SEAWATER DEAERATION AT VERY LOW STEAM FLOW-RATES IN THE STRIPPING SECTION

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Abstract

The model discussed in a previous work [1] for the simulation of a deaerator to remove the oxygen from seawater in desalination plants has been improved. The conservative character of that model has been revised on the basis of the experimental industrial experience of FISIA Italimpianti S.p.A.

In particular, the behaviour of the stripping section packing when very low, or null, steam flow rates are used has been investigated.

An interesting interpretation of the stripping efficiencies has been used to demonstrate how the typical technical requirement of 0.02÷ 0.03 p.p.m. as maximum oxygen content in the outlet-water stream can often be achieved also operating without steam feeding in the stripping section.

Keywords*: seawater, desalination, deaerator, modelling, stripping.*

INTRODUCTION

In a previous paper [1], deaeration efficiency was studied and discussed with regard to oxygen removal from the seawater fed to desalination plants. A simulation code was set up to describe deaerator behaviour and parametric studies were carried out to analyse efficiency sensitivity on several process parameters.

The results suggested a possible deaerator over-design in some existing plants, as the calculated outlet oxygen concentration appeared to be acceptable also without the contribution of the stripping steam.

A particularly interesting feature of the study was that simulation data could confirm the FISIA Italimpianti experience of operating the stripping section without steam feeding while still meeting the requirements in terms of oxygen content at the deaerator outlet. However, the proposed description appeared to be too conservative, so that, in some cases, the simulation results for correctly-operated plants remained somewhat above the typical technical requirement of 0.02 ÷ 0.03 p.p.m. as oxygen content in the outlet-water stream. An apparently obvious conservative assumption was that a stripping section with no steam feed had no deaeration effect (i.e. a null efficiency). In the present paper further theoretical ideas have been developed to investigate the behaviour of the stripping section packing when very low, or null, steam flow-rates are used. The focus is on the dissipation of the potential energy of the water falling over the packing that, under adiabatic conditions, is converted into a minor vaporisation effect, so that the steam flow rising through the packing increases from the inlet value (that could be zero) at the bottom to the outlet value (always greater than zero) at the top.

If stripping efficiencies are calculated with reference to these very low and variable steam flow-rates, values are obtained which are still sufficient to bring the total outlet oxygen concentration to acceptable values in most of the possible operating cases.

The new model approach and its application to real desalination plants built by FISIA Italimpianti will be presented and discussed in the following chapters.

1. *THE DEAERATOR*

In multi-stage flash (MSF) desalination plants the make-up, representing the fresh seawater input to the distiller, is taken from the heated seawater discharged from the heat reject section and then sent to the deaerator.

Inside this equipment the sea water make-up oxygen content is significantly reduced to minimise corrosion problems in the heat recovery tube bundles section. Antifoam injection into sea water make-up upstream the deaerator is provided to prevent foam formation, which would badly affect the non condensable gases release.

In the early MSF plants the deaerator was a separate equipment (usually a vertical, cylindrical packed tower), while nowadays the usual design is a chamber directly connected to the last reject stage of the evaporator unit.

According to this solution, the deaerator is a rectangular cross section column, divided into an empty upper part, where the flash takes place, and a lower one, packed with Pall rings, where the stripping process occurs. The seawater is fed from the top, through a spray pipe whose design allows the liquid to expand and then flash only at the pipe-nozzle exit, being the pipe always kept under pressure. The drops of seawater fall onto a corrugated plate that evenly distributes the liquid over the packing.

In case of stripping-steam provision, another spray pipe is provided at the bottom of the deaerator. In this way the steam fed from below comes into contact with the counter-current liquid flow in the packing, enhancing dissolved gas removal.

The steam and released-oxygen flows are sent to the vacuum system through extraction pipes equipped with orifices.

2. *STRIPPING SECTION MODEL*

2.1 Steam-Rate Calculation

In order to theoretically simulate a deaeration stripping section, the water and steam flow rates spatial invariance is usually assumed.

