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Thermal desalination in GCC and possible development

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ABSTRACT

The Water Desalination and Reuse Center in King Abdulla University of Science and Technology, in Saudi Arabia, held a workshop on thermal desalination on the 11th and 12th of March, 2013. This paper was presented as part of a lecture at the workshop. It presents the status and possible developments of the two main thermal desalination systems processing large quantities of seawater in the Gulf Cooperation Council, multi-stage flash, and thermal vapor compression systems. Developments of these systems were presented to show how these systems are competing with the more energy-efficient seawater reverse osmosis desalting.

Keywords: Multi-stage flash desalting systems; Desalination; Multi-effect; Thermal vapor compression; Falling film evaporators; Top brine temperature; Pumping energy; Gain ratio; Performance ratio; Equivalent work; Feed salinity; Brine salinity

1. Introduction

The main thermal desalting systems used in Gulf Cooperation Council (GCC) countries are multi-stage flash (MSF), and thermal vapor compression (TVC), combined with multi-effect (ME), or Multi-effect thermal vapors compression (ME-TVC) desalting systems. The GCC includes Saudi Arabia (SA), United Arab Emirates (UAE), Kuwait, Oman, Bahrain, and Qatar. The MSF system continues to be built only in the GCC, with no MSF units having been recently built outside the GCC. The numbers of desalting plants and their daily capacity in millions (M) m^3/d in GCC are given below [1]:

Country	SA	UAE	Kuwait	Qatar	Oman	Bahrain	Total
No. of desalting plants	128	98	24	13	19	12	294
Capacity (Mm^3/d)	12.5	9.5	1.7	1.9	1.6	1.4	28.6

Desalinating water trends in the rest of the world have been shifted towards seawater reverse osmosis (SWRO), which has been the standard outside the GCC for decades. This trend has recently been adopted in a number of places in GCC. These include a SWRO plant at the Palm Jumeirah plant in Dubai, with 32,000 m^3/d, 38% recovery ratio, and seawater with TDS levels at 45,000 mg/L; and the Addur SWRO plant in Bahrain, with a 218,208 m^3/d (48 MIGD) capacity, as well as 1,234 MW of power production capacity.

When it was commissioned in the early 1990s, Addur was one of the largest SWRO plants in the world. It was built to show that SWRO was a credible desalination technology in the GCC and to help overturn old orthodoxies that promoted using only distillation systems. Addur, now the second largest SWRO plant (after Barka II in Oman), is equipped with robust feed seawater pretreatment that enables reliable, flexible delivery of water at a reasonable price.

Examples of this technology from outside the GCC include an SWRO plant planned in Sydney, featuring four modules capable of processing 125,000 m^3/d each (with a total capacity of 500,000 m^3/d); and the Ténès SWRO plant in Algeria that has a 200,000 m^3/d capacity relies on a pressure exchange energy recovery system [2].

As it stands, in the Gulf, MSF and ME-TVC represent 68% while SWRO represents only 32% of the total capacity [1].

An in-depth comparison among the distillation desalting systems such as MSF, ME distillation, mechanical vapor compression (MVC), and ME-TVC is needed to show the future trend in these systems, in light of the rising competition from SWRO.

In the Gulf, the MSF desalting system has been the predominant method since 1960. It is the most reliable, mature desalting process, representing more than 50 years of experience in design, operation, material selection, and maintenance. It has the largest unit capacity among all desalting systems, exemplified by the 20 MIGD plant recently installed in Ras al Khair in Saudi Arabia (Fig. 1(a)) [3].

Recently, large numbers of high-capacity ME-TVC desalting units have been built in the GCC. The ME-TVC units operate at lower top brine temperature (TBT < 70˚C). This limits the risk of scale formation and corrosion, and encourages the use of cheaper material such as aluminum. The difference between TBT and the last effect temperature (T_n) decreased compared to that of MSF, and thus the ME-TVC requires heat transfer surface area which can be increased compared to that of the MSF system. However, the high heat transfer coefficients by film boiling on the tubes of ME-TVC evaporators is much higher than that of the condenser in the MSF units, and this keeps the amount of the used heat transfer areas in both systems very close. Further merits of the ME-TVC system over MSF include better response to steam supply variation, less footprint area, less pumping energy, and an ability to operate at different modes than the design. The MSF and ME-TVC methods are the two main thermally operated desalting techniques producing large quantities of desalinated water in the GCC.

2. The MSF system

The MSF system is the most widely used desalination method in the GCC. It is the only method used in a country like Kuwait, for example, where all its MSF units have a gain ratio (GR) = 8, (kg distillate product/kg of supplied steam) when operated at TBT = 90.5˚C, and a GR of about 8.6 at TBT = 110˚C.

Fig. 1(b) shows a schematic diagram of an MSF unit consisting of n stages, including three heat rejection stages (HJS), shown in the right side of the figure, and ($n = 3$) heat recovery stages (HRS), as well as a brine heater (BH) at the left side of the figure. Seawater (M_c) at temperature t_c enters the HJS, and leaves at temperature T_n. In the MSF and ME-TVC terminology, the temperatures are expressed by lower case for the

Fig. 1(a). One of 8 MSF units at Ras Al Khair plant, SA. of $US1.76 billion total cost billion, capacity/unit = 91,000 t/d (20 MIGD) or $11 M/MIGD, 123 m (l) × 33.7 m (w), and 4,150 t [3].

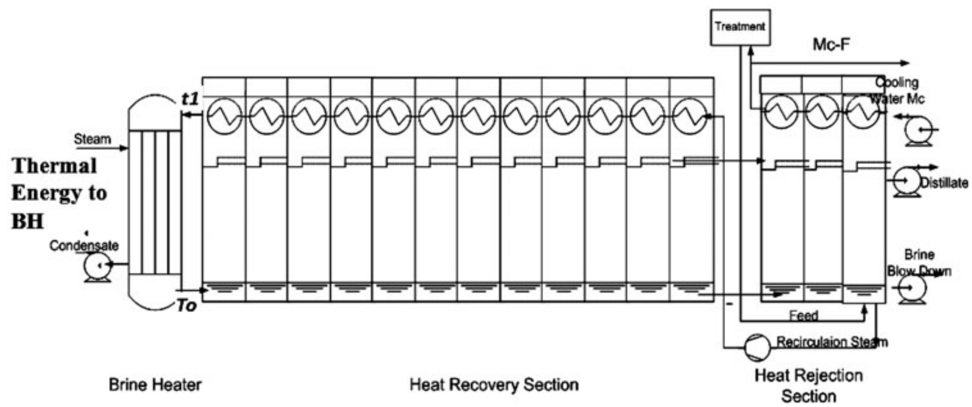

Fig. 1(b). Schematic diagram of an MSF unit consisting of n stages.

heated fluid, while the higher case for the heating fluid. Part of the exiting cooling seawater is taken as a feed (F), where it is pretreated before it flows to the last stage (n), and the balance ($M_c - F$) is rejected back to the sea. Recirculation brine (R) at temperature T_n is pumped from the last stage to the last stage of the HRS, and is heated in the condensers located at the top of these stages as it moves and is heated successively from the stage (n-3) of the HRS to first stage ($n = 1$). Recirculation brine leaves the first stage at temperature t_1, and enters the BH where it is heated to the TBT equal to T_o. It then enters the bottom (flashing chamber) of the first stage as the flashing vapor at T_o is higher than the saturation temperature of the first stage (T_1). Upon entering the bottom of the first stage, part of the flashing brine is flashed to stabilize the flow, which drops its temperature from T_o to T_1. The flashing vapor is moved upward to the condenser, which is located in the upper volume of the stage. The flashing brine moves from the first stage ($n = 1$) to the last stage (n), while it is partially flashed by a spontaneously temperature decrease of ΔT in each stage, until it ends in the last stage at T_n, and at a flow rate $R - D$, where D is the total vapor flashed from R.

The brine $B = R - D$ in the last stage is mixed with F entering this stage. The mixed stream ($R - D + F$) leaves this stage as R to recirculate in the recovery stages and ($F - D$) as blow-down brine rejected back to the sea.

The heat loss by the mean flashing brine stream is:

$$[R + (R - D)]/2 \times C(T_o - T_n)$$

This heat is used to evaporate D, or

$$DL = [R + (R - D)]/2 \times C(T_o - T_n)$$

This gives,

$$\frac{R}{D} = 0.5 + \frac{L}{C(T_o - T_n)} \tag{1}$$

$$1 = \frac{D}{2R} + \frac{D}{R}\frac{L}{C(T_o - T_n)}$$

$$\frac{D}{R} = \left(1 - \frac{D}{2R}\right)\frac{C(T_o - T_n)}{L}$$

The heat supplied to the BH (Q) is used to heat the recirculation (R) from t_1 to T_o, or

$$\frac{Q}{D} = \frac{R(T_o - t_1)/\eta_s}{D} \tag{2}$$

η_s is the heat efficiency of the BH $= RC(T_o - t_1)/DL_s$.

The gain ratio (GR) or distillate output per kilogram of steam consumption,

$$\frac{D}{S} = \frac{D}{Q/L_s} = \frac{L_s D}{RC(T_o - t_1)/\eta_s}$$

$$\frac{D}{S} = \frac{D}{Q_s/L_s}$$

where Q_s is the heat supplied by steam, and L_s is the steam latent heat,

Then

$$\frac{D}{S} = \eta_s \frac{L_s}{L}\left(1 - \frac{D}{2R}\right)\frac{(T_o - T_n)}{(T_o - t_1)}$$

$$\frac{D}{S} = \eta_s \frac{L_s}{L}\left(1 - \frac{D}{2R}\right)\frac{n}{(\delta t_l/\Delta T_r) + \gamma + [\eta_r/(e^{\alpha} - 1)]} \tag{3}$$

where

$$\alpha = \frac{U_r A_r}{RC},\ \Delta T = (T_o - T_n)/n, \quad \text{and} \quad \delta t = (T_1 - t_1) \tag{4}$$

The maximum unit capacity increased from 1 MIGD in 1960 to 15 MIGD in 2002, and again to 20 MIGD in 2012, [3]. The evaporator width increased from 23 m in 2002 to 33.5 m in 2012.

The MSF is a mature process, with very limited room for improvement. The main problem with MSF is its energy inefficiency and thus excessive energy consumption. The consumed thermal energy is in the range of 250–330 MJ/m^3 per unit distillate (D). The main pumping energy is that consumed by the R stream as it flows from the last stage to exit from the BH. It depends on the flow rate and pressure drop across the flow. The recirculation to the distillate rates ratio (R/D) is high and is a direct function of (T_o−T_n) as shown in Eq. (1). So, the increase of T_o decreases R/D and thus the specific pumping energy as shown in Fig. 2. For the same MSF unit and fixed R, the distillate D would increase by increasing T_o. The increase of T_o means improvement in the distillate output, and decrease of the specific pumping energy (per unit distillate). The selection of TBT is limited by the temperature to which the brine can be heated before serious scaling occurs and depends mainly on the feed (F) pretreatment method.

