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# Steam, electricity and water costs evaluation of power-desalination co-generation plants

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**ABSTRACT**

This paper addresses the effect of the oil prices on the cost of low pressure heating steam for thermal desalination systems, electricity and desalinated water. Two methods of calculating the low pressure steam, and electricity are developed and compared with a typical power/water cogeneration plant commissioned in 2009, by the King of Saudi Arabia. The first method is based on the calculation of the monetary cost of all streams of power cycle and charging the levelized capital and operating and maintenance costs based on exergy. The cost of bleeding steam is allocated based on the steam quality. In the second method we calculate the low pressure steam of the back pressure turbine based on equivalent cost of power loss due to the steam withdrawal for desalination plant. Visual Design and Simulation Program (VDSP) previously developed by the authors is used. The effect of a wide range of the oil price (5–100 \$/bbl) is investigated. The results showed that the second method overestimates the electricity cost by 8% and underestimates the low pressure steam cost by 25% at oil price of 70 \$/bbl. The results showed also that, the product water, electricity, and steam costs are significantly affected by the variation in oil price. The product water cost will be less than 1.0 $\$/m^3$, only if oil price is subsidized to near to 20 \$/bbl. However, the water unit cost may jump to 4 $\$/m^3$ when the oil price increased to 100 \$/bbl.

*Keywords:* Desalination; Electricity; MSF; Water cost; Exergy; Cogeneration

## 1. Introduction

Thermal desalination technology, particularly Multi stage flash (MSF), has proven to be the most reliable and mature technology for producing high quality distilled water on large scale and for high feed water salinity. Due to its advantages and accumulated operational experience gained (from over 50 years of operational feedback), MSF still dominates the thermal desalination market particularly in the Gulf region.

There is still a continuous need for large capacity units to satisfy the growing population and developments in the region. In addition, thermal technologies match the need to combine water production with electricity generation plant, and the use of available low grade thermal energy from thermal power plants for water production. According to the present state of technology, commercial thermal seawater desalination plants (MSF/MED) are built together with power plants to utilize the low pressure steam. When the water/power ratio is high, the back pressure turbine is utilized to provide low pressure steam to the MSF plant.

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The Gulf feed water problems of high salinity, high temperature, high turbidity and high marine growth militate against the choice of competing reverse osmosis (RO) technology. The significant draw back of the thermal desalination, as compared to other cost effective technologies (as RO) is the high specific energy consumption (i.e., heating steam energy and electrical power for pumping). This drawback addresses a real challenge to MSF to improve the unit's performance, reduce energy consumption and hence reduce the water production cost.

Real cost calculation of the low pressure steam and electricity used in the thermal desalination plant is important for many reasons including: (i) proper evaluation of the economics, efficiency and capacity-improvements of existing or new proposed process configuration (If the calculated cost is not accurate, many good energy projects may be rejected), (ii) proper evaluation of the proposed cogeneration projects for minimum cost.

In the recent times, crude oil costs in the world market are varying from 70–90 US$/bbl. However many tenders of governmental projects compare the water and energy cost based on the subsidized value of 5 US$/bbl (0.75 $/MBTU) [1]. With this low fuel cost, the desalinated water cost is calculated at around 1–1.5 US$/m$^3$. With today's world market prices for crude oil, the real specific desalted water cost will exceed the lowest value.

Technical and economical analysis to evaluate the integration of RO processes with existing thermal desalination processes and power generation is presented in [2]. It is reported that for all plant capacities, hybrid systems resulted in the most cost effective desalination system. The unit cost of low pressure steam is considerably high (60 US$/ton) when the steam is generated using direct firing (single purpose). However, if the fuel is used to generate power and steam (double purpose), the unit cost of steam may be assumed to be zero [2]. The capital and operational costs in typical thermal desalination plants is outlined in [3] and showed that the cost of energy represents the main part of the total operating cost (OPEX). The steam cost represented 38% of operation cost while electrical power represented 14%. This high percentage of energy cost share and eventual evaluation of its methods of calculation should be addressed.

Evaluation of real power cost based on theoretical study is presented in [4], where the technical and environmental evaluation study included four nuclear power plants and three fossils fuelled power plants (circulating fluidized bed coal-fired, oil-fired, gas turbine combined). The oil, gas and coal price were 60 $/bbl, 7 $/MBTU, and 65 $/t respectively. Among the fossil fuelled power plants, the cost of circulating fluidized bed coal fired would be the lowest and the oil fired plant would be the highest. The nuclear energy system gave the lowest cost of electricity. The study has also focused on the environmental costs power (fossil fuelled based and nuclear) and MED and RO desalination systems.

