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"HUMIDIFICATION - DEHUMIDIFICATION DESALINATION PROCESS USING WASTE HEAT FROM A GAS TURBINE"

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ABSTRACT

This paper presents a humidification-dehumidification desalination process using waste heat from a gas turbine power plant. In this process, the air is used as the operating fluid instead of water. The process has many advantages over many desalination processes which use waste heat from gas turbines.

The amount of fresh water produced and the mass of air-gas mixture leaving the desalination plant were found to decrease with decreasing gas turbine load and with increasing mixing temperature. The air-gas mixture can be safely used in space air conditioning. The specific power consumption of the proposed process was found to be less than that of either single or dual Multi-Stage Flash (MSF) plants. It is greater however, than that of all types of Reverse Osmosis (RO) plants. The specific capital investment of the process is found to be less than that for any other desalination process considered. The ratio of water production to power generation in the present process is nevertheless very small.

1. INTRODUCTION:

Energy cost and capital amortization are the major cost elements in the production of fresh water from saline water resources. The purpose of the use of waste heat from any power or industrial plant in salt water desalination is the reduction of fresh water cost. Gas turbine power generation plants are widely used in many places which are suffering from shortage of fresh water supply. The average thermal efficiency of gas turbine power generation plants is low and the majority of input energy is rejected through the exhaust gas. A number of investigations [1,2,3,4,5] indicate that the use of high temperature exhaust gas to supply the required energy for different desalination plants significantly reduces the cost of fresh water. For example, Foster and Pegg [5] reported that the use of gas turbine waste heat in water desalination reduces total water production cost to approximately 30% of that for a single purpose water plant.

The main difficulties in using waste heat from a gas turbine to supply energy for conventional desalination processes is the conversion of the thermal energy of the flue gas to a suitable form for a given desalination process. In thermal desalination processes, the flue gas energy is used to produce low pressure steam. In the electrodialysis process, energy is required in the form of electrical energy. In a reverse osmosis process, the energy is required as mechanical energy. Energy conversion processes will result in an eventual increase in plant capital investment, in addition to high energy losses associated with the energy conversion process due to low efficiencies of the energy conversion units.

In the present work, the humidification-dehumidification salt water desalination process is coupled with a gas turbine power plant. In the humidification-dehumidification processes [6,7,8] the salt water is heated by solar energy or by some other means. Hot water is then sprayed counter current to relatively cold and dry air in the humidifier tower. Fresh water is obtained by cooling the saturated air-vapour mixture to a temperature below its dewpoint temperature. The major disadvantages of the process are as follows:

1. Low thermal efficienty due to high percentage of heated water which is rejected to keep the water salinity below the scale formation limit inside the tower.

2. Inability to supply relatively dry and cold air.

3. Need for a large surface area to heat water, or use of expensive solar collectors.

To overcome the above mentioned difficulties, the proposed desalination process uses air as operating fluid instead of water. This is achieved by directly heating the available atmospheric air when mixed with the flue gas from the gas turbine power plant.

2. PROCESS DESCRIPTION:

The proposed humidification-dehumidification desalination process operating on waste heat from a gas turbine power plant is shown as a flow diagram in Figure (1), and is represented on the psychrometric chart in Figure (2). The flue gas is sprayed in wet scrubber using lime, sodium carbonate and diethanolamine solution as scrubbing media. In this process the main pollutant components such as sulfur and carbon oxides are removed. The removal efficiency of such a wet scrubbing system can be as high as 90% to 95% with flue gas containing up to 5000 ppm SO_2 [9]. The remaining traces of the oxides are condensed in the humidification units which operate at temperatures below the dew points of these oxides. Other pollutants such as nitrogen oxides are reduced in modern gas turbine power plants by modifying the rate and location of combustion air admission [10]. The amount of ash in fuel oil is very small and is usually a problem primarily inside the combustion chamber of the gas turbine. It is worth mentioning that the removal of air pollutant components by wet scrubbing from the flue gas is used in some power plants and is widely recommended in many places [11]. This measure is taken to protect the environment from harmful effluents of power plants and to satisfy some governments presable legal standards for ambient air quality. Moreover, the use of wet scrubbing process will reduce the cost of using the high stack necessary for controlling atmospheric air pollution. For these reasons the cost of the wet scrubbing process should be added to power cost.