Eq. (3) and Fig. 3 show that the gain ratio can be increased (specific consumed energy decreases) by increasing both the stage number (n) and heat transfer unit ($\alpha = UA/RC$) which depends on the overall heat transfer coefficient (U), and heat transfer area (A). However, the increase of n decreases ΔT, and thus the pressure difference across the stages (ΔP). The value of ΔP decreases as the temperature decreases for the same ΔT (Fig. 4). Enough ΔP is needed to move the flashing brine flow from one stage to another. Moreover, boiling point elevation also restricts the maximum number of stages, and ΔP can significantly decrease in the last cold stages. Also, the minimum inter-stage temperature drop must be greater than the boiling point elevation for flashing to occur.

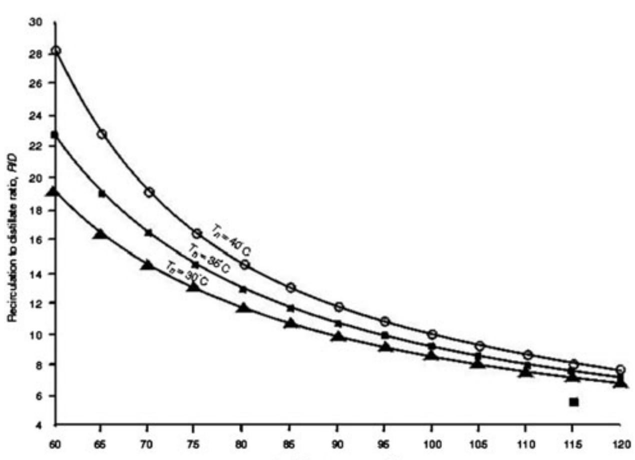

Fig. 2. The effect of the TBT on R/D ratio in a multi-stage desalting system having GR = 8 and at different blowdown temperature, T_n.

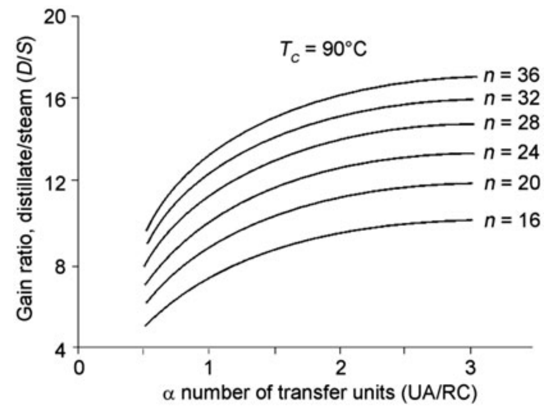

Fig. 3. The effect of the number of transfer units UA/RC on the gain ratio D/S for different number of stages, n, at T_o = 90 ℃.

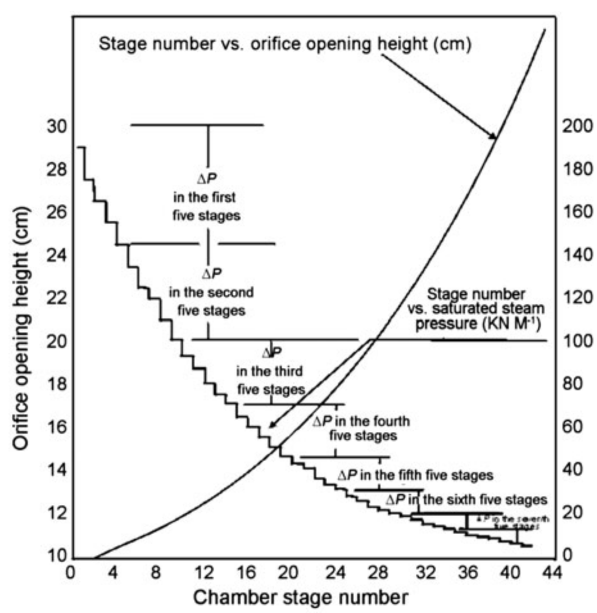

Fig. 4. Pressure drops across the stages of the opening heights of the orifice.

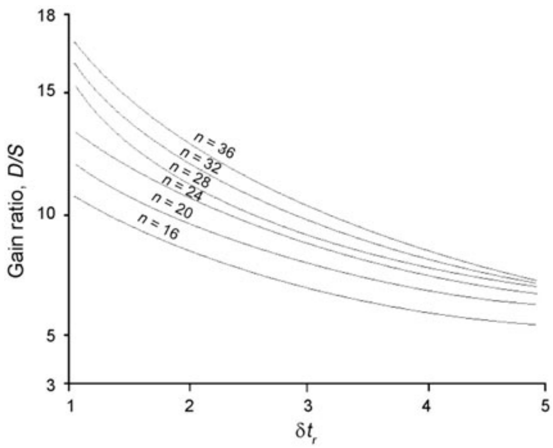

Fig. 5. The effect of the average ($T_i - t_i$), Δt_r (i.e. inversely proportional to the heat transfer area) on the gain ratio D/S for different number of stages n.

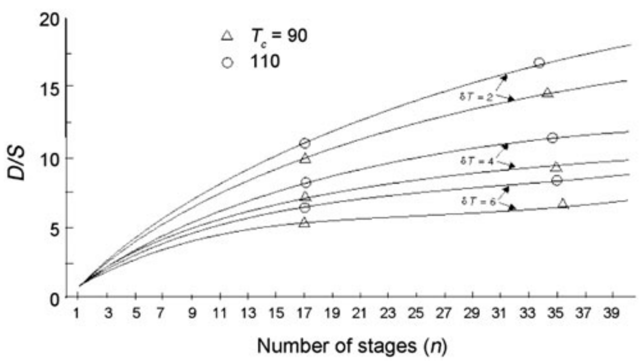

Fig. 6. The effect of the TBT on the gain ratio D/S for different stages and average temperature approach $T_i - t_i$. $\Delta T = T_i - t_i = 2$, $T_i - t_i = 4$, and $T_i - t_i = 6$.

Fig. 5 shows the effect of the temperature approach (T_i-t_i) on the GR (=D/S) for several numbers of stages. As the heat transfer area increases, the value of the temperature approach (T_i-t_i) is decreased. Similarly, Fig. 6 shows the effect of a number of stages on GR for different temperature approaches.

2.1. Energy consumed by MSF units

Extensive pumping energy is used by an MSF unit to move its streams, e.g. steam condensate from the BH to its steam source, cooling seawater from the sea, blow-down from the last stage back to sea, distillate, and recirculation (R), as well as other small pumps for chemical dosing. The stream R consumes about 66% of the total pumping energy. Typical specific pumping energy is 4 kWh/m^3 for 90℃ TBT and 3.5 kWh/m^3 for TBT = 110℃.

The condenser tubes in the stages can be arranged in cross-tube type (Fig. 7) or long tube type (Fig. 8). Most large MSF units are arranged in cross-tube type with many flow directional changes, which results in high-pressure drop and high pumping energy. It is unfortunate that this situation cannot be changed for easier tube detection, blocking and changing, relative to the long tubes.

2.2. Methods of thermal energy supply to MSF (or ME-TVC) systems

All large, thermally operated desalination plants are combined with power plants to secure their thermal energy needs in the form of relatively low temperature and pressure steam (2–3 bar). It is

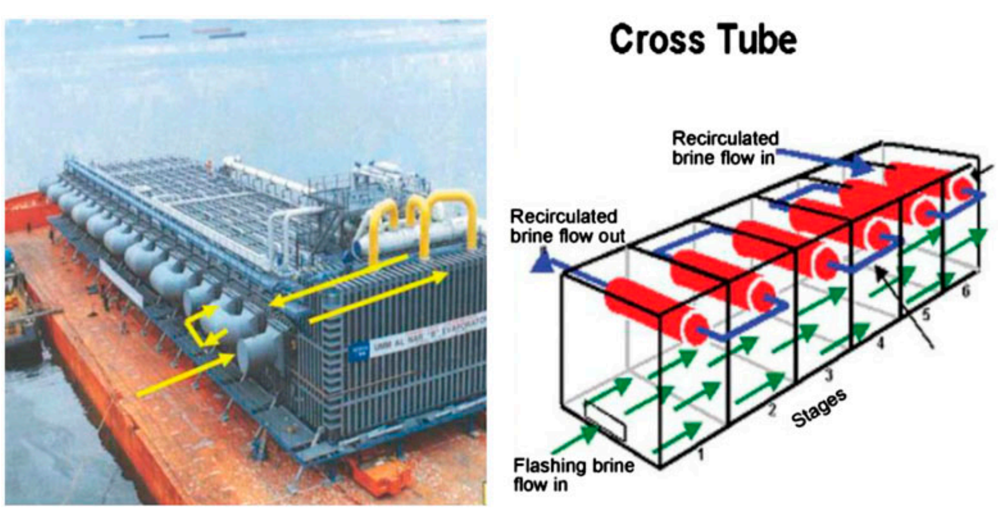

Fig. 7. Cross-tube MSF plant, [4].

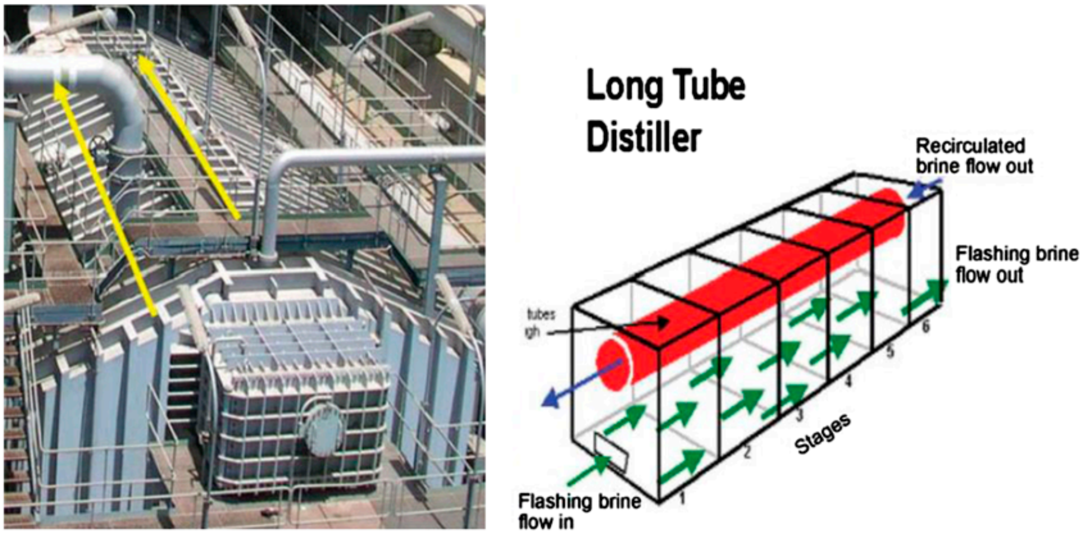

Fig. 8. Long-tube MSF plant, [4].