This is certainly acceptable when reference is made to the usual values of the ratio V/L (that is about 10^{-3}). On the contrary if the ratio is much lower (about $10^{-6} \div 10^{-5}$), the variations of V can be better evaluated by the following total energy and mass-balance balances, both written with reference to steady state:

$$d\,[L\,(v_L^2 / 2 + g\,z + h_L) - V\,(v_V^2 / 2 + g\,z + h_V)] - q\,p\,dz = 0 \qquad (1)$$

$$d\,(L - V) = 0 \qquad (2)$$

In Equation (1) kinetic energies, related to rather low velocities, are negligible with respect to the liquid potential energy. Similarly, when V<<L, the potential energy and the sensible heat of the steam are negligible too.

Under these assumptions, the combination of Equation (1) and Equation (2) can be written as:

$$L\,cpL\,dT + (L\,g - q\,p)\,dz - \Delta H\,dV = 0 \qquad (3)$$

where only the sensible heat and the potential energy of the liquid, the heat losses and the vaporisation effects are retained as significant terms. Moreover, the coefficients $L\,c_{pL}$, $(L\,g - q\,p)$ and ΔH can be considered as constants. Equation (3) states that the dissipation of the potential energy of the water flowing onto the packing is converted into i) heating of the water, ii) heat losses and iii) evaporation.

The assumption of thermodynamic equilibrium between the two phases as expressed by the Clausius–Clapeyron equation for pure water:

$$dP / dT = P\,\Delta H\,M / R\,T2 \qquad (4)$$

and the evaluation of steam pressure drops through the packing, assuming linear terms are prevailing:

$$- dP / dz = \rho G\,g + kV \qquad (5)$$

allow to write Equation (3) as a differential equation in the variables V and z.

In Equation (4) the effect of the dissolved non-condensible can be neglected being the pre-flashed water handled in the stripping operation, so that the oxygen levels are typically of the order of 0.03 ppm or lower. On the other hand, Equation (5) is appropriate since we have to consider a range of very low steam velocities, where head losses are proportional to velocity (and then to the mass flow-rate at almost constant P and T) and comparable to the piezometric head losses.

Combining Equations (3), (4) and (5), and assuming:

$$\alpha = V / L \qquad \zeta = z / H \qquad \chi = k L / \rho G g \tag{6}$$

$$\alpha 1 = \rho G g R T2 cpL H / P M (\Delta H)2 \qquad \alpha 3 = g H / \Delta H \tag{7}$$

$$\alpha 4 = q p H / L \Delta H \qquad \alpha 2 = \alpha 3 - \alpha 4 \qquad \beta = \alpha 2 - \alpha 1 = \alpha 3 - \alpha 4 - \alpha 1 \tag{8}$$

the following differential equation is obtained:

$$d\alpha / d\zeta = - \chi \alpha 1 \alpha + \beta, \quad \text{with} \quad \alpha = \alpha o \quad \text{for } \zeta = 0 \tag{9}$$

In Equation (9) the change of the vapour-liquid ratio α=V/L along the dimensionless height ζ=z/H of the packing takes into account the following effects, listed according to their significance:

- an increasing term α_3 due to the dissipation of the potential energy of the liquid;
- a decreasing term α_4 due to the heat losses;
- a decreasing term α_1 due to the heating of the liquid due to an increase in the total pressure;
- a decreasing term $\chi\ \alpha_1\ \alpha$, proportional to α, due to the head losses.

The first three terms are algebraically put together in the term β, where α_3 is prevailing, while the last term, $\chi\ \alpha_1\ \alpha$, is almost negligible (see the following discussion and §3).