A calculated amount of atmospheric air, depending on gas turbine load and mixing temperature, is admitted to the bottom of packed bed cooling tower. At the same time fresh water is sprayed from top of tower. In the cooling tower, water is cooled by furnishing part of the latent heat required to vapourize some of the water into the air stream. The cooling tower process is represented by a constant wet bulb temperature line on the psychrometric chart (line 1−2). The saturated air-vapour mixture leaving the cooling tower is mixed adiabatically with the controlled amount of flue gas nearly free of pollutant components. The mass of flue gas depends on the quantity of humid air and required maximum operating temperature in the humidification units. The absolute humidity of the gas mixture is a function of humid air and flue gas masses, humidity of air, and water content of flue gas. The process is represented on the psychrometric chart by approximately a horizontal line (line 2−3).

The gas mixture of high temperature is introduced to the bottom of the first humidifier where it intimately contacts the recirculated salt water sprayed at the top of tower. In the humidification process, the water temperature is constant and is equal to the adiabatic saturation temperature. The gas mixture is cooled and humidified, following the path of the adiabatic saturation line on the psychrometric chart. This line passes through the entering gas conditions. The deviation from the adiabatic process resulting from the variation in the make up water temperature can be ignored, because this water mass is small compared to the total circulated water inside the tower. The rejection of a small amount of water is necessary to keep the concentration of salts in the recirculated water within a controlled limit to prevent the formation of scale inside the humidification units. The saturated gas mixture is heated by mixing it with another portion of flue gas and humidified in the next humidification unit. The whole process is repeated in a series of humidifiers until all flue gas is consumed.

The humidified warm water vapour-gas mixture is then dehumidified as it comes in contact with the cold and fresh water coming from the cooling tower. In the dehumidifier, water vapour condenses from the gas phase and both sensible and latent heats are transferred to the liquid phase. The gas phases is cooled and the liquid water is warmed. The process is represented on the psychrometric chart by the 100% saturation curve.

Fresh water is obtained from the bottom of the dehumidification unit. The air-flue gas mixture coming out from the top of the dehumidifier has a lower dry bulb temperature than that of the atmospheric air. The gas mixture is then mixed with dry air and can then be used in space air conditioning. This mixture is clean because all pollutant components are removed in the wet scrubbing process.

The proposed desalination process has many advantages over any other desalination process which uses waste heat from a gas turbine power plant.

1. The use of direct contact in all units eliminates the need for expensive heat transfer surfaces. This will result in an appreciable reduction in equipment capital investment, since the cost of heat transfer surfaces

represents about 40% to 60% of total capital costs of the thermal desalination plants [12].

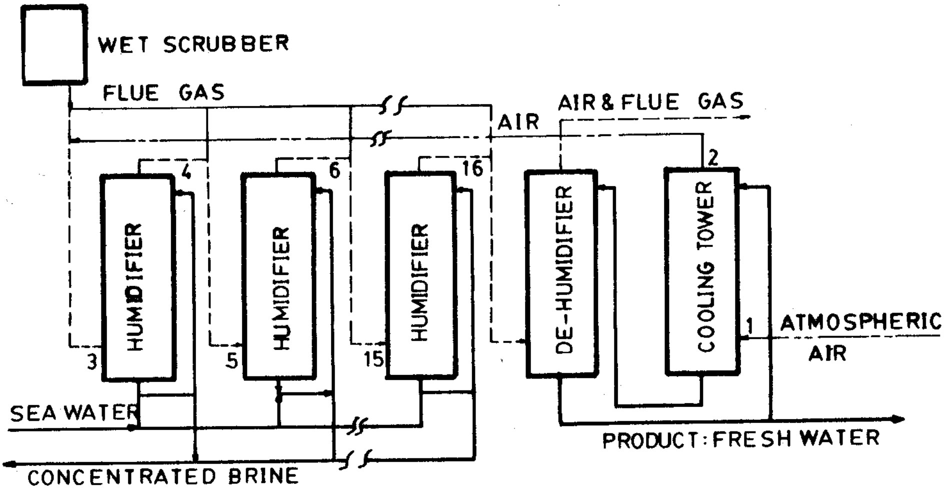

Fig (1) – Process flow sheet.

