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Oil Shale Utilization in Seawater Distillation and Electricity Generation

استخدام الصخر الزيتي لتقطير مياه البحر و توليد الكهرباء

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خلاصة

تمت دراسة جدوى حرق الصخر الزيتي في حارقة تمييع لتوليد الكهرباء وتقطير مياه البحر. يتم توليد طاقه بقدره 12.55 ميجاوات عندما يحرق 10.09 كغم \ث من الصخر الزيتي مع 25.93 كغم \ث من الهواء على درجة حرارة حوالي 500°م في حارقه تمييع بقطر 5م. استخدم حوالي 3.45 ميجاوات من الطاقة لتشغيل ضاغطة تزويد الهواء الى الحارقة و بسرعة 28.4 م \ث، و 0.6 ميجاوات لتزويد مضخة المحطة الحرارية بالطاقة اضافة الى مضخات دفع الماء المختلفة في النظام بما فيها مضخات مبخر التقطير المتعدد المراحل، و عليه يبقى حوالي 8.5 ميجاوات لتوليد الطاقة الكهربائية (حوالي واحد بالمائة من استهلاك الأردن من الطاقة الكهربائية). ينتج ايضا 30.3 كغم \ث ماء مقطرا عند تمرير 350 كغم \ث من مياه البحر (60 بالمائة منها معاد تدويره) عبر مبخر التقطير ذو العشرة مراحل.

Abstract

A feasibility study and investigation of a pilot plant for power generation and seawater distillation utilizing the energy generated by the combustion of oil shale in a fluidized bed is carried out. About 10.09 kg/s of oil shale is combusted by about 25.93 kg/s of air at a temperature of around 500 °C in the fluidized bed of 5 m in diameter to generate power of about 12.55 MW. About 3.45 MW of that power is used to run the feed air compressor to the combustion bed at a velocity of around 28.4 m/s. Although about 0.18 MW is estimated to be consumed by the power plant water pump, a total of 0.6 MW is assumed to accommodate the running power required by all the pumps for circulating the water in the power plant and the multistage flash evaporator. This leaves about 8.5 MW (about one percent of current total Jordanian electricity demand) to generate electricity.

A ten-stage flash evaporator makes use of the energy in the power plant condenser to heat up the brine water to produce about 30.3 kg/s (about 109 m^3/h) of distillate water when 350 kg/s of seawater part of which 60% re-circulated brine water is forced through the stages of the flash evaporator.

Introduction

Water scarcity is a critical problem since the increase in the global population and the exploding water needs for industry. This has placed a tremendous strain on a limited water supply. The situation in Jordan is even more critical. The need for water is increasing at high rates mainly because of the high rate of population growth, the rise in water consumption per capital and the growing demand for water to meet the needs of socio-economic growth for a better quality of life. Jordan does not possess world wide known scale rivers. Jordan, Yarmouk and Zerka rivers are very small water resources. Other surface water resources in Wadis like Karak, Mujib, Hasa, Yabis and El-Arab, and ground water are scarce and saline. Several waste water treatment plants were constructed to treat sewage from major cities in Jordan, but this source of water is not sufficient to meet the agriculture demand in addition to its not amenable for human usage. Therefore, to meet the

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water demand in Jordan, we turn our eyes towards those large bodies of water in the Red and Dead seas. Lots of research was performed on sea water distillation plants run with petroleum or nuclear fuels, which concluded that it was very expensive to run due to the lack of both fuels in Jordan. However, Jordan possesses a very large resource in its vast reserves of oil shale amounting to about 40 billion tones.

Oil shale is one of the most important prospective sources of energy in the world that is found intimately mixed with either large proportion of sand and sedimentary rock. The compacted mineral matter consists of carbonates, sand and clay associated with an organic material. Therefore a feasibility study of utilizing oil shale for concurrent seawater distillation and electricity generation was conducted. The goal of this work was, therefore, to propose a complete system incorporating a shale oil fluidized bed burner providing the heat to a multi-stage flash-evaporator distillation plant of seawater, and to a steam power plant to generate electricity for the complete system and general use.

Oil shale reserve in Jordan

Oil shale is one of the most important prospective energy sources in the world. Reserves of energy accumulated in oil shale are 2.5 fold more than those of coal and 30 fold more than those of petroleum. Large oil shale deposits are situated in almost every country and continent. There are in fact about 190 oil shale deposits in 62 countries, and about 17 trillion bbl (oil equivalent) of oil shale resources exist which produce more than 104 L/t of shale oil. Oil shale is a fine-grained, impervious sedimentary rock containing an organic matter, a solid hydrocarbon, kerogen, which is the first stage of the formation of oil and gas deposits. When heated to a temperature of 490°-530°C the kerogen decomposes and releases hydrocarbons in the form of vapor, which when cooled, becomes liquid oil and gas.

Oil shale, in Jordan, is a potential source of energy, abundant and wide spread. Major occurrences are divided into two groups according to depth of burial; near surface deposits (about 30m below the surface) that are exploitable by open-pit mining, and, deep subsurface deposits (present at a maximum depth of 784m) that are exploitable by underground mining or in-situ retorting techniques. Jordanian oil shale is of high quality, since they are kerogen-rich bituminous argillaceous limestone, the origin of which is the fossil remains of plants and animals that accumulated in prehistoric seas and lakes that covered most of Jordan some 80 million years ago (Abu-Ajamieh, 1989).

Jordan possesses a very large resource in reserves of oil shale. The shallow near-surface deposits of oil shale in Jordan amount to 40 billion metric tons that are estimated to contain about 4 billion metric tons (30 bbl) of recoverable oil. This amount of oil would supply Jordan's energy requirements (at the present rate of consumption, equivalent to about 400 million US$ worth yearly of petroleum oil) for approximately a thousand years. In addition the by-products of oil shale are sulfur, ammonia, building stones, carbon dioxide, and carbon black. Other recoverable substances that are present only in trace amounts are rare earth elements, metals (e.g. copper and nickel), vanadium, and uranium. Moreover, shale oil can be used in the production of synthetic fibers and polymers.

There are 17 known surface and near surface oil shale deposits in Jordan. Six of those deposits have been extensively studied (Abu-Ajamieh et. al., 1988), but three of these are regarded as commercially attractive, El-Lajjun, Sultani and Jurf Ed-Darawish, about 121 km, 100 km and 115 km, respectively away from Amman. These three deposits are located south of Amman and are easily accessible from the desert highway and substantial part of the infrastructure required for development is already in place. Those deposits are regarded as the richest organic bituminous marl and limestone which occur at shallow depth and are homogeneous and uniform in character with an oil yield of 10 -18%, and the heating value ranges between 1000-2000 kcal/kg.

