Exergy Analysis of the Optimized MSFD Type of Brackish Water Desalination Process

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ABSTRACT: *This paper presents a detailed thermodynamical exergy analysis of an optimized MSF distillation plant based on the latest published thermodynamics properties of water and seawater software of the Massachusetts Institute of Technology by using design and optimized plant operation data. Exergy flow rates are evaluated throughout the plant and the exergy flow diagram is prepared in both cases. The rates of exergy destruction and their percentages are indicated on the diagram so that the locations of each exergy destruction can easily be identified. The study concludes that as a result of an optimization, making the MSFD unit once-through cooling system to recirculating type by using cooling tower system, the unit's exergy destruction pattern changes meaningfully. Besides, in the three exist thermal desalination plants up to 53 percent of feed water, i.e.; 667 m^3/h and the same amount of reject water can be conserved. Though, with this modification, the unit steam consumption has been increased up to 13 ton/h, about 50 percent of design. Moreover, the detail of the study showed that the exergy destruction can be reduced more than 39% in the pumps, and 30% in blowdown and around 29% in distillate streams. For brine heater, an enhance as small as 0.37% is also achieved. In the other hand, the rate of destruction of exergy increased around 25% in the cooling process and above 5% in the evaporators.*

KEYWORDS: *Multi-Stage Flash Distillation (MSFD); Exergy destruction; Gain Output Ratio (GOR); Heat rejection.*

INTRODUCTION

Desalination systems are divided into two main types; thermal and non-thermal. Thermal type desalination plants such as Multi-Stage Flash (MSF), Vapor Compression (VC), solar distillation and freeze desalination uses heat either direct heating or heat moving. Other systems are classified as non-thermal system such as Reverse Osmosis (RO), Capacitive Deionization Technology (CDT). Actually cost for each method depends mainly on type of physical process of salt removal (i.e. evaporation, filtration, freezing or electrostatic potential difference). The efficiency of each type depends on the total energy required to remove the salt particles which depends to some extent on the method of operation and also on the purity of the required water [1].

Seawater desalination is increasingly useful in several regions around the world because the population is growing in places where there is not enough fresh water to support that growing. It is expected that by 2025, more than 70%

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of world's population will have water shortages [2]. Desalination is an energy-intensive process that currently is made by means of membranes or via thermal processes. The most common desalination processes are energized directly or indirectly by fossil fuels [3]. On the other hand, desalination of seawater has become one of the most important commercial processes to provide fresh water for many communities and industrial sectors. Distillation processes produce about 50% of the worldwide desalination capacity, and 84% of this is produced by MSF technology [3]. Currently, there are more than 18,000 desalination plants in operation worldwide producing 22.9 billion gallons of water per day, 60% are located in the Middle East [4]. The large-scale MSFD plants are among the region's most important commercial processes, as they play a crucial role in providing fresh water for many communal and industrial sectors, especially in areas with a high density of population.

Thermal systems are analyzed traditionally through energy. However, exergy analysis has increasing acceptance as a powerful diagnostic tool, [5], in thermal system design, evaluation, optimization, and improvement. The efficiency of thermal systems such as desalination can be estimated by both the first and second laws of thermodynamics. While the first law focuses on the quantity of energy, second law analysis (exergy analysis) introduces quality as well as quantity. Exergy analysis allocates the irreversibilities in the system and suggests economical modification and enhancement. However, only a limited number of studies have analyzed seawater desalination exergy, due to the complexity of the determination of the seawater stream exergy [11].

As *Fitzsimons* declares in her Ph.D. dissertation [8], one important issue that comes to attention in the literature is the number of different models and approaches that have been used by various research groups. Table 1 shows exergy models which have been used for desalination purposes and the literature reviewed indicates that the model differences typically relate to the chemical exergy terms. One body of research has used the Drioli aqueous solution model. Another approach has used a different model, termed the Cerci ideal mixture model. In her opinion, the variable amount of different exergy models in the literature presents a difficult challenge to the increased utilization of exergy analysis as a tool for desalination energy optimization. Tsatsaronis [9] advocated the need for symbol and nomenclature uniformity in exergy analysis, although this consensus is desirable, the difference in exergy calculation models evident in Table 1 is a more complex issue. In a recent paper, John H. Lienhard and his team [5], found that this ideal mixture model gives unrealistic flow exergy values and a second law efficiency that differs by as much as 80% from the correct value. This shows that exergy calculations and analyses performed using that model are comparatively far from the correct values.

This paper represents a detailed thermodynamical exergy analysis of an optimized MSF distillation plant in a petroleum refinery as a case study, based on the latest published thermodynamics properties of water and seawater of the Massachusetts Institute of Technology by using design and actual plant operation data after optimization. In this paper the effects of the water desalination process optimization as a case study in a petroleum refinery, where for the first time; the result of this study has been executed, is investigated by exergy analysis. The optimization of the unit, which is altering from once-through cooling to circulating type cooling results in preventing the return of treated water to the river with around 14°C temperature rise. This is done by using the cooling tower system [6]. The climatic conditions in Abadan include a very hot summer season where the ambient temperatures can reach temperatures of around 50°C. However, as shown in Fig. 1, monthly average differences of dry and wet bulb temperature of Abadan city from 1951 to 2005 based on Data Processing Center of Islamic Republic of Iran Meteorological Organization, is high from 15 to 25 °C. It is obviously clear that contrary to the first judge, using a cooling tower is suitable and feasible.

Exergy flow rates are evaluated throughout the plant, and the exergy flow diagram is prepared for both cases. In this research, the rates of exergy destruction and their percentages are indicated on the diagram so that the locations of each exergy destruction can easily be identified.

EXPERIMENTAL SECTION

MSFD Plant

The Desalination unit studied in this paper, commissioned in late 1988, is located at Abadan Oil

Table 1: Desalination exergy model equations [8].

Relevant Exergy or specific exergy equation variations	Specific nomenclature
$e=(h-h_o)-T(s-s_o)+R_W T_o\left[\Phi\frac{x_s x_{s0}}{x_s-x_{s0}}\ln\frac{x_s}{x_{s0}}-\Phi x_{s0}\right]$	Φ is a factor which accounts for the electrolytic nature of the salt, R_w is the gas constant for water.
Exergy calculations based on Leyendekker, Thermodynamics of Seawater, Part 1, 1976	Not applicable
$e=e_p\left(T-T_o-T_o\ln\frac{T}{T_o}\right)+\frac{P-P_o}{\rho}-N_{sol}RT_o\ln x_{sol}$ where $N_{sol}=\left[1000-\Sigma\frac{Ci}{\rho}\right]\Big/MW_{sol}$ *and* $x_{sol}=N_{sol}\Big/\left[N_{sol}+\Sigma\frac{\beta iCi}{\rho MWi}\right]$	c_p is the specific heat capacity
Variations of the following equation $e=(h-h_o)-T_o(s-s_o)$ where $h=mf_s h_s+mf_w h_w$ and $s=mf_s S_s+mf_w S_w-R_{im}(x_s\ln x_s+x_w\ln x_w)$	Not applicable
$e=c_{pw}\left(T-T_o-T_o\ln\frac{T}{T_o}\right)+\upsilon_W(P-P_o)+\Sigma x_i(\mu_i-\mu_{io})$ *where* $\Sigma x_i(\mu_i-\mu_{io})=RT_o\Sigma x_i\ln\frac{ai}{aio}$	c_{pw} is the specific heat capacity of pure water