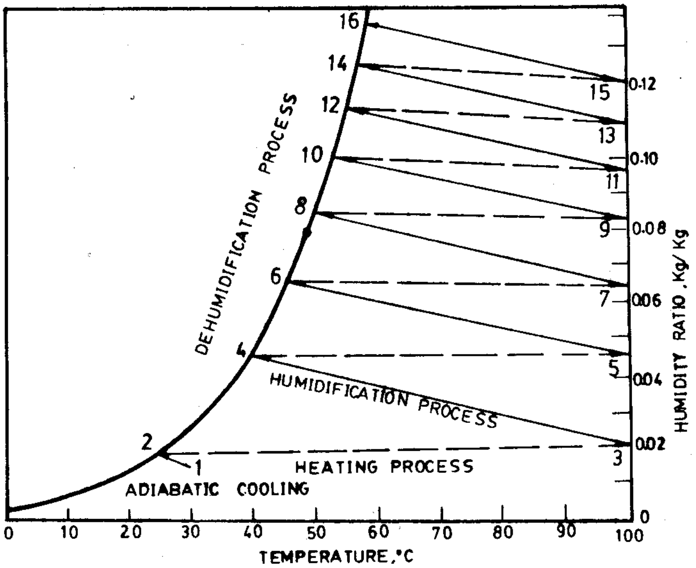

Fig (2) – Representation of Process on the Psychrometry Chart.

2. No feed water deaeration or chemical pretreatment equipment is required, since there is no corrosion of metallic heat transfer surfaces or fouling of membranes. This will reduce fixed and operating costs of desalination plants. The cost of sea water chemical pretreatment is about 5% to 12% of the water production cost in conventional desalination processes [13].
3. The process is simple and requires no auxiliaries such as steam jet ejectors, external decarbonators, vacuum deaerators, acid storage tanks, and caustic antiform systems.
4. The unit can be locally produced, easily and economically. Its operation would not call for any foreign market or expert training. This point is very important since most desalination plants are usually built in deserts where trained personnel are scarce.
5. The energy of flue gas is used as it is without any conversion.
6. It renders flexibility in load variation of the gas turbine power plant. This is achieved by controlling the amount of atmospheric air introduced to the cooling tower.
7. The relatively cool gas mixture after mixing with a controlled amount of dry air to reduce its relative humidity can be used in space air conditioning which is advantageous for most water short areas which also have hot climates.

3. CASE STUDY

The proposed process is assumed to be coupled with a 5 Mwatt type-3 Sulzer gas turbine power plant. This low capicity gas turbine is very suitable for use in places which suffer from water shortage. Table (1) shows the turbine flue gas temperature and mass flow rate as a function of the power plant load.

Table (1)
Temperature and mass of flue gas as a function of gas turbine load.

Load %	100 %	92 %	75 %	70 %	50 %	40 %	25 %	21 %
M_g (Kg/s)	26.9	27.13	27.13	27.18	27.18	27.19	27.20	27.20
T_g (°k)	818.2	788	733.7	715.5	658.2	629.2	577.3	565.9

Assume this type of gas turbine is operating on a liquid hydrocarbon fuel such as kerosene with an average molecular formula of $C_{12}H_{28}$ with a specific air to fuel ratio of 50. The composition of the fuel gas was calculated to be: 5.15 w% CO_2 : 2.29 w% H_2O : 17.12 w%O_2: and 75.47 w% N_2. The flue gas mean specific heat at constant pressure was calculated as a function of the temperature using available data [14].

$$C_p = 0.9394 + 0.212 \times 10^{-3} T_g \tag{1}$$

The following procedure is used to calculate the main process parameters. The mass flow rate of dry air (M_d) introduced to the cooling tower is initially assumed:

$$M_d = 1 \tag{2}$$

The state of atmospheric air (point 1) is located on the psychrometric chart according to its dry bulb temperature and relative humidity. The mass of air-vapour mixture (M_a) admitted to the cooling tower is given by:

$$M_a = 1 + w_1 \tag{3}$$

in which w_1 is the absolute humidity of the atmospheric air.

The state of air-vapour mixture leaving the cooling tower (point 2) is determined by drawing an adiabatic line, or constant wet bulb temperature line (line 1–2). Temperatures of air and water leaving the cooling tower are considered to be identical and equal to the wet bulb temperature of the inlet air.