Since 1985, the government of Jordan has been investigating the possibility of economically exploiting oil shale for direct combustion for power generation using circulating fluidized bed technology. Direct burning tests have been carried out in different international research centers to determine the suitability of the Jordanian oil shale as fluidized bed's fuel. Combustion efficiencies exceeded 98% for all tests and CO emissions were very low throughout the tests. Gaseous pollutant emissions, SO_2 and NO_x were very low, although the remainder amount of sulfurous and the relatively great amount of phosphorous gases emissions need to be cleaned using modern technologies, so that the emission of poisonous gases can be avoided, and the relatively high sulfur content in all bituminous limestone can be used as a by-product.

Processing of oil shale

A few new oil shale processing technologies have been studied in many parts of the world. Direct combustion of shale oil has been successfully dealt with. Another technique is the Stuart project in Australia employs the Alberta Taciuk process which is based on rotary kiln technology instead of a fluidized bed combustor. The process here, however, is based on direct burning in a fluidized bed combustor which is one of the most efficient methods to burn oil shale directly. The fluidization principle is quite simple, air is introduced into a chamber through a perforated plate at the bottom of the bed. The chamber is filled with fine-grained particles of oil shale. The velocity of the air flow exceeds that required to support the particles against the force of gravity but is not sufficient to carry them out of the chamber. The particles are then suspended in the air in a state of turbulent motion, and there will be intimate contact of the particles with the air. As a result, chemical reaction and heat transfer between the oil shale particles and the air are greatly enhanced. But if the velocity goes too high, the particles are thrown out of the chamber. The bed is brought to the ignition temperature using hot air (or burner). It is found that the combustion efficiency will exceed 90% (assuming suitable size particles) when using a preheated air at about 500°C, and also it is not necessary to use catalyst's to start ignition.

System Proposed

The proposed system is comprised of three major sub-systems, which are a fluidized bed combustor unit, a power plant, and a multistage flash evaporator as shown in Fig. 1.

Fluidized bed combustor unit

The major parts of the fluidized bed combustor unit used are (see Fig. 2):

I- Combustion chamber

A vertical cylindrical combustion chamber of 5m diameter is assumed according to the required energy input. The combustor, fabricated from steel, is connected by an exhaust section (a conical inlet connected with 90° elbow followed by a straight portion connected to a cyclone and stack system).

The auxiliary parts of the combustion chamber are:

- Air distributor plate

The air distributor plate is used to introduce the air to the bed, so that uniform and stable air fluidization is achieved. It prevents the downward losses of the bed material, and supports the bed granules when it is de-fluidized. The grid plays an important role in controlling the bubble size because the bubbles are formed initially at the grid holes and rise up through the bed and burst at the surface. Relatively small bubbles will lead to smooth fluidization, while large bubbles cause non-uniform fluidization (Yavuzkurt et al., 1979).

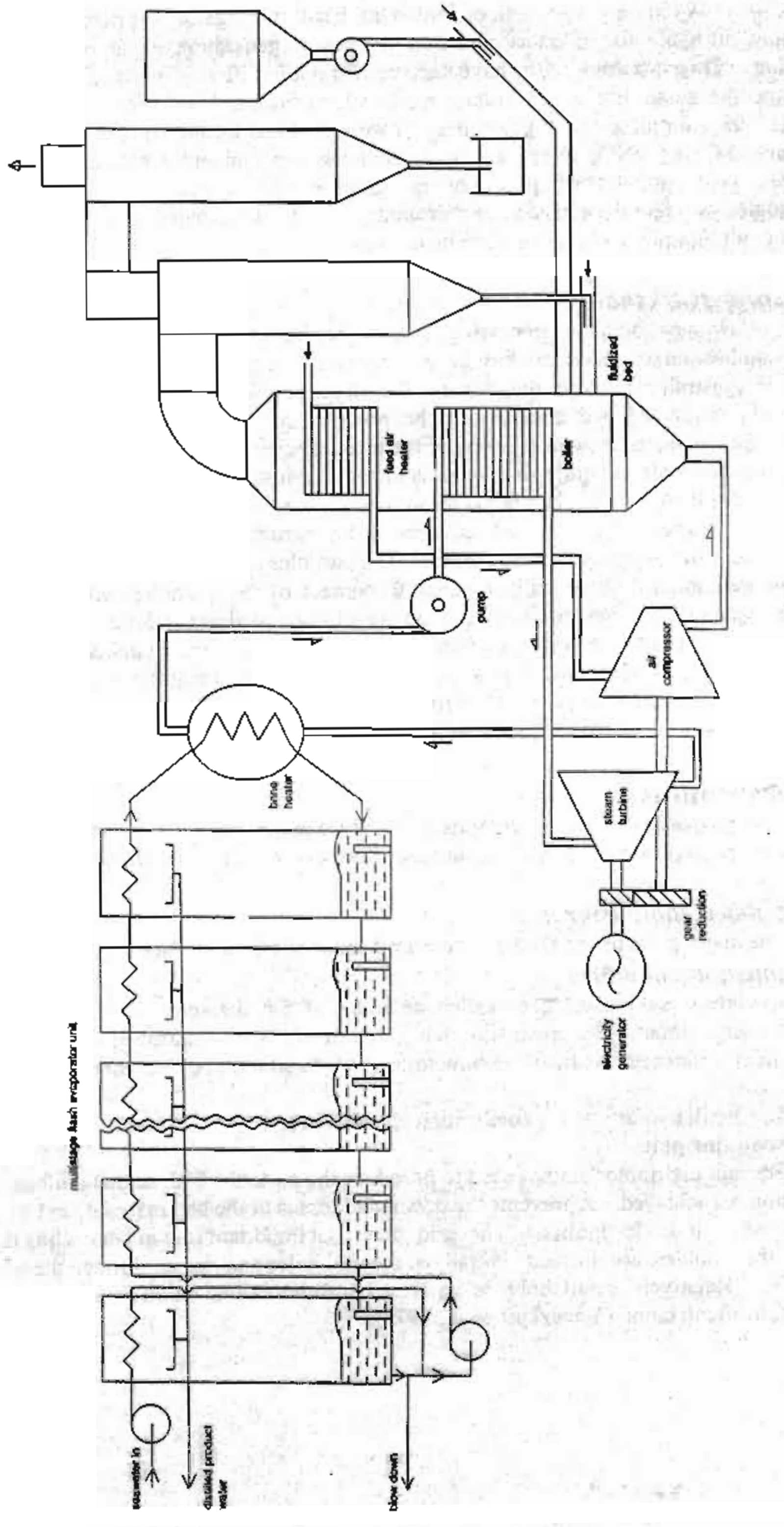

Fig. 1 Layout of complete system.

