DESIGN OF *AN* ABSORPTION COLUMN FOR STEAM WASHING

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ABSTRACT

The presence of HCl in geothermal steam may cause generalized corrosion in trasportation pipelines and stress corrosion cracking of turbine blades. This has led to the common practice of steam washing with an alkaline solution when the chloride ion exceeds **15-20** ppm. In the present paper the results obtained in a pilot plant operating with simulated fluids are used to design an absorption column operating in cocurrent flow. The design procedures which have been developed take into consideration the simultaneous heat transfer process and an attempt is made to minimize the power losses related to the washing operation.

1. INTRODUCTION

In geothermal power plants steam washing to eliminate acid compounds, particularly HCI, is usually done by injecting a cold solution of 0.2-0.3 N soda directly into the steam pipeline. The washing process is followed by the abatement of entrained droplets, first with a centrifugal separator and then with an impact separator.

In the case of good separating efficiencies, the main negative consequence of the washing operation consists in a power loss associated with steam desuperheating. In addition, there are corrosion problems in the pipes used for the washing operation.

The working hypothesis followed in this research activity is to perform the washing operation at a sufficiently low steam velocity to avoid the formation of dispersed droplets in appreciable amounts. For this purpose it is possible to use a structured packing column operating in cocurrent. This type of column is characterized by high packing surface areas, and low pressure drops. As we shall see later on in the work, the possibility of operating in an intermediate range of steam velocity (U_V = 6-8 m/s) permits sizable reductions in equipment dimensions without appreciable decreases in HCI removal efficiency for a given column height. It also allows to operate without complex separating systems, since the column itself can act as separator for any solid particles or droplets entrained by the steam fed into the washing plant.

Besides appreciable savings on plant costs, the adoption of structured packing columns for steam washing allows significant energy savings, linked mainly to a reduction of the flow rate of wash liquid. In addition, smaller savings can be made in connection with the lower pressure drops and lower desuperheating that it is possible to obtain or achieve in these apparatuses.

The energy savings connected with a more careful choice of the operating conditions at which the washing is performed were also evaluated in this study. In this regard it should be noted that even if the equipment and methods currently in use are maintained, it is possible to supply the washing liquid at a temperature near the boiling point relative to the plant operating pressure. This can be done either by sending to the washing the condensate formed in the first turbine stages, if it is sufficient, or utilizing the liquid coming out of the washing system to preheat the incoming liquid. The energy savings connected with these process modifications are significant, at least as much as the savings that can be made by reducing the flow rate of wash liquid. On the other hand the plant modifications required, even in the present configuration, are reduced to the installation of a small **(8-10** m^2) heat exchanger that must operate with noncorrosive, weakly alkaline solutions.

For the evaluation of the energy transfer connected with the washing operation and of the HCI removal efficiency, it was necessary to describe the behaviour of a column operating with simultaneous transfer of mass (HCI, steam), sensible heat (from superheated steam) and latent heat. The complexity of the transfer phenomena that must be considered is attenuated by the fact that the available data and correlations show that a large part of the resistances to mass and heat transfer are in the gas phase. This makes it possible to develop the design using only the experimental results obtained by Viviani **(1993).** These data regard the determination of the mass transfer coefficient (HCI) in columns with structured packing in the case of cocurrent flow of air-water mixtures.

2. PROCESS ANALYSIS

In washing superheated steam with alkaline solutions one has to do with interacting mass and heat transfer phenomena in which the mass transfer (absorption of HCI) takes place parallel to the condensation/evaporation of water vapor and the heat transfer regards both sensible heat and latent heat in the presence of noncondensables. In this work for the sake of convenience it was assumed that all the resistances to mass and heat transfer are contained in thin **films** at the two sides of the liquid-vapor interface. This hypothesis is the basis of the two-film theory that will later be used to determine the transfer coefficients.