In order to determine the cost of the low pressure steam of the back pressure turbine outlet used as a heating source for thermal desalination plants, more detailed data about the steam conditions are required. The prorating (based on costs in single purpose plant, based on power generated, and based on exergy) and credit (power credit or water credit) methods have been proposed to allocate the total annual cost of dual purpose plants to desalted water and electricity. [5–6]. In these methods, the fuel cost used to generate steam was charged to the two products (water and power). However none of these methods calculates the cost of low pressure steam used in the thermal desalination plants. Also, these methods, assume the total cost of dual purpose plant is given as a lump sum. However, consortium of two partners is candidate to build up and transfer the power/water project, power side partner and water side partner offer their own price based on Tariff calculation system. Therefore, the allocation of the common systems at the interface between power and water side must be calculated accurately.

The details of prorating on the basis of power generation or "power loss" method is presented in [5–6]. This method is based on evaluating the power consumed due to withdrawal low pressure heating steam for the thermal desalination plants (see Fig. 1.). The annual costs for the turbo generator and the evaporator plant are attributed to power and water, respectively, whereas the costs for the high pressure seam generator (including fuel cost) have to be divided between the two products. The water and electricity costs are calculated as follows [5].

$$\dot{C}_{water} = \dot{C}_{Evaporator} + \dot{C}_{steam-generator}\left(\frac{\dot{W}_{SP} - \dot{W}_{DP}}{\dot{W}_{SP}}\right) \quad (1)$$

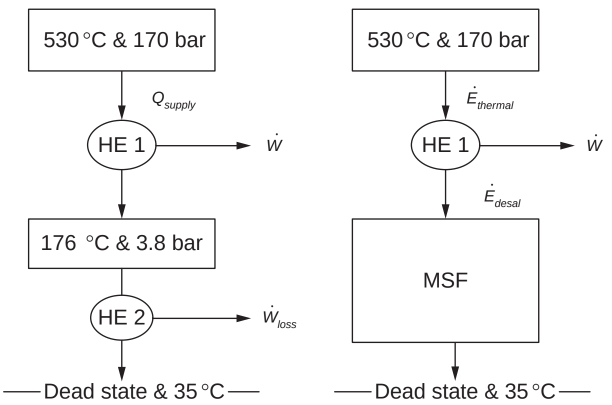


Fig. 1. Equivalent power loss of the LP steam for desalination plant.

$$\dot{C}_{electricity} = \dot{C}_{Turbine} + \dot{C}_{steam-generator}\left(\frac{\dot{W}_{DP}}{\dot{W}_{SP}}\right) \tag{2}$$

The disadvantage of this method is that it is by no means certain that a power-only and dual purpose plant would have the same type of steam generator or the same steam condition at the turbine inlet [5].

On the other hand, the prorating on the basis of exergy is also presented in [5–6]. This method proposed a simple exergy procedure that can be applies to any co-generation process at any design point or operational state. In this method, three main systems characterizing exergies: $E_{fuel}$, $E_{thermal}$, and $E_{desal}$. The first, $E_{fuel}$, is the fuel input chemical exergy. The second, $E_{thermal}$, is the converted thermal exergy from the fuel chemical exergy. The third, $E_{desal}$, is the exergy of the low pressure steam directed to MSF desalination system. The fuel cost assigned to water is proportion to the $E_{desal}/E_{thermal}$, however the cost of fuel assigned to power is proportional to $1 - E_{desal}/E_{thermal}$.

The water and electricity costs are calculated as follows:-

$$\dot{C}_{water} = \dot{C}_{Evaporator} + \dot{C}_{steam-generator}\left(\dot{E}_{desal}/\dot{E}_{thermal}\right) \tag{3}$$

$$\dot{C}_{electricity} = \dot{C}_{Turbine} + \dot{C}_{steam-generator}\left(1 - \dot{E}_{desal}/\dot{E}_{thermal}\right) \tag{4}$$

However this model did not consider the bleeding steam or low pressure steam costs.

In the present work, two methods are developed for calculating the cost of low pressure heating steam for desalination plant and electricity of power plant. The first method is based on exergy analysis, and the second method is based on power loss. The results of the two methods are compared under a wide range of oil prices. The effect of the oil price variation on steam, electricity and water costs is studied based on one world largest power and water cogeneration plant.