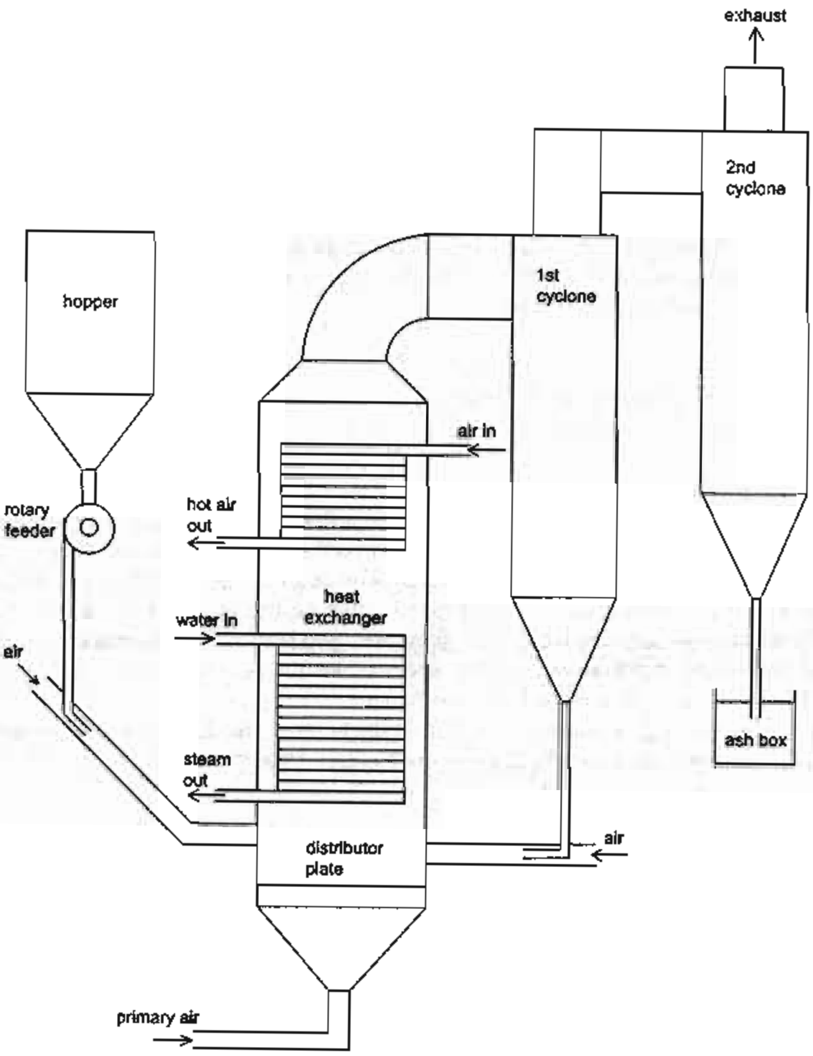

Fig. 2 Fluidized bed oil shale combustor with heat exchangers.

- **Discharge pipe**

It has two major functions; first it controls the bed height by controlling the amount of grains in the bed, and second it is used to remove and discharge the ash outside the bed, so that it is located at the bed center (Yavuzkurt et al., 1979).

Exhaust gas system

The exhaust system used consists of two cyclones, stack, and fly ash box. The combustion products are the exhaust flue gases, fly ashes and fine solids, entrained by air drag. The gases flow out of the combustor by the airflow and pass through the first cyclone-separator, which is used to re-circulate the particles entrained out of the bed back to the bed by integrating the cyclone directly to the combustion chamber. The second cyclone is used for higher efficiency of the recovery process, in which the fly ash and fine solids are extracted from the bottom of the cyclone under the action of centrifugal forces to the fly-ash box. The flue gases are then exhausted through the long stack that is located above the second cyclone.

II-The air supply

The air is supplied to the combustion chamber by a centrifugal compressor. The pressure drop across the fluidized bed, and the minimum fluidization velocity are estimated as follows:

- **Pressure drop in fluidized bed**

The fluid particle system begins to behave like a fluid and flows under hydrostatic head when fluidization state is achieved. The pressure drop across the bed is equal to the weight of mass of bed particles. Although the pressure drop would be exceeded just prior to the attainment of fluidization. The packing and interlocking of particles within the bed must first be broken down at relatively low flow rates in a packed bed, the pressure drop is then linearly proportional to gas velocity. Beyond minimum fluidization velocity and despite the increase in the bed (particles) expansion, the pressure drop remains, practically, unchanged.

The fluidization density is normally higher at the bottom and decays toward the top. The pressure drop across the fluidized bed, ΔP_f, is given by (Geldart, 1986),

$$\Delta P_f = \frac{Mg}{A}$$

Substituting for the bed mass yields,

$$\Delta P_f = (\rho_p - \rho_g)(1 - \varepsilon_{mf})gH$$

where ρ_p is the density of fluidized solid, ρ_g is the density of fluidization gas, and ε_{mf} is the voidage at minimum fluidization velocity.

- **Minimum fluidization and terminal velocities**

The bed is fluidized when the pressure drop across the bed is enough to balance the weight of the particles, and the velocity is then called minimum fluidization velocity, U_{mf}. The drag force on the particle due to the motion of the fluid to the weight of the particle is (El-Wakil, 1985),

$$C_D A_p \rho_g \frac{U_{mf}^2}{2} = \left(\frac{4}{3}\pi r^3\right)\rho_p g - \left(\frac{4}{3}\pi r^3\right)\rho_g g$$

where C_D is the drag coefficient, r is the particle radius, and A_p is the cross-sectional area of the particle = πr^2.

Then the minimum fluidization velocity is,

$$U_{mf}^2 = \left(\frac{\rho_p}{\rho_g} - 1\right)\left(\frac{8}{3}\right)\left(\frac{r}{C_D}\right)g$$

When the particles tend to be thrown from the bed by the airflow, the velocity is then called terminal, U_t. For particle Reynolds number <0.4, the terminal velocity is given by (Sapurov, 1978),

$$U_t = \frac{d_p^2(\rho_p - \rho_g)g}{18\mu}$$

For $0.4 < Re_p = \rho_p U_t d_p/\mu < 500$,

$$U_t = \left[\frac{4}{225}\frac{(\rho_p - \rho_g)^2 g^2}{\rho_g \mu}\right]^{1/3} d_p$$

III-Oil shale feeding system

Rotary vane feeders and screw feeders are most commonly used volumetric feeders, since they assure uniform and continuos flow of oil shale particles (powdered or lumped) to the bed. Rotary vane feeders are provided in either opened or closed designs. The open unit is best suited for handling small lumps materials, whereas the closed unit is used to act as an air lock in handling the materials. Rotary vane feeders include blow-through, side entry, or vertical construction. Normally V-belt or chains from motor reducer combination provide drive power with a variable speed. Rotary vane feeder is used here since it is favored over screw feeder as the former handles fine particles better than the later.