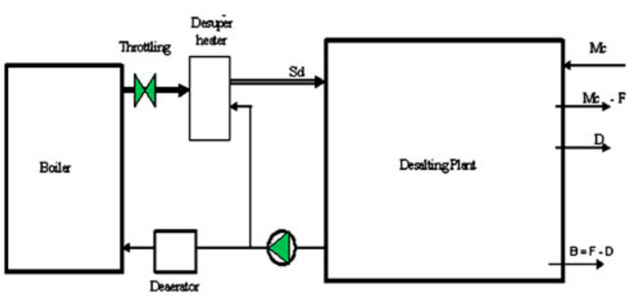

Fig. 9. Direct boiler operated desalting system (MSF or ME-TVC).

wasteful to generate steam at this low pressure (LP) and supply it directly to the desalting plant, as shown in Fig. 9. So steam is generated at high pressure and expanded in steam turbine to produce electric power (EP), before its extraction to the desalting unit at the required pressure. In the simple steam cycle, steam can be extracted through the extraction condensing steam turbine (ECST) to supply the desalination plant with its needed pressure (Fig. 10); or from a back-pressure steam turbine (BPST) discharging all its steam to the desalination plant (Fig. 11). These plants produce both EP and desalted seawater (DW) and are

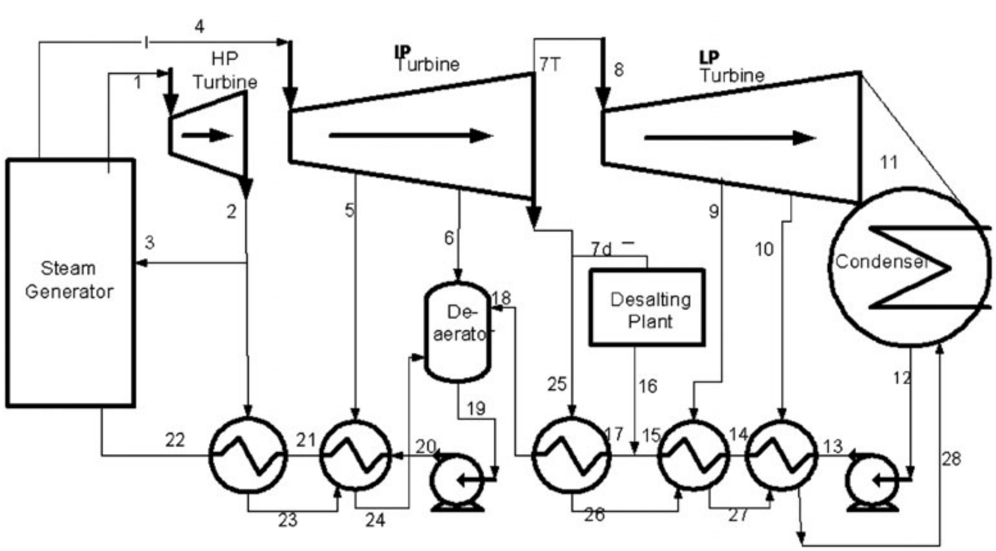

Fig. 10. CPDP using condensing extraction turbine (ECST).

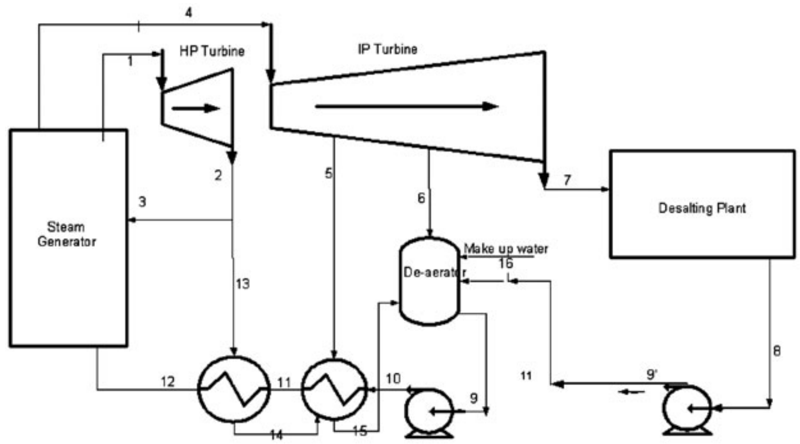

Fig. 11. Cogeneration power desalting plant using BPST.

thus called cogeneration power desalting plants (CPDP). Also, power plants using gas turbine (GT) can be equipped with heat recovery steam generators (HRSG). The HRSG utilizes the exhaust gases leaving the GT to produce steam, which can be supplied to the desalting plant (Fig. 12). In combined GT cycle, the steam leaving the HRSG is directed to a steam turbine to produce more EP without fuel addition, as shown in Fig. 13.

Fig. 14 shows the specific fuel energy consumption per m^3 of DW for boiler operated MSF systems and for CPDPs [5].

2.3. *Equivalent mechanical energy of thermal energy consumed by MSF (or ME-TVC)*

Most newly installed CPDP in the GCC are using GT combined cycle (CC) integrated with MSF or ME-TVC desalting plants. Examples are the Shuaiba CPDP in Kuwait (Fig. 15), which is using MSF units, and Ras Laffan in Qatar, which is using ME-TVC units.

The plant in Shuaiba Kuwait has three GTs of 3×215.5 MW capacity with three HRSGs and one BPST discharging its steam to three MSF units of 15 MIGD each. The hot gases exhausted from the GT are at around 600°C. The gases from each GT are supplied to single pressure HRSGs to generate steam. The steam generated from the three HRSGs is supplied to one BPST in the bottoming cycle. This BPST has a power output capacity of 215.7 MW, and all its discharged steam supplies the three MSF units at 2.8 bar. The CC plant net output capacity is ($3\text{GT} \times 215.5 + 1$ $\text{BPST} \times 215.5) = 819.7$ MW of EP, and 45 MIGD of DW. The thermal energy consumed by the desalting units is expressed in terms of equivalent mechanical energy as given in the next section and using the data given in Fig. 14.

Steam leaves the turbine to the desalting plant at a flow rate of $S = 293.6$ kg/s and at 2.8 bar, 158°C, and 2781.5 kJ/kg specific enthalpy (h). It is then de-superheated before its entrance into three MSF units, at 2.5 bar, 135°C and $h = 2{,}733$ kJ/kg. Steam enters the

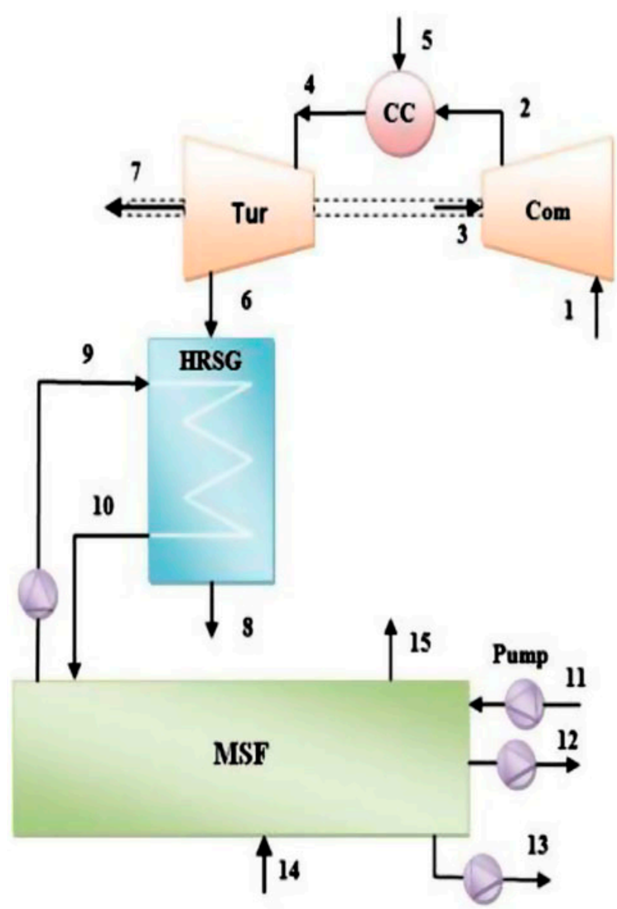

Fig. 12. GT cogeneration power desalination plant.

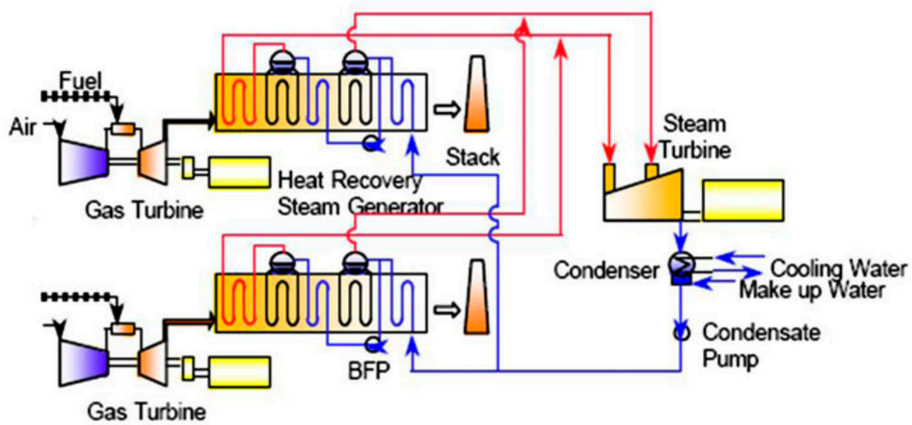

Fig. 13. CC power plant.

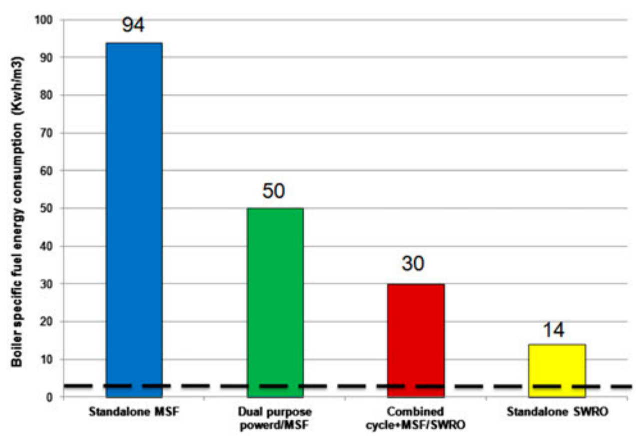

Fig. 14. Comparison of specific fuel energy consumption [5].

three MSF units at $S = 293.6\,\text{kg/s}$ to produce DW equal to 2,368 kg/s (45 MIGD from the 3 MSF units) and at gain ratio $GR = D/S = 8.06$. The heat gain by the desalting units is $Q_d = 657\,\text{MW}$.

If this steam was expanded in LP turbines to normal power plant condensers, as shown in Fig. 16, it would produce work W_{de} which is called the equivalent work (or work loss due to its supply to the desalting units rather than expanding to the condenser).

This work is equivalent to the thermal energy Q_d supplied to the desalting units. The W_{de} is expressed by:

$$W_{de}(\text{lost work or equivalent}) = S(h_{MSF} - h_{cond})$$

where h_{MSF} is the enthalpy of the steam extracted from the turbine and h_{cond} is the steam at the inlet of the condenser if it was expanded in the LP turbine to the condenser, then

$$W_{de} = 293.7(2,781 - 2345.5)/1,000 = 127.8\,\text{MW}$$

This 127.8 MW is the equivalent work to the heat $Q_d = 657\,\text{MW}$ supplied to the desalting unit. This gives specific consumed heat in kJ per kg distillate as 277.5 kJ/kg, or 14.6 MW/MIGD. The work equivalent to the consumed heat can be expressed by specific equivalent work = 54 kJ/kg (15 kWh/m^3) or 2.84 MW/MIGD. Additional pumping work equals 4 kWh/m^3. Thus, the total consumed equivalent work for the MSF units is 19 kWh/m^3.

