Researches Concerning the Intake of Carbon Dioxide Resulted from the Fossil Carbon Burning

TÂNASE DOBRE*, OCTAVIAN FLOAREA
University Politehnica of Bucharest, Chemical Engineering Department, 1-3 Polizu, 011061, Bucharest, Romania

The research by modelling has been particularized at the case of removal of carbon dioxide from hot burning gases. The gases washing with specific solution in mobile packed bed expresses the basic of proposed removal procedure of carbon dioxide. A large assembly of differential equations give the mathematical model of the considered procedure. The given simulations show that the carbon dioxide and sulphur dioxide removal occurs with high intensities depending of the concentration level of the liquid active species.

Keywords: mobile packed bed, carbon dioxide removal, mathematical modelling

The best technological solution adopted to intake the carbon dioxide resulted from the fossil carbon burning is that where the burning gases are washed at first with a calcium carbonate suspension and then with a potassium carbonate solution. The obtained results allow to design and to simulate an industrial pilot wherein the phases contacting occurs in two mobile packed columns (the first for washing with calcium carbonate suspension and the second for washing with potassium carbonate solution).

This solution allows the simultaneous realization of the cooling of burning gases, usually with an initial temperature of about 100°C, together with the inter-phase mass transfer of water vapours, carbon dioxide and sulphur dioxide in the both columns. The employed solution creates conditions to establish:

- the influence of process factors (gases temperature, compounds concentrations in gaseous phase, calcium carbonate and potassium carbonate concentrations in washing solutions etc) on the yield of taken carbon dioxide;
- the optimum hydrodynamic conditions for each column;
- the optimisation problems concerning the pilot set-up working;
- the impact of burning gases washing on the environment.

The used solution has also the advantage that the processed gases can contain dust or have a smoke aspect, because the mobile packed bed is able to retain all the solid impurities from gases.

Experimental part

The solution proposed for cooling and removal of carbon dioxide and sulphur dioxide from burning gases is that wherein a direct contact occurs between burning gases and calcium carbonate suspension in the first packed column and between partially purified gases and potassium carbonate solution in the second packed column (fig. 1). The selected solution is advantageous from point of view of a simple operating and small contactor dimensions. It must be specified that into vessels V₁ and V₂ are periodically added water (instead of vaporized water), calcium

![Fig. 1. Technological scheme concerning the processing of raw burning gas by mobile bed washing with calcium carbonate suspension (C₁) and potassium carbonate solution (C₂).](image_url)
The mathematical model describing the gas-liquid transfer of three molecular species (water vapour, carbon dioxide and sulphur dioxide) together with the cooling or heating of the liquid phases (calcium carbonate suspension in C1 and potassium carbonate in C2) in a mobile packed bed consists of relationships (1)-(5):

\[
\frac{dY_1}{dZ} = -K_{12}\sigma_{g1-s1}S \left( \frac{Y_1}{Y_1^*} \right) \tag{1}
\]

\[
\frac{dY_2}{dZ} = -K_{22}\sigma_{g2-s2}S \frac{Y_2}{Y_2^*} \tag{2}
\]

\[
\frac{dY_3}{dZ} = -K_{s3-g3}\sigma_{s3-g3}S \frac{Y_3}{Y_3^*} \tag{3}
\]

\[
\frac{dL}{dZ} = -k_{g-s}\sigma_{g-s}S \left( l_g - l_t \right) \tag{4}
\]

\[
\frac{dY_1^*}{dt} = -L_{M1}\frac{1}{\alpha_t} \frac{dL}{dZ} + \frac{k_{g-s}\sigma_{g-s}M_1}{c_t(\varepsilon_{L1} + \varepsilon_t)D_{p1}} \left( t_g - t_t \right) + \frac{k_{g-s}\sigma_{s3-g3}M_1}{c_t(\varepsilon_{L1} + \varepsilon_t)D_{p1}} \left( t_g - t_t \right) \tag{5}
\]

