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THE MSF/FBE: AN IMPROVED MULTI-STAGE FLASH DISTILLATION PROCESS

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ABSTRACT.

A new type of flash evaporator is discussed which avoids certain limitations of the conventional horizontal flash evaporator. The main characteristic of the plant is its vertical concept with short flash chambers accommodating a large number of parallel tubes. Heat transfer in the tubes takes place by means of a fluidized bed. Due to this fluidized bed the heat transfer coefficient is enhanced in particular during operation at low superficial fluid velocity. Heat transfer data, stability of the multiple parallel fluidized beds, flash-off, entrainment separation, allowable vapour space loading of the flash chamber as well as the possibilities to design the plant for lower specific heat consumption and to operate at higher maximum brine temperatures are discussed. Delft University of Technology developed a 50 m^3/day plant for the production of feed water for the University's central boiler house. Construction and operating experiences of this plant are presented. Results of easy start-up and part-load procedures due to the non-flooding precautions in this design will be discussed. A comparative cost study between a 10,800 m^3/day horizontal flash evaporator of the Rotterdam Municipal Water Department and a MSF/FBE for the same production as the conventional plant will be mentioned in this paper.
Further extensive studies on determination of heat transfer coefficients, fouling and scale removing capabilities of the fluidized bed are briefly mentioned.

1.0. INTRODUCTION.

Over the past years Delft University of Technology has developed a new type of flash evaporator. Started more or less as an attempt to improve a vertical upward flashing evaporator (see reference 1 and 2), the final result was a completely new type of flash evaporator in which the vertical design is the only feature still resembling the earlier development in the United Kingdom. This newly developed vertical flash evaporator, referred to as Multi-Stage Flash/Fluidized Bed Evaporator (MSF/FBE), has been extensively described in references [2] and [3].
Figure 1 shows the simplified flow diagram of a vertical MSF; the plant consists of a flash column and a condensation column including the brine heater. The stage geometry is strongly determined by the condenser design; normally there are two possibilities:

- cross flow; where flow through the condenser tubes is always perpendicular to the flashing brine flow.
- long-tube design with parallel flow.

The lay-out of the plant shown in figure 1 would benefit from a long-tube design as this would both reduce the number of waterboxes required and the number of tube-tubesheet connections. However, the long-tube condenser requires the flash chamber height to become equal to the length of the condenser tubes. As the stages are planned on top of each other, the long-tube condenser may therefore result in a total column length which is unpractical.
Condenser tube length depends on the various process parameters, a simplified expression being:

$$L = \frac{\rho_l . C_l . D_o . \bar{V}_l}{4K\left(\frac{n}{g_r} - \frac{1}{2}\right)} \qquad \text{reference [2]} \qquad (1)$$

where L = length of condenser tubes

D_o = outer diameter of condenser tubes

ρ_l = density of liquid in the condenser tubes

C_l = specific heat of liquid in the condenser tubes

$\bar{V}_l$ = superficial liquid velocity in the condenser tubes

K = overall heat transfer coefficient for the condenser tubes

n = total number of stages

g_r = gain ratio

For commonly used tube diameters, liquid velocities and overall heat transfer coefficients the length of the condenser tubes will usually exceed 2.5 m per stage. Taking into account that a flash evaporator may contain 35 stages, it is easy to understand that the long-tube condenser design does not fit the plant shown in figure 1 as it would yield a total colomn height of approx. 85 m. Application of a long-tube condenser design will therefore require a heat exchange mechanism yielding a high overall heat transfer coefficient at very low liquid velocities, so that stage heights much less than 1.0 m can be achieved.

It will be shown that a fluidized bed heat exchanger, consisting of a large number of vertical parallel condenser tubes partly filled with particles fluidized by the liquid flowing upward through the tubes, permits the use of a long-tube condenser design with an extremely short stage height.