## 2. Thermoeconomic model development

The thermoeconomic model is verified using a case study of power and MSF desalination cogeneration plant of 3 × 400 MW power and 12 × 16.2 MIGD MSF desalination plant. This plant is commissioned on 2009 in Shuiabah phase 3, Saudi Arabia [1]. Two companies shared the plant work. Siemens Company supplied three steam turbine-generator units, each rated at 400 MW. Doosan Heavy Industries and Construction manufactured, shipped and commissioned the desalination plant (194.4 MIGD) [1]. It was the largest worldwide desalination plant to date [10] and consists of twelve evaporators of 16.2 MIGD each. Each four evaporators are connected to one power cycle. Fig. 2 illustrates the process flow diagram of one power cycle with four evaporators. The boiler is supplied with fuel energy, in order to raise the availability (exergy) of the generated steam.

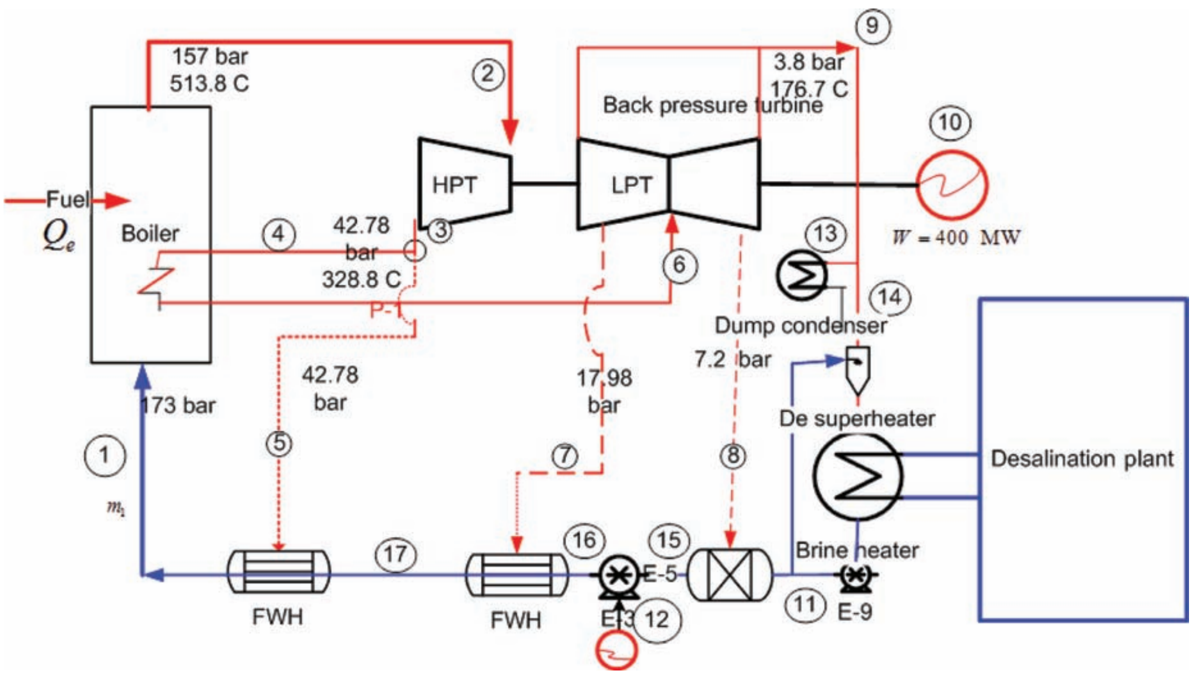


Fig. 2. Power and water cogeneration plant, (400 MW & 4 × 16.2 MIGD MSF).

The generated steam, at a high pressure and temperature (157 bar and 513 °C), is utilized to produce mechanical power in a back pressure turbine, before it is used as a heat source for the MSF, at 3.6 bar and 176 °C. The top brine temperature (TBT) for the MSF is TBT = 111 °C.

The newly developed Visual Simulation Program (VDSP) by the authors is used to perform process design calculations of MSF evaporator [7–10]. In the VDSP software, operating conditions of the heating steam operating conditions coming from the back pressure turbine (pressure, temperature), the target capacity by evaporator (Distillate rate per hour), top brine temperature (TBT), sea water conditions ( temperature, salinity), make up flow rate, brine recirculation salinity, reject brine temperature are defined. Some design parameters including the number of stages, tube length, diameters, material type, price of tube and shell material used in evaporator manufacturing are considered. Using VDSP, all process streams are determined (mass, temperature, pressure, entropy, and rated cost) also the heat transfer surface area (number of tubes), evaporator size, internal dimensions and pumps are sized. Then after, the detailed CAPEX analysis is estimated. Also the VDSP calculates the heating steam consumption rate, the consumed chemicals (anti scales, anti foam, chlorination) as well as the pumping power are determined (OPEX items). In the present study, the detailed cost of material used in the evaporator and purchased equipment cost such as pumps, valves, controls are estimated. The evaporator and power cycle equipments are levelized based on the interest rate of 7% along plant life cycle of 20 years. So the power and water unit tariff cost could be estimated.