The solution of Equation (9) leads to:

$$\alpha = (\beta / \chi\ \alpha_1)\ [1- \exp(-\chi\ \alpha_1\ \zeta)] + \alpha_o \exp(-\chi\ \alpha_1\ \zeta) \tag{10}$$

If reference is made to the typical conditions of desalination stripping, the orders of magnitude of the parameters in Equation (10) can be evaluated. Usually, as discussed in § 3, it is possible to assume $\chi\alpha_1 << 1$, so that Equation (10) can be reduced to the simple linear trend

$$\alpha = \beta\ \zeta + \alpha_o\ (1 - \chi\ \alpha_1\ \zeta) \tag{11}$$

which highlights that the steam flow rate tends to decrease as it rises only when α_3 is lower than $\alpha_4 + \alpha_1$ (that is $\beta < 0$) or when a relatively high steam-rate is fed to the bottom, so that:

$$\alpha o > \beta / \chi\ \alpha_1 \tag{12}$$

On the contrary, when α_3 is greater than $\alpha_4 + \alpha_1$ (that is $\beta < 0$) and α_o is of the order of β or lower, the steam flow-rate, although very low, tends to increase. In such instances, moreover, head losses are shown to be negligible in comparison to piezometric head. So, it simply results:

$$\alpha = \beta\zeta + \alpha o \tag{13}$$

When the stripping section is operated without using stripping steam ($\alpha_o = 0$), Equation (11) or (13) states that a steam flow-rate, of the order of some parts per million of the water flow-rate, may still rise from the top of the packing.

Above the packing, additional vapour, rising from the distribution plate and from the void zone just behind the flash zone, will be added. Then the flash vapour, being one or two orders of magnitude greater that the spontaneous stripping vapour, will also be added.

1.2. Stripping Efficiency Calculation

In the flash section, the oxygen content of the seawater is usually reduced to less than 0.1 ppm. In several cases the oxygen content can also be lower, so that the subsequent stripping operation could be unnecessary. In other cases, the stripping section is required to guarantee the reduction of oxygen to below 0.02 ÷ 0.03 p.p.m. [1]. This goal is quite easy to achieve if some steam is fed to the stripping section. Generally, a satisfactorily accurate evaluation of the stripping efficiency can be obtained from the equation:

$$1 - E = x_F / x_1 = \exp(-N) \qquad (14)$$

where N is the number of transfer units of the packing section. For these applications N is almost independent from the vapour flow-rate, with values typically around 2.

Equation (14) is correct when the vapour flow rate is constant through the packing ($\alpha_o >> \beta$) and the oxygen content of the vapour is negligible throughout the packing ($y / m << x$). In the range of low and reasonably constant vapour flows, Equation (14) can be written:

$$(1 - E)^{-1} = x_1 / x_F = (e^{N(1-s)} - s) / (1-s) \qquad \text{with } s = 1 / \alpha_o m \qquad (15)$$

taking into account also the effect of the progressive vapour saturation.

This equation tends to become inappropriate when the vapour rate, in the range of very low values, changes according to Equation (11).

As a matter of fact, in the mass balance equations for the packing

$$L\, dx / dz = a\, K_x\, S_t\, (x - y / m) \qquad (16)$$

$$L\, (x - x_F) = V\, y \qquad (17)$$

the vapour flow rate V has now to be considered as a variable, according to Equation (13).

In terms of dimensionless variables Equations (16), (17) and (13) lead to the following differential equation:

$$dx / d\zeta = N\, [x - (x - x_F) / m\, (\alpha o + \beta\, \zeta)] \qquad (18)$$

with $\zeta = 0$ for $x = x_F$ and $\zeta = 1$ for $x = x_1$

where the number of transfer units of the packing is defined as:

$$N = a\, K_x\, S_t\, H / L \qquad (19)$$

Equations (18) and (19) can be solved in terms of a definite integral using a series expansion:

$$x1 / xF = 1 + (N + \varphi)^{-\nu} e^{N+\varphi} \int_{\varphi}^{N+\varphi} \zeta^{\nu} e^{-\zeta} d\zeta =$$