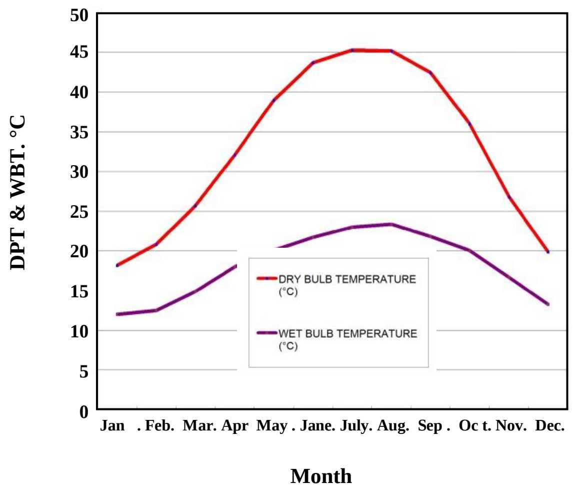

Fig. 1: Abadan Mean Dry and Wet Bulb Temperature during 1951-2005.

Refining Company, Abadan city, 53 km from the coast of Persian Gulf. This plant is capable of desalination of the Arvand river brackish water and producing distillate suitable for use as feed water in the high-pressure boilers**.** The plant comprises three units each capable of producing 197 ton per hour of distilled product water. All unit plants are of the multi-stage flash distillation type.

The multistage condensers have two sections; the heat-recovery (18 stages) and the heat-rejection (4 stages). In each of these stages, arranged in series, distillation takes place progressively. Boiling water in the sump of each stage is kept boiling in successive stages by reduced pressure at a lower temperature. Cooling water flows through the stages in counter flow through appropriate tube arrangement in order to condense the distillate for collection as product water. The 22 stages

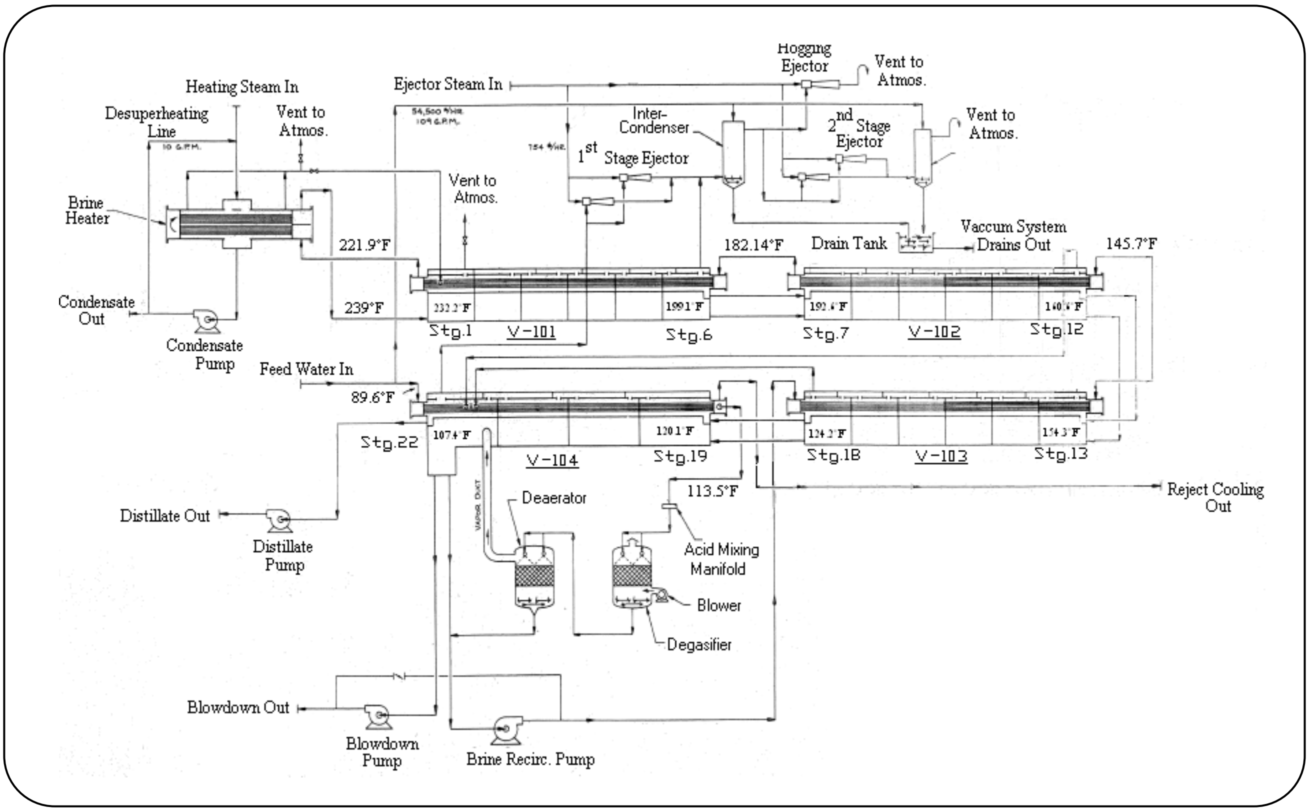

Fig. 2: Process Flow Diagram of MSFD Plant.

of each unit are provided as sub – divisions of the four main desalination vessels or evaporators, named V-101, 102, 103 and 104 [10]. As shown in the PFD, Fig. 2, river water, after clarification, brought to the plant, first passes through filters and then chlorinated before passing as cooling water through the stages of the final main vessel.

A part of this water continues to the higher temperature stages, is first dosed with sulphuric acid then degasified and deaerated, respectively. This makeup water joins recirculated water which is pumped around the system, first through distillate cooling tubes at increasing temperature, and then finally through the brine heater before joining the sump of the first stage in the first main vessel. In this stage, the water is boiled at slight positive pressure and passes from the stage at reducing pressure. Table 2 summarizes the material and heat balance of the unit before optimization and Table 3 specifies design data of existing MSFD Plant.

Cooling Water Optimization

As shown in the Fig. 3 and Table 2, in each unit, 1256.5 m^3 per hour of design rate of feed water with a temperature of 32 °C enters to stage no.22 of evaporator 104 as the last stage. About 667 m^3 per hour; i.e. 53 percent of this feed with a temperature of 45.3 °C in the heat rejection section and as once-through cooling water leaves the system. This water adsorbs the latent heat of vapors of the recirculating boiling brine in the heat rejection evaporator and leads to produce distillate water. In case of all three units in service, it reaches to 2000 m^3 per hour of clarified water.

According to design data, Gain Output Ratio which is the ratio of distillate to condensate rate is 7.5. Makeup water rate is 589.5 m^3 per hour and blowdown rate, which is the difference between makeup and distillate water, is 392.9 m^3 per hour. The cycle of concentration regarding feed water and blowdown conductivities is equal to 1.5. The rate of brine recycle water, in brine recycle pump discharge is calculated 1571.3 m^3 per hour with a temperature of 43 °C and salt concentration of 1.2 percent. This brine recycle water is sent to condenser tube of stage 18 in evaporator no. 103 as first section of heat recovery process and exits from stage 13 with an increased temperature of 63.2 °C. Then it enters to stage 12

Table 2: Material and heat balance of MSFD Plant.