### *2.1. Methods*

#### *2.1.1. Allocation based on exergy analysis*

Each of the plant components (e.g., steam generator, turbine, pump, etc) is represented by algebraic equation of cost balance which related outlet streams and inlet streams as follows:

1. Applying cost balance on boiler unit, referring to Fig. 2:

$$\left(\dot{C}_2-\dot{C}_1\right)+\left(\dot{C}_6-\dot{C}_4\right)=\dot{C}_{fuel}+\dot{z}_{boiler}^{CI+O\&M} \tag{5}$$

This equation means that the cost of the generated steam (point 2) plus the cost of reheat seam (point 6) is balance to the cost of input streams (1 and 4) plus the fuel cost and the levelized cost of the steam generator. The levelized cost of the steam generator includes capital investment and operating and maintenance, $\dot{z}_{boiler}^{CI+O\&M}$. Since the number of the outlet streams is more than the inlet stream by one, an auxiliary equation is invoked as follows:

$$\frac{\dot{C}_2-\dot{C}_1}{\dot{E}_2-\dot{E}_1}=\frac{\dot{C}_6-\dot{C}_4}{\dot{E}_6-\dot{E}_2} \tag{6}$$

This equation means that the specific cost per unit exergy of the generated steam is equal the specific cost per unit exergy of the re-heated steam.

A typical processed oil barrel produces 5.71 GJ of thermal energy of complete combustion. The fuel cost can then be calculated as follow:

$$\dot{C}_{fuel}=\dot{C}_{barrel}\times\frac{Q_{f,boiler}(MW)}{1000\times 5.71}\times 3600 \tag{7}$$

where the consumed thermal energy, $Q_{f,boiler}$, is calculated from heat balance of steam generator as follows:

$$Q_{f,\,boiler}=\frac{\left[\dot{M}_2\times(h_2-h_1)+\dot{M}_4\times(h_6-h_4)\right]}{\eta_b} \tag{8}$$

2. Applying cost balance for the back pressure turbine unit, Fig. 2

$$\left(\dot{C}_{10}+\dot{C}_9\right)=\left(\dot{C}_6+\dot{C}_2-\dot{C}_3-\dot{C}_7-\dot{C}_8+\dot{Z}_{turbine}\right) \tag{9}$$

Auxiliary equations

$$\frac{\dot{C}_3}{\dot{E}_3}=\frac{\dot{C}_2}{\dot{E}_2} \tag{10}$$

$$\frac{\dot{C}_7}{\dot{E}_7}=\frac{\dot{C}_6}{\dot{E}_6} \tag{11}$$

$$\frac{\dot{C}_8}{\dot{E}_8}=\frac{\dot{C}_6}{\dot{E}_6} \tag{12}$$

$$\frac{\dot{C}_9}{\dot{E}_9}=\frac{\dot{C}_6}{\dot{E}_6} \tag{13}$$

3. Aggregation the feed water heaters and de-aerator, referring to Fig. 2.

$$\dot{C}_1-\dot{C}_5-\dot{C}_7-\dot{C}_8-\dot{C}_{11}-\dot{C}_{12}=2\times\dot{Z}_{FWH}+\dot{Z}_{pump}+\dot{Z}_{de-aerator} \tag{14}$$

$$\frac{\dot{C}_{11}}{\dot{E}_{11}}=\frac{\dot{C}_9}{\dot{E}_9} \tag{15}$$

$$\frac{\dot{C}_{12}}{\dot{E}_{12}} = \frac{\dot{C}_{10}}{\dot{E}_{10}} \tag{16}$$

$$\dot{E}_{10} = \dot{W}_T = \begin{bmatrix} \dot{m}_2 \times (h_2 - h_3) + \left(\dot{m}_2 - \dot{m}_5\right) \times (h_6 - h_7) \\ + \left(\dot{m}_2 - \dot{m}_5 - \dot{m}_7\right) \times (h_7 - h_8) \\ + \left(\dot{m}_2 - \dot{m}_5 - \dot{m}_7 - \dot{m}_8\right) \times (h_8 - h_9) \end{bmatrix} \times \eta_g \tag{17}$$

$$\dot{E}_{12} = \dot{W}_{FWPUMP} = \frac{\dot{m}_1 \times \Delta\left(\dot{P}_1 - \dot{P}_{11}\right)}{\eta_p} \tag{18}$$