If the ME-TVC is assumed to have the same consumed heat and pumping energy equal to 2 kWh/m^3, the total consumed W equivalent for the ME-TVC is 17 kWh/m^3. The numbers for the MSF were compared with those of Wangnick [6] who reported 4 kWh/m^3 for pumping and 14 kWh/m^3 for thermal, totaling 18 kWh/m^3. Hamed [7] reported inherited energy loss in the range of 15.2–23.7 kWh/m^3.

Similar practical analysis was carried out by Marzook on the Shuaibah CPDP in Saudi Arabia [8]. The plant, shown in Fig. 17 has 391 MW of EP capacity and 64.7 MIGD (3,375 kg/s or 291,600 m^3/d). The reference plant—when the same steam generator HP and IP turbines as the MSF units are replaced by LP turbines—is shown in Fig. 18, and produces 665.4 MW of EP. This means that the extra 665.4 − 391 = 274.4 MW is the equivalent to 830 MW of thermal energy, producing 64.7 MIGD, or 4.24 MIGD/MW. This is the equivalent (eq) work to thermal energy as $W(\text{eq})/D = 81.3$ kJ/kg (22.6 kWh/m^3). When 4.18 kWh/m^3, reported as specific pumping energy, is added to 22.6 kWh/m^3

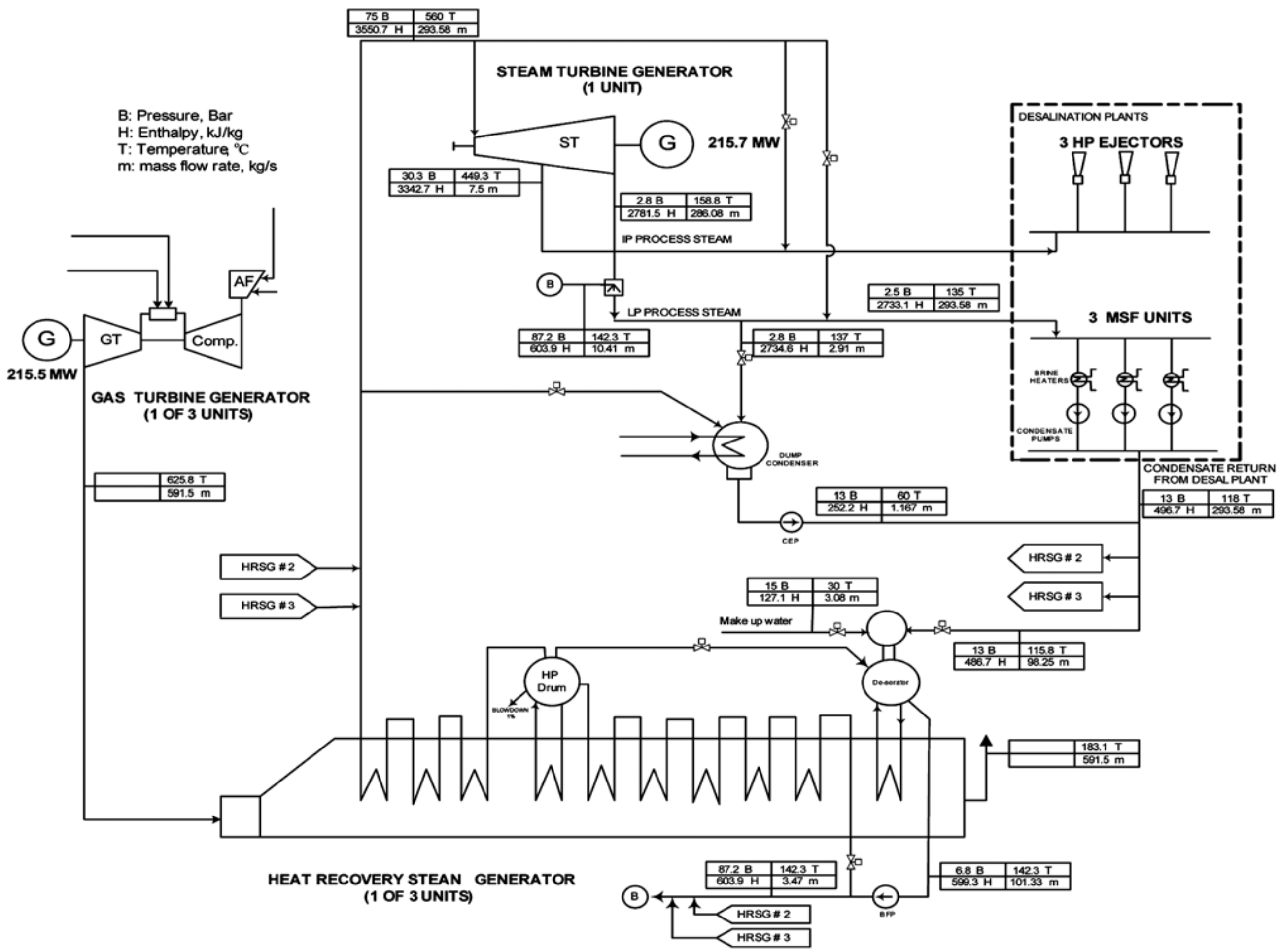

Fig. 15. CPDP using GT combined cycle and three MSF desalting plant in Shuaiba, Kuwait.

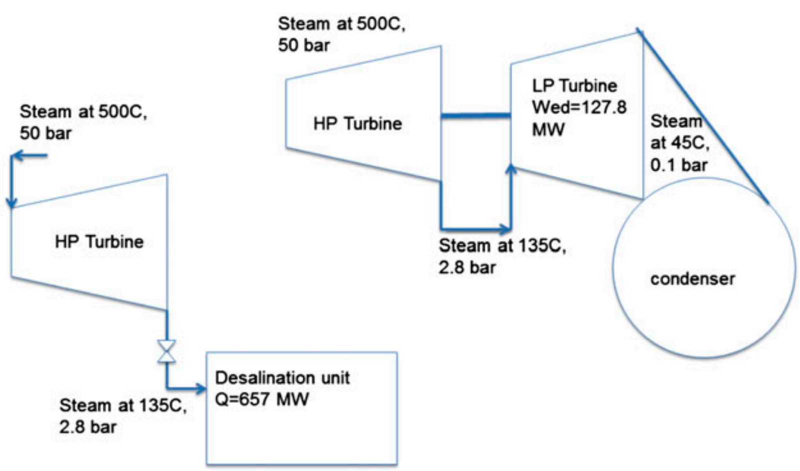

Fig. 16. Illustration of the heat added to the desalination plant is equivalent to the work obtained from LP turbine replacing the desalination plant.

specific W(eq), the total consumed energy is 26.78 kWh/m^3.

This shows that extracted steam to desalting units can produce power if expanded to the surface

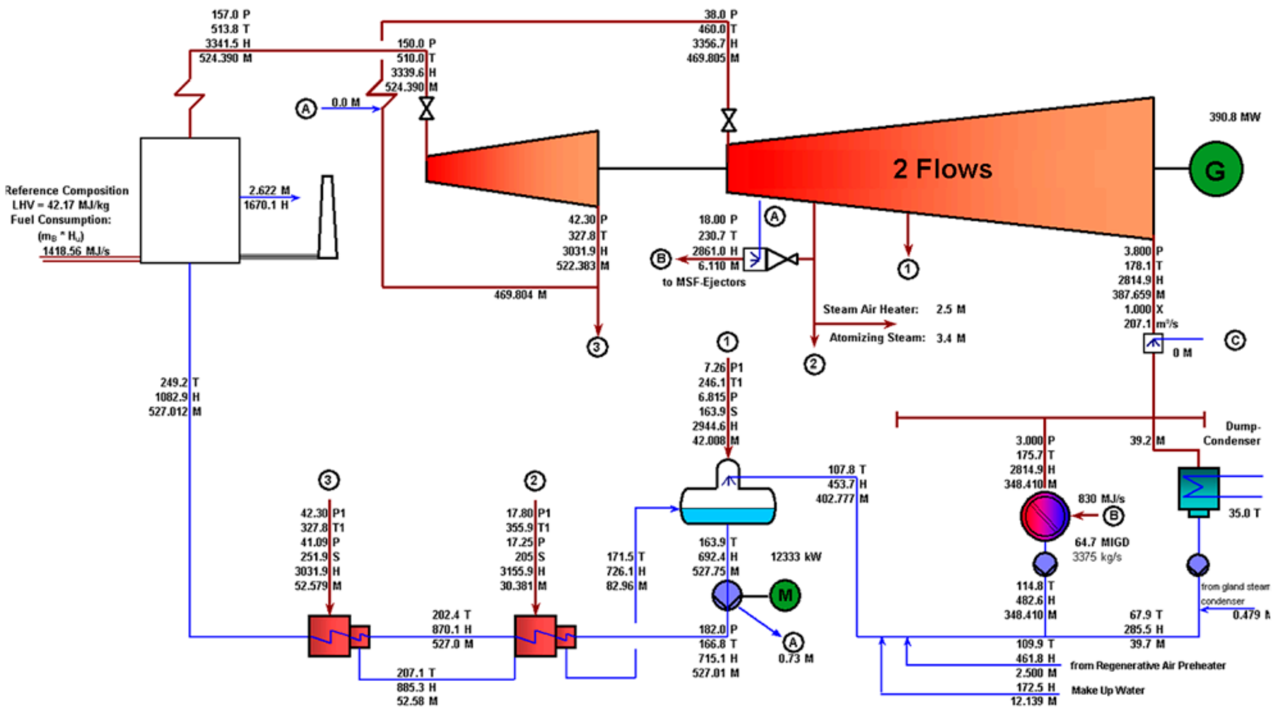

Fig. 17. Shuaibah steam CPDP of 391 MW, and 64.7 MIGD [8].

condenser, with no cheap or wasted energy as claimed. Although coupling MSF with steam turbines reduces energy by 50% compared with boiler operated MSF, it is still very high compared with SWRO.

2.4. Prospects of improving the MSF system by NanoFiltration (NF) pretreatment to raise TBT

Based on the exergy analysis, practical maximum limits on the MSF performance ratio have been determined, [5], and are shown in Fig. 19. The PR is defined by the volume of distilled water that can be obtained by supplying 2,330 kJ of heat (standard latent heat of steam). It was considered that the maximum number of stages that can be considered for TBT = 120˚C, and last stage temperature (T_n) would be 40 stages. This gives the minimum ΔT per stage = 2˚C after considering the temperature losses in the stages, including the boiling point elevation. Fig. 19 shows that the maximum PRs that can be obtained for the limiting number of stages of 40 is 13, 12.25, 11.25, and 10.5 for TBT equal to 120, 112, 100, and 90˚C, respectively. It is noticed here that 40 stages exceed the possible number of stages for a TBT of 100 and 90˚C, as it would give ΔT per stage = 1.5 and 1.25˚C, respectively, which are less than temperature losses that may occur in the stages, especially last stages. Under the maximum TBT = 120˚C set by acid feedwater pretreatment, the maximum PR that can possibly be achieved is about 13. Fig. 19 shows that the typical PR in all MSF plants in Saudi Arabia is 7.8–8.7, and all Kuwait plants have a PR number of 8 as mentioned before. The only way to increase the GR (or PR) of MSF units is to raise the TBT.