2.0. FLUIDIZED BED HEAT EXCHANGER.

2.1. Heat transfer coefficient.

The expression for the overall heat transfer coefficient K, based on the outer heat exchange surface of the condenser tubes, reads:

$$K = \frac{1}{\frac{1}{\alpha_c} + \frac{D_o}{2\lambda_t}\ln\frac{D_o}{D_i} + \frac{D_o}{D_i}\cdot\frac{1}{\alpha_l} + f_o + f_i} \qquad (2)$$

where α_c = heat transfer coefficient for condensation
λ_t = thermal conductivity of the tube material
α_l = wall-to-liquid heat transfer coefficient (based on the inner diameter)
D_i, D_o = inner and outer tube diameter, respectively
f_i, f_o = fouling factor for the inner and outer surface, respectively

For computing the heat transfer coefficient α_c the correlation of Nusselt for condensation on a vertical wall and laminar flow of the condensate film should be used:

$$\alpha_c = 0.943 \frac{\lambda_c^3 \cdot \rho_c^2 \cdot h_{lg}}{H \cdot \Delta T_c \cdot \eta_c} \tag{3}$$

where λ_c = thermal conductivity of the condensate
ρ_c = density of the condensate
h_{lg} = latent heat of evaporation
H = undisturbed length of the condensate film
ΔT_c = temperature difference over the condensate
η_c = dynamic viscosity of the condensate

For the wall-to-liquid heat transfer coefficient of a fluidized bed heat exchanger the correlation of Ruckenstein and Shorr [4] was found by the author to agree best with his experimental results. This correlation reads as follows:

$$Nu = 0.067 . Pr^{0.33} Re_{d_p}^{-0.237} Ar^{0.522} \tag{4}$$

$$\text{if } Re_{d_p} . Ar^{-0.50} > 0.09 \tag{5}$$

In the above equations the dimensionless numbers of Nusselt, Prandtl, Reynolds and Archimedes are composed as follows:

$$Nu = \frac{\alpha_l . d_p}{\lambda_l}; \quad P_r = \frac{\eta_l . c_l}{\lambda_l};$$

$$Re_{dp} = \frac{\rho_l . V_l . d_p}{V_l}; \quad Ar = \frac{g d_p^3}{V_l^2} \left(\frac{\rho_s - \rho_l}{\rho_l}\right)$$

where d_p = particle diameter
λ_l = thermal conductivity of the liquid
ρ_s, ρ_l = density of the solid and liquid phase, respectively
V_l = kinematic viscosity of the liquid

Figure 2 compares the results obtained experimentally for the wall-to-liquid heat transfer coefficients of a fluidized bed with the correlations found in literature. From this it can be seen that the wall-to-liquid heat transfer coefficients for a fluidized bed exceed the values given by the Dittus-Boelter correlation, i.e. without fluidized bed, by a factor of six to seven; this is caused by the break-down of the laminar thermal boundary layer at the wall of the tube due to the action of the fluidized particles. The fluidized bed heat exchanger only requires a superficial velocity* of 0.10 to 0.20 m/s-depending on the particle size and density and tube diameter-for obtaining the same value of wall-to-liquid heat transfer coëfficiënt as a conventional heat exchanger having a liquid velocity of 1.8 m/s. A fluidized bed heat exchanger thus offers the possibility to obtain high heat transfer coefficients at much lower liquid velocities. According to equation (1), this makes it possible to apply the long-tube condenser configuration by reducing the stage height of the vertical flash evaporator to very low values.