The mathematical model of vessel V1 consists of differential equations (12)-(14) and initial condition (15):

\[
V_1 \frac{dC_{pc1}}{dt} = -G_{in}(1 + Y_1 + Y_2)(Y_{2N1} - Y_2)M_{CaCO_3} - G_{in}(1 + Y_1 + Y_2)(Y_{3N1} - Y_3)M_{CaCO_3} \tag{12}
\]

\[
Z = 0, \quad \frac{dZ}{dZ} = 0, \quad t_g = t_{in1}, \quad Y_1 = Y_{2N1}, Y_2 = Y_{2N1}, Y_3 = Y_{2N1} \tag{6}
\]

\[
Z = H_g, \quad D_1 \frac{dt}{dZ} = \omega_{1g} \tag{7}
\]

\[
Z = 0, \quad \frac{dZ}{dZ} = 0, \quad t_g = t_{in1}, \quad Y_1 = Y_{2l1}, Y_2 = Y_{2l1}, Y_3 = Y_{3l1} \tag{8}
\]

\[
Z = H_g, \quad D_1 \frac{dt}{dZ} = \omega_{2g} \tag{9}
\]

This model is singularized by means of univocity conditions for C1 ((6) and (7)) and C2 ((8) and (9)), respectively.

\[
Y_1^* = \frac{f_{g1}P_s}{p - f_{g1}P_s} \frac{A - B}{t_g + C} \tag{10}
\]

\[
p_s = 10 \tag{11}
\]

It can be considered that the total molar flow rate of liquid, L, passing through each column has a constant value.

The numerical integration of proposed model is difficult. Accordingly, the integration begins at Z = 0 (gas entrance into the mobile bed) and ends at Z = H, (gas exit from mobile bed), where the limit condition must be accomplished. If it is not, the values of Y and t are kept and the value of t must verify the limit condition and then a new integration from Z = H to Z = 0 is performed. If the initial conditions are found at Z = 0 then the integration is correct, otherwise the procedure is repeated.

Adding to the two columns model the models characteristic of each of vessels V1 and V2, the complete model of experimental set-up is obtained. It is observed that both vessels are equipped with cooling systems, so that the temperature of the liquid fed into the columns to be constantly maintained (t_{in} in vessel V1 and t_{in} in vessel V2). It is assumed that the liquid phase in each vessel is perfectly mixed. The species existed in vessel V1, namely calcium carbonate, bicarbonate and bisulphite, keep their identity (calcium bisulphite would be turned into calcium sulphate, but it is assumed that the low concentration of oxygen in gases does not favour this reaction). In the case of vessel V2, the initial concentration of calcium carbonate in suspension is C_{CaO}, and the momentary concentrations of calcium carbonate, bicarbonate and bisulphite are C_{CaCO_3}, C_{CaCO_3}, and C_{CaCO_3}. In the case of vessel V1, the initial concentration of potassium carbonate in solution is C_{K_2CO_3}, and the momentary concentrations of potassium carbonate, bicarbonate and bisulphite are C_{K_2CO_3}, C_{K_2CO_3}, and C_{K_2CO_3}. The mathematical model of vessel V1, consists of differential equations (12)-(14) and initial condition (15):

\[
V_{in1} \frac{dC_{pc1}}{dt} = -G_{in}(1 + Y_1 + Y_2)(Y_{2N1} - Y_2)M_{CaCO_3} - G_{in}(1 + Y_1 + Y_2)(Y_{3N1} - Y_3)M_{CaCO_3} \tag{12}
\]
Similarly, differential equations (16)-(18) and initial condition (19) characterize the mathematical model of vessel $V_2$:

$$\frac{dC_i}{dt} = G_m(1 + Y_1 + Y_2)(Y_{s+1} - Y_i)M_{CO_2}CO_2 - G_m(1 + Y_1 + Y_2)(Y_{s+1} - Y_i)M_{SO_2}SO_2$$  

$$\frac{dC_{ph}}{dt} = G_m(1 + Y_1 + Y_2)(X_{s+1} - Y_{s+1})M_{SO_2}SO_2$$  

$$\tau = 0, C_m = C_{ph}, C_{m} = 0, C_{ph} = 0$$

The system of equations and restrictions of (1)-(19) describes the mathematical model of processes occurred in the experimental set-up. Because it is difficult to integrate numerically this model, it was turned into a simplified form which allows simulating the set-up working.