3. Raising the TBT by nanofiltration (NF) for full or partial feedwater pretreatment

A suggested system improvement for MSF is to pretreat the feed seawater (fully or partially) through nanofiltration (NF) to remove some of its scale constituents such as sulfate, calcium, carbonate, and magnesium [9–14]. This allows for an increase in its TBT, T_o, and its flashing range ($T_o – T_n$), and thus the unit output. The distillate output is directly proportional to ($T_o – T_n$). The NF is also sometimes used as a SWRO pretreatment to remove scale constituents and raise its recovery ratio. The use of NF pretreatment for both systems was suggested and extensively studied in Saudi Arabia [10–12]. Related work was presented by Awerbuch [13], who showed the benefit of using NF membranes in the removal of scale elements from seawater. It was suggested that a mixture of NF permeate and seawater as feed (partial feed pretreatment) to the MSF unit would reduce the cost of NF pretreatment. Al-Rawajfah et al. [9] reported influence

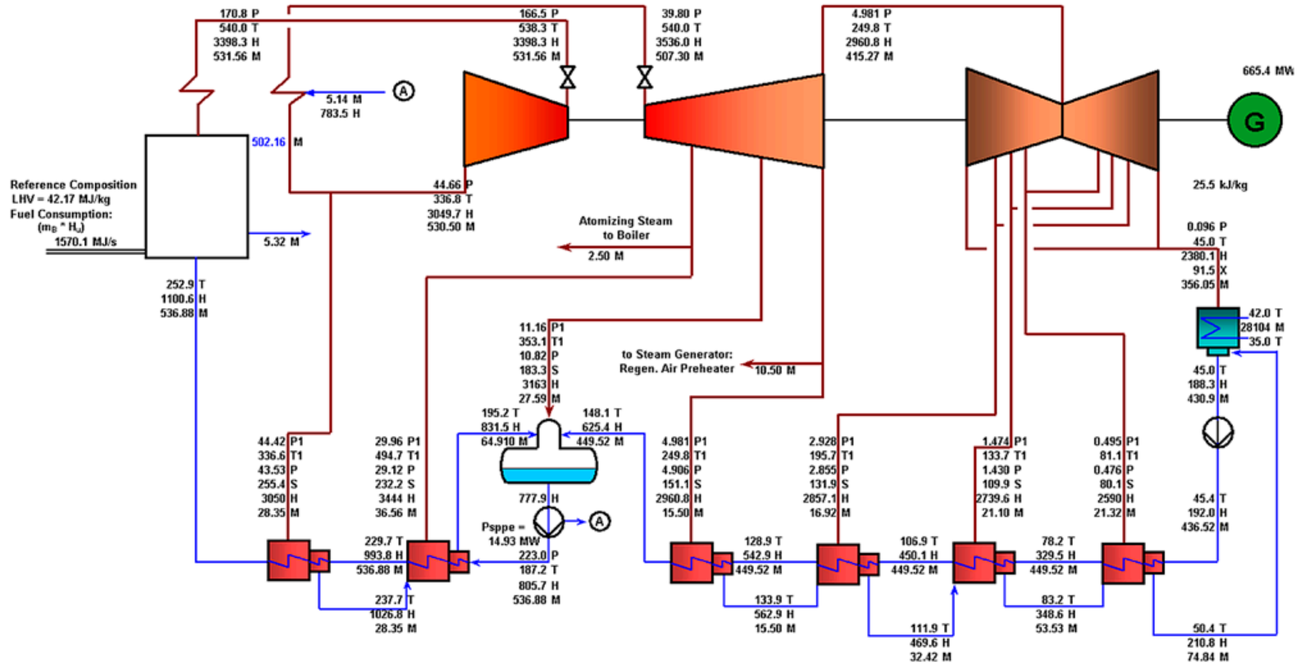

Fig. 18. Shuaibah single purpose power plant [8].

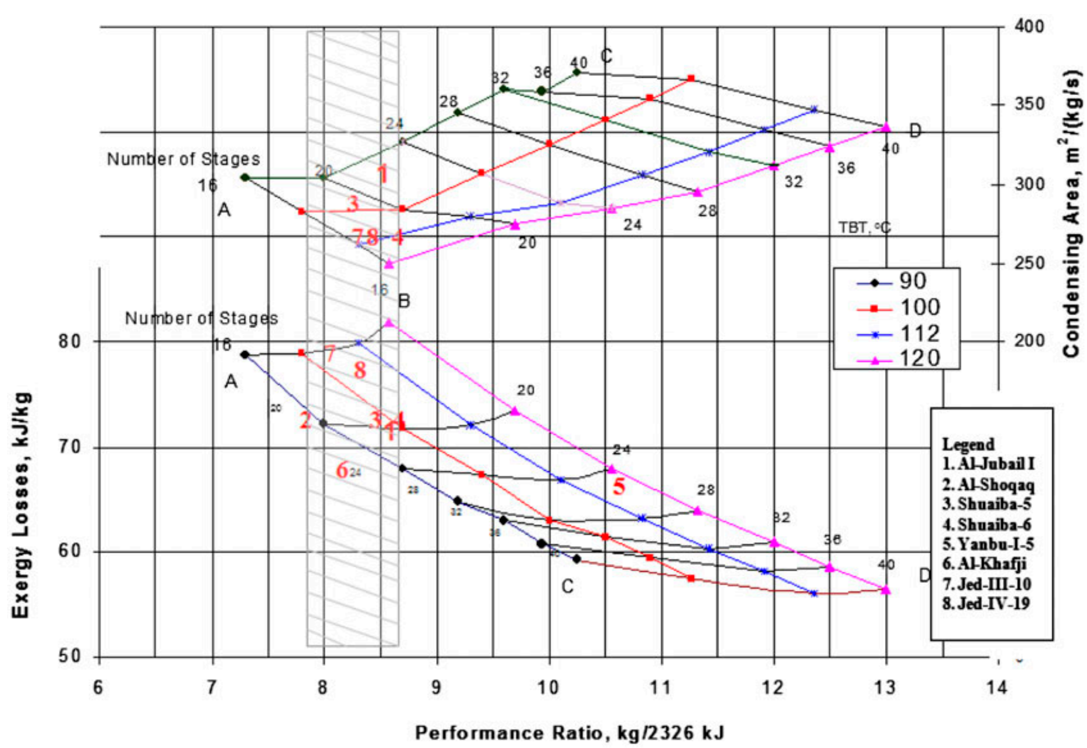

Fig. 19. Dependence of performance ratio, exergy losses and specific condensing area on TBT and number of stages (TTD = 2˚C), [5].

of NF on sulfate scale potential in recirculation MSF plants, as shown in Fig. 20. This figure shows the sulfate scale potential, expressed by Skillman index (SI), for seawater with 0, 10, 30, 50, and 100% NF-treated make-up in a recirculation MSF reference plant. The SI is a simple sulfate solubility index for estimating the likelihood of calcium sulfate scaling—it is a ratio between the actual concentrations of either calcium or sulfate theoretical or equilibrium concentration, whichever is the limiting species [14].

Fig. 20 shows that the scale potential *increases* with the increase of TBT and *decreases* with the percentage of NF-treated feed. For seawater with no NF feed pretreatment, the scale can start to deposit at 115˚C. The maximum TBT, at which sulfate scale begins to precipitate, is shifted to 120, 135, and 145˚C when the NF-treated portion of the makeup water increased from 10, 25, and 50%, respectively.

The MSF modifications to raise its TBT—and thus its capacity—by using NF, is a viable solution that has already been put into practice in Sharjah, UAE. The combination of the MSF unit with NF pretreatment is not free as the following points show:

(1) The NF feed requires the same stringent pretreatment system as the SWRO system, in addition the NF membranes system and pumps. This eliminates MSF's main advantage of having very simple pretreatment.
(2) The MSF unit also has to be modified to deal with the anticipated production increase. This includes increasing the recirculation stream flow rate and its delivery pressure, adjusting the weirs between stages, raising the saturation temperature of the steam supplied, and dealing with the increasing vapor generated in stages, especially the last stages.
(3) The increase of generated vapor in a stage without the ability to condense it completely can build higher pressure in the stage, decreasing the flashing brine flow, and resulting in capacity decreases instead of increases. Venting systems should be modified, and the heat transfer areas should be checked. The increase of vapor velocity in any stage increases the entrained brine droplets with the generated vapor and deteriorates the product quality. So, demisters should be modified to increase their areas and efficiency.

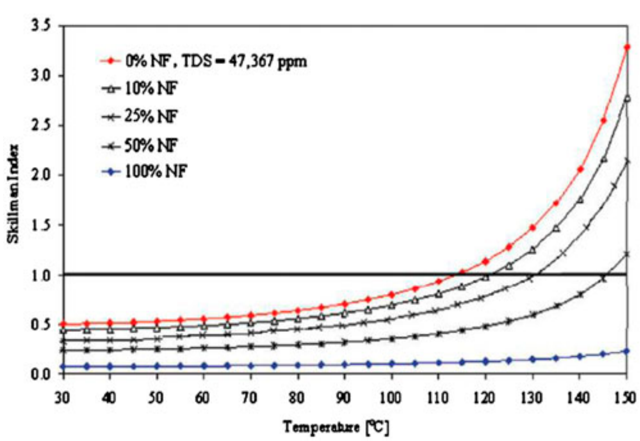

Fig. 20. Influence of NF on sulfate scale potential in BR-MSF plant [9].

4. Case study

A study was conducted in Kuwait to improve one of its operating MSF units by increasing the TBT from 110 to 135 ˚C, and thus increasing its capacity from 7.2 to 9.72 MIGD (44,187 m^3/d) or a 35% increase [2]. This can be done by pretreating 37.5% of its feed with NF. This case was compared with the case of obtaining the same additional capacity of 2.52 MIGD by simply adding SWRO train of that additional capacity of 2.52 MIGD.

The case of adding SWRO is called Case 1 here, and is shown in Fig. 21. The case of modification of the MSF unit itself to increase its capacity from 7.2 to 9.52 MIGD, adding NF to treat 37.5% of the feed to the modified MSF unit is called Case 2 and is shown in Fig. 22. These two cases are compared here in terms of energy and cost.

The increase of MSF capacity by NF (Case 2) is justified only if its running cost, in terms of energy, is less than in Case 1 and the capital costs for retrofitting the MSF unit to deal with the additional capacity, and for adding NF treatment, are less than the cost of adding SWRO train, as in Case 1.

For Case 1, the addition of a SWRO train of 2.52 MIGD would cost $11.456 M since the cost for SWRO is in the range of $1,000/($m^3$/d). The DW annual output would be 11.947 Mm^3/y by the MSF unit, and 4.18 Mm^3/y by SWRO. By taking the consumed energy of the MSF as 19 kWh/m^3 and by SWRO as 5 kWh/m^3, the annual consumed energy is 226.99 GWh/y for an MSF unit and 20.907 GWh/y for SWRO, with a total 247.9 GWh/y, and a cost of $24.79 M/y for $0.1 kWh.

For Case 2, the costs of energy, adding the NF pretreatment, and modifying the MSF unit are discussed here.