BASIS h=0 & 0°C	Material balance		T(°C)	P(kPa)	C_p(kJ/kg.°C)	h(kj/kg)	Heat balance	
	ton/h-in	ton/h-out					MJ/h-in	MJ/h-out
Feed water in	1281	-	32.0	377.1	4.15	132.8	170.400	-
Heating steam in	26	-	260.0	1482.4	-	2948.0	77.370	-
Ejector steam in	0	-	260.0	1482.4	-	2948.0	1.009	-
Reject water out	-	667	45.3	101.4	4.15	187.9	-	125.513
Condensate out	-	26	121.1	377.1	4.20	508.4	-	13.346
Distillate out	-	197	41.3	377.1	4.19	173.1	-	34.070
Blow down out	-	393	41.9	101.4	4.11	172.2	-	67.070
Vaccum drains out	-	25	41.2	101.4	4.16	171.2	-	4.291
Radiation & vent loss	-	-	-	-	-	-	-	3.748
Total	1308	1308					248.779	248.779

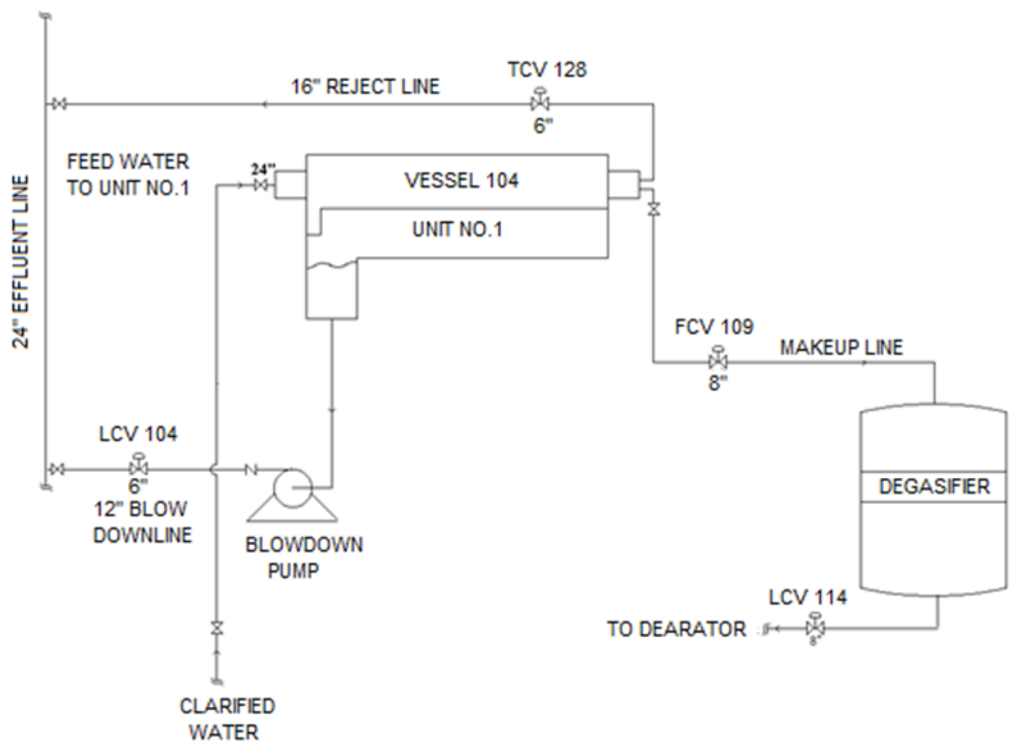

Fig. 3: Heat Rejection Section before modification.

Table 3: Design data for exist MSFD Plant [10].

unit	Abadan Refinery Water Desalination Plant
Manufacture	Aqua Chem.
Year of commissioning	1988
Type	Multi-stage flash distillation
Number of distillation units	Three
Heat –input section (Brine Heater)	
Dimensions (H, W, L), m	2.18, 1.75, 7.62
Number of passes	Two
Number of tubes	1060 per pass
Tube size, mm	19.1
Tube length, mm	6,248
Heat transfer area, m^2	782
Tube material:	70-30 Cu-Ni
Brine velocity (Max), m/s	1.91
Brine rate, kg/s	1,573,800
Heat recovery section (V-101, 102& 103)	
Dimensions (H, W, L), m	2.83, 3.02, 22.1
Number of stages	Eighteen
Number of tubes	2000
Tube size, mm	15.88
Tube length, mm	19,888
Heat transfer area, m^2	5,928
Tube material:	70-30 Cu-Ni
Brine velocity (Max), m/s	1.48
Brine rate, kg/s	1,573,800
Heat rejection section (V-104)	
Dimensions (H, W, L), m	2.83, 3.02, 22.1
Number of stages	Four
Number of tubes	1566
Tube size, mm	15.88
Tube length, mm	19,875
Heat transfer area, m^2	1,547
Tube material:	90-10 Cu-Ni
Brine velocity (Max), m/s	1.58
Brine rate, kg/s	1,283,401

Table 4: The heat rejection, recovery and input sections of exist MSFD Plant.

	Stages		T_{in}(°C)	T_{out}(°C)	Heat capacity, °C		R (m^3h)	Meat absorbed (MJ/h)
	Stage-in	Stage-out			kJ/kg.°C-in	kJ/kg.°C-out		
Vessel 103	18	13	43.2	63.2	4.18	4.18	1571.3	131614.0
Vessel 102	12	7	63.2	83.2	4.18	4.20	1571.3	133505.8
Vessel 101	6	1	83.4	105.5	4.20	4.22	1571.3	146373.7
Brine heater	-	-	105.5	115.0	4.22	4.24	1571.3	63265.5
Vessel 104	22	19	32.0	45.3	4.18	4.18	1256.8	69814.7
Total								