2.1 HCI transfer

The resistance to HCI transfer is exclusively in the gas phase, since the high reactivity of the acid in aqueous solutions of NaOH leads to a **null** value of HCI concentration at the interface. It must be noted that the diffusion of HCI in the gas film occurs parallel to the diffusive/convective flow of the steam and other gases (assumed to be noncondensable for the sake of simplicity). The HCI transfer can thus be described by the equation

$$n_A = \rho_G(-D_A \frac{d\omega_A}{dy} + \omega_A v) \qquad (1)$$

where n_A is the mass flow of HCl towards the interface, D, the diffusion coefficient of the HCl in the steam-CO, mixture, ω_A the weight fraction of HCl, ρ_G the density of the steam phase-noncondensables and lastly v the flow (by volume) of steam and HCl towards the interface. Taking into account that the transfer phenomenon occurs in stationary conditions, we have

$$\frac{dn_A}{dy} = 0 \qquad (2)$$

Eq. 2, with the boundary conditions

$$y = o : \quad \omega_A = 0$$

$$y = s : \quad \omega_A = \omega_A^b \qquad (3)$$

where ω_A^b is the average concentration of HCl in the steam at the given axial position, can be easily integrated in the case v = constant, ρ_G = constant. As we shall see later on, only the first of these two assumptions is verified with certainty, while the second proves to be an acceptable simplification. The result of the integration can be set in the form

$$n_A = k_A \, \rho_G \, \omega_A^b \frac{\alpha_A}{1-\exp(-\alpha_A)} \qquad (4)$$

where k, is the transfer coefficient relative to HCl, which in the case of diffusion in the absence of convective flow, v=0, is equal according to the two-film theory to

$$k_A = \frac{D_A}{\delta} \qquad (5)$$

and α_A is defined as

$$\alpha_A = \frac{v}{k_A} \qquad (6)$$

It should be noted that, in general, the flow v can be either positive or negative and in both cases the HCl transfer can increase or decrease significantly.

2.2 Gas-phase steam transfer

In practice the mass flow v that appears in Eq. 1 corresponds only to the steam flow, since on the one hand HCl is present only in traces, and on the other it can be assumed that the net flow of CO, towards the gas/liquid interface is negligible with respect to the steam flow. However, the presence of noncondensables in the vapor phase causes a resistance to the condensation/evaporation of steam which must be appropriately estimated. For this purpose, it can be pointed out that for all the species present in the gas film the transport equation is

$$n_i = j_i + \rho_G \, v \, \omega_i \qquad (7)$$

where n_i, j_i and ω_i are the global flow, the diffusive flow and the fraction by weight of the species i. In practice, $\rho_G v = \sum n_i \cong n_W$ and from Eq. 7 one gets

$$v(1-\omega_W) = -D_W \frac{d\omega_W}{dy} \qquad (8)$$

for the steam and

$$0 = -D_C \frac{d\omega_C}{dy} + v\,\omega_C \qquad (9)$$

for the CO„ where D_C, the diffusivity of CO, in the steam, is equal to D_W, diffusivity of the steam in CO,.

By integrating these equations in the gas film with the boundary conditions

$$y = 0: \quad \omega_W = \omega_W^i, \quad \omega_C = 1-\omega_W^i \qquad (10)$$

$$y = \delta: \quad \omega_W = \omega_W^b, \quad \omega_C = 1-\omega_W^b$$

one gets

$$v = k_W \ln\frac{1-\omega_W^i}{1-\omega_W^b} \qquad (11)$$

with $k_W = \frac{D_W}{\delta}$

It should be noted that at the steam-liquid interface equilibrium is admitted. It thus follows that the steam fraction by weight at the interface, ω_W^i, is linked to the interface temperature by a relation of thermodynamic equilibrium

$$\omega_W^i = f(T^i) \qquad (12)$$

Eq. 12 can be derived from thermodynamic data, taking into account the presence of dissolved salts in the wash liquid.