2.2. Hydraulic behaviour of a fluidized bed.

Stable operation of multiple parallel fluidized beds is achieved by proper setting of the throttling devices required at the inlet of the beds. If the stability criterion is satisfied all individual fluidized beds will become fluidized up to the outlet box of the bundle. In case of unstable operation some fluidized beds will eject most of their particle content while other beds will contain an excess of particles. For stable operation of a bundle all the individual beds have to satisfy the equation:

$$\left.\frac{\Delta P_{thr}}{\Delta P_w}\right|_{operating} > \frac{1}{2n}\left(\frac{Ex}{P} - 1\right) \quad [2] \qquad (6)$$

where ΔP_{thr} = pressure loss across the throttling devices
ΔP_w = hydrostatic pressure drop across the bed
n = empirical factor = 2.39
Ex = expansion factor of the fluidized bed
P = volume fraction of the particles in the fixed bed (~ 0.60)

The pressure loss ΔP_w can be expressed by:

$$\Delta P_w = g.L_f\ (1 - \varepsilon)\ (\rho_s - \rho_l) \qquad (7)$$

where L_f represents the height over which the particles are fluidized and ε the porosity

$$\varepsilon = 1 - \frac{P}{Ex} \qquad (8)$$

A sizeable part of the pressure loss across the throttling tubes must be due to wall friction in the tubes in order to allow relatively wide cross sections of these tubes and thus to prevent clogging.

* superficial velocity = the velocity of the liquid in the tube in fluidized state, related to the cross-section of this tube.

The expression for the pressure loss reads:

$$\Delta P_{thr} = (\xi + \lambda \frac{L_i}{d_i}) \cdot \tfrac{1}{2}\, \rho_1 . V_1^2 \, (\frac{D_i}{d_i})^4 \qquad (9)$$

where ξ = pressure loss coefficient

λ = wall friction factor

L_i = length of the throttling device tube

D_i,d_i = inner diameter of the condensor resp. inlet tube

2.3. Self-cleaning and descaling capability.

The self-cleaning and descaling capability of the fluidized bed heat exchanger is based on the abrasive action of the fluidized particles on the wall of the heat exchanger. Also because of this abrasive effect the fouling factor of the fluidized bed is drastically reduced compared to conventional heat exchangers; as confirmed both by experiments and by one year's operating experience with the pilot plant described in chapters 3 and 4. Actually, the heat exchanger can be kept clean from fouling, which results in a lower average specific heat consumption than for a conventional design with comparable heat transfer surface. Due to the abrasive power of the fluidized particles a fluidized bed heat exchanger can also remove part of the scale precipitated from the liquid. The heat transfer surface remains free from scale if the abrasive power of the particles exceeds the rate of precipitation.

3.0. PROCESS PROPERTIES.

Flash-off and entrainment separation determine to a large extent the configuration of a flash chamber and the maximum flash chamber loading.

3.1. Flash-off.

Figure 3 shows a flash chamber interior. Here the brine passes through a slit from the upper chamber to the lower one after which it diverts and is intercepted by a vertically placed perforated plate, made of stainless steel. Because of the metastable state of the brine when entering the lower flash chamber, the brine is torn apart and passes through the upper part of the flash chamber as thin film liquid filaments and relatively large droplets. Afterwards, it flows down the perforated plate as a highly turbulent liquid film.
Breakdown of the brine into thin film liquid filaments and droplets and the subsequent formation of a turbulent liquid film both contribute to a better interfacial heat transfer which causes a complete flash-off of the brine.

3.2. Entrainment separation.

As can be seen from figure 4 showing the steam paths the removal of most of the droplets in the vapour may be expected by centrifugal effects due to the change of direction of the vapour in the flash chamber. Experiments proved that for this type of flash chamber very good entrainment separation can be achieved even without the use of a conventional wire mesh demister.

The above information about flash-off and entrainment separation becomes even more striking when it is realized that this result was obtained at flash chamber loadings more than twice those of a conventional horizontal flash evaporator.