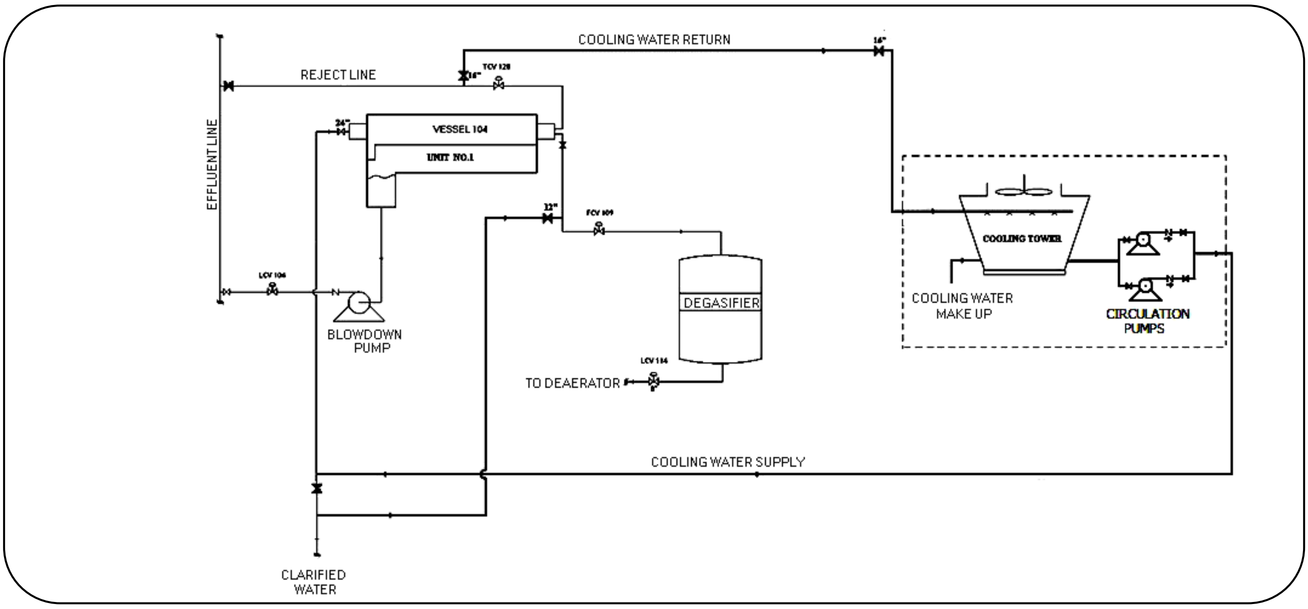

Fig. 4: Heat Rejection Section after modification.

of the second section of heat recovery in the evaporator no. 102 and after heat absorption comes out from stage 7. Here its temperature reaches 83.4 °C. In the evaporator no. 101, recycle water is conducted to condenser tube of stage 6 and at the end of this vessel exits with a temperature of 105.5 °C and is sent to brine heater. In the brine heater, by using heating media, here steam, the water temperature rises to 115 °C. To calculate heat capacity of brine water, we use specific water heat capacity C_{Pd} of the equation (1). The correlation of brine specific heat (kJ/kg K) is obtained by applying a factor dependent upon the solid concentrations and temperature to the heat capacity of pure water CP_d at the desired temperature (Sharqawy et al., 2010) [12]:

$$C_{Pd} = A + BT_b + CT_b^2 + D^3T_b^3 \quad (1)$$

Where,

$$A = 5.328 - 9.76\times10^{-2} w_s + 4.04\times10^{-4} w_s^2 \quad (2)$$

$$B = -6.913\times10^{-3} + 7.351\times10^{-4} w_s - 3.15\times10^{-6} w_s^2 \quad (3)$$

$$C = 9.6\times10^{-6} - 1.927\times10^{-6} w_s + 8.23\times10^{-9} w_s^2 \quad (4)$$

$$D = 2.5\times10^{-9} + 1.666\times10^{-9} w_s - 7.125\times10^{-12} w_s^2 \quad (5)$$

T_b is the brine temperature (273.15< T_b < 453.15 K) and w_s is the salinity of solution (0 < w_s < 180 g/kg). The amount of heat transfer in the three different sections of heat rejection; evaporator no. 104, heat recovery; evaporators no. 101, 102 and 103 and heat input; brine heater are calculated and shown in Table 4.

Fig. 4 shows the optimized MSFD plant, where required cooling water of heat rejection section is altered from once through to circulating type using the cooling tower.

As the result, for each of three exist thermal desalination plants up to 53 percent of feed water, i.e.; 667 m^3/hr and the same amount of reject water has been conserved. Nevertheless, in the new situation, the makeup water temperature drops to inlet feed water temperature. So in order to keep the plant production load and maintain its process conditions, following changes should be taken into consideration. New water temperature in the recycle pump discharge which is derived from mixing of 32°C make up water and last stage water temperature of 42°C, decrease to 38.2°C. With the assumption of the same heat loads of all evaporators, the modeling gives new water temperature in the outlet of each evaporator and even required supplementary heat to overcome makeup temperature loss in the brine heater. The results are shown in Table 5 [6].

Accordingly, recycle water, after exiting from condenser tubes of stage 1 of evaporators no. 101 with a temperature of 100.5 °C, is sent to brine heater. To reach the desired temperature of 115°C, an increase in brine heater heat load is a need. In this case, total heat transfer in brine heater is 96,115 MJ per hour. This shows that the rate of required steam increases up to 13.1 ton per hour [6].

A simple economic evaluation based on Abadan refinery local utility cost table for financial year of 2015 for the investigated modifications shows that the amount of saving from clarified water conservation is around 5,500,000 $ per year:

667 (m^3/hr) × 24 (hr/day)× 350 (day/year) × 3 units × 11,478 (Rials/$m^{3)}$ / 34500 ($/ Rials) = + 5,592,081 $ / year

In the other hand, the unit steam consumption has been risen to 13 ton/hr:

13 (ton/hr) × 24 (hr/day)× 350 (day/year) × 3 units × 215,828 (Rials/ton) / 34500 ($/ Rials)= - 2,049,427 $ / year

And we required same amount of cooling water instead of clarified water in lower price for heat rejection section:

667 (m^3/hr) × 24 (hr/day)× 350 (day/year) × 3 units × 1,777 (Rials/$m^{3)}$ / 34500 ($/ Rials) = - 865,754 $ / year

So, total saving excluded the cost of installation of new cooling tower and required piping is:

Total saving = + 5,592,081 - 2,049,427 - 865,754 = + 2,676,900 $ / year

So, it is clear that the project from point of financial aspect is economical and feasible.

Exergy Analysis model

Exergy is defined as the maximum obtainable useful work when a system is moved to equilibrium from the initial state of the environmental state (Dead State). The total exergy (E_T) of any stream is defined as:

$$E_T = E_{PH} + E_{CH} + E_{PO} + E_{KE} \quad (6)$$

where E_{PH}, E_{CH}, $E_{PO,}$ and E_{KE}, are the total physical exergy, total chemical exergy, total potential exergy and total kinetic exergy, respectively. Specific exergy is total exergy divided by the mass flow rate of the stream:

$$e_T = E_T / \dot{m} \quad (7)$$

Therefore, the specific exergy is a sum of the specific exergies of defined stream:

$$e_T = e_{PH} + e_{CH} + e_{PO} + e_{KE} \quad (8)$$

In which e_{PO} and e_{KE} are considered negligible since the stream is assumed to be at rest relative to the environment [11].

In the MSF process, the streams are pure water, seawater, and heating steam. The physical and chemical exergy of the water and seawater streams is calculated by correlations suggested and validated (with a maximum deviation of 1.5%) by Sharqawy et al. [10]. Physical exergy (e_{PH}) of the fluid stream is:

$$e_{PH} = \left(h - h_o\right) - T_o\left(s - s_o\right) \quad (9)$$

Where h_0, T_0, s_0, are the enthalpy (kJ/kg), temperature (K) and entropy in (kJ/(kg K)) of the stream at the dead state. For the water and seawater the enthalpy is given by; the constants presented in Table 5:

$$h_{sw} = h_w - w_s\left[b_1 + b_2 w_s + b_3 w_s^2 + b_4 w_s^3 + b_5 T + b_6 T^2 + b_7 T^3 + b_8 w_s T + b_9 w_s^2 T + b_{10} w_s T^2\right] \quad (10)$$

where the water enthalpy is:

$$h_w = 141.355 + 4202.070T - 0.535T^2 + 0.004T^3 \quad (11)$$

The effect of the stream pressure on the enthalpy of the stream is then added:

Table 5: The new calculated temperatures.