Power Plant

The power plant in this system is used to generate electricity, heat the brine of the multistage flash evaporator in the power cycle condenser, and power the feed air compressor to the fluidized bed. Steam is used in the plant as the working fluid, and oil shale is used as the fuel to supply heat to the steam in a large heat exchanger immersed in the fluidized bed chamber. The power cycle consists of the four major components of the Rankine cycle, the ideal cycle for vapor power plants (pump, boiler, turbine connected to an electric generator and the feed air compressor of the fluidized bed, and condenser which is actually a large heat exchanger that rejects heat to a cooling medium that is the brine of the multistage flash evaporator in this system). Energy analysis, however, was carried out on the basis of actual vapor power cycles (see Fig. 3).

Multistage flash evaporator

One of the methods for water desalination is a distillation process employing multistage flash evaporators, as shown in Fig. 4. In this proposed system ten stages of flash evaporators were used, however, more stages could be employed if required.

Feedwater is preheated in a condenser in each stage before it flows to a brine heater where a low-pressure steam is introduced from an external source (the power plant condenser). The feedwater is maintained under pressure conditions that do not permit vapor formation. No boiling takes place in the pipe leading to the flash chamber. Hot feedwater from the brine heater is introduced into the flash chamber, which is maintained under vacuum by an ejector. Water can be made to flash (boil) just as effectively by reducing the pressure as by raising the temperature.

The temperature in each stage is slightly below the boiling point (or saturation temperature) of the feed pressure. When feed enters the first stage, it is already at the saturation temperature for a higher pressure. It becomes superheated and has to give off vapors that causes brine to cool down to the saturation temperature equivalent to the first stage pressure.

The vapors are condensed on the condenser tubes. The heat of condensation supplies a large part of the heat required to raise the feed to its boiling point. Distillate is collected in the distillate collection pan. It then flows into the second stage collection pan and so on.

The pressure in each stage is lower than pressure in the preceding stage. The minimum pressure and temperature in the last stage are fixed by vapor volume and heat rejection considerations. This temperature is usually in the 36°-40°C range. The addition of multiple stages reduces the amount of heat that has to be removed from the process.

Brine from the first stage is sent to the second stage. When brine enters the second stage it flashes again. Vapor is condensed on the condenser at the top of stage two. The temperature of unflashed brine drops to a value corresponding to the stage pressure This brine flows into the third stage where it again undergoes flashing (as in the first two stages). This continues till the last stage. Concentrated brine from this stage is rejected (once through process) or recycled (re-circulation process). Unevaporated brine is finally rejected to the sea. Fresh feed water is added continuously.
In each stage distillate is produced and collected in connected pans to form salt free product water.

Multistage flash with re-circulation

To reduce the amount of fresh feed and to have better heat recovery, part of the brine coming out of the last stage is recycled (see Fig. 4). Recycling permits selection of a desired feed to produced ratio, which is directly related to the concentration ratio. Usually 50-75% of brine from the last stage is mixed with fresh feed (makeup) and re-circulated through the condensers and brine heater. The balance is rejected as blow down. Makeup flow is equal to the sum of the distillate and blow down flows.

The ratio of makeup to blow down flow is called the concentration ratio and usually varies in the 1.3-2 range. It is effected by the raw seawater salinity and the maximum temperature selected for flashing brine. In general a value of 1.7 should not be exceeded. To keep the concentration ratio within safe limits seawater makeup is continuously added to brine in the stage.

Compared to the once through process that comprises only two sections (brine heater and heat recovery section) the re-circulation process consists of three sections. The additional third section is the heat rejection section, the function of which is to remove excess heat from flashing brine. It does this by heat exchange with the cold seawater that is pumped to the inlet of the last stage of the heat rejection section.

System Analysis and Results

Fluidized bed combustion analysis

The ultimate analysis of El-Lajjun oil shale as reported by the natural resources authority (Abu-Ajamieh et.al., 1988) are as shown in Table (1)

Table 1 El-Lajjun oil shale contents

Constituent	% by mass	Average %
Moisture content	(3.00 – 5.78)	4.39
Ash	(52.00 – 57.29)	54.68
CO2	(16.62 – 21.13)	18.88
Total sulfur	(2.73 – 3.50)	3.12
C	(12.96 – 16.79)	14.88
H2	(1.41 – 1.87)	1.64
N2	(0.34 – 0.41)	0.38
O	(1.34 – 2.4)	1.87

The average mass of combustible materials (organic materials) which are C, H, N, and S is 20.02 % of total. Their mass and mole fractions are calculated and shown in Table 2.

Table 2 Mass and mole fraction of constituents of combustible materials.

Constituent	Mass fraction	Molecular weight (kg/kmol)	kmol/kg of organic mixture	Mole fraction
C	0.7433	12	0.061938	0.5466
H_2	0.0819	2	0.040950	0.3164
N_2	0.0189	28	0.000675	0.006
S	0.1559	16	0.009744	0.086
Sum			0.113307	1.0

- Combustion analysis and stoichiometric air/fuel ratio

The combustion analysis is curried out using chemical equations of combusted elements and products of combustion. The stoichiometric air/fuel ratio (A/F) is obtained through the following relation,

A/F = (Mole of Air × M.W)/(Mole of organic material× M.W)

Using the values in Table 2 yields,

A/F =12.837 kg air / kg of organic material in oil shale

For an average mass of combustible material of 20.02% by mass of unit of oil shale,

A/F = 0.2002 * 12.837 = 2.57 kg air / kg of oil shale

i.e. $\dot{m}_a = 2.57\ \dot{m}_{fuel}$

- Mass flow rate of oil shale and air

The mass flow rate of oil shale is computed based on the amount of heat required to the boiler, brine heater and air heater which were determined to be,

The amount of heat required to the boiler = 44.1615 MW

The amount of heat required to the feed air heater = $\dot{m}_a C_{pa} \Delta T$

The amount of heat released by burning the oil shale = $\dot{m}_{fuel}$ *C.V.

where C.V. is the calorific value of oil shale = 5442 kJ/kg.

The amount of heat released by burnt oil shale is equal to the amount of heat acquired by the boiler, feed air heater, and combustion losses (assumed negligible). Assuming the exit temperature from the air heater to be 500 K, with an inlet air temperature of 298 K, and the efficiency of combustion to be 90%, then,

$\eta_{comb} * \dot{m}_{fuel} * C.V. = Q_{boiler} + Q_{airheater}$

$0.9 * \dot{m}_{fuel} * 5442 = 44.1615*1000 + \dot{m}_{air}*1.005*202$

With $\dot{m}_a = 2.57\ \dot{m}_{fuel}$, gives the mass flow rate of oil shale, $\dot{m}_{fuel}$ = 10.09 kg/s, and the mass flow rate of air $\dot{m}_a$ = 25.93 kg/s.