Nomenclature used in the above relationships was:

- $Y_1$, $Y_2$, $Y_3$ - molar ratios of water vapour, carbon dioxide and sulphur dioxide in processed gas;
- $G_m$ - molar flow rate of inert gas (nitrogen);
- $S$ - cross section area of mobile packed column;
- $\sigma_g-l$ - gas-liquid contacting surface area in mobile packed column;
- $\varepsilon_s$, $\varepsilon_l$, $\varepsilon_g$ - fractions of solid, liquid and gas in mobile packed bed;
- $K_Y$, $i = 1...3$ - mass transfer total coefficient of $i$ species;
- $D_{l}$ - axial mixing coefficient in mobile bed;
- $L_t$ - total molar flow rate of liquid in mobile packed column;
- $M_l$ - mean molar mass of liquid in mobile packed column;
- $\rho_l$ - liquid density;
- $t_l$ - liquid temperature;
- $t_g$ - gas temperature;
- $M$ - molar mass;
- $k_{if}$ - inter-phase heat transfer coefficient.

Results and discussion

It is considered the case of mobile bed washing of hot gases with a flow rate of 5000 m$^3$/h, temperature of 100 $^\circ$C and composition (% vol): 2% water vapour, 14 % CO$_2$, 0.8% SO$_2$ and 83.2 % N$_2$, using specific flow rates of washing water of 3, 6, 9, 12, 15, 18 m$^3$/m$^2$ h and assuming that both mobile bed columns work identically.

The column cross section area, $S$.

The minimum mobility rate, in absence of liquid phase, of a mobile packed bed consisting of empty polyethylene spheres with a diameter $d = 0.03$ m and a density $\rho = 300$ Kg/m$^3$, is calculated with the following correlation:

$$w^0_{min} = 0.9(d_{p} \rho / \rho_s)^{0.5} = 0.9(0.03 - 300/0.91)^{0.5} = 2.85 \ m/s$$

Accordingly, it results:

$$D = \left( \frac{4G_s}{nw_g} \right)^{0.5} = \left( \frac{4 \cdot 5000 \cdot 273}{273 \pi \cdot 3} \right)^{0.5} = 0.855 \ m; D = 0.8 m$$

The column diameter is determined considering a value $w_g = 3$ m/s:

$S = \frac{\pi D^2}{4} = \frac{\pi \cdot 0.8^2}{4} = 0.5 \ m^2$ and $w_g = \frac{G_s}{S} = \frac{5000}{273} = 3.2 \ m/s$

The minimum mobility rate calculated with the below given relationship leads to the values presented in table 1.

$$w^0_{min} = w^0_{min} \ exp(-83q_{v_l})$$

The bed expanding is calculated by means of the relationship $H_{e} = \frac{1 + 0.18(w - w_{min})q_{v_l}^{0.35}}{H_{0}}$ where $q_{v_l}$ is expressed in m$^3$/m$^2$.h.

The bed expanding of liquid, $e_l$, gas, $e_g$, and solid, $e_s$, in mobile bed are determined with the following set of correlations [4,5]:

$$e_l = 0.216 \left( \frac{30H_e + 8}{30H_e + 6} \right) q_{v_l}^{-0.85} d_{p}^{-0.5}$$