	Stages		T_{in}(°C)	T_{out}(°C)	Heat capacity, °C		R (m^3h)	T calculated (°C)
	Stage-in	Stage-out			kJ/kg.°C-in	kJ/kg.°C-out		
Vessel 103	18	13	38.2	58.2	4.18	4.18	1571	58.2
Vessel 102	12	7	58.2	78.4	4.18	4.19	1571	78.4
Vessel 101	6	1	78.4	100.5	4.19	4.22	1571	100.6

Table 6: Constants used to calculate the enthalpy and entropy of seawater [9].

$b_1 = -2.348\times10^4$	$b_6 = -4.417\times10^1$
$b_2 = 3.152\times10^5$	$b_7 = 2.139\times10^{-1}$
$b_3 = 2.803\times10^6$	$b_8 = -1.991\times10^4$
$b_4 = -1.446\times10^7$	$b_9 = 2.778\times10^4$
$b_5 = 7.826\times10^3$	$b_{10} = 9.728\times10^1$
$c_1 = -4.231\times10^2$	$c_6 = -1.443\times10^{-1}$
$c_2 = 1.463\times10^4$	$c_7 = 5.879\times10^{-4}$
$c_3 = -9.880\times10^4$	$c_8 = -6.111\times10^1$
$c_4 = 3.095\times10^5$	$c_9 = 8.041\times10^1$
$c_5 = 2.562\times10^1$	$c_{10} = 3.035\times10^{-1}$

$$h_{sw}(T,p,ws)=h_{sw}(T,p_o,ws)+\nu(p-p_0) \quad (12)$$

For the water and seawater, the entropy is given by;

$$s_{sw}=s_w-w_s\left[c_1+c_2w_s+c_3w_s^2+c_4w_s^3+ c_5T+c_6T^2+c_7T^3+c_8w_sT+c_9w_s^2T+c_{10}w_sT^2\right] \quad (13)$$

The constants presented in Table 6.

The pure water entropy is:

$$s_w=0.1543+15.383T-2.996\times10^{-2}T^2+ 8.193\times10^{-5}T^3-1.370\times10^{-7}T^4 \quad (14)$$

The chemical exergy of a pure water and seawater stream is produced when the stream has a salt concentration that is different from the dead state concentration. The chemical exergy is obtained by:

$$e_{CH}=\sum_{i=1}^{n}w_i\left(\mu_i^*-\mu_i^o\right) \quad (15)$$

Where μi^* and μi^o are the chemical potentials of (i) component at (T_0, P_0, ws*) and (T_0, P_0, ws_0) respectively? In the case of a mixture of pure water and seawater the chemical potential can be obtained by differentiating the Gibbs function as follows:

$$\mu_w=G_{sw}/m_w=g_{sw}-w_s\left(g_{sw}/w_s\right) \quad (16)$$

$$\mu_s=G_{sw}/m_s=g_{sw}+w_s\left(1-w_s\right)\left(g_{sw}/w_s\right) \quad (17)$$

Where g_{sw} is the specific Gibbs function at *T* (°C) given by:

$$g_{sw}=h_{sw}-(T+273.15)s_{sw} \quad (18)$$

Differentiation of the Gibbs function gives:

$$\left(g_{sw}/w_s\right)=\left(h_{sw}/w_s\right)-(T+273.15)\left(s_{sw}/w_s\right) \quad (19)$$

The partial derivatives of enthalpy and entropy with respect to the salt concentration are obtained from the following correlations (c, b constants listed in Table 5:

$$-\left(h_{sw}/w_s\right)=b_1+2b_2w_s+3b_3w_s^2+4b_4w_s^3 \quad (20)$$
$$b_5T+b_6T^2+b_7T^3+2b_8w_sT+3b_9w_s^2T+2b_{10}w_sT^2$$

$$-\left(s_{sw}/w_s\right)=c_1+2c_2w_s+3c_3w_s^2+4c_4w_s^3 \quad (21)$$
$$c_5T+c_6T^2+c_7T^3+2c_8w_sT+3c_9w_s^2T+2c_{10}w_sT^2$$

The working range of the above correlations does not cover heating of the steam. Nevertheless, the steam enthalpy and entropy can be extracted from steam tables or calculated by equations from *Badr*, *Probert,* and *O'Callaghan* (1990), from formulations by *Keenan* and *Keyes* (1955, 1969) and conveniently expressed for computer calculation (*Schnakel*, 1958). The temperature and pressure range is valid below the critical point [13].

$$h_{stm} = F + 101.3155 F_o \left\{ \frac{p}{101325.0} + \right. \quad (22)$$

$$\frac{Bo}{2}\left(\frac{p}{101325.0T}\right)^2 - B_6 +$$

$$\left. B_o\left(B_2 - B_3 - B_o B_7\left[\frac{p}{101324.0T}\right]^2\right)\right\}$$

$$S_{stm} = 1472.626 \ln T - 461.4874 \ln p + \quad (23)$$

$$0.7557174T + 3830.4065 - \frac{47845.076}{T} -$$

$$101.31344\beta$$

where;

$$B_0 = 1.89 - B_1 \quad (24)$$

$$B_1 = \frac{2641.62}{T} 10^{\left(80870/T^2\right)} \quad (25)$$

$$B_2 = 82.546 \quad (26)$$

$$B_3 = \frac{162470}{T} \quad (27)$$

$$B_4 = 0.21828T \quad (28)$$

$$B_5 = \frac{126970}{T} \quad (29)$$

$$B_6 = B_0 B_3 - 2F_0\left(B_2 - B_3\right) \quad (30)$$

$$B_7 = 2F_0\left(B_4 - B_5\right) - B_0 B_5 \quad (31)$$

$$F_0 = 1.89 B_1\left(372420/T^2 + 2\right) \quad (32)$$

$$F = 1804036.3 + 1472.265T + 0.37789824T2 \quad (33)$$

$$47845.137 \ln T$$

$$\beta = \frac{1}{T}\left\{\left(B_0 - F_0\right)\frac{P}{101325} + \right. \quad (34)$$

$$\frac{B_0}{2}\left(\frac{p}{101325}\right)^2\left(B_6 + \frac{1}{2}\left(\frac{BoP}{101325}\right)^2\right)\times$$

$$\left.\left[B_0\left(B_4 - B_5\right) - 2B_7\right] - 2B_7\right\}$$

MSF overall exergy efficiency is defined as the ratio of the minimum separation work required to the total input exergy:

$$\eta_{min} = W_{min} / E_{input} \quad (35)$$

In the previous correlations, the authors have given correlations for seawater thermophysical properties as functions of temperature and salinity, but only for near atmospheric pressures. It should be said that *Sharqawy et al.* has recently reported a reliable method of estimating the effect of pressure on seawater properties [14]. In this work, we use this method and new correlations for seawater thermophysical properties by using visual basic package software of the Massachusetts Institute of Technology [15] that are valid within these ranges:

T = 0 - 120 °C, S = 0 - 120 g/kg, and P = 0 - 12 MPa.