- Calculation of mass of flue gases

The total mass of flue gases, $\dot{m} = \dot{m}_{product} + \dot{m}_{moisture} + \dot{m}_{Co2(mineral)}$

The mass of the products (CO_2+H_2O+NO_2+SO_2+N_2) in kg / kg of combusted material is obtained from the following equation,

$\dot{m}_{product}$ = total mass of products (kg / kmol comb. mat.) * 0.113307 (kmol / kg comb. mat.)
= 13.8375 kg / kg comb. mat

The mass flow rate of oil shale is 10.09 kg/s. A 20.02% combustible material of total of 10.09 kg/s oil shale is 2.02 kg/s of combustible material, therefore,

$\dot{m}_{product}$ = 13.8375*2.02 = 27.952 kg/s

Using the data in Table 1, we have,

$\dot{m}_{moisture}$ = 10.09*4.39/100 = 0.443 kg/s

and, $\dot{m}_{mineral\ CO2}$ = 10.09*18.88/100 = 1.905 kg/s

Therefore, the total mass of flue gases is,

$\dot{m}$ = 27.952+0.443+1.905 = 30.3 kg/s

- Minimum fluidization velocity

The minimum fluidization velocity is obtained using the following relation (El-Wakil, 1985),

$$U_{mf}^2 = \left(\frac{\rho_p}{\rho_g} - 1\right)\left(\frac{8}{3}\right)\left(\frac{r}{C_D}\right)g$$

By substitution of the oil shale density, $\rho_p = 1720$ kg/m^3, with r = 0.5 mm, and $C_D = 0.6$, we have,

$U_{mf}^2 = [(1720/\rho_g)-1][0.0218]$

Applying the continuity equation to the fluidization bed,

$\dot{m}_{air} = 25.93$ kg/s $= \rho_g \cdot U_{mf} \cdot A_f$

With a fluidization bed diameter of 5 m, $A_f = (\pi/4)\,5^2 = 19.635$ m^2, hence,

$\rho_g = 1.32 / U_{mf}$, and, $U_{mf}^2 = [(1720 * U_{mf} / 1.32) - 1][0.0218]$

yielding, $U_{mf} = 28.4$ m/s, and $\rho_g = 0.0465$ kg/m^3

- Feed air compressor power requirement

The schematic of the feed air compressor system is shown in Fig. 5.

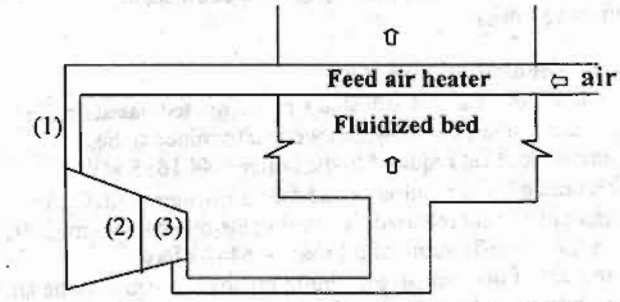

Fig. 5 Schematic of the feed air compressor system.

The energy absorbed by the compressor is determined by the conditions of the air at the inlet and outlet of the compressor impeller. If $(T_{o3} - T_{o1})$ is the stagnation temperature rise across the whole compressor then, since no energy is added in the compressor diffuser, this must be equal to the stagnation temperature rise $(T_{o2} - T_{o1})$ across the impeller alone. It would therefore be equal to the temperature equivalent of the work done on the air given by,

$T_{o3} - T_{o1} = \psi\sigma U^2 / Cp$

Where, ψ = power input factor, a typical value for the power input factor is 1.04,

σ = slip factor, a typical value for the slip factor is 0.9, and

U = impeller tip speed. For an assumed impeller overall diameter of 0.6 m and a rotational speed of 200 rps, U = 377.0 m/s,

$T_{o3} - T_{o1} = \psi\sigma U^2 / Cp = 132.37$

Therefore, the power required by the air feed compressor $= \dot{m}_a Cp\,(T_{o3} - T_{o1})$

$= 3450$ kW $= 3.45$ MW

Power plant analysis

Typical analysis of the steam power plant was conducted taking into account the irreversibilities occurring within the pump and turbine, but neglecting the heat loss from the steam to the surroundings and the pressure drop due to friction in the boiler, the condenser, and piping between the various components. A schematic of the power plant and the corresponding T-s diagram are shown in Fig. 3.

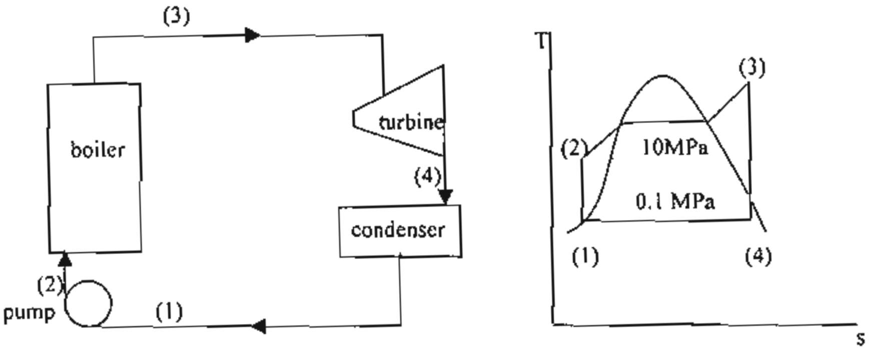

Fig. 3 Schematic of a power plant and its T-s diagram.

The design assumptions were,

- Superheated steam enters the turbine at 10 MPa and 500°C.
- Steam is condensed in the condenser at a pressure of 100 kPa.
- The mass flow rate of steam is 15 kg/s.
- The turbine and pump isentropic efficiencies are, $\eta_t = \eta_p$ = 85 percent.

The powers associated with this power plant were computed therefore to be,

turbine power = $\dot{m} * w_{t,out,act}$ = 12.55 MW

pump power consumption = $\dot{m} * w_{p,in,act}$ = 0.18222 MW

boiler heat input = $\dot{m} * q_{boiler,in}$ = 44.1615 MW

condenser heat rejection = $\dot{m} * q_{condenser,out}$ = 31.795 MW

The thermal efficiency of the power cycle, $\eta = w_{net}/q_{boiler,in}$ = 28 percent

Flash evaporator analysis

The analysis of a flash evaporator is based on the condensing temperature, T_{sat}, in each chamber.