First, the energy cost will be considered. In this case, the steam supply to the MSF unit should have high saturation temperature (143˚C compared to 117˚C when NF was not used). The plant gain ratio GR (D/S) increased to 10. The availability of the heat

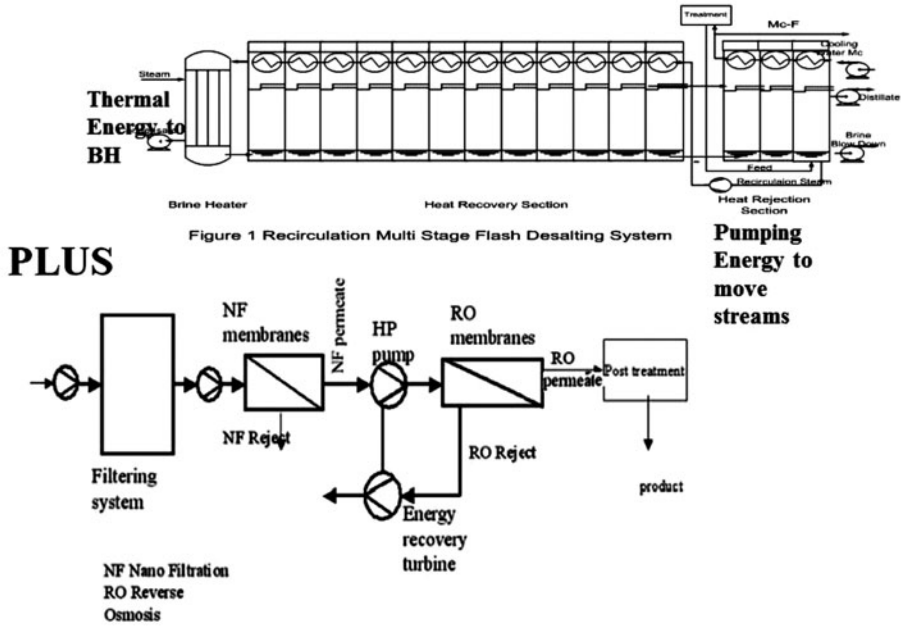

Fig. 21. Adding SWRO train to produce additional 2.52 MIGD to unmodified MSF unit of 7.2 MIGD capacity (Case 1).

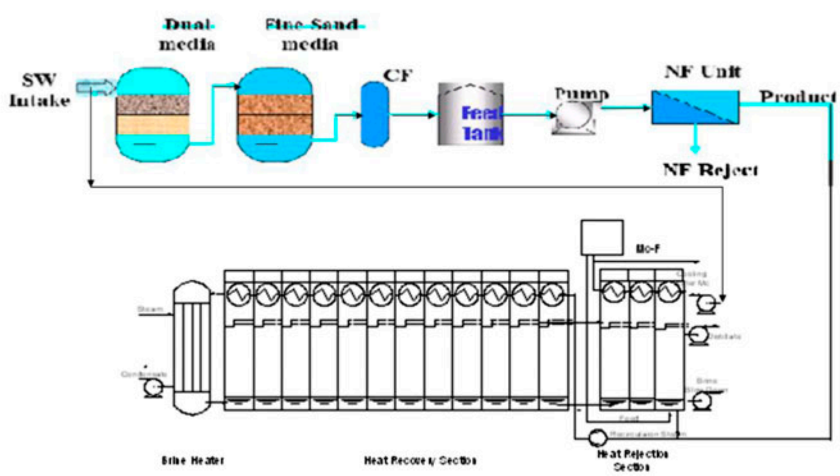

Fig. 22. Modified MSF unit to raise its TBT by adding NF unit to remove scale constituents (Case 2).

supplied at TBT = 135℃ is 20.23 kWh/m^3 of distillate which is higher than that at TBT = 110℃ (16.03 kWh/m^3). Consequently, the work loss due to steam extraction per 1 kg of expanded steam to the desalting plant would increase by about 20%. For 110˚C operation, the W_{ed}/D was 54 kJ/kg of distillate (15 kWh/m^3), and for a gain ratio of 8 the work loss per kg of extracted steam was 432 kJ/kg. For 135˚C, the work loss per kg of steam would be 518.4 kJ/kg, and the work loss per kg of distillate would be 51.82 kJ/kg (14.4 kWh/m^3) for GR = 10. This means that the W (eq)/D is decreased by only 4%. As indicated in Table 1, the recirculation ratio (R/D) decreased from 9.48 (when TBT = 110℃) to 6.99 (when TBT = 135℃). However, the recirculation stream for both cases is almost the same due to the increase of D in Case 2. Meanwhile, the feed to distillate ratio (F/D) increases from 3.037 (at TBT = 110℃) to 3.335 (at TBT = 135℃), and this increases the required feed from 1150.1 kg/s (for TBT = 110℃) to 1,706 kg/s (for TBT = 135℃), i.e., a 48% increase. Part of the feed (37.5% of it equal to 640 kg/s or 168.9 US gallons/s or 14.6 MGD) is to be treated by NF. The balance 1,066 kg/s would be exposed to the conventional MSF treatment.

More mechanical energy is consumed by pumping 37.5% of the feed through the NF system. The feed to

Table 1
Comparison between Case 1 (MSF and SWRO) and Case 2 (modified MSF with NF pretreatment)

	Case 1		Case 2	
	MSF1	SWRO	NF	MSF2
Output	378.83	132.59	640	511.425
R/D	9.48	NA	NA	6.99
R	3591.34	NA	NA	3574.86
Xr/Xb	0.8945	NA	NA	0.857
Xr (g/l)	62.616	NA	NA	59.99
F/D or 1/RR	3.037	3	1/0.7	3.3352
F (kg/s)	1150.61	397.775	914.286	1705.7
S (kg/s)	47.354	NA	NA	51.14
Weq/D (kWh/m^3)	19	5	19.27	
Mm^3/y	11.947	4.1814	16.13	
GWh/y	226.99	20.907	310.8	
Energy cost ($M/y)	22.699	2.0907	31.08	
Additional capital cost ($M)		11.456	26.4	

Note: RR: recovery ratio in NF (permeate/feed *F*) or SWRO; Xr: MSF recirculation stream salinity; *F*: feed; *S:* steam supply to the MSF unit; Xb: maximum brine salinity (70 gm/l). The cost of specific energy is considered $0.1/kWh.

the MSF unit is in the range of 3.35 times the desalted water output (*D*), then ($0.375 \times 3.35=$) 1.256 D will be pumped to the NF unit. For each cubic meter (m^3) of *D*, and reported feed pressure to the NF membrane of 20 bar, the specific pumping energy (per 1 m^3 of distilled water) consumed by NF for pumping efficiency of 0.8 is:

$$\begin{aligned} Wp(\mathrm{NF}) &= 1.256 \times 2,000/0.8 \\ &= 3,140\,\mathrm{kJ/m^3}(0.872\,\mathrm{kWh/m^3}) \end{aligned}$$

Pumping energy increases from 4 kWh/m^3 (for TBT = 110°C) to 4.872 kWh/m^3 (for TBT = 135°C) to account for the NF pumping. The specific mechanical energy for the modified case is 19.27 kWh/m^3. The total annual consumed energy for Case 2 is 310.8 GWh/y and the cost is $31.08 M/y. This is $6.29 M (or 25%) more than Case 1.

Second, the cost of the NF pretreatment addition is considered. The amount of pretreated feedwater by the NF is 640 kg/s. It was reported [15] that an 11 million US gallons per day (MGD) (or 482.5 kg/s) NF system would cost US$19.9 million and thus the NF pretreatment for Case 2 would cost US$26.4 million. It is noticed here that the NF permeate is 4.8 times that of the SWRO of Case 1, and the feed to the NF is more than doubled that of the SWRO of Case 1.

Third, the cost of retrofitting the original MSF unit to handle the additional capacity is negligible compared with the difference in energy cost with adding the NF unit. The cooling water, recirculation, and condensate steam streams and pumps would be almost the same, while the feedwater pump will involve 50% increase in costs.

Table 1 indicates clearly that adding SWRO to existing MSF units to increase desalting capacity is more economical than modifying the existing MSF by using NF to raise itself TBT.

5. The multi-effect thermal vapors compression (ME-TVC) system

The GCC has seen a recent increase in the combination of TVC with conventional (ME); or (ME-TVC), forming what is called multi-effect distillation (MED) desalting units. The system is not as well known or as developed as the MSF. Although the name of MED is misleading, it is the name used in the industry and will be used here interchangeably with ME-TVC.

Developments in MED are presently concentrated on increasing the GR and capacity per unit and improving its integration into power plants. These should lead to a reduction in the desalted water cost. Before discussing these developments, the MED system is explained with examples given for some operating units.

A schematic diagram of a typical unit, Umm Al-Nar in the UAE, is given in Fig. 23. This unit has an average capacity of 3.5 MIGD and produces the following data.

The number of effects is six, product capacity is 184.2 kg/s, gain ratio (D/S) is 8.7, supplied with

21.6 kg/s motive steam extracted from a turbine at $P = 2.8$ bar, and 130˚C. It has two parallel sections (A and B) with three effects each, operating in TVC mode. Section A (or B) has three evaporators A1, A2, and A3, (or B1, B2, and B3), and a thermal compressor (TC). The other three effects E4, E5, and E6 form conventional ME sections. Other data are given in Table 2. Vapor generated in A3 (or B3) is partially compressed (13.38 kg/s out of 22.68 kg/s) from $P_3 = 15.4$ kPa to $P_d = 25$ kPa (heating side of A1 or B1). The balance is used as heating vapor to E4, (first of three conventional ME effects) and heats feed F in feed heater 2 (FH2). Cooling water M_c is preheated to 33˚C in a heat exchanger by distillate output D at 43˚C (not shown in Fig. 23). M_c is then heated in the end condenser to 41˚C by vapor from E6. Part of M_c is used as feed F, and the balance (M–F) is rejected back to sea. Feed F6, and F5 (parts of F) enter E6 and E5 at 41˚C. The balance (F-F6-F5) is heated to 49.2˚C in FH 1 by vapor from E4, and part of it (F4) is fed to E4. The feed (F-F6-F5-F4) to A1, A2, and A3 (or B1, B2, and B3) is heated again to 52˚C in FH2 by vapor from A3 (and B3) at 54.7˚C. The feed F2 and F3 enter A2 and A3 (and B2 and B3). The balance (F1) is fed to E1 after being heated finally in the steam ejector condenser. Note that [T_{vi} (condenser)$-T_{bi}$] in an effect, given in Table 2, is the heat transfer temperature difference in that effect and has an average of 2.6˚C. The limited capacity of the TC is the reason for using this arrangement and not simple TVC. Each TC compresses 13.38 kg/s at 54.6˚C and specific volume $v_g = 9.8$ m^3/kg from P3 = 15.4 to $P_d = 25$ kPa. So, the compressed vapor by single TC has a 131 m^3/s volume flow rate and $P_r = 1.63$, [16].

The effects are usually arranged in a circular or rectangular vessel along with TC, which is shown in Figs. 24(a) and 24(b), respectively. Both vessels are connected in parallel with a third vessel in the middle, which contains a number of effects along with the end condenser.