2.3 Heat transfer

In the case of heat transfer between the superheated steam and the aqueous solution, the controlling transfer coefficient is in the vapor phase, although in this case the liquid phase resistance cannot be neglected. In the gas film the heat flux, q, can be expressed as

$$q = -\alpha_G \frac{dT_G}{dy} + H_G \, n_W \qquad (13)$$

here α_G is the thermal conductivity coeficient, T_G the temperature, and H_G the enthalpy of the steam. If it is taken into account that at steady state q and v are constant, and further that

$$dH_G = c_p \, dT_G \qquad (14)$$

Eq. 13 can be easily integrated with the boundary conditions

$$y = 0 \quad H_G = H_G^i \qquad (15)$$

$$y = \delta \quad H_G = H_G^b$$

This gives

$$q = \frac{H_G^b - H_G^i \exp(-\alpha_H)}{1-\exp(-\alpha_H)} n_W \qquad (16)$$

with

$$\alpha_H = \frac{\bar{c}_p \, n_W}{[IC} \qquad (17)$$

where $h_G = \alpha_G/\delta$ is the sensible heat transfer coefficient for gas-side convection in the absence of simultaneous mass flow.

If we suppose

$$H_G = H_G^i + \bar{c}_p(T_G - T^i) \qquad (18)$$

where $\bar{c}_p$ is the mean specific heat in the temperature range considered, Eq. 16 can be rewritten as

$$q = h_G(T_G^b - T^i)\frac{\alpha_H}{1-\exp(-\alpha_H)} + H_G^i \, n_W \qquad (19)$$

The liquid phase heat flux can be more simply expressed as

$$q = h_L(T^i - T_L^b) + H_L^i \, n_W \qquad (20)$$

where h_L is the liquid phase heat transfer coefficient.

From the equality of Eqs. 19 and 20 we get

$$h_G(T_G^b - T^i)\frac{\alpha_H}{1-\exp(-\alpha_H)} = h_L(T^i - T_L^b) - \lambda \, n_W \qquad (21)$$

since $H_G^i - H_L^i = \lambda$, with λ being the latent heat at the interface temperature. Eq. 21 clarifies how the sensible heat deriving from the steam desuperheating is spent to heat the liquid and partly vaporize it (negative n_W). In the case that the temperature difference $T^i - T_L^b$ is high (feed liquid cold), n_w is positive. Generally it can be stated that in the top part of a column operating in cocurrent there is steam condensation and in the bottom part there is evaporation.

Eqs. **4**, **11**, **19** and **20** make it possible to calculate, for each position inside a transfer apparatus, the mass and heat flows between the vapor and liquid phases. The use of these equations requires knowledge of the transfer coefficients, in addition to the surface area across which the transfer occurs. It will be noted that, while the equations considered in this section are valid for any equipment used in steam washing, the values assumed by the transfer coefficients depend on the particular apparatus that is used.

3. WASHING COLUMN DESIGN

The physical model developed to describe the elementary mass and heat transfer phenomena was applied to the case of cocurrent washing in a packed column equipped with structured packing. The column was discretized into a series of elementary sections in which the physical properties of the fluids were supposed constant. For each of these sections the mass flow and energy flux were estimated and the corresponding balances were performed. The pressure drops were also calculated. The estimation of the pressure drops and the calculation of the holdup, effective velocities, wetted area and transfer coefficients were based on the work by Bravo *et al* **(1991)** and Viviani **(1993).** The design was developed on the basis of the following hypotheses:

- The gas is constituted by superheated steam containing only CO, (7%) and HCl **(20** ppm).
- The HCl reaction in the liquid phase is instantaneous and its concentration at the interface is null. Absorption of CO, is neglected.
- The reaction has negligible thermal effects.
- The boiling point elevation of the liquid phase due to the presence of ions in solution is neglected.
- The physical properties of the fluids were estimated as a function of pressure and temperature. The properties of the mixture were estimated as weighted average of the properties of the single components, neglecting the effects of mixing.