4. Applying cost balance for splitter of point, 3, Fig. 2.

$$\dot{C}_4 + \dot{C}_5 = \dot{C}_3 \tag{19}$$

$$\frac{\dot{C}_5}{\dot{E}_5} = \frac{\dot{C}_4}{\dot{E}_4} \tag{20}$$

5. Applying cost balance for splitter of point, 9, (dump condenser)

$$\dot{C}_9 - \dot{C}_{13} - \dot{C}_{14} = 0 \tag{21}$$

$$\frac{\dot{C}_{13}}{\dot{E}_{13}} = \frac{\dot{C}_{14}}{\dot{E}_{14}} \tag{22}$$

Equations 5–22 are solved simultaneously using the VDSP software [6–9]. The cost flow rate of each plant stream is calculated such that the cost of the electricity ($C_w = C_{10}$) and low pressure steam ($C_{LP,steam} = C_{14}$).

### 2.1.2. Allocation based on power loss method

In this method, it is assumed that the back steam is expanded in another turbine until the saturation pressure of (0.112 bar and 48 °C) and condensed using condenser as shown in Fig. 3. Then the condensate preheated again using bleeding steam (1.86 bar).

Heat balance for the additional condensate feed water heater

$$m \times \left(h_{24} - h_{24f}\right) = M \times \left(h_{11} - h_{22}\right) \tag{23}$$

The power generation due to steam expansion through the addition LP turbine is calculated as follows:

$$\dot{W}_{loss} = \eta_g \left[ m \times (h_9 - h_{24}) + (M - m) \times (h_{24} - h_{20}) \right] \tag{24}$$

The total power of the single purpose plant is calculated as follows:

$$\dot{W}_{SP} = \dot{W}_{DP} + \dot{W}_{loss} \tag{25}$$

Cost of the electricity of single purpose 'power-only' plant is

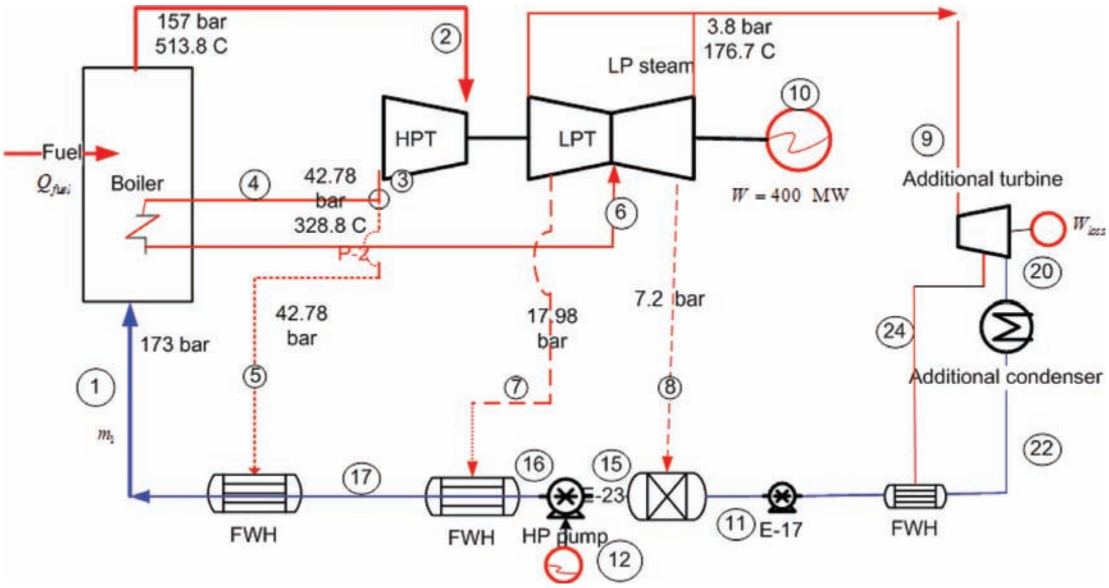


Fig. 3. Single purpose power & power loss.

$$\dot{C}_{SP,power} = \dot{C}_{fuel} + \dot{z}^{CI+O\&M}_{equipment+add.-turbine}, \$/\text{hr} \quad (26)$$

The specific cost of electricity is

$$C_{SP,power} = \frac{\dot{C}_{fuel} + \dot{z}^{CI+O\&M}_{equipment+add.turbine}}{\dot{W}_{SP}}, \$/\text{kWh} \quad (27)$$

Cost of lost power

$$\dot{C}_{lost,power} = \frac{\dot{C}_{fuel} + \dot{z}^{CI+O\&M}_{equipment+add.turbine}}{\dot{W}_{SP}} \times W_{lost}, \$/\text{hr} \quad (28)$$

$$C_{steam} = C_{lost,power,\$/\text{hr}} \quad (29)$$

By solving Eqs. 23–29, the cost of electricity and MSF low pressure steam are calculated.

