MASS TRANSFER IN LIQUID IN AN APPARATUS WITH MOBILE PACKING. APPLICATION OF A DISPERSION MODEL

Zdeněk PALATÝ

Department of Chemical Engineering,
University of Chemical Technology, 532 10 Pardubice

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The paper deals with the mass transfer in a liquid on a plate with mobile packing. A procedure has been suggested which enables estimation of the mass transfer coefficients from experimental data considering the dispersion flow of the liquid. The results obtained from the desorption of CO_2 from water are presented graphically and in a form of empirical equation.

Apparatusses with mobile packing have been known for a number of years and have been successfully adopted for the purposes of absorption and cooling of gases, rectification, and dust collection. They work on the principle of three-phase fluidized layer in which a random and intensive motion of spherical particles and high velocities of both phases to be contacted create favourable conditions for heat and mass transfer. The high intensity of mixing of the liquid on the plate, however, can significantly affect the driving force of the diffusion or heat process taking place.

In the field of mass transfer in liquid on a plate with mobile packing, only few papers have been published, most papers are directed to problems of mass transfer in gaseous phase. The mass transfer coefficients in liquid were obtained from experimental measurements of desorption of carbon dioxide[1 – 3] or oxygen[4] from water by a stream of air. A common feature of all the papers cited is the presumption of ideal mixing of the liquid on the plate with mobile packing. This presumption, however, can represent a considerable idealization of the conditions really existing.

The aim of the present work is experimental estimation of the mass transfer coefficient in a liquid considering a dispersion flow of the liquid and verification of the above-mentioned presumption.

THEORETICAL

In an apparatus with mobile packing the flow of liquid on a plate is described by a dispersion model[5 – 10]. Under such conditions at a steady state, the dependence of concentration of the component desorbed on the height can be expressed by Eq. (*1*)

$$\frac{1}{Pe}\frac{\mathrm{d}^2 c_A}{\mathrm{d}Z^2} - \frac{\mathrm{d}c_A}{\mathrm{d}Z} - k_L a\,\tau_m (c_A - c_{Ai}) = 0 \tag{1}$$

with the initial conditions $c_A = c_A|_{Z=1}$ and $\frac{\mathrm{d}c_A}{\mathrm{d}Z}|_{Z=1} = 0$.

The Z coordinate ($Z = z/h_d$) is oriented in the direction of the flow of liquid with the origin at the place of inlet of the liquid into the layer. In Eq. (*1*) it is $Pe = \bar{u}_L h_d / D_L$ (the Péclet number) and $\tau_m = h_d/\bar{u}_L$ (the mean residence time of liquid in the layer).

Using the equation of mass balance of the component A in the layer, and with the presumption of negligible mass transfer resistance in the gas phase, it is possible to obtain the solution of the differential equation (*1*) in the form

$$c_A = \frac{1}{2M}\left\{ c_A\big|_{Z=1} + \frac{H_A\left[c_A\big|_{Z=1} \cdot RT\,\frac{u_L}{u_G} - p_A\big|_{Z=1}\right]}{1 - H_A RT\,\frac{u_L}{u_G}} \right\} \cdot$$

$$\cdot \left\{ (M-1)\exp\left[\frac{Pe}{2}(M+1)(Z-1)\right] + (M+1)\exp\left[-\frac{Pe}{2}(M-1)(Z-1)\right] \right\} -$$

$$- \frac{H_A\left[c_A\big|_{Z=1} \cdot RT\,\frac{u_L}{u_G} - p_A\big|_{Z=1}\right]}{1 - H_A RT\,\frac{u_L}{u_G}}\,, \tag{2}$$

where

$$M = \sqrt{1 + 4 k_L a\,\tau_m \left(1 - H_A RT\,\frac{u_L}{u_G}\right) Pe^{-1}}\,. \tag{3}$$

If the limit concentration of component A for $Z \to 0^+$ is known, then Eq. (*2*) can be used for calculation of the mass transfer coefficient $k_L a$. This concentration can be estimated from the Danckwerts boundary condition

$$c_A\big|_{Z\to 0^+} - \frac{1}{Pe}\frac{\mathrm{d}c_A}{\mathrm{d}Z}\bigg|_{Z\to 0^+} = c_A\big|_{Z\to 0^-} \tag{4}$$

in which the derivative dc_A/dZ for $Z \to 0^+$ is expressed from Eq. (2)

$$c_A\Big|_{Z\to 0^+} = c_A\Big|_{Z\to 0^-} - \frac{M^2-1}{4M}\left\{c_A\Big|_{Z=1} + \frac{H_A\left[c_A\Big|_{Z=1} \cdot RT\,\dfrac{u_L}{u_G} - p_A\Big|_{Z=1}\right]}{1 - H_A RT\,\dfrac{u_L}{u_G}}\right\} \cdot$$

$$\cdot \left\{\exp\left[\frac{Pe}{2}(M-1)\right] - \exp\left[-\frac{Pe}{2}(M+1)\right]\right\} . \qquad (5)$$