3.1 Estimation of transfer coefficients

Some semiempirical relations are available for the calculation of transfer coefficients for washing operations performed in towers with structured packing elements. In particular, the relation proposed by Bravo ***et al*** **(1991) for** estimating the gas phase mass transfer coefficient was experimentally verified by Viviani **(1993)** for the system air-HCl-aqueous solution of NaOH:

$$Sh = \frac{k_G\, s}{D_G} = 0.054\left[\frac{u_G\,\rho_G\, s}{\mu_G}\right]^{0.8}\left[\frac{\mu_G}{\rho_G\, D_G}\right]^{0.33} \tag{22}$$

where **Sh** is the Sherwood number, s the characteristic packing dimension, D_G the diffusivity of the component considered in the gas phase and u_G is the relative velocity between the gas and liquid. In practice this last value is very close to the effective gas velocity, which is determined once the column diameter is chosen and the fraction of area occupied by the gas is estimated.

The liquid holdup, ε_L, can be estimated with the correlation proposed by Bravo, Rocha and Fair (**1991**)

$$\varepsilon_L = \left[4\frac{F_T}{s}\right]^{2/3}\left[\frac{3\,\mu_L\, u_{SL}}{\rho_L\, g\, \varepsilon \sin\theta}\right]^{1/3} \tag{23}$$

where F_T is expressed as

$$F_T = \frac{29.12(We_L\, Fr_L)^{0.15}\, s^{0.359}}{\mathrm{Re}_L^{0.2}\;\varepsilon^{0.6}(\sin\vartheta)^{0.3}(1-0.93\,\cos\gamma)} \tag{24}$$

Re_L, We_L and Fr_L are the Reynolds number, Weber **number** and Froude number, defined as $\mathrm{Re}_L=\rho_L u_{SL} s/\mu_L$ $\mathrm{We}_L=\rho_L u_{SL}^2 s/\sigma$ $\mathrm{Fr}_L=u^2_{SL}/gs$ ε is the void fraction of the packing, θ is the corrugation angle of the packing, σ, μ_L, γ are respectively the surface tension, viscosity and contact angle of the liquid, and U_{SG} and U_{SL} are the vapor and liquid superficial velocities.

The effective velocities of the gas and liquid, U_G^e and U_L^e, are calculated as:

$$u_G^e = \frac{u_{SG}}{(1-\varepsilon_L)\,\varepsilon \sin\theta} \tag{25}$$

$$u_L^e = \frac{u_{SL}}{\varepsilon\, \varepsilon_L \sin\theta} \tag{26}$$

Along with the transfer coefficient, for absorption column sizing it is necessary to know the effective interfacial area. According to the **same** authors (Bravo ***et al*, 1991)** the specific area of the packing (surface per unit volume) can be estimated by means of the correlation

$$\frac{a}{a_g} = 0.350\cdot\frac{29.12\cdot(We_L\cdot Fr_L)^{0.15}\cdot s}{\mathrm{Re}_L^{0.2}\cdot\varepsilon^{0.6}\cdot(\sin\vartheta)^{0.3}\cdot(1-\cos\gamma)} \tag{27}$$

where a_g is the geometric area and the numerical coefficient **(0.350)** is relative to the particular packing considered (FLEXIPAC **II**, Sulzer, see Bravo *et al*, **1991**),

The experimental measurements made by Viviani **(1993) on** an ambient temperature system confirm the validity of Eqs. **22** and **23** for the calculation of the gas-side mass transfer coefficient.

Knowledge of the gas-side mass transfer coefficient, relative to the transfer of HCl, makes it possible to determine the **mass** transfer coefficients relative to water vapor and CO, using the appropriate diffusivity values. In addition, at least in the gas phase, the analogy between mass transfer and heat transfer can be considered valid, by which it can be assumed that for the Schmidt number ($\mu_G / D_G\, \rho_G$) equal to the Prandtl number ($c_p\, \mu_G / \alpha_G$) we have

