \text{K} \)
- \( W \) = electrical power, \( \text{W} \)

Greek Symbols

- \( \Delta \) = difference
- \( \eta \) = efficiency
- \( \pi \) = osmotic pressure, \( \text{bar} \)
- \( \rho \) = density, \( \text{kg/m}^3 \)
- \( \Xi \) = exergy flow rate, \( \text{W} \)

Subscripts and Superscripts

- \( \text{avg} \) = average
- \( b \) = baseline
- \( \text{desal} \) = energy being used for desalination, as opposed to being
  sent to the grid
- \( \text{dest} \) = destroyed
- \( c \) = with extraction
- \( \text{eq} \) = equivalent
- \( \text{eff} \) = in the effect
- \( \text{df} \) = driving force
- \( \text{fh} \) = feed heater
- \( \text{grid} \) = related to power sent to the grid
- \( \text{H} \) = high temperature
- \( i \) = index of effect

Appendix A: Conversion Factors

Without performing numerous complex simulations to deter-
mine how primary exergy is distributed between electrical exergy,
LP steam exergy, and medium pressure (MP) steam exergy, we
can estimate conversion factors between primary exergy and each
grade of derived exergy using existing data \[4\]. The electrical
exergy requirement for a new MED plant is assumed constant, as
the only changes being made are in the design of the effects.
Therefore, 2.56 kWh/m\(^3\) of primary exergy is required to satisfy
electrical needs, the same as in Ref. \[4\].

Dealing with thermal exergy becomes slightly more compli-
cated. We can use the data of Altmann et al. to infer conversion
factors from primary exergy entering the power plant into MP
steam and LP steam. The membrane distillation (MD) system uses
the same grade of LP steam as the MED system but no MP steam.
The approximate conversion factors can be found by substituting
data from from Table 2 into Eqs. (A1)–(A3):

\[
\Xi_{\text{primary,th}} = \Xi_{\text{MP}} \times F_{\text{MP}} + \Xi_{\text{LP}} \times F_{\text{LP}} \quad (A1)
\]

\[
\text{MED: } 6.48 = 1.3 \times F_{\text{MP}} + 8.2 \times F_{\text{LP}} \quad (A2)
\]

\[
\text{MD: } 5.55 = 0 \times F_{\text{MP}} + 8.5 \times F_{\text{LP}} \quad (A3)
\]

Here, \( \Xi_{\text{primary,th}} \) is the primary exergy consumed to make the
derived steam exergy (in kWh/m\(^3\)). \( F \) is the conversion factor for
the noted grade of steam, and \( \Xi_{\text{MP}} \) and \( \Xi_{\text{LP}} \) are the derived exergy
consumption of the noted grade of steam. Solving, we find \( F_{\text{MP}} = 0.92 \) and \( F_{\text{LP}} = 0.65 \). We use these conversion factors to deter-
mine how much LP exergy can be consumed by the new MED
plant.
Appendix B: Multiple Effect Distillation Performance Equations

A simplified model of MED performance has been used in this paper based on data reported by Mistry et al. [32]. We adopt an expression for PR proposed by El-Sayed and Silver [2]

$$ PR = \frac{h_{fg} \cdot S}{N \cdot \frac{C_p}{RR} (TTD_{fh} + BPE_{avg}) + \frac{N - 1}{2N} c_p AT_e} $$

(B1)

In the above equation, \(h_{fg}\) is the enthalpy of vaporization of steam, \(h_e\) is the mean latent heat, \(N\) is the number of effects, \(c_p\) is the specific heat, \(RR\) is the recovery ratio, \(TTD_{fh}\) is the terminal temperature difference in the feed heaters, \(BPE_{avg}\) is the average boiling point elevation in all effects, and \(\Delta T_{ef}\) is the temperature difference between two effects.

To understand the system’s sensitivity to design parameters such as the number of effects, \(N\), and overall heat transfer coefficient, \(U\), we must relate these design variables to the parameters occurring in the above expression, such as the temperature difference between effects, \(\Delta T_e\). In the baseline system, the terminal temperature difference in the feed heaters is set as \(TTD_{fh} = 5^\circ C\), following Mistry et al. [32].

Using a seawater property package [33,34], the boiling point elevation, specific heat, and enthalpy of evaporation are evaluated at various combinations of temperature and pressure, corresponding to various effects in the MED system. These are then averaged to obtain \(BPE_{avg}\), \(c_p\), and \(h_{fg}\).

The overall heat transfer coefficient \(U\) in the effects and in the feed heaters is obtained by taking an average of the heat transfer coefficients over the range of operating temperatures. Correlations for \(U(T)\) from El-Dessouky and Ettooney [22] are used to obtain these average values.

The driving force for heat transfer at any given effect can be written as \(\Delta T_{ef} = T_i^{-1} - BPE^{-1} - T_i^{-1}\). Since water boiled in effect \(i\) condenses in effect \(i + 1\), we assume that the heat transfer rate, \(\dot{Q}_e\), is constant across the effects. Due to changes in \(h_{fg}\) as a function of effect temperature, the amount of water evaporated in each effect is different \(\dot{m}_i = \frac{\dot{Q}_e}{h_{fg}}\). At each effect we can write

$$ \dot{Q}_e = \frac{UA}{(T_i^{-1} - BPE^{-1} - T_i)} $$

(B2)

If we assume \(\langle UA \rangle\) is constant, the driving temperature difference is also constant across all the effects: \(\Delta T_{ef} = \Delta_T\). Combining the equations for all \(N\) effects, we can estimate \(\Delta_T = \frac{(T_{steam} - T_0)}{(N + 1) + BPE_{avg}/(N + 1)}\). The temperature difference between effects therefore is \(\Delta_T = \Delta T_{ef} + BPE_{avg}\).

The specific area required for heat transfer surfaces is estimated as:

$$ A = \frac{h_{fg}}{U_e \Delta T_{ef}} + \left(1 - \frac{\Delta T_{ef} - TTD_{fh}}{TTD_{fh}}\right) \frac{c_p(\Delta T_{ef} - T_0)}{RR \cdot TTD_{fh}} $$

(B3)

and flux can be obtained as \(J_{MED} = \frac{3600}{A/m_r}\).

Figure 6 shows a comparison of the results of Eqs. (B1) and (B3) with the results of a more detailed componentwise model of the MED process described in Mistry and Lienhard [32]. Data for up to 19 effects are presented in the literature, and the deviation of PR and specific area are less than 0.8% and 6%, respectively. These equations are used in the present study to understand the impact of overall heat transfer coefficients and \(N\) on MED performance measured in terms of PR and flux, and they are used to identify the level of improvement in \(U\) required for MED to outperform RO energetically.

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