$$e_g = 0.19P_{g}^{0.11} d_{p}^{0.22}$$

$$e_s = 1 - e_g - e_l$$

<table>
<thead>
<tr>
<th>Table 1</th>
<th>MINIMUM MOBILITY RATE DEPENDING ON SPECIFIC FLOW RATE OF WASHING WATER</th>
</tr>
</thead>
<tbody>
<tr>
<td>No</td>
<td>1</td>
</tr>
<tr>
<td>$q_{v_l}$ (m$^3$/m$^2$ s)</td>
<td>0.00084</td>
</tr>
<tr>
<td>$w^0_{min}$ (m/s)</td>
<td>2.65</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Table 2</th>
<th>MOBILE BED HEIGHT AND BED EXPANDING DEPENDING ON SPECIFIC FLOW RATE OF WASHING LIQUID (H$_0$ = 0.4 m)</th>
</tr>
</thead>
<tbody>
<tr>
<td>No</td>
<td>1</td>
</tr>
<tr>
<td>$q_{v_l}$ (m$^3$/m$^2$ s)</td>
<td>3</td>
</tr>
<tr>
<td>$H_{e}/H_{0}$</td>
<td>1.15</td>
</tr>
<tr>
<td>$H_{e}$ (m)</td>
<td>0.46</td>
</tr>
</tbody>
</table>
Here $q_0$ is expressed in m$^3$/m$^2$s and Weber and Froude numbers, $W_e$, $F_r$, are calculated with the relationships where the characteristic length is the diameter of the particles.

$$W_e = d_p \frac{w_g^2 \rho_b}{\sigma_s}$$

$$F_r = \frac{w_g}{(g d_p)^{0.5}}$$

The packing form factor for spherical particles is $\Psi = 1$. Table 3 presents the values of liquid, gas and solid fractions calculated for each value of specific flow rate of washing water.

Axial mixing coefficient, $D_l$, is determined according to the procedure proposed in our previous studies [3,5] by means of the following set of correlations:

$$d_p (H_d) = -(2H_d + 6H_s^2)$$

$$\alpha_s (H_d) = -(8H_d + 6H_s^2 + 1.2H_s^3)$$

$$\alpha_s (H_d) = 3 + 3H_d + 1.2H_s^2$$

$$\varphi = \frac{\alpha_s (H_d) \cdot \alpha_s (H_s) - 2\alpha_s (H_s) \cdot \alpha_s (H_d)}{\alpha_s (H_d)}$$

$$\varphi = \frac{\alpha_s (H_d) \cdot \alpha_s (H_s) - 2\alpha_s (H_s) \cdot \alpha_s (H_d)}{\alpha_s (H_d)}$$

$$\sigma^2 = \varphi_0 (0, H_d) \frac{\varphi_0 (0, H_d)}{\sigma^2}$$

where $H_d$ is expressed in dm [5].

The dependence between axial mixing coefficient and specific flow rate of washing water is emphasized in table 4.

Mass transfer coefficient of water vapour, $K_{y1}$, is calculated considering that mass transfer resistance of vapour is concentrated in gaseous phase, accordingly $K_{y1} = k_{y1}$. Calculus procedure and criterial relationships employed to calculate the mass transfer partial coefficient, $k_{y1}$, are according to our previous studies [2,3].

The values of diffusion coefficient of water vapour in gaseous phase used in the below relations was $D_v = 1.4 \cdot 10^{-5}$ m$^2$/s.

$$k_{y1} = \frac{D_v}{(1 + \gamma)^2}$$

Table 3 presents the values of liquid, gas and solid bed fractions depending on specific flow rate of washing liquid.

Table 4 presents the values of liquid, gas and solid fractions calculated for each value of specific flow rate of washing water.

Table 5 presents the values of mass transfer partial coefficient of water vapour depending on specific flow rate of washing liquid.
Mass transfer coefficients of carbon dioxide and sulphur dioxide, $K_{y_2}$ and $K_{y_3}$, are calculated in the same way as mass transfer coefficient of water vapour. Considering the values $D_{g_2} = 1.1 \times 10^{-5}$ m²/s and $D_{g_3} = 0.95 \times 10^{-5}$ m²/s for diffusion coefficient of carbon dioxide and sulphur dioxide, the values of $K_{y_2} = k_y^2$ and $K_{y_3} = k_y^3$ reported in Table 6 and Table 7 have been obtained.