$$Nu = \frac{h_G S}{\alpha_G} = Sh \tag{28}$$

where Nu is the Nusselt number, or, taking Eq. **22** into account

$$\mathrm{Nu} = 0.054(\mathrm{Re}_G)^{0.8}(\mathrm{Pr})^{0.33} \tag{29}$$

Analysis of the elementary phenomena of heat **and** mass transfer led to identification of the gas phase coefficients as controlling coefficients. The generally accepted hypothesis of analogy between **mass** and heat transfer makes it possible to use the experimental measurements **made** by Viviani **(1993)** to design sufficiently accurately a column with structured packing, with the advantage that, thanks to the analogy between these two physical phenomena, any possible uncertainty in the estimate of one of the coefficients is reflected on the other in a proportional way, It can thus be concluded that the estimation of the transfer processes in a column can be assumed to be reliable at least with regard to the relative size of the mass and heat transfer.

Also the liquid phase mass and heat transfer coefficients can be estimated on the basis of knowledge of one of the two by applying the principle of analogy between transfer process. In this case, however, it must be noted that because of the large difference between the Prandtl and Schmidt numbers, the validity of the analogy is less forceful. It must also be remembered that the liquid phase transfer coefficient has no influence at all on the HCl transfer and is only of limited significance for what regards the heat transfer.

In the literature there seem to be practically no correlations relating to direct contact heat transfer in columns with structured packing operating in cocurrent. To estimate the heat transfer coefficient h_L it is therefore necessary to go back to the analogy principle and arrive at it from the correlation for the mass transfer coefficient proposed by Bravo *et al* (1991) and based on the penetration theory:

$$k_L = 2 \cdot \sqrt{\frac{D_L\, u_L^e}{\pi\, s\, \varepsilon}} \tag{30}$$

In this case the principle of analogy between heat and mass transfer leads us to determine h_L as

$$h_L = k_L \sqrt{\frac{\bar{c}_p\, \rho_L\, \alpha_L}{D_L}} \tag{31}$$

where α_L is the thermal conductivity in the liquid

3.2 Steam velocity in the column

A packed column fed in cocurrent makes it possible to operate with high steam velocities since the only limits to this velocity are set by possible constraints on the pressure drops and by the necessity of not creating a finely dispersed spray in the outflowing steam. From the standpoint of heat and mass transfer, the efficiency of the operation is determined by the values of the transfer coefficient, the liquid area exposed to the steam and the contact times. Of these parameters, the transfer coefficients increase with the steam velocity raised to an exponent of about 0.8; the transfer surface stays approximately constant; the contact time decreases inversely with the increase of u_{SG}. As found also by Viviani (1993), the overall result is that the HCl removal efficiency for a given column height is weakly influenced by the steam velocity.

On the basis of these considerations, the steam velocity in the column can be chosen by first taking into account the size of the droplets generated by the steam and then checking the value of the resulting pressure drops. The aim of this type of dimensioning is to avoid excessive droplet entrainment by the steam, which would necessitate the use of considerably large separators.

The size of the droplets generated by the steam flow and their volume distribution can be determined by resorting to correlations available in the literature relating to droplet entrainment in the case of cocurrent flow of a gas on a thin film of liquid. The most recent of these correlations is by Ambrosini et *al* (1991) and is based on a data bank of considerable size which includes measurements obtained by varying the physical properties and characteristic dimensions of the conduit over a very wide range. According to this correlation Sauter's mean diameter, d_{32}, is given by the expression

$$\frac{d_{32}}{m} = 22.0\left(\frac{\sigma}{\rho_G\, f_i\, {u_G^e}^2\, m}\right)^{0.5}\left(\frac{\rho_G}{\rho_L}\right)^{0.83} \exp\left(0.60\frac{G_{LE}}{\rho_L {u_G^e}^2}\frac{s}{d_{32}}+\frac{99.0}{We_G}\right) \tag{32}$$

where G_{LE} is the mass flow of liquid transported by the gas, f_i is the friction factor on the surface of the liquid film, m is the thickness of the liquid film, and $We_G = \rho_G {u^e_G}^2 s/\sigma$ is the Weber number based on the steam velocity. Andreussi and Azzopardi (1983) found that, in first approximation, the volume distribution of the droplets entrained by the gas can be determined by the equation

$$V_d = \exp\left[-\left(\frac{d_p}{b \cdot d_{32}}\right)^n\right] \tag{33}$$

where V_d is the volume fraction of droplets of dimensions greater than d_p. The two parameters that characterize this distribution, b and n, are very weak functions of the operating conditions and can both be assumed equal to **2**.