## 3. Analysis of the results

Table 1 shows the power cycle streams mass flow rates, exergy flow rates, cost flow rates, and the cost per unit exergy and specific cost of steam. The oil price is subsidized in this project at 5 $/bbl (0.75 $/MBTU) as appeared in recent bidding. The process calculation showed that the steam generator produce 534 kg/s (stream 2) as shown in Fig. 2, and the low pressure steam at the back pressure turbine is 454 kg/s (stream no. 9). The amount of the heating steam flows to the desalination plans is 406 kg/s. However 48 kg/s flows to the process heat or dumped condenser. The third column of this table shows that the exergy of the high pressure steam (stream 2) is 773 MW in the cycle. The exergy of stream No. 10 which represents the power output of 367 MW. The fourth column of this table presented the cost flow rate for each stream. Each stream cost represents the capital and fuel invested per hour. The cost of steam at point 2 (boiler outlet) represent the highest value of 10,429 $/h. the cost of the steam directed to the desalination plants is 3662 $/h as shown in the table.

The highest exergetic and cost stream is achieved at stream 2 (boiler outlet). This is due to the input chemical energy associated with fuel consumption in the boiler. This leads to a highest specific steam cost (5.42 $/ton of steam). However, the calculated lower specific cost 2.5 $/ton is for the low pressure steam (stream 14). The highest specific exergy unit cost is obtained at stream 10 (turbine shaft) of 4.79 $/GJ. This is because the generated energy is converted to a shaft power. The price of electricity is calculated by dividing the value of stream cost (6331.98 $/h) by the generated power of 367.6 MW which gives the electricity cost by 0.017 $/kWh.

However, if the oil price is 70 $/bbl, the steam cost in each point in the power cycle and the low pressure steam (stream 14) are calculated as shown in Table 2. The cost of high quality steam (stream 2) is 25.04 $/ton). However, the lower pressure steam (stream 14) is calculated by 11.48 $/ton as shown in Table 2. The highest exergy unit cost is obtained at stream 10 (turbine shaft) of 21 $/GJ. This is because the generated energy is converted to a shaft power. The price of electricity is calculated by dividing the value of stream cost (27,847 $/h) by the generated power of 367.6 MW which gives the electricity cost by 0.076 $/kWh.

Table 1
Thermo economic results of power cycle of 400 MW at (5 $/bbl)

| Stream No. | Mass flow rate, kg/s | Exergy, MW | Cost, $/h | Cost per unit exergy, $/GJ | Specific cost, $/ton |
|---|---|---|---|---|---|
| 1 | 534 | 141.36 | 2106 | 4.14 | 1.1 |
| 2 | 534 | **773.31** | **10,429.00** | 3.75 | **5.42** |
| 3 | 534 | 617.56 | 8328 | 3.75 | 4.33 |
| 4 | 482 | 557.78 | 7522 | 3.75 | 4.33 |
| 5 | 52 | 59.78 | 806 | 3.75 | 4.33 |
| 6 | 482 | 593.83 | 7996 | 3.74 | 4.61 |
| 7 | 28 | 27.93 | 376 | 3.74 | 3.76 |
| 8 | 28 | 22.07 | 297 | 3.74 | 2.94 |
| 9 | 454 | 303.61 | 4088 | 3.74 | 2.5 |
| 10 | | **367.58** | 6332 | 4.79 | |
| 11 | 454 | 25.46 | 343 | 3.74 | 0.21 |
| 12 | | 11.09 | 191 | **4.79** | |
| 13 | 48 | 32.07 | 426 | 3.7 | 2.47 |
| 14 | 406 | 275.23 | 3662 | 3.7 | **2.5** |