The combination of Eqs (2) for $Z \to 0^+$ and (5) gives the equation

$$c_A\Big|_{Z\to 0^-} - \frac{1}{4M}\left\{c_A\Big|_{Z=1} + \frac{H_A\left[c_A\Big|_{Z=1} \cdot RT\,\dfrac{u_L}{u_G} - p_A\Big|_{Z=1}\right]}{1 - H_A RT\,\dfrac{u_L}{u_G}}\right\} \cdot$$

$$\cdot \left\{(M+1)^2 \exp\left[\frac{Pe}{2}(M-1)\right] - (M-1)^2 \exp\left[-\frac{Pe}{2}(M+1)\right]\right\} +$$

$$+ \frac{H_A\left[c_A\Big|_{Z=1} \cdot RT\,\dfrac{u_L}{u_G} - p_A\Big|_{Z=1}\right]}{1 - H_A RT\,\dfrac{u_L}{u_G}} = 0 . \qquad (6)$$

If the information about the mixing of the liquid, i.e. the quantities Pe and τ_m and the concentrations at the boundary of the layer in both phases, is available, then it is possible to obtain the values $k_L a$ by solving the nonlinear Eq. (6). A suitable method for the solution is the secant method which does not necessitate the calculation of derivatives and converges sufficiently rapidly.

EXPERIMENTAL

The same experimental apparatus as that used in ref.[11] was adopted for the measurements of the mass transfer coefficient in liquid after modifications predominantly involving the circuit of liquid. The scheme of the modified experimental apparatus is given in Fig. 1.

The liquid phase was distilled water saturated with carbon dioxide from a pressure cylinder. The water was saturated in a thermostat which maintained its constant temperature at 20 °C. The CO_2 concentration in the inlet and outlet streams of liquid was determined by means of a method by Winkler[12]. The CO_2 concentration in the inlet stream of liquid varied from 33 to 39 mol m^{-3}. The ranges of gas and liquid flow rates and of the static bed height were the same as those in ref.[11]. Each measurement was repeated at least 5 times at constant hydrodynamic conditions. The values of the Péclet number and of the mean residence time of the liquid estimated for different hydrodynamic conditions in the same apparatus can be found in our preceding paper[11].

RESULTS AND DISCUSSION

The k_La values obtained from Eq. (*6*) represent a product of the mass transfer coefficient in the liquid, k_L, and specific interfacial area a, defined as the ratio of the interfacial area and volume of the liquid hold-up in the system. For a recalculation of these coefficients to the currently used coefficients in which the interfacial area is referred to the volume of the contact space it is necessary to know the hold-up of the liquid. This quantity can be estimated from the volume of the liquid passed and from the mean residence time of the liquid[13] according to the relation

$$V_L = \dot{V}_L \tau_m \,. \qquad (7)$$

For the recalculation of coefficients then it is

$$k_La' = \varepsilon_L k_La \,, \qquad (8)$$

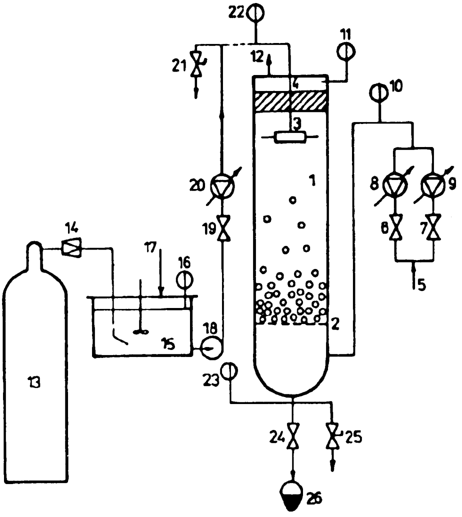

FIG. 1

Scheme of experimental apparatus: 1 column; 2 plate with mobile packing; 3 distributor of liquid; 4 mist eliminator; 5 air inlet; 6, 7, 19, 24 valves; 8, 9, 20 rotameters; 10, 11, 16, 22, 23 thermometers; 12 air outlet; 13 cylinder with CO_2; 14 cylinder pressure regulator; 15 thermostat; 17 distilled water inlet; 18 pump; 21, 25 cocks for taking samples; 26 waste where ε_L is dimensionless hold-up of liquid

$$\varepsilon_L = u_L \tau_m / z_a \tag{9}$$

and $z_a = 0.45$ m is the height of contact space – see ref.[11]. An exact recalculation of k_La to k_La' needs the expression of ε_L on the basis of the height of expanded packing h_d ($h_d < z_a$). However, a precise estimation of h_d is difficult. In addition to it, the recalculation of the mass transfer coefficients according to Eqs (*8*) and (*9*) enables an estimate of the mass transfer coefficient, k_L, since, for the same experimental apparatus, ref.[14] presents the values of interfacial area referred to the volume of contact space.