The progress of ME-TVC in terms of increasing both the unit capacity and GR is given in Table 3, based on data given [17,18]:

The status of the presently used ME-TVC system indicates that the capacity increase can lower the specific investment cost. However, very large capacity units can present an availability problem in case they are rendered out of operation for any reason. So, it seems that the maximum TVC/ME unit capacity would remain in the range of 15 MIGD. The standard GR is little above 10, e.g. Ras Laffan C, Qatar, has a GR equal to 10.9 with motive steam pressure at 3 bar, and Az Zour North in Kuwait has a GR equal to 11.2 with steam pressure at 3 bar. The GR can be increased

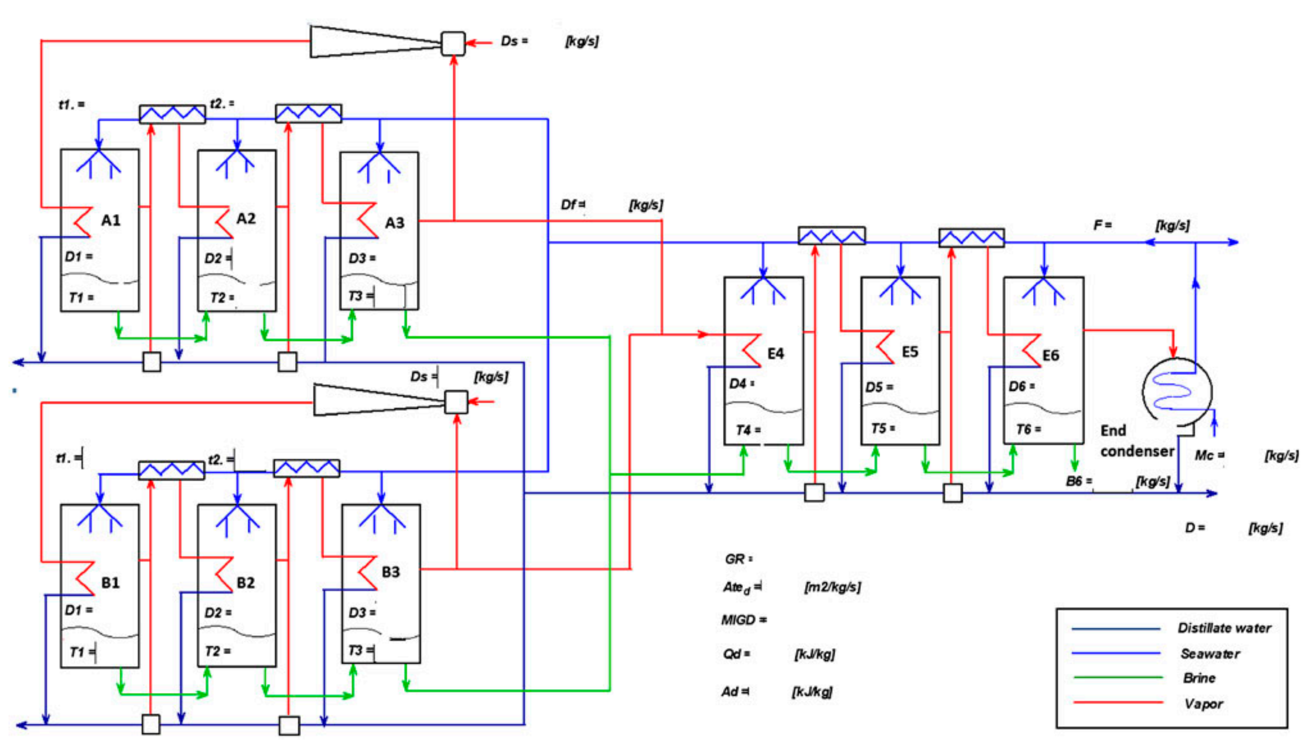

Fig. 23. Umm Al-Nar ME-TVC unit, [17].

Table 2
Operating data of Umm Al Nar ME-TVC unit

Effect No.	Feed, t (°C)	T_{bi} (°C)	T_{vi} condensate (°C)	t_f (°C)	Vapor out (kg/s)	B_i (kg/s)	U_i (kW/m^2 °C)	A_i (m^2)	BPE (°C)	P_v (kPa)	vg (m^3/kg)
Condenser	33		43	33							
6	41	44	46.2	41	17.3	555	3.23	4,453	0.8	8.65	16.8
5	41	48	49.7	41	14	380	3.28	4,453	0.8	10.2	13.9
4	40	51	54	49.2	17.3	380	3.15	4,453	0.9	12.3	11.2
3A	52	56	57.8	53.2	22.7	172	3.59	6,724	0.9	15.02	9.9
3B	52	59	57.8	53.2	22.7	172	3.59	6,724	0.9	15.02	9.9
2A	56	59	61.4	53.2	22.3	115	3.57	6,724	0.9	18	8.5
2B	56	63	61.4	53.2	22.3	115	3.57	6,724	0.9	18	8.5
1A	60	63	65	57.1	22.9	56.9	3.63	6,724	0.9	21	7
1B	60	63	65	57.1	22.9	56.9	3.63	6,724	0.9	21	7

Note: $D = 184.2$ kg/s = 3.5 MIGD; GR = 8.9; heat transfer areas: $A_{(effects)} = (6\times6{,}724) + (3\times4{,}453) = 53{,}703$ m^2, $A_{(condenser)} = 2{,}874$ m^2, $A_{(feed\ heaters\ between\ E4–E5)} = 1{,}156$ m^2, $A_{(feed\ heaters\ between\ E3–E4)} = 693$ m^2; $A_{(total)} = 58{,}426$ m^2; $M_{cw/D} = 2.88$; specific area $A/D = 310.7$ m^2/(kg/s).

Fig. 24(a). Six effects–13,333 m^3/d unit with TVC in Tobruk, Libya, in circular vessel along the TC.

to 12–13 if ΔT is decreased from 2.4 to 2.2°C with 3 bar motive steam pressure. If this pressure is increased to 6–10 bar, the GR can reach 16, [18].

The three main advantages of the ME-TVC system over the MSF are: low TBT to reduce the problems of corrosion and scale formation; small foot print, and lower pumping energy compared with MSF where recirculation stream high flow rate exists.

Concerning the energy consumption, the ME-TVC has almost the same specific heat as the MSF for the same GR, but this is hindered by the requirement of higher availability steam than the MSF. The ME-TVC system has a higher specific heat transfer area—less than 300 m^2/(kg/s) in MSF and higher for ME-TVC, e. g. 335 m^2/(kg/s) in Umm Al-Nar, and 452 m^2/(kg/s) in Al-Jubail, see Table 4.

All large capacity ME-TVC (or MSF) are supplied by steam extracted from steam turbines in power plants. The steam pressure and flow rates vary with the power demand imposed on the turbine.

At low power demand, the decreased turbine power output is accompanied by the decrease of its steam flow rate and pressure along the turbine due to steam throttling at the turbine inlet. Another back pressure turbine control is to raise the pressure at the turbine discharge while keeping the same steam flow rate. This will increase the pressure of the motive steam to the ME-TVC unit and can increase its GR.

Also, as pointed out by Baujuat [18] the performance of the ME-TVC thermo-compressors can be improved by using variable throat steam nozzle (Figs. 25(a) and 25(b)). In winter, when the power demand is low, the increased back pressure of the turbine will result in higher velocities in the nozzle and thus a higher suction ratio for the thermocompressor and higher GR for the unit. This allows MED to be fed with increased pressure and to achieve a GR that is higher by 5 to 8% in winter.

Fig. 24(b). Marafic Yanbu 2 × 6.07 MIGD ME-TVC units having rectangular vessel along the TC, shipped on 29 July 2012, from DOOSAN.

Table 3
Progress in unit capacity and GR of ME-TVC units

Plant name (TBT ˚C)	Capacity/unit (MIGD)	GR	Time
Trapani, Sicily, (62.2)	2	16	About 2006
Curacao Island	2.6	13.4	About 2006
Umm Al-Nar, (62.8)	3.5	7.9	About 1998
Layyah, Sharja	5	8.4	About 2001
Ras Laffan C, Qatar	6.3	10.9	About 2008
Al Hidd, Bahrain	6	9.07	About 2007
Marafic, Jubail, SA	6.59	9.85	About 2007
Layyah, Sharja,	8	8.4	About 2007
Fujairah ll	8.5	10.3	About 2008
Azzour North, Kuwait	10.8	11.2	About 2011

Table 4
ME-TVC desalination plants comparison

Operating and design parameters	Desalination plants					
	ALBA		UMM Al-NAR		AL-JUBAIL	
	Model	Actual	Model	Actual	Model	Actual
No. of effects	4		6		8	
Motive pressure (bar)	21	21	2.8	2.8	2.7	2.7
TBT (˚C)	63	63	63	62	63	NA
Minimum brine temperature (˚C)	48	48	44	43	42	NA
Feed sea water temperature (˚C)	43	43	40	40	40	NA
Motive steam flow rate (kg/s)	8.4 × 2	8.3 × 2	11 × 2	10.65 × 2	15.5 × 2	NA
Temperature drop per effect (˚C)	5	5	3.8	3.8	3	NA
Compression ratio	1.57	NA	1.7	NA	1.75	NA
Expansion ratio	120	NA	18.11	NA	18.7	NA
Motive to entrained vapor ratio	0.58	NA	0.885	NA	0.98	NA
Distillate production (kg/s)	123	127	184.2	184.38	340.4	342.22
Gain output ratio	7.23	7.5	8.37	8.6	10.9	9.8
Specific heat consumption (kJ/kg)	348.4	NA	292.1	287.5	223	NA
Specific heat transfer area (m^2/kg/s)	244.2	NA	335.6	310	452.2	NA
Specific exergy destruction (kJ/kg)	84.65	NA	54.24	NA	41.16	NA

6. Thermal and MVC

The question arising from the use of ME-TVC is why energy-inefficient thermal vapor compressors are used when mechanical vapor compressors (MVC) are more efficient. Thermal compressors have relatively low adiabatic efficiency, due the irreversible mixing of vapor streams, having two different pressures, thus limiting the potential of the MED to increase the gain ratio.

Due to the recent increase in energy costs, replacement of the conventional inefficient TC has been considered through using a novel, large centrifugal compressor driven directly by an auxiliary steam turbine, powered by extraction steam. The centrifugal compressor and auxiliary turbine have a much higher efficiency than the thermo compressor, resulting in significant energy savings, thus lowering the desalination costs [19].

If 1 kg of steam at the extraction point of the turbine id fed to the ME-TVC at 3 bar and saturated condition (133.55˚C temperature, 2,725.3 kJ/kg specific enthalpy, and 6.9919 kJ/(kg.K) specific entropy) continued to expand in the turbine to a condenser at 10 kP and an 86% dryness fraction, its specific enthalpy at the condenser inlet would be 2249.7 kJ/kg, and the specific entropy would be 7.1 kJ/kg K.

Fig. 25(a). The TC used in ME-TVC [18].

Fig. 25(b). The TC used in ME-TVC with moving nozzle [18].

The work output/kg of this steam would be (2,725.3 – 2,249.7=) 475.6 kJ/kg. For GR 8.6 as in Umm Alnar plant, or GR = 9.8 as in Al-Jubail plant, the work loss per kg of the produced distillate is 55.3 kJ/kg (15.36 kWh/m^3) or 48.5 kJ/kg (13.48 kWh/m^3), respectively. This is separate from the consumed energy required for pumping (2 kWh/m^3). This is almost twice the energy consumed by the MVC, as reported by a leading MVC desalting unit manufacturer of 8 kWh/m^3 for units producing 3,000 m^3/d, and is expected to be 7.5 kWh/m^3 for 4,000 and 5,000 m^3/d for newly designed units [19].