Eqs. 32 and 33 make it possible to determine, for given operating conditions and for the flow configuration encountered in a column with structured packing, the volume fraction of droplets with a diameter smaller than a given threshold value, d_p, beyond which the droplets are entrained by the gas. If a typical value of the steam velocity, 20 m/s, is taken as reference, it is found that in a column with structured packing the diameter of the droplets generated is such that 1.6% by volume is made up of droplets smaller than 30 μm in diameter. For a 50% reduction of this velocity, the percentage is 0.1%, with a much more than proportional decrease of the load of fine droplets to be separated.

These estimations are undoubtedly subject to a good deal of uncertainty due to the lack of field measurements of the characteristics of the entrained liquid and of the separating efficiency. However, it should be noted that a sizable decrease in entrained liquid with decreasing steam velocity is a fact that systematically emerges in the technical and scientific literature, and if nothing else, it makes it possible to expect considerable advantages from the standpoint of a decrease in entrained liquid from reducing the steam velocity. On the basis of these considerations, a value of about 6.4, corresponding to an effective velocity of 10 m/s, is chosen for the superficial velocity of the steam in the column.

To calculate the pressure drops in the column the correlation proposed by Bravo *et al* (1985) can be adopted:

$$\frac{dP}{dx} = \left[C_1 + \frac{C_2}{\mathrm{Re}_G}\right]\left[\frac{\rho_G\, u_{SG}^2}{s}\right]\left[\frac{1}{1 - C_3\, Fr^{0.5}}\right]^5 \tag{34}$$

where dP/dx is the pressure gradient. The values of the constants C_1 and C_2 relative to the friction factor, were obtained experimentally by Viviani (1993).

$$C_1 = 0.171 \qquad C_2 = 92.7$$

For the constant C_3 a dependance on the nominal size of the packing is found. For the packing utilized $C_3 = 3.08$. At the velocity chosen for the stem the pressure gradient is 1900 Pa/m.

3.3 Balance equations

The mass balances relative to the liquid phase and to the steam can be expressed as

$$\frac{dL}{dx} = \rho_G\, v\, a\, A \tag{35}$$

$$\frac{dG}{dx} = -\rho_G\, v\, a\, A \tag{36}$$

where L and G are the liquid and steam flow rates and A is the column section. In these equations the steam flow, v, is calculated using Eq. 11.

The HCl balance in the vapor phase is given by

$$\frac{d(G\,\omega_A)}{dx} = -n_A a\, A \tag{37}$$

with n_A given by Eq. 4.

The enthalpy balances relative to the two phases are given by:

$$\frac{d(L\,H_L)}{dx} = q\, a\, A \tag{38}$$

$$\frac{d(G\,H_G)}{dx} = -q\, a\, A \tag{39}$$

where q is given by Eq. 19. The boundary conditions of the system of Eqs. 30-34 are

$$x=0 \qquad \begin{aligned} L&=L^o\\ G&=G^o\\ \omega_A&=\omega^o A\\ H_L&=H^o{}_L\\ H_G&=H^o{}_G \end{aligned} \tag{40}$$

To solve the system it is necessary to determine the conditions at the interface along the column. These conditions are obtained by solving Eqs. **11, 12** and **21.** These three equations constitute a nonlinear system in the three unknowns v, ω_w^i and T^i.