3.3. Lower specific heat consumption.

The MSF/FBE can apply very small temperature differences over the stages, because the driving force for the interstage brine transport is now largely obtained from the brine height in the stages. For the conventional horizontal flash evaporator with a maximum brine temperature of 120^{o}C the number of stages which can be inserted in the given flash range without the risk of operating instability due to inadequate temperature differences over the stages is limited to about 40; this corresponds with a maximum gain ratio of about 12, i.e. a specific heat consumption of about 190 KJ/kg.
For the MSF/FBE, this limitation does not exist as a brine height in the stages of 10 to 15 cm already guarantees a sufficiently stable transport between the stages independent of the interstage temperature differences. Actually, this makes it possible to design the MSF/FBE for a far greater number of stages than 40 and consequently for much higher gain ratio's or lower specific heat consumption than can be applied in the horizontal evaporator.

4.0. THE 50 m^3/DAY MSF/FBE.

For the development of a MSF/FBE a number of studies were carried out in the Laboratory for Thermal Power Engineering, Delft University of Technology. The first step towards application of the MSF/FBE on a more industrial scale has been the construction of such a plant for the University's boiler house for the production of boiler feed water from highly polluted canal water.

4.1. Construction.

Figure 4 shows the configuration of the five stage once-through MSF/FBE designed for a production of 50 m^3/day with a specific heat consumption of approx. 700 KJ/kg distillate. The evaporator can best be described as a countercurrent heat exchanger; the raw feed is heated while flowing upward through the heat recovery section consisting of fluidized bed condenser tubes. The brine, after obtaining a final heat supply in the heat input section arranged on top of the heat recovery section proper, flashes downward through all stages via orifices in the interstage walls. The gradual drop in saturation pressure and temperature in the stages results in partial evaporation of the brine in each flash chamber. The vapour thus produced, after being demisted, reaches the surface of the condenser tubes cooled by the upward flowing feed. The distillate cascades down in the same way as the brine. Vacuum is maintained by means of steam ejectors. Tables 1 and 2 summarize the main plant data such as distillate production, temperatures, porosity, heat transfer surface, etc.

4.1.1. Flash chamber geometry.

The plant consists of a circular tower structure with stacked stages. In this configuration of the MSF/FBE only one column is used with the condenser located in a central shell and surrounded by an annular flash chamber.

4.1.2. Brine heater.

The brine heater of the plant is integrated with the highest temperature stage of the recovery section. Therefore it is not necessary to install extra tube plates, waterboxes and the accessory piping.

4.1.3. Stage sealing.

The interstage walls are not welded to the shell; the sealing between the interstage wall and the shell is provided by a rubber strip as shown in figure 5. In this way it is possible to pre-assemble the complete condenser bundle, the interstage walls, the necessary siphons, the orifices and the flash basket prior to final erection.

4.1.4. Brine siphons.

The fluidized bed heat exchanger requires the beds to be expanded up to the outlet water box for all operating conditions. Since any fluidized bed heat exchanger originally contains a given amount of particles, the above requirement leads to a minimum allowable flow for the expanded beds to reach the outlet water box. During normal operation there will be an excess of particles in the outlet water box.
During start-up procedure this minimum flow must be established while there are as yet no temperature differences and thus no pressure differences over the stages (the driving force for the interstage brine transport consists only of the hydrostatic pressure from the brine heights in the stages).
Flooding of the stages is unavoidable unless special devices are applied to increase the interstage brine transport during start-up. Possible devices are interstage valves or siphons. In this pilot plant we choose for siphons; if the brine level in the stage reaches the siphon inlet, an additional flow area becomes available. During normal operation the siphon forms a water seal between the stages.

4.1.5. Stirring device.

The particles in the outlet water box will behave as a fixed bed; during normal operation this excess of particles enables the tubes of the heat exchanger to remain completely filled with fluidized particles in case of a slight reduction in flow. A fixed bed of particles in the outlet box could cause bridging across the outlet opening of the tubes. However, this problem is easily solved by applying a stirring device in the outlet water box as shown in figure 6.

4.1.6. Distillate siphons.

Distillate is cascading down the plant and the interstage distillate transport takes place via distillate siphons. These distillate siphons act as water seals against vapour short circuiting between the stages and have about the same design as the brine siphons.