The amount of water (M_1) picked up by l kg of dry air in the cooling tower is given by:

$$M_{11} = w_2 - w_1 \tag{4}$$

and the mass of air-vapour mixture leaving the tower (M_{a2}) by:

$$M_{a2} = w_2 + 1 \tag{5}$$

The mass of flue gas (M_g3) required to raise the temperature of gas mixture to the maximum operating temperature (T_m) depends on gas turbine load, required maximum temperature T_m, and air-vapour mixture temperature (T_a). This mass can be calculated from the following relation:

$$M_{g3} = (w_2 + 1)\ C_{pa}(T_m - T_a)/C_{pg}(T_g - T_m) \tag{6}$$

The mass of gas mixture (M_{a3}) introduced to the first humidifier is given by:

$$M_{a3} = (w_2 + 1) + M_{g3} \tag{7}$$

and the mass of dry gas mixture introduced to humidifier by:

$$M_{d3} = 1 + 0.9771\ M_{g3} \tag{8}$$

The mass of water vapour (M_{13}) in the gas mixture would read:

$$M_{13} = w_2 + 0.0229\ M_{g3} \tag{9}$$

The absolute humidity of the gas mixture (W_3) is calculated from the relation:

$$w_3 = M_{13} / M_{d3} \tag{10}$$

$$= (w_2 + 0.0229\ M_{g3})/(1 + 0.9771\ M_{g3})$$

For other humidification towers, same parameters can be evaluated by same procedures.

For any humidifier, the amount of make up water required (M_f) is determined from the relation:

$$M_f = M_e + M_r \tag{11}$$

in which M_e is the mass of water evaporated in the humidifier, viz:

$$M_e = M_d (w_o - w_i) \tag{12}$$

and the mass of rejected concentrated water M_r is related to the make up water M_f through the salt balance inside the unit.

$$X_f M_f = X_e M_e + X_r M_r \tag{13}$$

in which X_f, X_e and X_r are respective salt concentration in the make up water, in the evaporated water, and in the rejected brine. Since X_e is assumed to be equal to zero,

$$M_f = M_r (X_r / X_f) \tag{14}$$

The amount of recirculated water (M_c) is related to the amount of gas flowing in the tower by the relation:

$$M_c = M_a R \tag{15}$$

R being the mass flow rate ratio of liquid to gas flowing in the tower: it depends on tower design.

The mass of fresh water (M_p) condensed in the dehumidifier is given by:

$$M_p = M_{a16} (w_{16} - w_2) \tag{16}$$

The above calculations are based on 1 kg of dry air. The actual masses of the flowing fluids in the towers can be calculated by determining a multiplying factor (L) viz:

$$L = M_g \,/\, \Sigma\, M_g \qquad (17)$$

M_g being the total mass of flue gas produced from the gas turbine; it is nearly constant for different loads, Table (1) and Σ Mg is the total mass of the flue gas used per 1 kg

4. PROCESS DESIGN OF MAIN COMPONENTS

The desalination unit is composed of three main components; the cooling tower, the dehumidifier, and the humidification units. In the present calculations, each of the above components is regarded as a packed tower.

In order to carry out a process design for any packed tower, the following parameters must be determined: (1) the operating velocity through the tower; (2) the depth and type of packing; (3) the rate of water circulated through the tower; and (4) the pressure drop of the gas side.

The limiting gas velocity in a packed tower is set by the flooding point condition, where liguid begins to accumulate or backs up at any level in the packing. The flooding point was reported to occur at 3.7 kg/s per m^2 air flow rate when the water flow rate is 4 kg/s.m^2 and the tower is packed with 2.54 mm chemical porcelain Rashig rings [14]. The design value for allowable velocity is usually estimated to be 0.7 of the flooding velocity [15]. The diameter (D) of the towers is given by:

$$D = \left[\frac{4 \times \text{gas flow rate}}{11 \times 0.7 \times \text{gas flux at flooding point}}\right]^{\frac{1}{2}}$$

$$= 0.7\, M_a^{\frac{1}{2}}$$

and the mass of circulated water M_c by:

$$M_c = \frac{\text{allowable water flow rate}}{0.7 \times \text{gas flux at flooding point}} \qquad (19)$$

$$= 1.544\, M_a$$

The pressure drop per meter of packing height was reported to be 0.816 KN/m^2 for a tower under these conditions [16]. Conventional methods for calculating the pressure drop are used to calculate the pressure drop due to the flow of gases and water in connecting pipes.