- Inter-phase heat transfer coefficient, $k_y$ is calculated similarly to $k_y^i$ with the following relationships:

$$Pr_y = \frac{c_p m_H}{\lambda_y} = \frac{0.244 \cdot 15 \cdot 10^{-6} \cdot 3.600}{0.027} = 0.488$$

$$Pr_y = \frac{c_p m_H}{\lambda_y} = \frac{1.08 \cdot 10^{-3} \cdot 3.600}{0.027} = 5.14$$

$$Pr_y = \frac{c_p m_H}{\lambda_y} = \frac{0.244 \cdot 15 \cdot 10^{-6} \cdot 3.600}{0.027} = 0.488$$

$$Pr_y = \frac{c_p m_H}{\lambda_y} = \frac{1.08 \cdot 10^{-3} \cdot 3.600}{0.027} = 5.14$$

$$Re_y = \frac{g_m \beta \rho_x}{\epsilon_x g_3600}$$

ex: $Re_y = \frac{3.046}{0.02-0.8 \cdot 10^{-3} \cdot 3.600} = 1.28 \cdot 10^4$

$$Nu_y = 3.75(1.11-0.11)(1.51^3)(1.51^3)(0.488)^{0.6} = 1.22 \cdot 10^3$$

$$Nu_y = 3.75(1.11-0.11)(1.51^3)(1.51^3)(0.488)^{0.6} = 1.22 \cdot 10^3$$

The obtained values of $k_y$ depending on the values of liquid specific flow rate are presented in Table 8.

Specific surface area of inter-phase transfer at contacting in mobile packing, $\sigma_{g-l}$, is determined by means of the following relationship [9]:

$$\sigma_{g-l} = 6 \left[ 1 - e_y - 0.6e_y \frac{1}{1 - e_y} \right] = 0.33 \rho_y + \frac{0.435}{\sigma_y}$$

ex: $\sigma_{g-l} = 0.03(1 + 0.0202/(1 - 0.7))^{0.33} + 0.033 + 0.435 = 0.071$

$= 86 \text{ m}^2 / \text{m}^3$

Table 9 contains the values of $\sigma_{g-l}$ calculated for various values of liquid flow rate.
Molar flow rate of inert gas, \( G_{in} \):

\[
G_{in} = \frac{G_{in}}{22.4} \left( 1 - \frac{1}{3600} \right) = 1 - 0.021 = 0.979 \text{ kmole}_g / s
\]

Total molar flow rate of gas, \( G_t \):

\[
G_t = G_a + \sum_i Y_{ei} = 0.061(1 + \sum_i Y_{ei}) = 0.062 \text{ kmole}_g / s
\]

Molar specific heat of gaseous phase, \( C_{pg} \):

\[
C_{pg} = C_{pg} \cdot M_g = 1000 \cdot 28 = 28000 \text{ J/kgmoleK}
\]

Total molar flow rate of liquid, \( L_t \):

\[
L_t = q_d \cdot S \cdot \rho / M_L; \text{ ex: } L_t = 3.05 \cdot 0.4 = 0.222 \text{ kmole/s}
\]

Molar specific heat of liquid phase, \( C_L \):

\[
C_L = c_L \cdot M_L = 4000 \cdot 18 = 72000 \text{ J/kgmoleK}
\]

Molar latent vapourisation heat of the liquid, \( L_v \):

\[
L_v = L _v = 3 m^3/m^2 h \text{ and } V_L = V_L = 3 m^3 \text{ (m3/m2 h)}
\]

Substituting the calculated parameters into the model equations, the following set of relationships is obtained (for \( q_d = 3 m^3/m^2 h \) and \( V_L = V_L = 3 m^3 \)):