Table 2
Thermo economic results of power cycle of 400 MW at (70 $/bbl)

| Stream No. | Mass flow rate, kg/s | Exergy, MW | Cost, $/h | Cost per unit exergy, $/GJ | Specific cost, $/ton |
|---|---|---|---|---|---|
| 1 | 534 | 141.36 | 9,253 | 18.18 | 4.81 |
| 2 | 534 | 773.31 | **48,138** | 17.29 | **25** |
| 3 | 534 | 617.56 | 38,442 | 17.29 | 20 |
| 4 | 482 | 557.78 | 34,720 | 17.29 | 20 |
| 5 | 52 | 59.78 | 3,721 | 17.29 | 20 |
| 6 | 482 | 593.83 | 36,939 | 17.28 | 21.3 |
| 7 | 28 | 27.93 | 1,737 | 17.28 | 17.4 |
| 8 | 28 | 22.07 | 1,373 | 17.28 | 13.6 |
| 9 | 454.5 | 303.61 | 18,886 | 17.28 | 11.54 |
| 10 | | 367.58 | 27,847 | 21.04 | |
| 11 | 454.5 | 25.46 | 1,584 | 17.28 | 0.97 |
| 12 | | 11.09 | 742 | **21.04** | |
| 13 | 48 | 32.07 | 2,083 | 18.06 | 12 |
| 14 | 406 | 275.23 | 16,804 | 18.06 | **11.5** |

Using a wide range of oil prices 5–100 $/bbl, the calculated electricity and heat steam are compared as shown in Figs. 4 and 5. The electricity and heating steam cost dramatically increase since the oil price increase. As shown in Table 3, the power loss method overestimated the electricity cost however underestimated the steam cost. The difference between the two methods is due to using different allocation of annualized cost of common systems. The power loss method use approximate value of the annualized capital investment of the additional equipment (turbine, condenser, feed water heater and pump). However, the exergy model is rather accurate since it deals with the annualized cost of the present case as well as it consider all power plant streams such as the cost of bleeding steam for both feed water heaters and

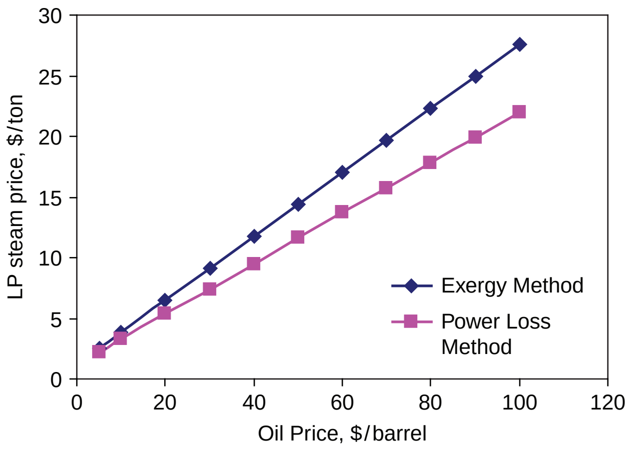


Fig. 5. Low pressure steam cost versus oil price.

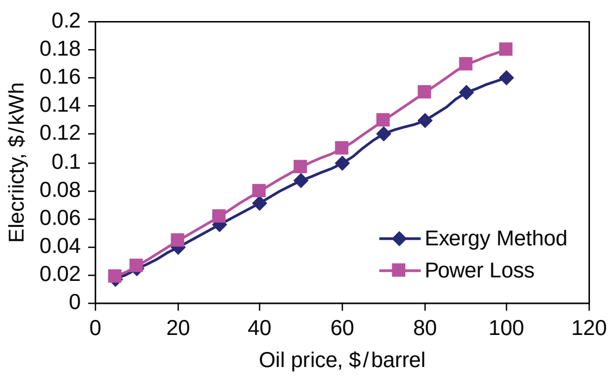


Fig. 4. Electricity cost versus the oil price.

Table 3
Comparison between exergy and power loss method

| Oil price, $/Barrel | Steam cost, % diff | Electricity, % diff |
|---|---|---|
| 5 | 12 | –7 |
| 10 | 16 | –7 |
| 20 | 21 | –11 |
| 30 | 23 | –10 |
| 40 | 24 | –10 |
| 50 | 24 | –10 |
| 60 | 24 | –9 |
| 70 | 25 | –8 |
| 80 | 25 | –13 |
| 90 | 25 | –12 |
| 100 | 26 | –11 |