It is better to direct the extracted steam to the ME-TVC system to run a steam turbine coupled with the mechanical vapor compressor, as shown in Fig. 26 [19]. The turbine and the compressor are mounted on a single shaft. The turbine is fed by the supplied extraction steam at a higher pressure and discharges it at a lower pressure of 35 kPa, which is directed to the first effect of the MED plant. The rotating turbine drives the compressor which, in turn, sucks water vapor from one of the effects and discharges it, also to the first effect. After the compressor suction point the remaining vapor, originating from the turbine discharge, continues to operate the rear effects, now lower in size, and eventually discharges its heat to the condenser.

It should be noted that, compared to the standard vapor compression units, no titanium plate heat exchangers are needed, which constitutes a major cost saving. Techno-economic evaluation of the turbo-compressor concept—a preliminary design of an MED plant of 15,000 m^3/d capacity—was carried out. For extraction steam of 4.5 bar at 330˚C, an economy ratio (product/steam) of 20 was obtained for the turbo compressor systems. The required compressor characteristics would be a volumetric flow of 170 m^3/s and a two compression ratio. The maximum volumetric flow that can be obtained so far per compressor is 320 m^3/s [20].

The main obstacle of using the MVC is the small capacity by single unit. However, the required capacities may be achieved by the installation of multiple units. The commercially available MVC units have many of the advantages of the ME-TVC such as the use of falling film horizontal tube evaporator/condenser and low temperature operation (a maximum brine temperature of 70˚C).

Finally, to improve the advances in the ME-TVC and MVC, the following would be required:

(1) Increase the heat transfer coefficients further by utilizing corrugated oval tubes as heat transfer surfaces instead of the round tubes used in the MED plants today.

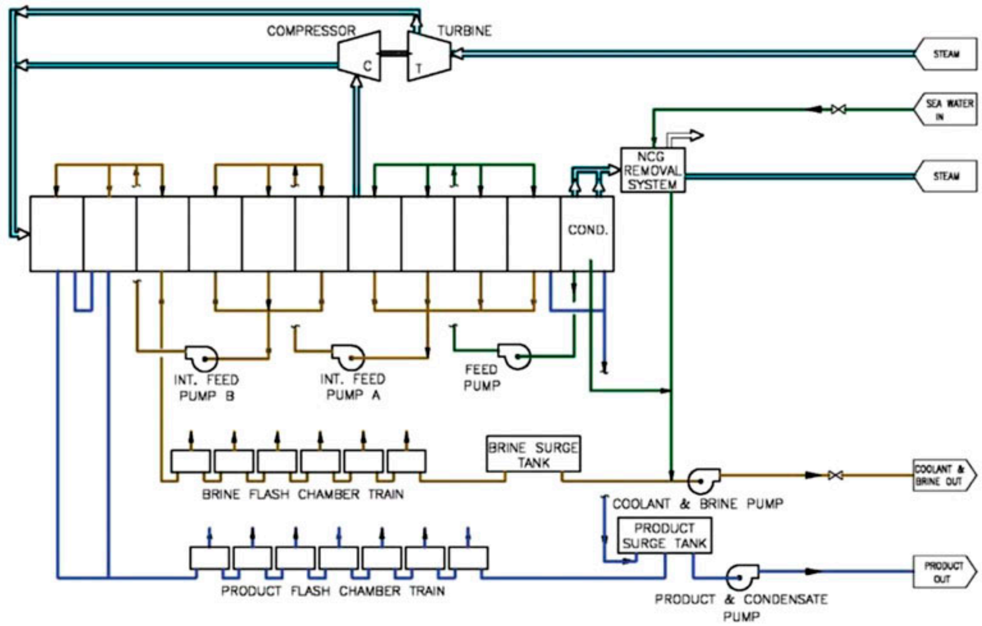

Fig. 26. Mechanically-operated vapor compression desalting system with compressor on shaft with steam turbine driven by extracted steam [20].

(2) Extend the operation range by increasing the top temperature from 70°C to about 85°C.
(3) Increase the economy ratio in the MED plants by replacing the relatively inefficient ejector with a mechanical compressor coupled to a steam turbine.

7. Using solar energy for desalting

The feasibility of using concentrated solar power (CSP) to produce DW by a thermally operated desalting system in large quantities is questionable, at least for the time being. This is illustrated by a simple example. The use of parabolic trough collectors to produce one MW(th) requires 2,000 m^2 aperture area, and about 8,000 m^2 land area. For a desalting plant of 10 GR, which represents high typical figure for GR in MSF and TVC/ME, the specific energy input is 233 kJ/kg. Mechanical pumping energy is also needed (14.4 kJ/kg for MSF or 7.2 kJ/kg for MED), which requires thermal energy input of 43.24 kJ/kg for MSF or 21.6 kJ/kg for MED. So, the desalted water output of TVC/ME from solar collectors producing one MW (th) is 313 m^3/d for MSF or 339 m^3/d for TVC/ME. If these solar collectors were used to produce power, the output would be 330 kWe. This power can operate mechanically driven desalting system like SWRO or MVC, consuming 5 and 8 kWh/m^3 respectively. The SWRO would produce 18.3 kg/s (1,584 m^3/d), and the MVC would produce 11.46 kg/s (990 m^3/d). The SWRO and MVC output would be 4.67 and 2.92 times that of TVC/MED respectively, or 5.06 and 3.17 times that of MSF respectively. The problem worsens as the number of hours operated per day with a nominal thermal output of less than 6 h and not 24 h, as calculated here. This means an average daily desalted water output from solar energy of one MW(th)—output is in terms of 396 m^3/d from SWRO, and 247.5 m^3/d from MVC when the solar collectors operate power Rankine cycle, and 84.8 m^3/d from TVC/ME, and 78.2 from MSF, when the solar collectors operate these desalting system directly. These numbers give the collector area in square meters required to produce one cubic meter $m^2/(m^3/d)$ of desalted water as 5.05 $m^2/(m^3/d)$ for SWRO, 8.08 $m^2/(m^3/d)$ for MVC, 23.58 $m^2/(m^3/d)$ for TVC, and 25.6 $m^2/(m^3/d)$ for MSF desalting systems.

The cost of the parabolic trough collectors and the heat transfer fluid required to produce one kW(th) is \$1,160/kw(th), or \$580/m^2 of collector area, [21]. Three kW(th) would produce one kWe. Knowing this, the cost of the collectors producing one MW(th) is 1.16 M, and the cost of the collectors required to produce one MIGD (4,546 m^3/d) is 13.317 M for SWRO, 21.31 M for MVC, 62.16 M for TVC, and 67.44 M for MSF. It is noted that the cost of solar collectors and its heat transfer fluid (HTF) represents about 50% of solar power plant cost as shown in Fig. 27. So, the capital costs of solar equipment (collectors in the case of MSF

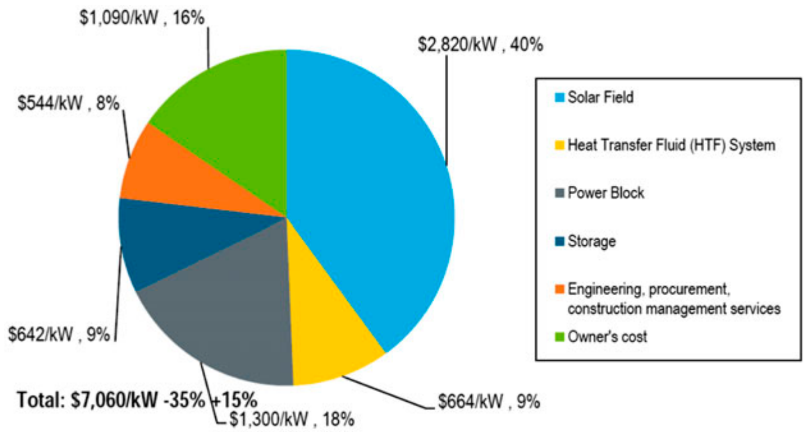

Fig. 27. Capital cost break of thermal CSP solar power plant using PTC [21].

or TVC/ME, and solar power plant in case of SWRO or MVC) required to produce one MIGD (not including the desalting units themselves) are 26.63 M for SWRO, and 42.62 M for MVC, 62.16 M for TVC, and 67.44 M for MSF systems.

8. Conclusion

This paper presents the two thermal desalting methods used in the GCC to desalinate seawater in large quantities, namely the MSF and TVC/ME systems. These two methods are competing with the much more energy-efficient SWRO desalting system. The MSF is a mature process with little room with improvement. The value of suggested increase of MSF unit capacity, TBT, and GR by using partial or full feedwater NF pretreatment is questionable. Increasing the GR, currently limited to 13, will decrease the amount of the steam consumed per kg of distillate, but the value of this steam represented by exergy is also increased. The decrease of consumed steam is balanced by its exergy increase. The use of NF is costly and complicated as illustrated herein. The consumed energy gap by MSF (about 19 kWh/m^3) and that of SWRO (4–5 kWh/m^3) is too wide to be closed by known measures. Installation of new MSF units is a waste of the GCC's prime energy (its real wealth), and should be ceased. The situation of the TVC/ME is better than that of MSF. There is room for development in order to increase its GR to 20.

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Symbols and abbreviations

AF	—	air filter
BH	—	brine heater
BFP	—	boiler feed pump
BPST	—	back pressure steam turbine
CC	—	combined gas/steam turbine cycle
CSP	—	concentrated solar power
CPDP	—	cogeneration power desalting plant
D	—	total vapor flashed from R
DW	—	desalted seawater
eq	—	equivalent
EP	—	electric power
ECST	—	extracting condensing steam turbine
F	—	feed
FH	—	feed heater
GR	—	gain ratio
GT	—	gas turbine
GCC	—	gulf cooperating countries
h	—	specific enthalpy
HP	—	high pressure
HJS	—	heat rejection stages
HRS	—	heat recovery stages
HRSG	—	heat recovery steam generators
IP	—	intermediate Pressure
KAUST	—	King Abdullah University of Science and Technology
LP	—	low pressure
M	—	millions

ME	—	multi-effect
M_c	—	cooling seawater mass flow rate to the MSF or ME desalting units
MSF	—	multi-stage flash
MVC	—	mechanical vapor compression
MIGD	—	million imperial gallons per day
n	—	no. of stages
NF	—	nanofiltration
PR	—	performance ratio
PTC	—	parabolic trough collectors
Q	—	heat gain by the desalting unit
R	—	recirculation
R/D	—	recirculation to distillate rates ratio
S	—	steam
SA	—	Saudi Arabia
SI	—	Skillman index
ST	—	steam turbine
SW	—	seawater
SWRO	—	seawater reverse osmosis
SWCC	—	Saline Water Conversion Corporation
TC	—	thermal compressor
TVC	—	thermal vapor compression
TDS	—	total dissolved solids, in part per million, or mg/l
TTD	—	terminal temperature difference
TBT	—	top brine temperature
U	—	overall heat transfer coefficient
UAE	—	United Arab Emirates
W	—	work
η_s	—	heat efficiency of the brine heater
α	—	heat transfer unit (= UA/RC)

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