RESULTS AND DISCUSSION

The main purpose of this paper is to investigate thermodynamically the effects of our provided solution for one famous bottleneck against more popularity of MSFD application among the world; high rate of feed water rejection. A new opportunity in the reduction of cooling water the MSFD has been presented and executed for the first time. It is revealed that as a consequence of this optimization, two big changes have been achieved in each unit: 1. A 53 percent decrease of feed water inlet, i.e.; 667 m3/h and the same amount of reject water; 2. An increase in the unit steam consumption up to 13 ton/h, about 50 percent of the design value. This optimization results in exergy change in following main components:

- All pumps including Seawater pump, Brine recycle pump, Blow down pump, Distillate pump and Condensate pump
- Brine heater
- Evaporators including in V-101, V-102, V-103 & V-104
- Cooling process
- Product disposal
- Blowdown disposal
- Condensate disposal

The effect of the studied modifications on the performance of the ejectors, deaerators, degasifiers and etc is categorized as “other equipment”.

Fig. 5 shows a schematic of the MSF desalination unit which first in design case is studied. Table 6 shows

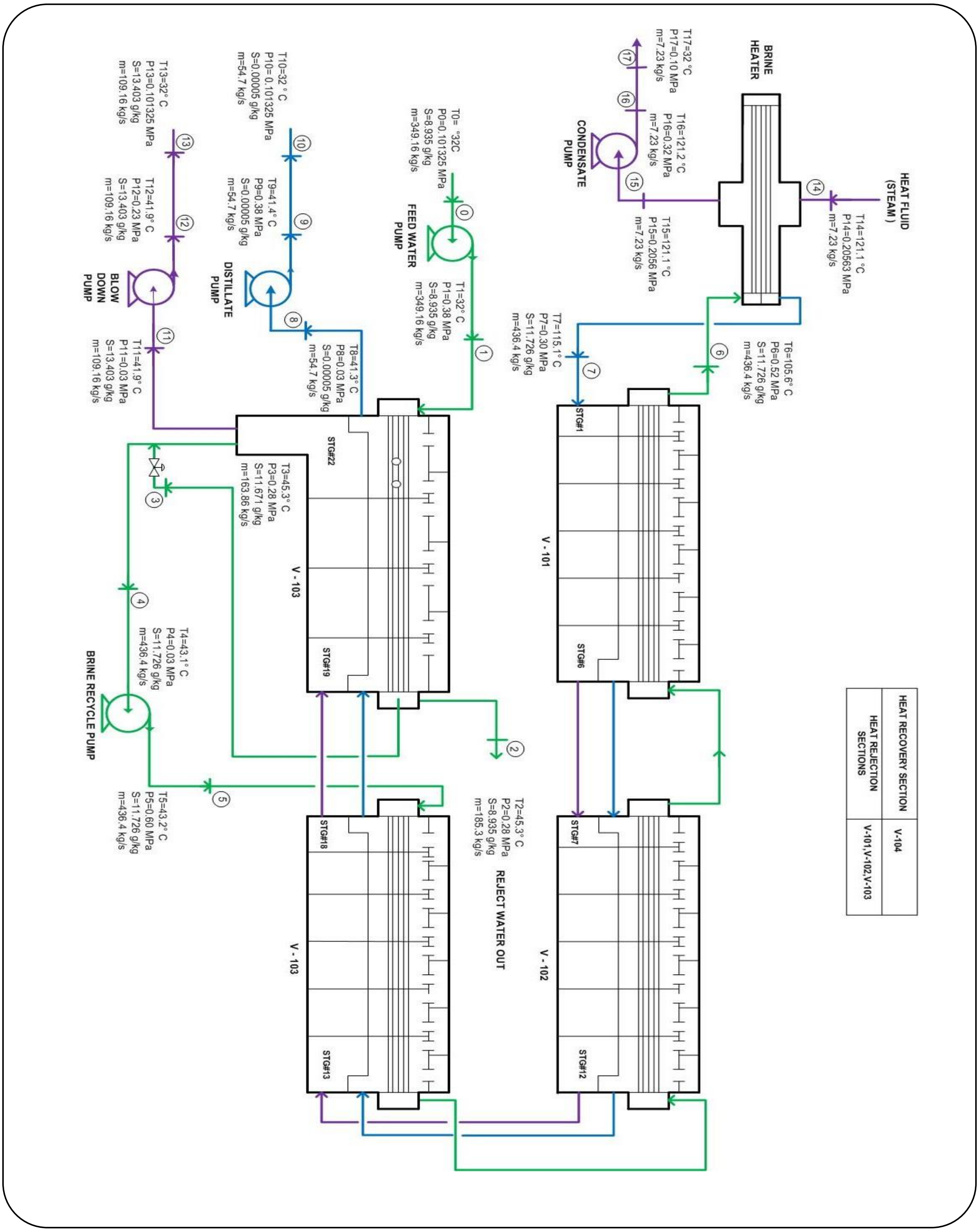

Fig. 5: Schematic of the unit studied in design case.

Table 7: Thermodynamic properties of the indicated streams in design case.

Stream	Temperature (°C)	Pressure (MPa)	Salinity w_3,(g/kg)	Mass flow (kg/s)	Specific exergy (kJ/kg)	Total exergy (kW)
0	32.0	0.101325	8.935	349.16	0.00	0.00
1	32.0	0.38	8.935	349.16	0.28	96.95
2	45.3	0.28	8.935	185.30	1.33	245.92
3	45.3	0.28	11.671	163.86	1.35	221.24
4	43.1	0.03	11.726	436.40	0.76	333.17
5	43.2	0.60	11.726	436.40	1.34	586.69
6	105.6	0.52	11.726	436.40	32.01	13970.29
7	115.1	0.30	11.726	436.40	39.83	17382.23
8	41.3	0.03	0.00005	54.70	1.16	63.36
9	41.1	0.38	0.00005	54.70	1.52	83.07
10	32.0	0.101325	0.00005	54.70	0.66	35.91
11	41.9	0.03	13.403	109.16	0.63	69.14
12	41.9	0.23	13.403	109.16	0.83	90.81
13	32.0	0.101325	0	109.16	0.07	7.83
14	121.1	0.20563	0	7.23	543.10	3926.65
15	121.1	0.20563	0	7.23	46.11	333.38
16	121.2	0.32	0	7.23	46.33	3334.96
17	32.0	0.101325	0	7.23	0.00	0.00

the calculated thermodynamic properties for all numbered streams in Fig. 4 using MIT software. Since all exit streams go from their exit state to the dead state for the purposes of exergy analysis, selection of the dead state varies according to the researcher's objectives. In this current exergy analysis, the dead state has been selected at P_0 = 101.325 kPa, $ws,_0$ = 8.935 g/kg and T_0 = 32 °C which matches the feedwater intake parameters; stream 0. The three streams including 10, 13 and 17 in Table 6 represent the residual exergy when they move to the dead state at (P_0, T_0) to calculate the minimum separation work (W_{min}).

Table 8 summarizes the exergy analysis. Eq. (32) is applied to calculate the overall exergy efficiency of the unit. The input exergy to the unit is the sum of the heating steam and pump work inputs (pump efficiency is assumed to be typically 75%). The output minimum separation work (for the exergy efficiency) is the sum of the discharge distillate and discharge blowdown relative to the entering exergy of the feed water to the unit. Exergy destruction of the unit components is evaluated by the difference between the input and output exergies of the individual components.