$$1 + (N + \varphi)^{-\nu}\, e^{N+\varphi}\, \Sigma_{k=0,\infty}\, (-1)^k\, [(N + \varphi)^{k+\nu+1} - \varphi^{\,k+\nu+1}] / k!\,(k + \nu + 1) \qquad (20)$$

where

$$\varphi = N\, \alpha_o / \beta \qquad\qquad \nu = N/m\beta \qquad (21)$$

If $\alpha_o >> \beta$ Equation (20) simplifies to Equation (15), while, for $\alpha_o = 0$ it is:

$$x_1 / x_F = 1 + e^N\, \Sigma_{k=0,\infty}\, (-1)^k\, N^{k+1} / k!\,(k + \nu + 1) \qquad (22)$$

For N = 2 and m β = 1.5 (ν = 2) the summation in Equation (22) is about 0.16, so that a ratio x_F / x_1 of the order of 0.46 is obtained, which is significantly lower than the requested one (about 0.67). This is also true in the case of a significant uncertainty on the parameters β and N.

In other words, according to Equation (20), the decrease in the steam inlet is accompanied only by a slight decrease in the efficiency and this is also true when the steam supply is led to zero, on the condition that a very low flow of rising steam is still present as a consequence of energy dissipation.

The sign of β could be a matter of discussion. With negligible head losses and an almost adiabatic operation, β is positive ($\beta \cong 8.3\ 10^{-6}$; β m $\cong$ 1.5 in our calculations).

Now, the assumption of negligible head losses is realistic: in a standard operation with α_o of the order of $10^{-3} - 10^{-4}$, we have losses of only a few mm H_2O, so that $\chi\, \alpha_1 << 1$.

On the other hand, heat losses towards external air could be great enough to change the sign of β. To compare the potential energy of the water with the heat losses the parameter:

$$\gamma = \alpha_4 / \alpha_3 = q\, p / g\, L \qquad (23)$$

can be used.

If $\gamma << 1$ the effect of thermal dispersions is negligible (adiabatic operation). On the contrary, and especially if $\gamma > 0.3$ as in our reference calculation, the effect of thermal dispersion can be so important that condensation rather than evaporation can occur throughout the packing. Condensation phenomena will surely be present when $\gamma > 1$.

To better calculate the value of each parameter, as well as to validate the overall model described here, verifying the effective capability of the stripping section to work efficiently also without a steam feed. The parameters characterising the operation of real plants have been taken into account

3. RESULTS

Fisia Italimpianti is involved in desalination plant design since the 70's. The company holds today the leadership in Multi Stage Flash technology and has developed fast-track methods to design, build and commission plants in the shortest possible time. This capacity has guaranteed good market success, enabling Fisia Italimpianti to acquire several huge projects in the Gulf area. [2].

In the following the above introduced theoretical hypothesis will be applied to two real plants, extremely different as far as the size is concerned.

The first one, built at Ruwais (Abu Dhabi - UAE) and in operation since 2000, consists of two units, each capable to produce 625 m^3/h of distillate water is

The second one, Jebel Ali L (Dubai – UAE) and now under commissioning, consists of 5 units, each capable to produce 2650 m^3/h of distillate water.

Pictures of the two plants are reported in the next, respectively in Figures1 and 2.

Fig. 1: FISIA Italimpianti desalination plant in Ruwais (Abu Dhabi).

Fig. 2: FISIA Italimpianti desalination plant in Jebel Ali L.

The specific geometrical data of the two plants deaerators together with the corresponding calculated model parameters are reported in Table 1.