For column 1 (at time \( \tau \))

\[
\frac{dY_1}{dZ} = \frac{1.65}{(1 + Y_1)(1 + Y_2)} (Y_1 - Y_1)
\]

\[
\frac{dY_2}{dZ} = \frac{1.29(C_{pg})}{(1 + Y_2)(1 + Y_1 + Y_2)} (Y_2 - f_{21}(C_{pg}) \cdot Y_2^2)
\]

\[
\frac{dY_3}{dZ} = \frac{1.12(C_{pg})}{(1 + Y_3)(1 + Y_1 + Y_2)} (Y_3 - f_{32}(C_{pg}) \cdot Y_3^2)
\]

\[
\frac{dt}{dZ} = \frac{1.978}{(1 + Y)} (t_g - t_l)
\]

\[
\frac{d^2t}{dZ^2} + 0.311 \frac{dt}{dZ} = \frac{4.4041}{(1 + Y)} (Y - Y^*) + 0.037(t_g - t_l)
\]

\[
Z = 0, \frac{dt}{dZ} = 0, t_g = t_{l+1}, Y_1 = Y_{1+1}, Y_2 = Y_{2+1}, Y_3 = Y_{3+1} = 0.025, Y_2 = Y_{2+1} = 0.185, Y_3 = Y_{3+1} = 0.05
\]

\[
Z = H_d, D_1 \frac{dt}{dZ} = \frac{q_d}{e_1}, t_{l+10}
\]

For vessel \( V_1 \)

\[
\frac{dC_{sc}}{d\tau} = -0.0163(1 + Y + Y_3)(Y_{sc+1} - Y_{sc}) - (C_{sc}) = -0.0163(1 + Y + Y_3)(Y_{sc+1} - Y_{sc}) - (C_{sc})
\]

\[
\frac{dC_{sh}}{d\tau} = 0.0163(1 + Y + Y_3)(Y_{sh+1} - Y_{sh}) - (C_{sh})
\]

\[
\frac{dC_{sc}}{d\tau} = 0.0163(1 + Y + Y_3)(Y_{sc+1} - Y_{sc}) - (C_{sc})
\]

\[
\tau = 0, C_{sc} = C_{sc} = 75 g/l, C_{sh} = 0, C_{sh} = 0
\]

For column 2 (at time \( \tau \))

\[
\frac{dY_1}{dZ} = \frac{1.65}{(1 + Y_1)(1 + Y_2 + Y_3)} (Y_1 - Y_1)
\]

\[
\frac{dY_2}{dZ} = \frac{1.29(C_{pe})}{(1 + Y_2)(1 + Y_1 + Y_2)} (Y_2 - f_{22}(C_{pe}) \cdot Y_2^2)
\]

\[
\frac{dY_3}{dZ} = \frac{1.12(C_{pe})}{(1 + Y_3)(1 + Y_1 + Y_2)} (Y_3 - f_{32}(C_{pe}) \cdot Y_3^2)
\]

\[
\frac{dt}{dZ} = \frac{1.978}{(1 + Y)} (t_g - t_l)
\]

\[
\frac{d^2t}{dZ^2} + 0.311 \frac{dt}{dZ} = \frac{4.4041}{(1 + Y)} (Y - Y^*) + 0.037(t_g - t_l)
\]

\[
Z = 0, \frac{dt}{dZ} = 0, t_g = t_{l+1}, Y_1 = Y_{1+1}, Y_2 = Y_{2+1}, Y_3 = Y_{3+1}
\]

\[
Z = H_d, D_1 \frac{dt}{dZ} = \frac{q_d}{e_1}, t_{l+10}
\]

The variable change \( dt/dZ = \) is accomplished for a numeric integration of this equations and restrictions system. By the system integration, the profiles of temperature and of compounds concentration in columns \( C_1 \) and \( C_2 \) and also concentrations dynamics in vessels \( V_1 \) and \( V_2 \) are obtained. Factors of mass transfer intensification by chemical reactions (\( I_{21}, I_{31} \) for column \( C_1 \) and \( I_{22}, I_{32} \) for column \( C_2 \)) and of equilibrium molar ratio correction of gaseous compounds implied in chemical reactions (\( f_{21}, f_{32} \)) have been introduced in the above model. Accordingly, the equilibrium molar ratios \( Y_2^* \) and \( Y_3^* \) are expressed by means of Henry law.