4.2. Operation.

4.2.1. Load variation.

Looking at the load variation of a conventional horizontal flash evaporator it is clear that both the brine flow and the top temperature of the brine have to be varied in order to maintain the brine levels in

the flash chambers within their required limits.
The MSF/FBE has the problem that brine flow must be maintained near the design value since the fluidized beds have to remain expanded over their total length. The only variation permitted is that of the maximum brine temperature. If this temperature is lowered, the brine levels in the flash chambers will increase. However, the siphons are used for by-passing part of the brine flow during off-design load. Consequently the load of an FBE can be reduced to zero, i.e. where the maxium brine temperature corresponds to the discharge temperature.

4.2.2. Fouling.

As explained in section 2.3 a greatly reduced fouling factor for the FBE plant can be used in the calculation for the heat transfer surface, compared to the conventional horizontal flash evaporator.

4.2.3. Vapour space loading.

Operating experience with this MSF/FBE and additional experiments proved that much higher vapour space loadings [defined as: kg distillate produced per hour and per m^3 vapour space in the flash chamber] can be obtained in this evaporator type. In this MSF/FBE, however, vapour space loading is relatively low, viz. 1.25, because reduction of the flash chamber dimensions is only justified for large plants. For the smaller plants the dimensions are mainly determined by accessibility considerations for inspection and repair. For larger plants this value will be increased up to 3.

5.0. COMPARATIVE COST STUDY.

A comparative cost study between an existing 10,800 m^3/day conventional horizontal flash evaporator of the Rotterdam Municipal Water Department and a MSF/FBE designed for the same capacity and operating conditions will be discussed in the following.[see ref. 5]

5.1. Introduction.

In 1967 the Rotterdam Municipal Water Department decided to install three flash evaporator units for the production of process water to serve the industrial area of the city of Rotterdam. The flash evaporator units are of the conventional horizontal type and of long-tube design. Each unit consists of a total number of 35 stages, 32 of which are in the recuperation section. The distillate production per unit is 450 m^3/hour with a steam consumption of 50 tons/hour at a steam pressure of 1.85 bara and a temperature of 119°C. In terms of specific heat consumption the plant required 246 KJ/kg distillate produced.
Two MSF/FBE designs will be compared with the horizontal plant, viz.:

- A low gain ratio plant exactly matching the specification of the horizontal plant;
- A high gain ratio plant, of which the specific heat consumption is only half of that of the horizontal plant, while its number of stages for both the recirculation section and heat reject sections has been increased by a factor 2.

5.2. Specific heat consumption.

In section 3.3 it has been explained that a MSF/FBE obtains a large fraction of its driving force for the interstage brine transport from

the brine height in the stages, while for a horizontal flash evaporator this driving force exclusively depends on the temperature differences over the stages. Hence a horizontal flash evaporator is severely limited in applying high gain ratio's; for the MSF/FBE this limitation does not exist, as a brine height in the stages of 10 to 15 cm is already adequate for providing stable brine transport between the stages, independent of the interstage temperature differences.
It will be shown that, while the low gain ratio MSF/FBE has already a distinct advantage over the horizontal plant, this advantage vastly increases for the high gain ratio plant. This is of course due to the steep increase in fuel cost over the last years, which made energy cost the decisive factor in total production cost.

5.3. The low gain ratio MSF/FBE.

Table 3 presents detailed data on this design. The actual flash chamber dimensions followed from the vapour space loading, required for complete flash-off and adequate entrainment separation [2].
Table 4 compares this MSF/FBE to the horizontal plant. The following brief comments appear in order:

- capital costs:
  savings are obtained due to
  a) the smaller temperature losses in the stages caused by the reduced fouling factor
  b) the absence of a separate brine heater
  c) the reduced flash chamber dimensions due to the high flash chamber loading factor and the absence of reinforcement of the vessel walls.
- running costs:
  production loss due to fouling is avoided thanks to the abrasive action of the particles on the tube wall.