The heat released in flashing brine is equal to (Porteous, 1983),

(rate of condensation of product water on distiller heat transfer surface) ×

(average enthalpy of condensation of product water)

i.e.

$$(\dot{m}_w - \frac{\dot{m}_p}{2}) Cp_w \Delta T = \dot{m}_p h_{fg}$$

The amount of product distillate water $\dot{m}_p$ in each of the ten stages of the flash evaporator (see Fig. 4) were therefore estimated using the relation above, and based on the assumption that the mass flow rate of brine water $\dot{m}_w$ in the first stage is 350 kg/s, and a drop in temperature of ΔT = 5°C in each stage for T_{sat}. The calculated results were as follows, with the details are displayed in Table 3,

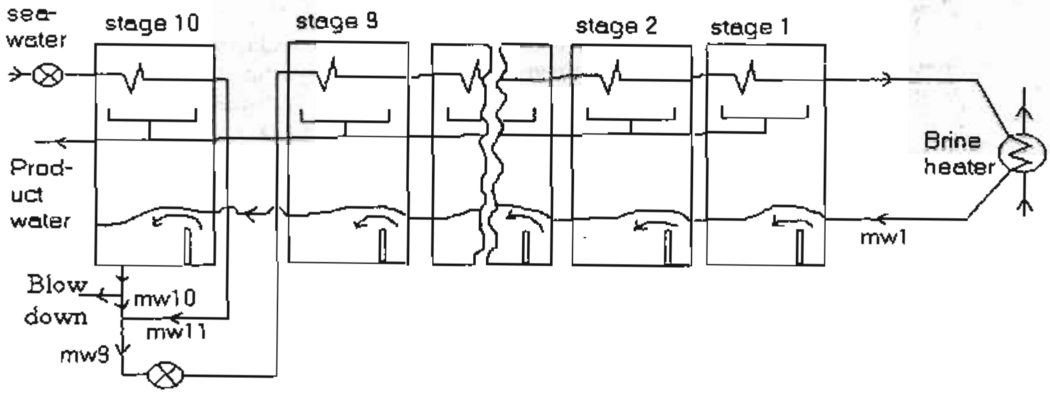

Fig. 4 Multistage flash evaporator system.

Total distillate water produced = $\sum \dot{m}_p$ = 30.286 kg/s = 109.03 m^3/h.

The amount of brine water leaving the last stage = 350 – 30.426 = 319.714 kg/s at T=50°C.

Usually 50 – 75 % of brine from the last stage is mixed with the fresh feed (make up) and re-circulated. Assuming that about 60% of the brine leaving the last stage is re-circulated, and making an energy balance for the mixed streams gives (see Fig. 4),

$$\dot{m}_{w10}\, h_{f10} + \dot{m}_{w11}\, h_{f11} = \dot{m}_{w9}\, h_{f9}$$

Since, $\dot{m}_{w10}$, $\dot{m}_{w11}$, and $\dot{m}_{w9}$ values are known, and T_{10}=50°C and T_{11}=20°C, the mixture temperature T_9 is evaluated to be 36.5°C.

As for the design of all heat exchangers, being the vapor condensers inside the flash evaporator, the power plant condenser (brine heater), the power plant boiler, and the compressor air feed heater is presented in Appendix A. The major findings of these heat exchangers, however, are given in Tables 3 and 4.

Table 3 Design outcome of multistage flash evaporator stages.

Stage	$\dot{m}_p$ kg/s	T_{sat} °C	T_s °C	h W/m^2.K	h_w W/m^2.K	U W/m^2.K	$\dot{Q}$ kW	LMTD °C	A m^2	n	L m
1	3.23	95	71	4806.43	21153.4	2116.8	7554.13	24.0	148.7	300	5.44
2	3.184	90	66.5	4728.67	20587.5	2073.54	7320.0	23.4	150.86	300	5.52
3	3.137	85	62.5	4742.59	20035.0	2090.74	7250.55	22.4	154.8	300	5.66
4	3.09	80	58.5	3925.0	19685.0	1911.13	7179.92	21.4	175.56	350	5.50
5	3.047	75	54.5	3895.67	19001.6	1896.49	7114.75	20.4	183.9	350	5.77
6	3.003	70	50.5	3859.59	18295.0	1879.53	7048.94	19.4	193.32	400	5.30
7	2.96	65	46.5	3822.5	17561.3	1861.47	6981.46	18.4	203.83	400	5.59
8	2.919	60	42.5	3786.49	17182.6	1847.9	6920.95	17.4	215.25	400	5.90
9	2.878	55	38.5	3790.25	16828.2	1843.95	6855.97	16.4	226.7	450	5.53
10	2.838	50	17.5	2973.49	13327.2	1580.0	6825.39	32.4	133.33	250	5.86

Power plant analysis

Typical analysis of the steam power plant was conducted taking into account the irreversibilities occurring within the pump and turbine, but neglecting the heat loss from the steam to the surroundings and the pressure drop due to friction in the boiler, the condenser, and piping between the various components. A schematic of the power plant and the corresponding T-s diagram are shown in Fig. 3.

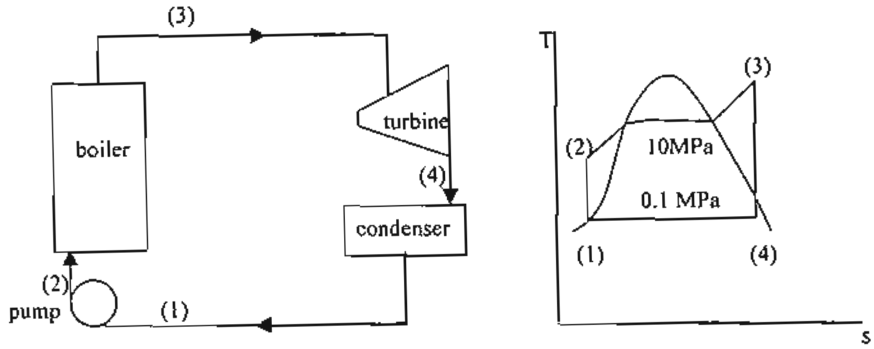

Fig. 3 Schematic of a power plant and its T-s diagram.

The design assumptions were,

- Superheated steam enters the turbine at 10 MPa and 500°C.
- Steam is condensed in the condenser at a pressure of 100 kPa.
- The mass flow rate of steam is 15 kg/s.
- The turbine and pump isentropic efficiencies are, $\eta_t = \eta_p$ = 85 percent.

The powers associated with this power plant were computed therefore to be,

turbine power = $\dot{m} * w_{t,out,act}$ = 12.55 MW

pump power consumption = $\dot{m} * w_{p,in,act}$ = 0.18222 MW

boiler heat input = $\dot{m} * q_{boiler,in}$ = 44.1615 MW

condenser heat rejection = $\dot{m} * q_{condenser,out}$ = 31.795 MW

The thermal efficiency of the power cycle, $\eta = w_{net}/q_{boiler,in}$ = 28 percent

Flash evaporator analysis

The analysis of a flash evaporator is based on the condensing temperature, T_{sat}, in each chamber.