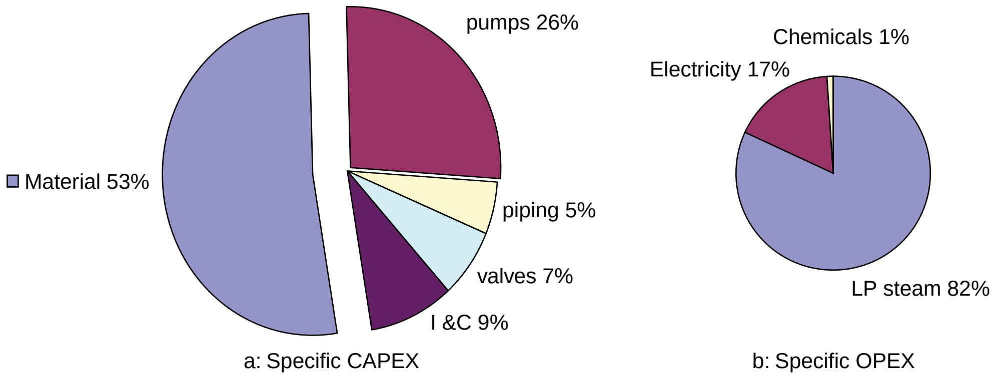


Fig. 6. Cost analysis of 16.2 MIGD MSF plant.

de-aerator. The exergy model takes also care of the levelized cost details and the cost allocation is based on the steam quality. The results showed that the power loss method overestimates the electricity cost by 8% and underestimates the low pressure steam cost by 25% at oil price of 70 $/bbl.

Fig. 6 shows the overall cost breakdown of MSF plant. The specific capital cost (CAPEX) percentage includes the evaporator material, pumps, valves, and instrumentation and control devices costs is shown in Fig. 6a. The cost analysis showed that the material cost of MSF represented 53% of the total specific CAPEX. The pumps with its facilities came as the second contributor of 26%. The specific operating cost (OPEX) is varying according to the oil price as explained above. The OPEX includes the low pressure steam, the electricity and the chemicals costs as shown in Fig. 6b. Based on oil price of 70 $/bbl, the low pressure steam cost represents also 82% of the OPEX. The electricity cost represents the second item cost 17%; however the chemicals cost represents a lower value of only 1% of the OPEX.

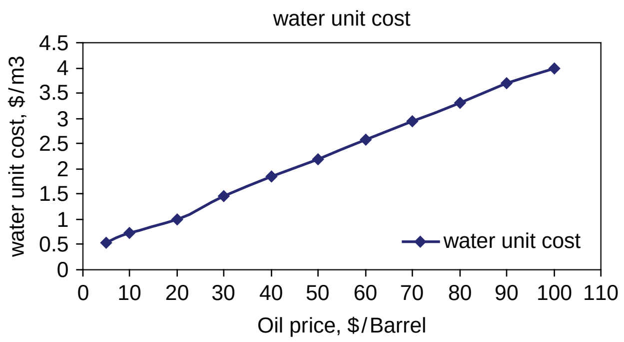


Fig. 7. Water cost at different values of oil price.

The effect of oil price increase on the water costs as shown in Fig. 7. It shows that the unit product water cost is significantly affected by the oil price variation. If the subsidized fuel cost of 20 $/barrel is used, the unit product water cost could be around 1.0 $/m$^3$. However for oil price of 100 $/barrel, the unit product water cost may jump to 4 $/m$^3$.

## 4. Conclusion

Two methods of calculating the cost of low pressure steam, electricity are developed and compared for a typical power-water cogeneration plant, recently commissioned on 2009, KSA. The first method (exergy) is based on calculating the monetary cost of all streams of power cycle and charging the levelized capital and operating and maintenance cost based on exergy. The second method is based on equivalent cost of power loss due to steam withdrawal to the desalination plant (power loss). The thermoeconomic (exergy based) method considered the cost of bleeding steam as well as allocating cost based on the steam quality. The effect of a wide range of the oil price (5–100 $/Barrel) is studied. The results showed that the power loss method overestimates the electricity cost by 8% and underestimates the low pressure steam cost by 25% at oil price of 70 $/bbl. The results show also that, the product water, electricity, and steam costs are significantly affected by the variation in oil price. The product water cost will be less than 1.0 $/m$^3$, if fuel cost is subsidized to near to 20 $/Barrel. However, if the fuel cost is not subsidized the water unit cost may jump to 4 $/m$^3$ if the fuel price increased to 100 $/Barrel.

## Symbols

$C$ — cost flow rate, \$/hr
$E$ — exergy flow rate, MW
$h$ — specific enthalpy, kJ/kg
$M$ — mass flow rate, kg/s
$Q$ — input energy, MW
$S$ — specific entropy, kJ/kg.K
$T$ — temperature, °C
$W$ — power, MW

### *Subscripts*

$F$ — fuel
MSF — multi stage flash
$O$ — dead state
$P$ — product

### *Greek letters*

$\eta_b$ — Efficiency

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