5.4. The high gain ratio MSF/FBE.

Differences in design between the high and low gain ratio MSF/FBE are the following:
- the number of stages of the recirculation and heat reject sections have been increased to 64 and 6, respectively.
- the recirculation section is distributed over two columns, while the brine heater forms an integral part of the high temperature recirculation column.
- heat transfer surface is approximately doubled.
- steel weight increases to 500 tons.
- total required pumping power increases by 40%.

Table 5 compares the horizontal plant with the high gain ratio MSF/FBE.

6.0. RESEARCH PROGRAMME.

In addition to the theoretical work and the erection and operation of the first pilot plant a number of test modules have been built to investigate the various mechanisms in a vertical flash evaporator.
Some of the main research items are listed below:

- heat transfer coefficients.
  due to the abrasive power of the fluidized particles the tube wall remains free from deposits. Hence there is a more urgent need to know the wall-to-liquid-heat transfer coefficient for a large range of temperatures, particle densities, tube diameters, particle diameters etc.

- self-cleaning and descaling capability.
  a research programme is under way to investigate the depositing and removal mechanisms for both organic materials and scale deposits.
- flash chamber loading.
  This part of the program aims at establishing the limits for flash chamber loadings.

7.0. CONCLUSIONS.

The vertical flash evaporator combined with a fluidized bed heat exchanger has a number of advantages over the conventional horizontal evaporator:

- smaller plant size, better flash-off, no demister losses and high vapour space loading factor.
- much lower specific heat consumption.
- the possibility to vary the load of the MSF/FBE from 0 to 100%.
- the possibility to increase the maximum brine temperature above temperature limits normally imposed by scale formation.

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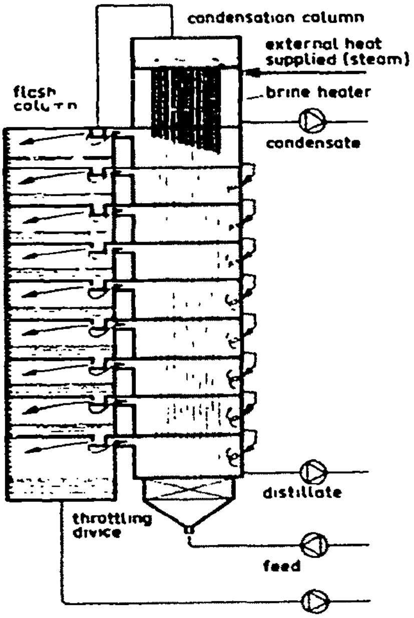