Figures 2-5 illustrate the concentration dynamics of calcium carbonate suspension in vessel \( V_1 \) and of potassium carbonate solution in vessel \( V_2 \) together with the concentration profiles of water vapours, carbon dioxide and sulphur dioxide in packed beds from column \( C_1 \) and column \( C_2 \), in the case of a liquid specific flow rate \( q_d = 3 m^3/m^2 h \) and a single packed bed \( (H_0 = 0.4 m) \) in each column (simulation 1). Figures 6-9 show the distribution of carbon dioxide and sulphur dioxide for simulation 1 at \( \tau = 60 min \) and \( \tau = 110 min \) (when the potassium solution is practically exhausted).

Figures 10-13 show the effects of increase of contact time between burning gas and liquid, using two mobile beds, each of them with a height of \( 0.4 m \), and also of liquid specific flow rate, which was increased at a value of...
15 m³/m²h (simulation 2). Accordingly, it is observed a more rapid exhausting of potassium carbonate solution and an important changing of concentration profiles of water vapour, carbon dioxide and sulphur dioxide than in the case of simulation 1.

Fig. 2. Concentration dynamics of calcium carbonate and potassium carbonate for simulation 1 (V₁=V₂=3m³, qᵥ=3m³/m²h, 1 bed with H₀=0.4m)

Fig. 3. Distribution of molar ratio of water vapour in mobile packing bed in C₁ and C₂ for simulation 1 at τ=10 min

Fig. 4. Distribution of molar ratio of carbon dioxide in mobile packing bed in C₁ and C₂ for simulation 1 at τ=10 min

Fig. 5. Distribution of molar ratio of sulphur dioxide in mobile packing bed in C₁ and C₂ for simulation 1 at τ=10 min

Fig. 6. Distribution of molar ratio of carbon dioxide in mobile packing bed in C₁ and C₂ for simulation 1 at τ=60 min

Fig. 7. Distribution of molar ratio of sulphur dioxide in mobile packing bed in C₁ and C₂ for simulation 1 at τ=60 min

Fig. 8. Distribution of molar ratio of carbon

Fig. 9. Distribution of molar ratio of sulphur
The imposed condition relating to constant values of feed temperatures of gas and liquid in columns $C_1$ and $C_2$ shows that the dynamics of phases temperatures in the both columns is characteristic of a stationary process (fig. 14).

An analysis of data presented in these figures emphasizes that the employment of mobile beds to intake the gases heat constitutes a facile and technologically advantageous technique. It was concluded that it is not recommendable to use more than two beds in a column due to an increase of pressure losses and of energy consumption for gases transport, respectively. It is specified that gases with high temperatures can be cooled with an acceptable increase of specific losses by vaporized water.

It is emphasized also that it is not recommendable to use high liquid flow rates, because it does not lead to an important effect concerning an increase of heat intake degree and a decrease of water specific losses by vaporisation; otherwise, an increase of this parameter determines an increase of pressure losses and of energy consumption for gases transport.

In comparison with classical solutions of cooling in towers with air natural convection, the gases cooling in mobile systems is advantageous from point of view of small apparatus dimensions. Accordingly, a mobile packed bed set-up of 0.8 m$^3$ can be used instead of a cooling tower of 25 m$^3$.

**Conclusions**

The use of mobile packed bed column as basic contacting unit between gasses from burning processes and an adequate washing liquid has analysed using modelling procedure.
As shown in figure 2 and figure 10, the high consumption of active components (calcium carbonate in the washing liquid from C₁ and potassium carbonate in the washing liquid from C₂) is the sensible point of this proposed procedure for removing of carbon dioxide from gases produced by the fossil carbon burning.

Acknowledgement: The authors thank Ph. D. Oana Parvulescu for her help in paper developing.

References
3. DOBRE, T., Ph.D. Thesis, Politehnica University from Bucharest, 1985
4. DOBRE, T., FLOAREA, O., Rev. Chim. (București), 35, nr. 9, 1984, p. 867
5. DOBRE, T., FLOAREA, O., Rev. Chim. (București), 36, nr. 11, 1985, p. 1021
6. DOBRE, T., Rev Chim. (București), 41, nr. 9, 1989, p. 812

Manuscript received: 20.08.2008