Data	Ruwais	Jebel Ali L
Width [m]	7.8	22.8
Length [m]	2.5	3.5
Packing Height [m]	2.1	2
α_1	$3.94\ 10^{-8}$	$4.79\ 10^{-8}$
α_2	$8.59\ 10^{-6}$	$8.14\ 10^{-6}$
$\beta = \alpha_2 - \alpha_1$	$8.55\ 10^{-6}$	$8.10\ 10^{-6}$
χ	$3.2\ 10^{3}$	$3.1\ 10^{3}$
$\chi\ \alpha_1$	$1.28\ 10^{-4}$	$1.49\ 10^{-4}$
$\beta / \chi\ \alpha_1$	$6.69\ 10^{-2}$	$5.42\ 10^{-2}$

Tab. 1: Plant deaerator data.

These values confirm the assumption $\chi\alpha_1 << 1$, allowing the reduction of Equation (10) to the simple linear trend, which is the basis the model is established upon.

An interesting feature is that this is true for plants of extremely different sizes, so giving a general validity to the discussion.

Dealing with the value of γ, for example with reference to Ruwais plant, we can assume $\Delta T = 20$ °K, p = 20 m, and L = 500 kg/s (minimum), while a largely conservative assumption on thermal exchange can be done taking into account only the external coefficient of free convection between the atmospheric air and vertical surfaces (< 4 W/m^{2}K). So, γ results to be lower than 0.3, well supporting the considerations developed on stripping efficiency in the absence of a steam feed (§ 2).

Figure 3 shows an analysis carried out using the specific operating values of the two above-mentioned plants (Table 2) and an inlet oxygen weight-fraction in the liquid (x_1) of 0.03 ppm. The final weight-fraction (x_F) was calculated as a function of the ratio α_0 between the steam and the liquid flow rates, according to Equation 20.

It can be observed that both plants behave in a very similar was despite their very different dimensions (see Table 2): the constraint of 0.02 ppm is fulfilled in both cases without feeding steam to the packing section. In addition, an asymptotic efficiency value is achieved with steam flow-rates much lower than the operating values, which usually involve $\alpha_0 \approx 10^{-3}$.

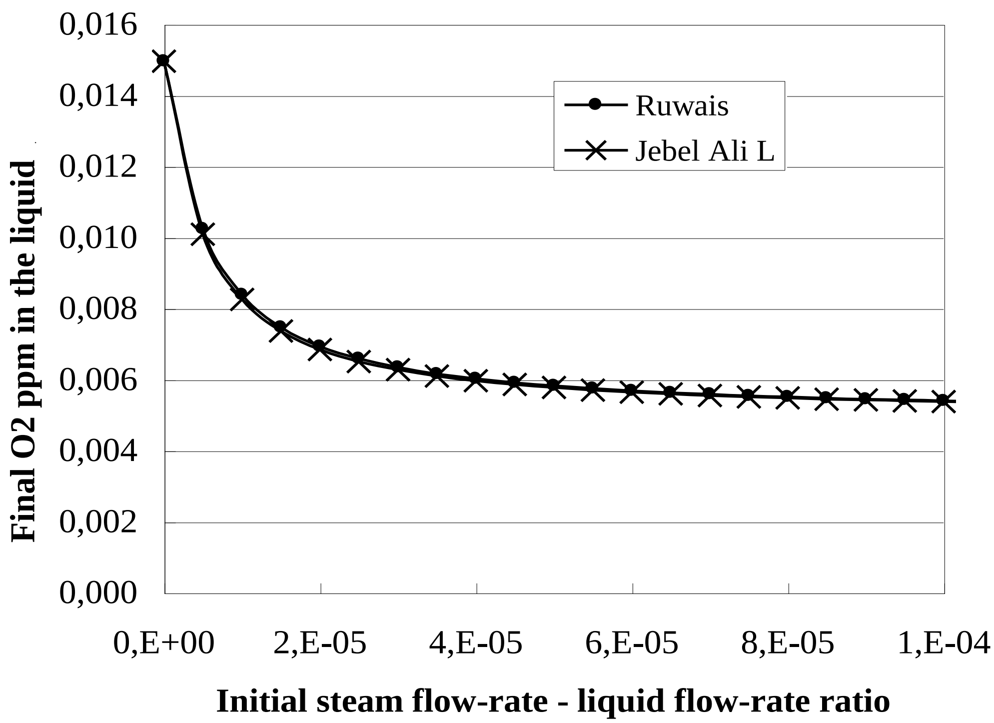

Figure 3: Calculated outlet oxygen weight fraction (x_F) as a function of the ratio between the steam and the liquid feed flow rates (α_0).