The heat released in flashing brine is equal to (Porteous, 1983),

(rate of condensation of product water on distiller heat transfer surface) ×

(average enthalpy of condensation of product water)

i.e.

$$(\dot{m}_w - \frac{\dot{m}_p}{2})Cp_w \Delta T = \dot{m}_p h_{fg}$$

The amount of product distillate water $\dot{m}_p$ in each of the ten stages of the flash evaporator (see Fig. 4) were therefore estimated using the relation above, and based on the assumption that the mass flow rate of brine water $\dot{m}_w$ in the first stage is 350 kg/s, and a drop in temperature of ΔT = 5°C in each stage for T_{sat}. The calculated results were as follows, with the details are displayed in Table 3,

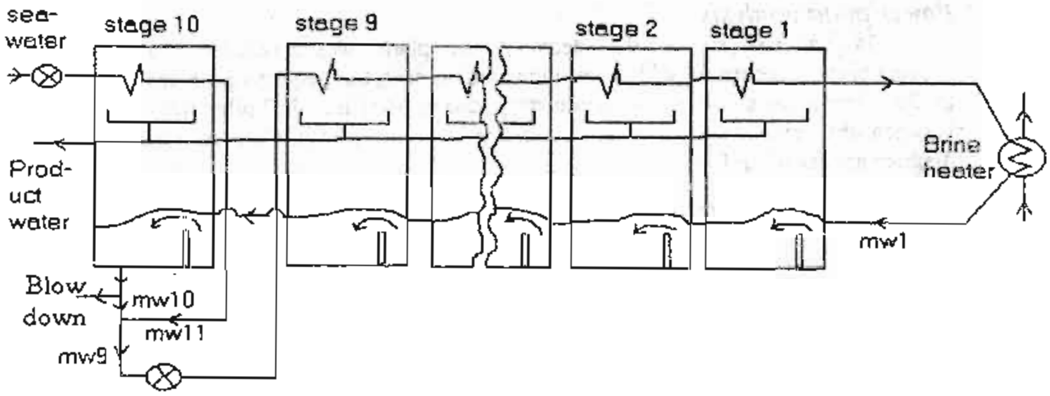

Fig. 4 Multistage flash evaporator system.

Total distillate water produced = $\sum \dot{m}_p$ = 30.286 kg/s = 109.03 m³/h.

The amount of brine water leaving the last stage = 350 – 30.426 = 319.714 kg/s at T=50°C.

Usually 50 – 75 % of brine from the last stage is mixed with the fresh feed (make up) and re-circulated. Assuming that about 60% of the brine leaving the last stage is re-circulated, and making an energy balance for the mixed streams gives (see Fig. 4),

$$\dot{m}_{w10}\, h_{f10} + \dot{m}_{w11}\, h_{f11} = \dot{m}_{w9}\, h_{f9}$$

Since, $\dot{m}_{w10}$, $\dot{m}_{w11}$, and $\dot{m}_{w9}$ values are known, and T_{10}=50°C and T_{11}=20°C, the mixture temperature T_9 is evaluated to be 36.5°C.

As for the design of all heat exchangers, being the vapor condensers inside the flash evaporator, the power plant condenser (brine heater), the power plant boiler, and the compressor air feed heater is presented in Appendix A. The major findings of these heat exchangers, however, are given in Tables 3 and 4.

Table 3 Design outcome of multistage flash evaporator stages.

Stage	$\dot{m}_p$ kg/s	T_{sat} °C	T_s °C	h W/m².K	h_w W/m².K	U W/m².K	$\dot{Q}$ kW	LMTD °C	A m²	n	L m
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10	2.838	50	17.5	2973.49	13327.2	1580.0	6825.39	32.4	133.33	250	5.86

Table 4 Design outcome of the various heat exchangers.

Component	h_i W/m².K	$\bar{h}_o$ W/m².K	$\dot{Q}$ MW	A m²	d (thin wall) mm	n	L m
Brine heater	3655	-	31.795	404.6	50	500	5.15
Power plant boiler	-	238.87	44.1615	541.9	50	750	4.60
Feed air heater	-	274.31	5.3112	80.1	50	150	3.40

Conclusions

This work dealt with seawater distillation using multistage flash evaporator accompanied with electrical power generation using oil shale as a fuel burned in a fluidized bed.

- Jordan possesses a very large resource in its vast reserve of oil shale. There are 17 known surface and near surface occurrences of oil shale containing 40 billions tons of oil shale. This huge amount of oil shale is sufficient for total national fuel consumption for about hundreds of years.
- A mass flow rate of oil shale of 10.09 kg/s is burned by 25.93 kg/s air at 500°C in a fluidized bed combustor of 5 m diameter which reached a temperature around 950°C. The energy of the combustion is utilized in the power plant boiler that required 44.16 MW and in the heater of the feed air to the compressor that required 5.26 MW. Flue gases exhaust from the fluidized bed at a rate of 30.3 kg/s with a temperature of 300°C. Such energy in the flue gases can be used after being cleaned in the cyclones in process heating or to drive a gas turbine.

The feed air single sided centrifugal compressor required power of about 3.45 MW which fed compressed air with a velocity of about 28.4 m/s to the fluidized bed.

- A power plant of turbine power of 12.55 MW is used to provide the power for the feed air compressor and water circulation pumps.. The remaining power that amounts to about 8.5 MW is utilized in electricity generation.
- A flash evaporator of ten stages of about 6 m in diameter each, that makes use of the energy in the power plant condenser which amounts to about 31.8 MW to heat the brine water was designed. The multistage flash evaporator collects a total of 30.3 kg/s (109 m³/h) when 350 kg/s of seawater, part of which 191.8 kg/s (60% of total) brine from the last stage is mixed with the fresh feed of seawater and re-circulated through the other stages and brine heater.

Nomenclature

A	Cross sectional area, or surface area, m²
A/F	Stoichiometric Air/Fuel ratio
C_D	Drag coefficient
C.V.	Calorific value of oil shale
Cp	Specific heat, kJ/kg.K.
d	Diameter, m
g	Gravitational acceleration, m/s²
H	Bed height, m
h	Enthalpy, kJ/kg, or convection heat transfer coefficient, W/m².K
Ja	Jakob number
k	Thermal conductivity, W/m.K
L	Length, m
LMTD	Log mean temperature difference, °C
M	Bed mass, kg
$\dot{m}$	Mass flow rate, kg/s

N	Number of tubes in column
n	Total number of tubes
Nu	Nusselt number
P	Pressure, Pa
r	Radius, m
Pr	Prandtl number
Q	Volume flow rate, m^3/s
$\dot{Q}$	Heat transfer rate, kW
Re	Reynolds number
T	Temperature, °C
T_o	Stagnation temperature, °C
U	Overall heat transfer coefficient, W/m^2K, or impeller tip speed of the compressor, or velocity, m/s
w	Work, kJ/kg
x	Quality
y	Mole fraction

Greek

ε	Packed bed voidage
η	Efficiency
μ	Absolute viscosity, $N.s/m^2$
ρ	Density, kg/m^3
σ	Slip factor of the compressor blades
ψ	Power input factor of the compressor