Despite the apparent complexity of the problem, the results show that, in first approximation, the problem can be put **in** the general form

$$\frac{d(G\,\Phi_n)}{dx} = -\lambda_n a(\Phi_n - \Phi_n^i)A \tag{41}$$

where Φ_n can be H_G or $\rho_G\,\omega_A$, λ_n **is** the transfer coefficient h_G or k_A and Φ_n^i represents the conditions at the interface. Eq. **41** can be easily integrated for constant G and Φ_n^i, with the result

$$\ln\frac{\Phi_n - \Phi_n^i}{\Phi_n^o - \Phi_n^i} = -\frac{\lambda_n\, a}{G} A\, x \tag{42}$$

In the case of **mass** transfer we have $\omega_A^i=0$ and the HCl removal efficiency, η_A, defined as

$$\eta_A = \frac{\omega_A^o - \omega_A}{\omega_A^o} \tag{43}$$

is equal to

$$\eta_A = 1-\exp\left(-\frac{k_A\, a\, A}{G} x\, \rho_G\right) \tag{44}$$

In the case of sensible heat transfer, we have $\Phi_n^i=H_G^i$ and the heat transfer efficiency, η_H, defined as

$$\eta_H = \frac{H_G^o - H_G}{H_G^o - H_G^i} \tag{45}$$

is equal to

$$\eta_H = 1-\exp\left(-\frac{h_G\, a\, A}{G\, C_p} x\right) \tag{46}$$

From this approximate treatment of the transfer phenomena, it is seen that the efficiency of the operation chiefly depends on the two dimensionless parameters $h_G aAx/GC_p$ and $k_A aAx\rho_G/G$. The ratio between the transfer efficiencies does not depend on the type of equipment, one finds:

$$\frac{1-\eta_A}{1-\eta_H} = \exp\left[(Le-1)X\right] \tag{47}$$

where $Le = \dfrac{h_G}{k_A\, \rho_G\, c_P}$ and X is defined as

$$X = x\frac{k_A\, a\, A\, \rho_G}{G} \tag{48}$$

3.4 Analysis of results

For a given set of initial conditions, Fig. **1** reports the efficiencies η_A and η_H as a function of the dimensionless coordinate X defined above, it can be noted that the mass transfer is slightly faster than the heat transfer; that the HCl removal is 90% complete for relatively short columns ($x \cong$ **1.2** m).

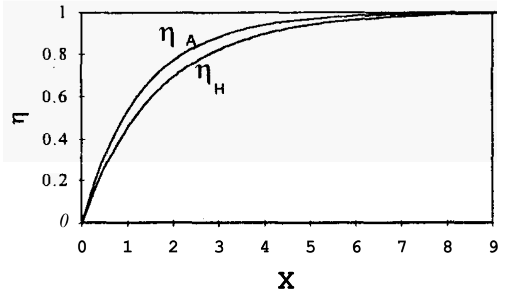

Fig. **1**

The design presented in **this** report is based on the direct measurement of the gas phase mass transfer coefficient. The fact that the analogy between heat and mass transfer phenomena is commonly accepted in the design of various apparatuses, beginning with cooling towers, constitute a confirmation of the fact that the experimental basis of the design presented in this work can be considered sufficient. However, any uncertainties on the possible scale-up effects and maldistribution of the wash liquid require increasing the column height, or carrying out field experimentation aimed at defining the optimum packing height that would allow substantial HCl removal yet keep steam superheating at an appreciable level.

The effect of operating parameters such as liquid flow rate and temperature has been also investigated. The wash liquid flow rate can be appreciably reduced with respect to the flow rate now used without significantly altering the efficiency of the operation. Negligible effects are also given by an increase of the liquid supply temperature. As we shall see later in this work, these results make it possible to optimize the washing operation for energy purposes.

It was already noted that the HCl removal efficiency for a column of a given height is substantially independent of the steam velocity. Since the steam velocity determines the column section and the pressure drops, the choice of this parameter is based solely on considerations of fluid dynamics.