The results shown in Fig. 3 have been calculated taking into account the effective number of packing transfer units in the considered plants, that is 1.8.

Figure 4 shows a parametric analysis of how the efficiency of the packing section is affected by the number of transfer units and the steam flow-rate, for a reference value of $\beta = 8.3 \cdot 10^{-6}$ and corresponding constant values of $\varphi / \alpha_0 = 2.4 \cdot 10^5$ and $\nu = 2$. The results also highlight a significant residual efficiency of the packing also without stripping steam ($\alpha_0 = 0$) and an asymptotic trend for very low steam rates ($\alpha_0 > 10^{-4}$).

According to these calculations, the studied deaerators seem to be slightly oversized also when very low, or null, steam flow rates are used. This conclusion suggests the usefulness of a further detailed validation of Equation 20 based on experimental data in order to allow a final optimised packing section and stripping steam flow design.

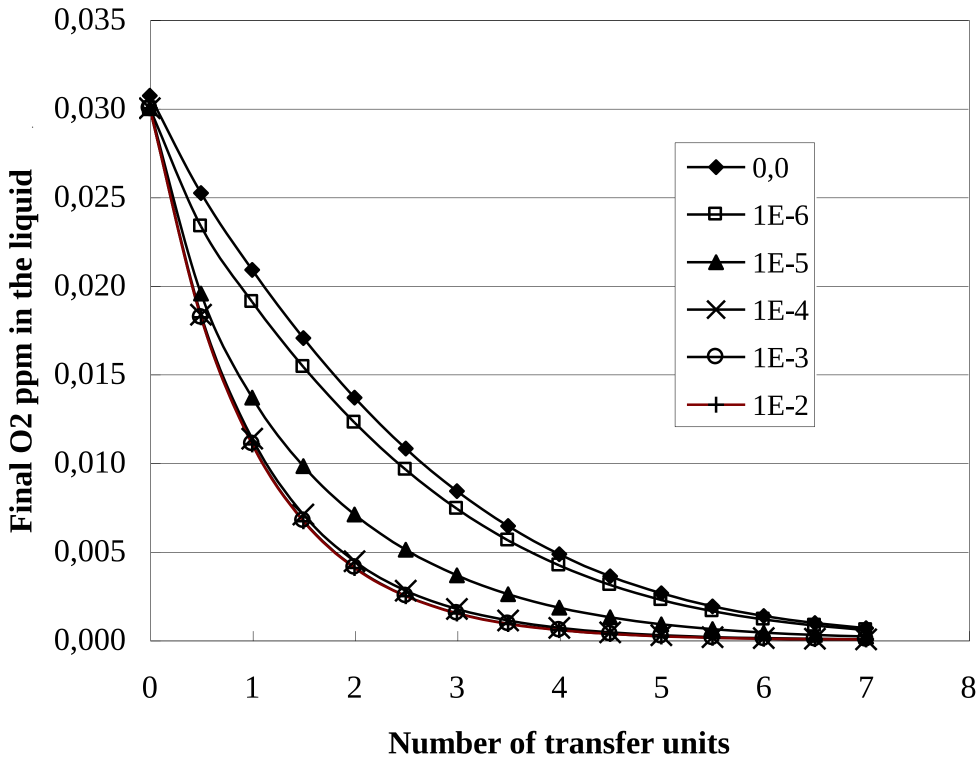

Figure 4: Outlet oxygen weight fraction (x_F) calculated as a function of the number of packing transfer units (N) at different ratios (α_0) of steam and liquid flow rates.