Consequently, Fig. 6 shows the percentage exergy destruction as a ratio of the component exergy destruction to the total exergy destruction of the MSF unit. The results show that the overall exergy efficiency of the MSF unit studied in design case is only 0.98%. More than 71% of the exergy destruction occurs in the MSF evaporators, shared between the heat recovery stages 1 to 18 of vessels 101, 102 & 103 and the heat rejection stages 19 to 20 of vessel 104. Pumps, brine heater and streams disposal to the dead state each contribute 2.98%, 4.11%, and 10.55%, respectively, to the overall exergy destruction, in general agreement with most published studies of exergy analysis for MSF desalination units.

In this case, the amount of exergy destruction in the cooling process is 5.58%.

Fig. 7 shows the schematic of the studied MSF desalination unit in optimized case. Table 9 shows

Table 8: MSF exergy analysis results in design case.

Equipment	Calculation method	Result	Unit
Seawater pump exergy in	E1-E0	96.95	kW
Brine recycle pump exergy in	E5-E4	253.52	kW
Blow down pump exergy in	E12-E11	21.67	kW
Distillate pump exergy in	E9-E8	19.70	kW
Condensate pump exergy in	E16-E15	1.57	kW
Pump input exergy in		524.56	kW
Heating steam exergy	E14	3926.65	kW
Exergy in		4451.1	kW
Minimum separation work	(E11+E8)-E0	43.74	kW
Exergy efficiency		0.98	%
Total exergy destruction		4407.46	kW
Exergy destroyed in pumps		131.14	kW
Exergy destroyed in brine heater	(E14+E6)-(E15+E7)	181.33	kW
Exergy destroyed in V-101, V-102, 103 & 104	(E1+E7)-(E2+E6+E8+E11)	3130.46	kW
Exergy destroyed in the cooling process	(E2-E0)	245.92	kW
Exergy destroyed in product	(E9-E10)	47.16	kW
Exergy destroyed in blowdown disposal	(E12-E13)	82.98	kW
Exergy destroyed in condensate disposal	(E16-E17)	334.96	kW
Exergy destroyed in other equipment		253.52	kW

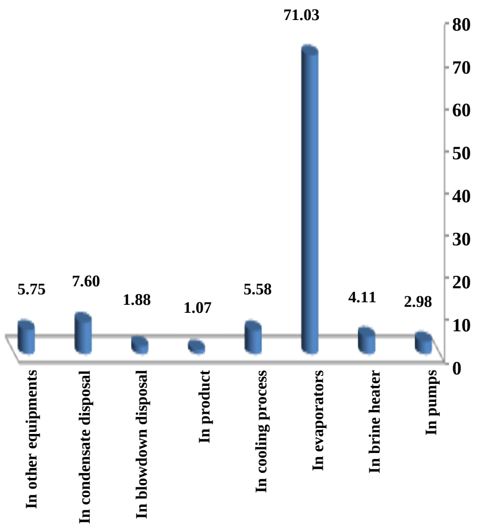

Fig. 6: Exergy destruction (%) in MSF desalination components in design case.

the calculated thermodynamic properties for all numbered streams in this case. The main difference between two cases, as mentioned in the "Cooling Water Optimization" section of this paper, can be found in flow rate stream 1 and the amount of reject water that now is recycled.

Similarly, in Table 10 the exergy analysis of the new case is presented.

The results as illustrated in Fig. 8, show that the overall exergy efficiency of the MSF unit in optimized case is reached to 0.69%, mostly due to needing to provide more heating steam. More than 74% of the exergy destruction occurs in the MSF evaporators vessels 101, 102 &103 and 104. Pumps, brine heater and streams disposal to the dead state each contribute 1.8%, 4.10%, and 10.05%, respectively, to the overall exergy destruction. The amount of exergy destruction in the cooling process is 6.96%.

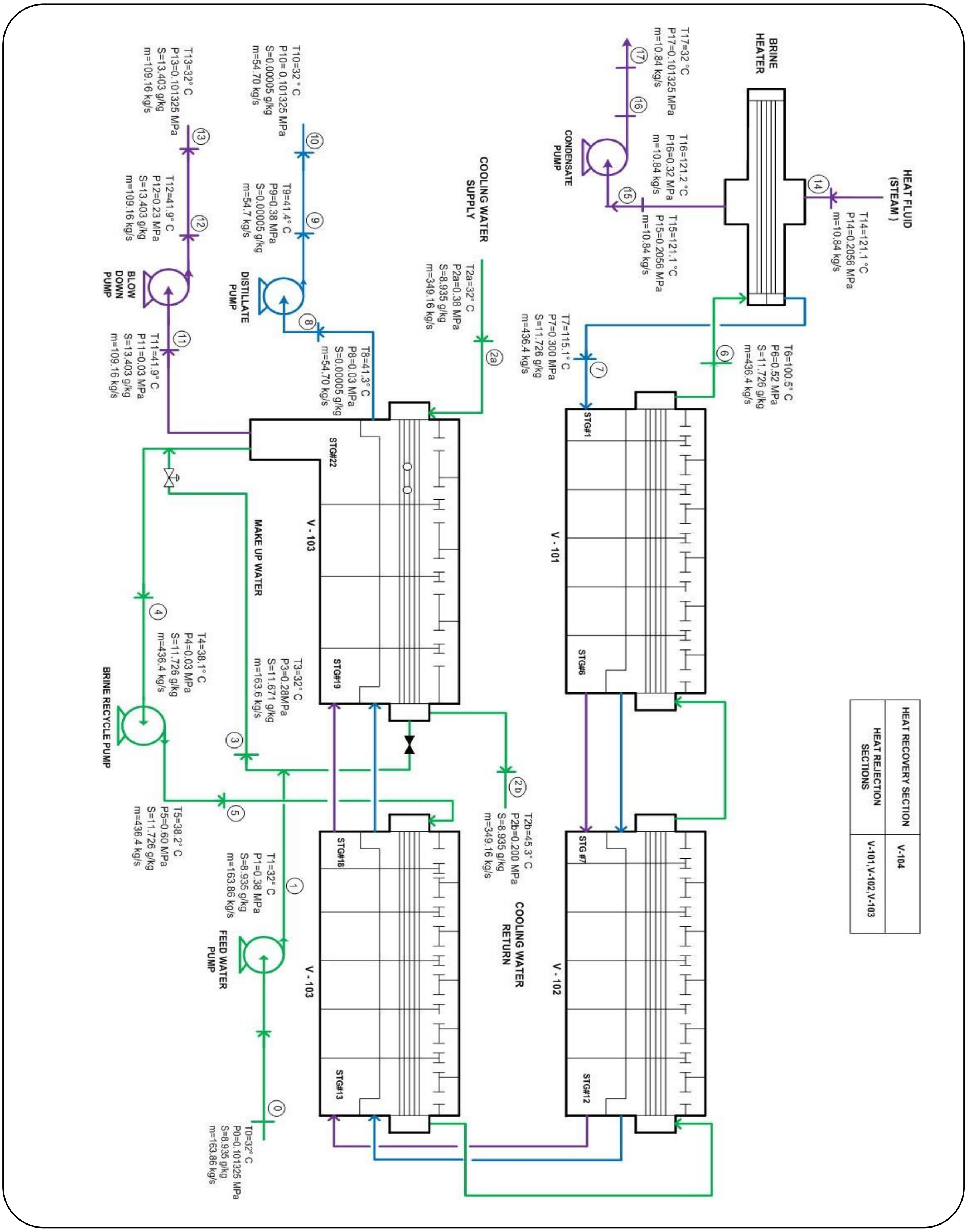

Fig.7: Schematic of the unit studied in optimized case.