4. OPTIMIZATION OF THE WASHING PROCESS

The washing process causes a net power loss linked to the steam desuperheating, offset only partly by the partial evaporation of the wash liquid. With reference to the steam characteristics considered in this report

Steam flow rate	100 **t/h**
Pressure	11 bar abs
Temperature	236°C
HCl concentration	20 ppm
Noncondensable gases	7%

and to the performance data of the Tosi turbines installed in ENEL's geothermal fields, Table 1 reports the percentage power loss in various operating hypotheses. **As** can be seen from this table, in the case of complete desuperheating of the steam with a liquid flow rate of 10 t/h and a liquid supply temperature of **25°C,** the power loss is 6.5%. These conditions correspond to the scheme presently in use in Italian geothermal fields, illustrated in Fig. **2,** scheme A.

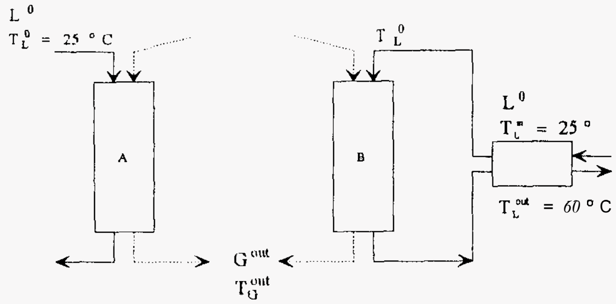

Fig. 2

L^0 t/hr	Scheme	η_A	G^{out} t/hr	T_G^{out} °C	T_L^0 °C	Power lost %
10	A	1	102916	182	25	6.4
7		1	103897	182	25	5.5
7		0.9	102465	194	25	4.7
10	B	1	104340	182	95	5.1
7		1	104515	182	68	4.9
7		0.9	103380	194	89	3.8

Table 1

An appreciable increase in generated power (on the order of 0.8-1%) would result from limiting the desuperheating to 78-80%, a value that corresponds to 90% HCl removal. Still more significant increases are obtained by reducing the liquid flow rate in the column or increasing the supply temperature. The latter operation can be achieved very simply by preheating the liquid sent to the washing column with the outgoing liquid as indicated in scheme B of Fig. 2. The maximum power saving achievable overall by acting simultaneously on the degree of steam desuperheating and on the supply liquid flow rate and temperature is 2.6%. It should be noted that this saving is only partly connected to the use of a column with structured packing, which with respect to the process currently in use makes it possible to reduce the liquid flow rate in the column thanks to the more efficient liquid-vapor contacting system. As indicated in Table 1, substantial power savings can also be achieved just by preheating the liquid fed into the column.

The process described in this work foresees the possibility of not pushing the washing to complete desuperheating of the steam, with an appreciable energy saving as a result. This choice is connected to the maximum concentration of chlorides that it is possible to admit into the turbine, considering that removal efficiencies on the order of 85-90% lead to a slightly lower steam desuperheating percentage.

5. CONCLUSIONS

The process presented in this report is based on the utilization of an apparatus operating at low steam velocity ($u_{SG} < 7.0$ m/s). According to preliminary estimates reported in the test, a decrease in steam velocity results in a much more than proportional decrease in the fine entrained liquid due to the combined effect of a decrease in the atomized liquid fraction and an increase in droplet size (large drops will not be entrained by the gas stream). The substantial decrease in entrained liquid and of small droplets in particular makes it possible to operate without a primary separator, with a sizable reduction in plant costs.

The columns with structured packing proposed for use are characterized by a high specific area and low pressure drops. This makes it possible to reduce the liquid flow rate used for the washing and the pressure drops associated with the operation.

If these modifications in the steam washing equipment are accompanied by heat recovery on the outflowing alkaline solution - such recovery can also be performed in the plants now in operation - the overall power loss in the washing stage can be reduced, for the typical case considered here, from 6.5-7% to 4-4.5%, resulting in a significant saving.

Finally, the analogy between heat transfer and mass transfer illustrated in the report and simple calibrations to be carried out in the field make it possible to propose a simple process control system based on the reading of the steam temperatures at the washing inlet and outlet and a consequent control of the flow rate of liquid to be sent to the process.

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