fig 1 flow diagram of the MSF/FBE

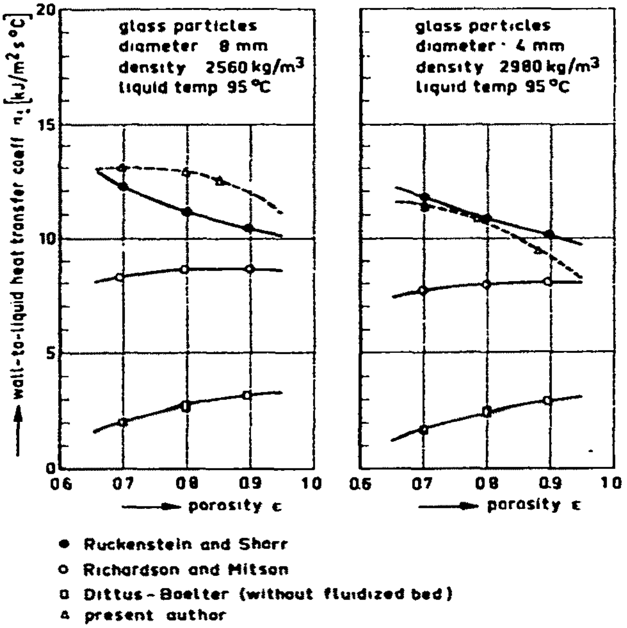

fig 2 wall-to-liquid heat transfer coefficients in a fluidized bed

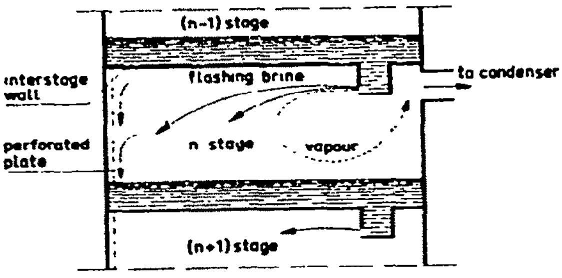

fig 3 an MSF/FBE flash chamber design

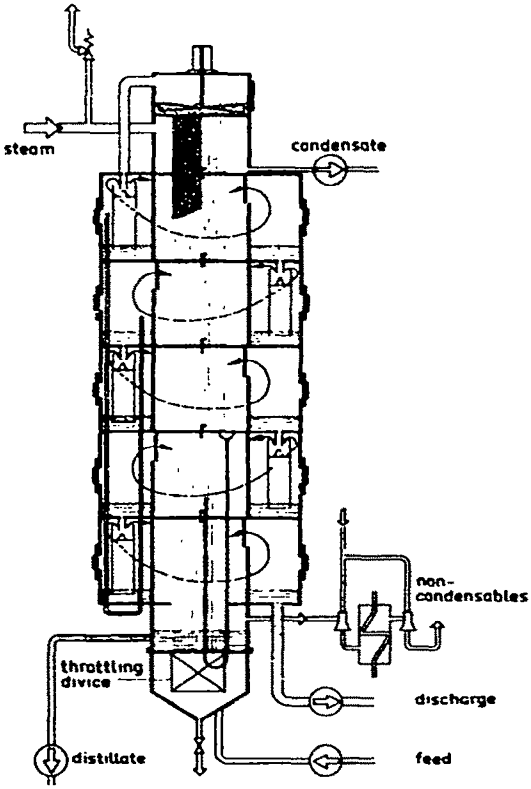

fig 4 the 50 m^3/day MSF FBE

Distillate production	2010	kg/hr
Steam consumption	586	kg/hr
Steam pressure	1.85	bara
Steam temperature	119.0	°C
Condensate temperature	105.0	°C
Gain ratio	3.43	
Specific heat consumption	643	kJ/kg
Maximum brine temperature	100.0	°C
Cooling water inlet temperature	18.0	°C
Blow-down temperature	35.3	°C
Distillate temperature	34.7	°C
Number of stages	5	
Concentration of the feed	1000	TDS
Brine flow (feed)	18286	kg/hr
Number of condenser tubes	288	
Condenser tube material	aluminum brass	
Condenser tube diameter	I.D.: 13.44	mm
	O.D.: 15.88	mm
Fluidized bed expansion factor	3.8	
Particle diameter	2.0	mm
Particle material density	2550	kg/m^3
total particle weight	94	kg
Fouling factor	0.0355	$\frac{m^2.s.^\circ C}{kJ}$

Table 1. Specification of MSF/FBE pilot plant.

Stage	brine temp. [°C]	distill. temp. [°C]	log. temp. difference [°C]	terminal difference of heate [°C]	ditillate production [kg/s]	brine flow [kg/s]
1	86.05	85.54	8.042	2.983	0.13	5.08
2	72.54	72.30	8.332	3.317	0.12	4.95
3	59.53	59.15	8.676	3.724	0.11	4.83
4	47.09	46.38	9.031	4.221	0.10	4.72
5	35.32	34.67	9.534	4.810	0.09	4.62

Stage	porosity	heat transfer coefficient (inside) kJ/m^2.°C.s.	heat transfer coefficient (outside) kJ/m^2.°C.s.	heat transfer coefficient overall kJ/m^2.°C s.	heat transfer surface [m^2]	stage height [m]
1	0.797	8.02	9.02	3.26	11.3	0.8
2	0.797	7.25	8.58	3.05	11.3	0.8
3	0.797	6.48	8.09	2.82	11.3	0.8
4	0.806	5.70	7.57	2.58	11.3	0.8
5	0.825	4.93	7.02	2.32	11.3	0.8

Table 2. Specification of MSF/FBE pilot plant (continued).