Table 9: Thermodynamic properties of the indicated streams in optimized case.

Stream	Temperature (°C)	Pressure (MPa)	Salinity w_3,(g/kg)	Mass flow (kg/s)	Specific exergy (kJ/kg)	Total exergy (kW)
0	32.0	0.101325	8.935	163.86	0.00	0.00
1	32.0	0.380000	8.935	163.86	0.28	45.50
2a	32.0	0.38	8.935	349.16	0.28	96.95
2b	45.3	0.20	8.935	349.16	1.26	438.59
3	32.0	0.28	11.671	163.60	0.21	33.70
4	38.1	0.03	11.726	436.40	0.20	87.82
5	38.2	0.60	11.726	436.40	0.78	338.64
6	100.5	0.52	11.726	436.40	28.08	12252.94
7	115.1	0.30	11.726	436.40	39.83	17382.23
8	41.3	0.03	0.00005	54.70	1.16	63.36
9	41.4	0.38	0.00005	54.70	1.52	83.22
10	32.0	0.101325	0.00005	54.70	0.66	35.91
11	41.9	0.03	13.403	109.16	0.63	69.14
12	41.9	0.23	13.403	109.16	0.83	90.81
13	32.0	0.101325	13.403	109.16	0.07	7.83
14	121.1	0.20563	0	10.84	543.10	5887.25
15	121.1	0.20563	0	10.84	46.11	499.84
16	121.2	0.32	0	10.84	46.33	502.20
17	32.0	0.101325	0	10.84	0.00	0.00

So, besides the return of more than fifty percent of inlet feed water as reject water, exergy destruction is reduced more than 39% in the pumps, and 30% in blowdown and around 29% in distillate streams mainly due to the reduction in feed water pumpage. For brine heater, an enhance as small as 0.37% is also achieved. In the other hand, the rate of destruction of exergy increased around 25% in the cooling process and above 5% evaporators as a result of the unit steam consumption rise and lower makeup water temperature.

CONCLUSIONS

The variation of the exergy and flow exergy with the system and environment properties are carefully examined both for design and optimized cases with the most up-to-date seawater properties, based on the latest published validated thermodynamic properties of seawater Visual Basic package software of the Massachusetts Institute of Technology. The main issue is in the industries like petroleum refineries, which are located beside the rivers, more than 50 percent of raw water after pretreatment and clarification with coagulants and other chemicals, now named clarified water, disposes to river as reject water. Nevertheless, by means of cooling tower system, it can be recycled in this process. In case of cooling tower installation, we make the once-through cooling system to recirculating type. It has been shown that for each of three exist thermal desalination plants up to 53 percent of feed water, i.e.; 667 m^3/hr and the same amount of reject water can be conserved. On the other hand about 50 percent of design value increase in the unit steam consumption up to 13 ton/hr, is seen.

A simple economic showed that our provided solution from point of financial view is economical and feasible.

It is demonstrated clearly that besides the return of more than fifty percent of inlet feed water as reject water, the exergy destruction can be reduced in these components of unit:

Table 10: MSF exergy analysis results in the optimized case.

Equipment	Calculation method	Result	Unit
Seawater pump exergy in	E1-E0	45.50	kW
Brine recycle pump exergy in	E5-E4	250.83	kW
Blow down pump exergy in	E12-E11	21.67	kW
Distillate pump exergy in	E9-E8	19.86	kW
Condensate pump exergy in	E16-E15	2.36	kW
Pump input exergy in		453.62	kW
Heating steam exergy	E14	5887.25	kW
Exergy in		6340.87	kW
Minimum separation work	(E11+E8)-E0	43.74	kW
Exergy efficiency		0.69	%
Total exergy destruction		6297.13	kW
Exergy destroyed in pumps		113.41	kW
Exergy destroyed in brine heater	(E14+E6)-(E15+E7)	258.12	kW
Exergy destroyed in V-101, V-102, 103 & 104	(E1+E7)-(E2+E6+E8+E11)	4700.65	kW
Exergy destroyed in the cooling process	(E2-E0)	438.59	kW
Exergy destroyed in product	(E9-E10)	47.31	kW
Exergy destroyed in blowdown disposal	(E12-E13)	82.98	kW
Exergy destroyed in condensate disposal	(E16-E17)	502.20	kW
Exergy destroyed in other equipment		153.88	kW

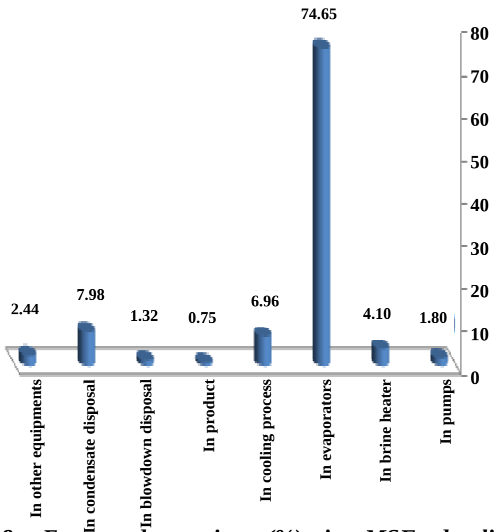

Fig. 8: Exergy destruction (%) in MSF desalination components in the optimized case.

- More than 39% in pumps
- Around 29% of distillate stream
- and only small change in brine heater; 0.37%
- and 30% in blowdown the disposal

However, it is found that the unit steam consumption has been risen to 13 ton/hr, about 50 percent of design and the destruction of exergy is increased in these components of the unit as follows:

- Around 25% in the cooling process
- Above 5% in evaporators

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Refining Company for cooperation and financially supporting the project (Ph.D. Research Contract no.7522).

Nomenclature

E	Rate of exergy flow in stream, kW
E_d	Rate of exergy destruction, kW
E_{input}	Rate of input exergy, kW
e	Specific exergy of stream, kJ/kg
G	Gibbs energy, kJ
g	Specific Gibbs energy, kJ/kg
h	Enthalpy of the stream, kJ/kg
H	Height
L	Length
P	Pressure of the stream, MPa
PFD	Process flow diagram
R	Recycle rate
s	Entropy of the stream, kJ/(kg K)
T	Temperature of the stream, °C
v	Specific volume, m^3/kg
w	Salinity of the stream, g/kg
W	Width
W_{min}	Minimum work of separation, kW
MP	Medium Pressure
MSFD	Multi Stage Flash distillation
m	Mass flow rate, kg/s
TBT	Top brine temperature, °C

Subscripts

o	Dead state
b	Brine
CH	Chemical
cw	Cooling water
d	Distillate product
f	Feed
KE	Kinetic
PH	Physical
PO	Potential
s	Salt
sw	Sea water
T	Total
w	Water

Greek

μ	Chemical potential, kJ/kg
η	Exergy efficiency, %

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