To overcome the effect of gas side pressure drop on the gas turbine power plant output, fans would have to be installed. The power consumption of the fans was calculated from the following relation [16]:

$$P = M_g \, \Delta P_g \;/\; \rho_g \tag{20}$$

power consumption of pumps circulating water in the towers is given as [17] by

$$P = M_l \, \Delta P_l \;/\; \rho_l \tag{21}$$

These pumps are necessary to circulate brine in the humidification units, remove concentrated brine from the towers, pump fresh water out of the dehumidification unit and circulate fresh water between cooling tower and dehumidification unit.

The packed bed tower, required for a given separation, can be obtained from either one of the following equations.

$$Z = (HTU)_g \; (NTU)_g \tag{22}$$

$$Z = (HTU)_l \; (NTU)_l \tag{23}$$

in which Z is the height of packing material in the tower, and (HTU) and (NTU) are the respective height and number of the transfer units. The choice of the appropriate equation depends upon the form in which the mass transfer coefficient is introduced.

The air-water system is a vapour phase controlling system [14]. The $(HTU)_g$ for tower packed with 25.4 mm chemical porcelain Raschig rings, gas flow rate of 2.59 kg/s.m^2, and liguid flow rate of 4 kg/s.m^2 is given as 0.35 m, [16]. The number of transfer units $(NTU)_g$, for the cooling tower is defined from the following relation [16]:

$$(NTU)_g = C_p \int_{T_{lo}}^{T_{li}} dt/(H^{*}-H), \tag{24}$$

T_{li} and T_{lo} being the water outlet and inlet temperatures respectitively. H* the enthalpy of saturated air vapour mixture in equilibrium with the bulk liquid, and H the enthalpy of the air vapour mixture.

In the present work, the above equation was integrated numerically using the Tschebycheff integration method, [14].

The number of transfer units $(NTU)_g$ for the humidification units and the dehumidification unit are given respectively by the following relations [16].

$$\text{Humid} \quad (NTU)_g = Ln\ (w_o - w^*)/(w_i - w^*) \tag{25}$$

$$\text{Dehumid} \quad (NTU)_g = Ln\ (w^* - w_i)/(w^* - w_o) \tag{26}$$

in which w* is the saturated absolute humidity at the adiabatic saturation temperature, and w_i are the absolute humidity of the gas at the tower inlet and outlet respectively.

5. DISCUSSION OF RESULTS

Results of the above mentioned calculations are based upon the following assumptions:

- Atmospheric air dry bulb temperature = 303 °K
- Atmospheric air relative humidity = 60%
- Number of humidification units = 7
- Number of cooling tower units = 1
- Number of dehumidification units = 1
- Mixing temperature T_m = 353, 363, 373 °K
- Feed water salinity X_f = 35000 ppm
- Rejected brine salinity X_r = 70000 ppm
- Salinity of product water X_e = zero ppm
- Length of the pipes and ducts = 4 Z
- Diameter of the pipes for water flow = 150 mm
- Diameter of the ducts for gas flow = 1.5 m
- Efficiency of pumps and fans = 80%
- Pipes and ducts are made of galvanised iron with absolute roughness = 0.15 mm [14].

- The performance ration of the Multi Stage Flash (MSF) plant }
- All dual purpose plants are coupled with a gas turbine.
- The capital cost includes only main equipment cost.
- All cost data are based on prevailing prices in 1988.

The production rate of fresh water from the desalination unit at different gas turbine loads and at different mixing temperatures is plotted in Figure (3). At any given mixing temperature, the fresh water production rate was found to decrease linearly with decreasing gas turbine load. This is interpreted as being due to the decrease in the exhaust gas temperature upon load reduction, as evident from Table (1). The dependence of flue gas mass flow rate on gas turbine load is very weak, and the product C_pM_g varies very little with load change. Accordingly, the required amount of air for mixing with the flue gas to attain the required mixing temperature will decrease as the gas turbine load is reduced. This means that, the amount of water diffused to this mixture and condensed from it in the dehumidifier, yielding the fresh water, is strongly a function of the flue gas temperature which in turn, depends mainly on the gas turbine power output.