Subscript

a	Air
c	Compressor
f	Fluidized bed
g	Gas
H	Brine heater
i	Inside
L	Liquid
mf	Minimum fluidization
o	Outside
p	Particle, or pump, or product
s	Tube surface
sat	Condensing condition in flash evaporator
t	Terminal, or turbine
v	Vapor
w	Water

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Appendix A

Design of Condensers, Brine Heater, Boiler, and Feed Air Heater

- Design of vapor condensers inside the flash evaporator

stage (1) condenser

The condensing heat transfer coefficient is determined from the following equation,

$$h = 0.729\left[\frac{g\rho(\rho-\rho_v)h_{fg}^{*}k^3}{\mu(T_{sat}-T_s)d_o}\right]^{\frac{1}{4}}$$

where h_{fg}^{*} is the modified latent heat of vaporization, which reflects the subcooling of the condensate in the analysis,

$h_{fg}^{*} = h_{fg} + 0.68\,Cp\,(T_{sat} - T_s)$

The effect of ripples is checked, so that if,

$$Re = \frac{h\pi\frac{d_o}{2}\left(T_{sat} - T_s\right)}{h_{fg}^{*}\mu} > 10$$

ripples do appear, and h is then increased by 20%.

If the number of tubes in a column, N, is more than 14, then the heat transfer coefficient is $h/N^{1/4}$.

To determine the water-side heat transfer coefficient, h_w, the Reynolds number is computed first,

$Re = \rho U_i d_i/\mu$

where, U_i = velocity = $Q/A = \dot{m}_w/\rho.A = 4\,\dot{m}_w/\rho.\pi.d_i^2$

Therefore, if the flow is turbulent,

$$Nu = \frac{h_w d_i}{k} = 0.023\,Re^{0.8}\,Pr^{0.4}$$

else, Nu = 4.36

from which h_w can be obtained.

The overall heat transfer coefficient, U, is therefore,

$$\frac{1}{U}=\frac{1}{h}+\frac{XA_o}{kA_{mean}}+\frac{A_o}{A_i}\frac{1}{h_{ff}}+\frac{A_o}{A_i}\frac{1}{h_w}$$

where, $\frac{XA_o}{kA_{mean}}$, is the resistance of the tube, and

$1/h_{ff}$ is the fouling factor = 0.000176 m^2.K/W (Incropera and DeWitt, 1990)

The heat transfer rate, $\dot{Q}$, is

$$\dot{Q}=\dot{m}_p h_{fg}^{*}=UA\,(LMTD)$$

where, LMTD is the log mean temperature difference.

Hence, $A = \dot{Q}\,/\,U\,(LMTD)$, and the length of each tube, L, is,

$$L = A\,/\,\pi\, d_o\, n$$

where n is the total number of tubes.

Using the same procedure, the length of each tube in the other stages is determined as shown in Table 3.

- Design of the brine heater

Seawater is first fed to the last stage of the flash evaporator, and then flows through the condenser tubes of all the stages. On leaving the first stage condenser tubes, the seawater is heated in the brine heater (the condenser component of the power plant).

Applying an energy balance over the brine heater (see Fig. 3), we have,

$$\dot{m}_w Cp_w \Delta T_w = \dot{m}_{steam}\left(h_4 - h_1\right)$$

from which, ΔT_w, the seawater temperature rise in the brine heater is determined.

The length of each tube in the brine heater is evaluated by referring to Fig. 7.

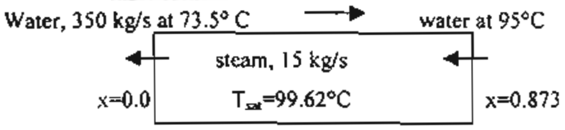

Fig. 7 Schematic of the brine heater tube section.

The water temperature inlet is = 73.5°C, and outlet 73.5 + 21.5 = 95°C

The Reynolds number, Re_v, based on the mean mass flow rate of water vapor in the condenser tube, $\dot{m}_v$, is,

$$Re_v = \frac{4\dot{m}_v}{\pi d_i \mu}\left(\frac{\rho}{\rho_v}\right)^{\frac{1}{2}}$$

The Reynolds number, Re_l, based on the mean mass flow rate of liquid water in the condenser tube, $\dot{m}_l$, is,

$$Re_l = \frac{4\dot{m}_l}{\pi d_i \mu}$$

For $20000 < Re_v < 100000$, and $Re_l < 5000$,

$$Nu=\frac{hd}{k}=0.1\,Pr^{\frac{1}{3}}\,Re^{\frac{2}{3}}\,Ja^{-\frac{1}{6}}$$

where, $Ja = Cp.\Delta T/h_{fg}$

Therefore, the heat transfer coefficient, $h = Nu.k/d$, is then computed, and since,

$$\dot{Q} = \dot{m}_{st}.h_{fg}^{*} = h.A.\Delta T$$

the length of each tube of the brine heater (power plant condenser) is then,

$$L = A / \pi\, d_i\, n$$

- Design of power plant boiler

Liquid water enters the boiler heat exchanger inside the combustor of the oil shale fluidized bed at 100°C and leaves at 500°C, whereas flue gases enter at 950°C and leave at 450°C.

The heat transfer rate acquired by the steam, $\dot{Q} = 44.1615$ MW as was determined by the analysis of the power plant.

An average heat transfer coefficient to the entire horizontal tube bundle in the bubbling fluidized bed combustor is required. The average convection heat transfer coefficient used, $\bar{h}_o$, was evaluated by using the correlation that was proposed by Zhukauskas (1972),

$$\overline{Nu}_d = \frac{\bar{h}_o d}{k} = C\, Re_{d.max}^{m}\, Pr^{0.36}\left(\frac{Pr}{Pr_s}\right)^{0.25}$$

where, C and m are given for each tube bank in cross flow configuration. The Reynolds number to be evaluated at maximum velocity between the tubes. The heat transfer rate is based on the log mean temperature difference, LMTD, so that,

$$\dot{Q} = \bar{h}_o\, A_o\, LMTD$$

the length of each tube of the power plant boiler heater is then,

$$L = A_o / \pi\, d_o\, n$$

- Design of feed air heater

The outside air enters at 25°C (298 K) and leaves at 227°C (500 K), and the flue gases enter at 450°C and leave at 300°C (see Fig. 5).

The heat transfer rate acquired by the feed air is,

$$\dot{Q} = \dot{m}_{air}\, Cp_{air}\, \Delta T_{air} = \bar{h}_o\, A_o\, LMTD$$

where $\bar{h}_o$ is the average convection heat transfer coefficient which is computed from the correlation of Zhukauskas (1972) using the same procedure that was carried out for the boiler. Hence, the length of each tube of the feed air heater is determined from,

$$L = A_o / \pi\, d_o\, n$$