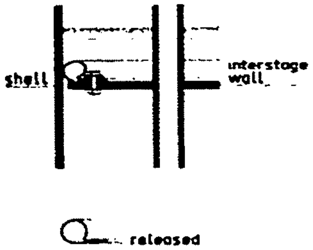

fig.5 rubber strip on the interstage wall

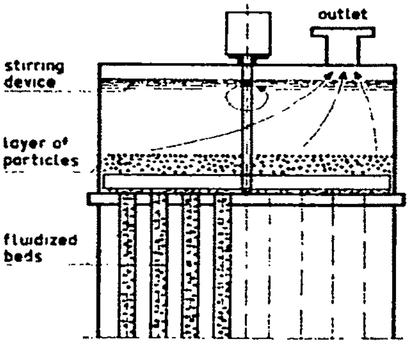

fig 6 stirring device in the outlet waterbox

Distillate production	450	tons/h
Steam consumption	50	tons/h
Specific heat consumption	246.0	kJ/h
Gain ratio	9	
Maximum brine temperature	113.0	°C
Cooling water inlet temperature	25.0	°C
Cooling water outlet temperature	35.0	°C
Blow-down temperature	34.5	°C
Total number of recirculation stages	32	-
Total number of heat reject stages	3	-
Concentration of the feed	18,500	ppm
Concentration of the blow-down	42,440	ppm
Condenser tube diameter, recirculation section and brine heater	90/10 1" x 20	CuNi iswg
Total number of condenser tubes, the recirculation section	10264	-
Porosity of the fluidized beds, the recirculation section	0.643	-
Particle material of the fluidized beds, recirculation section	glass (3000 kg/m^3) 4 mm diameter	
Condenser tube diameter, heat reject section	90/10 1¼"x 20	CuNi iswg
Total number of condenser tubes, heat reject section	7133	-
Porosity of the fluidized beds, heat reject section	0.70	-
Particle material of the fluidized beds, heat reject section	glass (3000 kg/m^3) 4 mm diameter	
Fouling factor of the brine heater, recirculation section and heat reject section	0.044	$\frac{m^2 s\,^oC}{kJ}$

Table 3. Specification of large low gain ratio MSF/FPE.

production cost breakdown	Conv. horiz. evap. (gain ratio 9:1) [Dfl./tonne]	MSF/FBE (gain ratio 9:1) [Dfl./tonne]
Capital costs	0.544	0.430
Operating costs:		
a) Steam	1.171	1.171
b) Electricity	0.386	0.289
c) Chemicals	0.056	0.050
d) Salaries, maintenance and administration	0.322	0.273
Total production cost	Dfl. 2.479 / tonne	Dfl. 2.213 / tonne

Table 4. Production cost comparison for conventional horizontal flash evaporator and low gain ratio MSF/FBE.

production cost breakdown	Conv. horiz. evap. (gain ratio 9:1) [Dfl./tonne]	MSF/FBE (gain ratio 18:1) [Dfl./tonne]
Capital costs	0.544	0.549
Operating costs:		
a) Steam	1.171	0.586
b) Electricity	0.386	0.405
c) Chemicals	0.056	0.050
d) Salaries	0.322	0.305
Total production cost	Dfl. 2.479 / tonne	Dfl. 1.895 / tonne

Table 5. Production cost comparison for conventional horizontal flash evaporator and high gain ratio MSF/FBE.