At the present, the efficiency of the stripping section without steam feeding is experimentally confirmed by the most recent FISIA plants. Among those also Shuweihat (Abu Dhabi) plant: its six desalination units are the biggest ones in the world and work without an external steam feed, still always respecting the requested make up oxygen content at deaerator outlet.

CONCLUSIONS

The paper proposes a new interpretation of the phenomena occurring in the deaerator stripping section of a seawater desalination plant with very low vapour flow rates.

In particular, with respect to a previously developed model assuming efficiency equal to zero in absence of steam feed, less conservative hypothesis have been now taken into account.

Assuming the possibility of steam release inside the deaerator vessel by evaporation process, this quantity often results is sufficient to guarantee proper oxygen concentration in the outlet stream.

This interpretation is reasonable for standard operating condition of specific deaerators. Experimental data collected on recent Fisia Italimpianti plants running without additional steam feed confirm in any case a good deaerator efficiency.

NOTATIONS

A	packing area per unit volume	[m^2 / m^3]
c_p	specific heat	[J/kg K]
g	gravity acceleration	[m/s^2]
h	specific enthalpy	[J/kg]
H	packing height	[m]
k	headloss constant – see Equation (5)	[$1/m^2s$]
K_x	overall mass transfer coefficient in terms of the liquid phase	[kg/m^2s]
L	liquid flow rate	[kg/s]
m	vapour – liquid distribution coefficient for oxygen	
M	molecular weight	[kg/kmol]
N	number of liquid phase transfer unit	
p	deaerator perimeter	[m]
P	pressure	[N/m^2]
q	heat losses	[J/m^2s]
R	gas constant	[J/kmol K]
s	see Equation (15)	
S_t	deaerator cross section	[m^2]
T	temperature	[K]
v	velocity	[m/s]
V	steam flow rate	[kg/s]
x	weight fraction of oxygen in the liquid phase	
y	weight fraction of oxygen in the vapour phase	
z	vertical spatial coordinate	[m]

Greek:

α, α_1, α_2, α_3, α_4	see Equations (6), (7), (8)	
β	see Equation (8)	
γ	see Equation (23)	
ΔH	heat of vaporisation [J/kg]	
φ	see Equation (21)	
ζ	dimensionless spatial coordinate – see Equation (6)	
υ	see Equation (21)	
χ	see Equation (6)	
ρ	density	[kg/m^3]

Subscripts:

L liquid
V steam
F final for liquid (deaerator bottom)
0 deaerator bottom
1 initial for liquid (packing inlet)

REFERENCES

[1] E. Ferro, B. Bosio, P. Costa, E. Ghiazza, *Modelling of Flash and Stripping Phenomena in Deaerators for Seawater Desalination*, Desalination, **142**, 2002, 171-180.

[2] FISIA ITALIMPIANTI main references for MSF plants:

Plant	**Owner**	**Capacity m^3/d (MIGD)**	**Commissioning**
Jebel Ali “L” (Dubai UAE)	DEWA	317.000 (70)	2005
Shuweihat (Abu Dhabi UAE)	S. C. I. P. CO.	454.000 (100)	2004
Ras Laffan (Quatar)	AES/KAHRAMAA	181.600 (40)	2004
MIRFA (Abu Dhabi UAE)	A.D.E.W.A.	102.000 (22.5)	2002
Jebel Ali “KII” (Dubai UAE)	D.E.W.A.	181.600 (40)	2003
Jebel Ali “KI” (Dubai UAE)	D.E.W.A.	90.960 (20)	2001
Jebel Ali “G” (Dubai UAE)	D.E.W.A.	34.080 (7.5)	2000
RUWAIS (Abu Dhabi UAE)	ADNOC	30.000 (6.6)	2000
Al Hidd (Bahrein)	M.E.W	136.000 (30)	1999
Al Taweelah B (Abu Dhabi UAE)	W.E.D.	345.000 (76)	1995