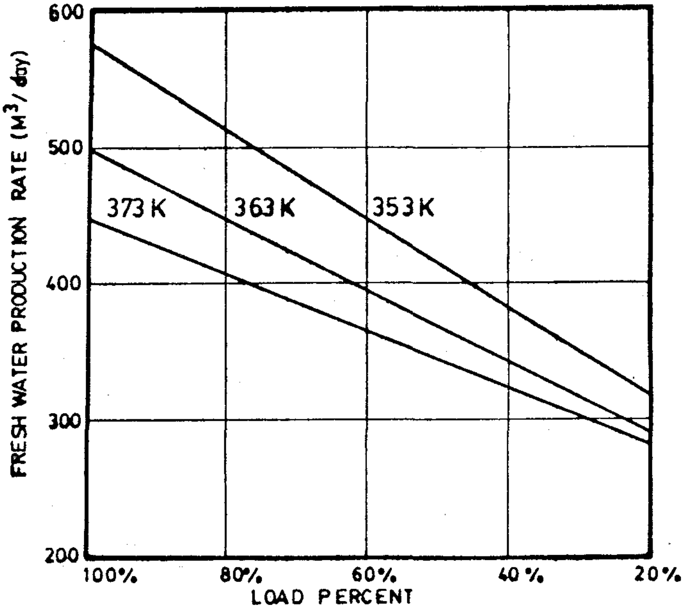

Fig (3) – **Fresh Water Production Rate at Different Values of Load and Mixing Temperature.**

Figure (3) also shows that, at fixed load, the rate of fresh water production rises as the mixing temperature decreases. Decreasing mixing temperature will increase the amount of air required for mixing with the flue gas which will increase the total mass of gas flowing inside the unit. Since the amount of water evaporated per unit mass of gas mixture depends on inlet and outlet conditions of the dehumidifier, which is independent of the mixing temperature, the eventual increase in gas mass flow rate will result in an increase in the rate of fresh water production.

Figure (4) displays the variation of the amount of air-gas mixture leaving at different mixing temperatures. The amount of gas mixture was found to decrease with decreasing gas turbine load and with increasing mixing temperature. This behaviour is also due to the dependence of the amount of air-gas mixture on load as well as on mixing temperature, as pointed out in the above.

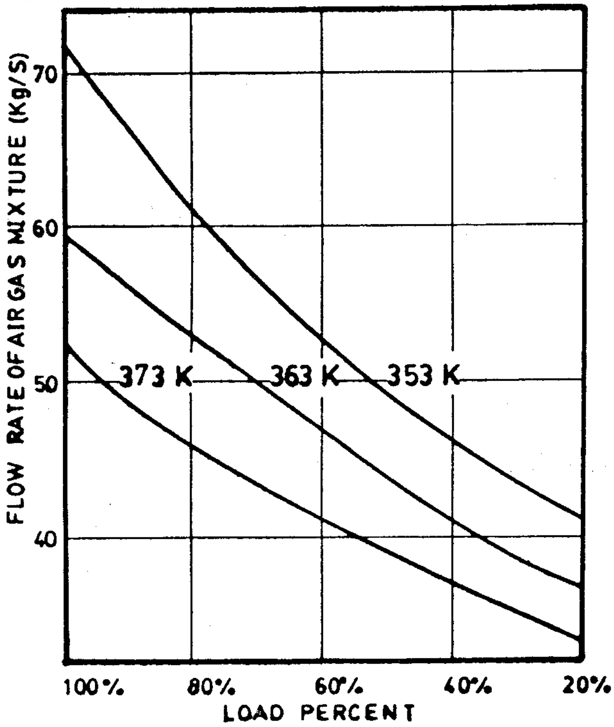

Fig (4) – Flow Rate of air gas Mixture at Different Values of Load and Mixing Temperature.

The air-gas mixture can be safely used in space air conditioning, since it is mainly air and is free of all pollutant constituents. Moreover, it has a lower dry bulb temperature than that of the atmospheric air, Figure (2). This air-gas mixture is considered free of all polluting constituents, since the majority of these constituents is removed in the wet scrubbing unit and the remaining traces are condensed in the desalination plant units operating at a lower temperature than the dew point of the traces.

Table (2) gives main design conditions for the different desalination plant units at maxixmum gas turbine load and a mixing temperature T_m of 100°C. The cost given is only the cost of main units (excluding auxiliaries), and is based on prevailing prices in 1979 [18] and updated by the use of the January 1988 equipment cost index [19].

Table (2)
Main design parameters of proposed process.

Unit	W_i	W_o	M_g	M_c	P_c	P_l	NTU	D	Z	Cost $
C.T. 1/2	0.017	0.02	21.8	32.7	11.8	0.52	1.83	3.2	0.64	3600
Hum. (1)3/4	0.02	0.05	25.1	37.6	25.1	1.1	3.36	3.4	1.18	5080
Hum. (2)5/6	0.046	0.07	28.3	42.5	26.9	1.18	3.19	3.7	1.16	4979
Hum. (3)7/8	0.067	0.09	31.8	47.7	30.2	1.13	3.19	3.8	1.16	5138
Hum.(4)9/10	0.082	0.10	35.1	53.1	32.9	1.46	3.13	4.1	1.1	6300
Hum(5)11/12	0.096	0.12	39.6	59.4	36.7	1.64	3.12	4.3	1.1	6610
Hum(6)13/14	0.110	0.13	48.5	72.7	44.4	2.0	3.09	4.8	1.1	7776
Hum(7)15/16	0.148	0.14	53.9	80.9	49.1	2.26	3.05	5.0	1.1	9245
Dehum. 16/2	0.144	0.02	53.9	80.9	66.1	3.04	4.13	5.0	1.5	12442

Table (3) provides the important bases for comparing the proposed desalination process with other processes. These bases are the specific capital cost (SC), the specific power consumption (SP), and the ratio of water production to power plant output (SR).

Processes used in the present comparison comprise: single and dual Multi Stage Flash (MSF), single and dual Reverse Osmosis (RO) without energy recovery, and single and dual Reverse Osmosis (RO) with energy recovery. The cost includes only the equipment capital investment. This cost data is based on 1987 prices [20, 21, 22, 23, 24] on a performance ratio of the Multi Stage Flash plant of 7, and on feed water for Reverse Osmosis plant of 43000 ppm. All dual purpose plants assume gas turbines.

Table (3)
Comparison of proposed process with other processes.

Parameter	**specific cost (SC)**	**specific power (SP)**	**water to power ratio (SR)**
process	**$\$/m^3$ day**	**KJ/kg**	**kg/KW.hr.**
Humidification/Dehumidification	237.	61.26	3.972
Single MSF	1451.18	294.21	—
Dual MSF	3647.4	96.519	23.919
Single RO without energy recovery	1022.8	50.420	—
Single RO with energy recovery	1127.4	33.619	—
Dual RO without energy recovery	1906.3	36.20	36.819
Dual RO with energy recovery	2225.8	27.20	49.19

6. CONCLUSION

The following conclusions may be drawn:

- The rate of fresh water production and mass of air-gas mixture increase with increase of gas turbine load and decreasing mixing temperature.
- The specific capital cost of the present process is much less than that of Multi Stage Flash and Reverse Osmosis desalination processes.
- The specific power consumption of this process is less than that of MSF plants but is greater than that of RO.
- The ratio of fresh water production to power generated, in the present process, is very small compared to other processes.

NOMENCLATURE

C_p	Specific heat at constant pressure, KJ/kg. ^{o}K
D	Diameter of the packed bed tower, m
H	Enthalpy of air-water vapour mixture, KG/Kg.
H^*	Enthalpy of a saturated air-vapour mixture in equilibrium with the bulk liguid, KJ/Kg.
HTU	Height of a transfer unit, m
L	Constant
M	Mass flow rate, Kg/s.
NTU	Number of transfer units.
P	Pressure drop, KN/m^2
P	Power required to overcome the pressure drop, KW.
R	Ratio of liquid to gas mass flow rate, kg/Kg.
SC	Specific capital investment, \$/m^3 of fresh water per day.
SP	Specific power consumption, KJ/Kg of fresh water.
t	Temperature, ^{o}K.
W	Absolute humidity, Kg of water vapour/Kg of ddry air.
X	Salt concentration, ppm.
Z	Packed bed with tower height, m.
	Efficiency.

Subscript

a	Air-vapour mixture
d	Dry air
e	Evaporated
f	Feed
g	Gas
i	Inlet
l	Liquid
m	Mixing
o	Outlet
r	Rejected

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