Figures 2 – 4 represent the dependences of the mass transfer coefficient k_La' calculated in the above-given way from the ca 300 measurements carried out. The individual dependences are constructed for constant static bed height, viz. 21, 32, and 47 . 10^{-3} m. The parameter of the lines in the individual cases is the flow rate of liquid. From these graphical representations it is obvious that the mass transfer coefficients k_La' increase with increasing gas flow rate, and this dependence can be considered linear. Moreover it is obvious that increasing of static bed height in the range of 21 – 47 . 10^{-3} m causes an increase in the k_La' value. In the same sense it is possible to evaluate also the effect of the flow rate of liquid on the mass transfer coefficient. The straight lines in Figs 2 – 4, and in Figs 6 and 7 as well, were obtained by linear regression of the given data sets for u_L = const and h_0 = const.

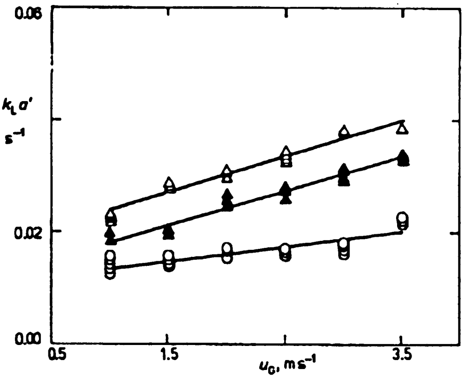

FIG. 2

Dependence of mass transfer coefficient in liquid on gas flow rate for h_0 = 21 . 10^{-3} m and u_L: ○ 5.36 . 10^{-3} m s^{-1}, ▲ 8.93 . 10^{-3} m s^{-1}, △ 12.5 . 10^{-3} m s^{-1}

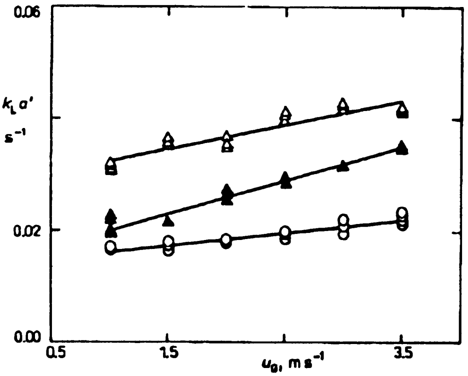

FIG. 3

Dependence of mass transfer coefficient in liquid on gas flow rate for h_0 = 32 . 10^{-3} m and u_L: ○ 5.36 . 10^{-3} m s^{-1}, ▲ 8.93 . 10^{-3} m s^{-1}, △ 12.5 . 10^{-3} m s^{-1}

For evaluation of the effect of the gas and liquid flow rates and of the static bed height on the mass transfer coefficient in the liquid, the results obtained were correlated with the help of the relation

$$k_L a' = D_1 \, u_G^{p_1} \, u_L^{p_2} \, h_0^{p_3} \, . \qquad (10)$$

The constant D_1 and the exponents p_i (i = 1, 2, 3) were estimated by means of nonlinear regression. The concrete form of Eq. (*10*) reads as follows:

$$k_L a' = 2.92 u_G^{0.268} \, u_L^{0.818} \, h_0^{0.284} \, , \qquad (11)$$

where $[k_L a'] = s^{-1}$, $[u_G] = m \, s^{-1}$, $[u_L] = m \, s^{-1}$, and $[h_0] = m$.

The suitability of Eq. (*11*) was evaluated on the basis of the mean quadratic relative error

$$\bar{\delta} = \sqrt{\frac{1}{n} \sum_{i=1}^{n} \left(\frac{k_L a'_{i\,\mathrm{exp}} - k_L a'_{i\,\mathrm{calc}}}{k_L a'_{i\,\mathrm{exp}}} \right)^2} \, . \, 100 \qquad (12)$$

which was 8.89%. The fitness of correlation can also be evaluated from Fig. 5 where the $k_L a'$ values calculated from Eq. (*11*) are plotted against the experimental ones.

When evaluating the mass transfer coefficients in liquid on a plate with mobile packing one often presumes ideal mixing of the liquid. The extent to which this

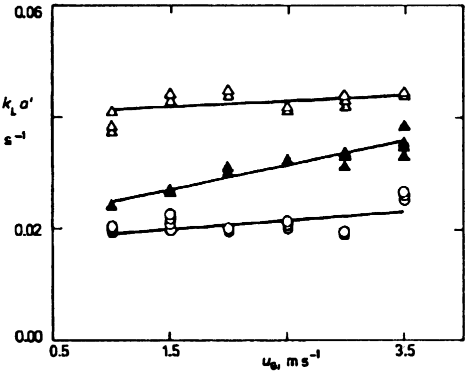

FIG. 4

Dependence of mass transfer coefficient in liquid on gas flow rate for $h_0 = 47 \, . \, 10^{-3}$ m and u_L: ○ $5.36 \, . \, 10^{-3}$ m s^{-1}, ▲ $8.93 \, . \, 10^{-3}$ m s^{-1}, △ $12.5 \, . \, 10^{-3}$ m s^{-1}

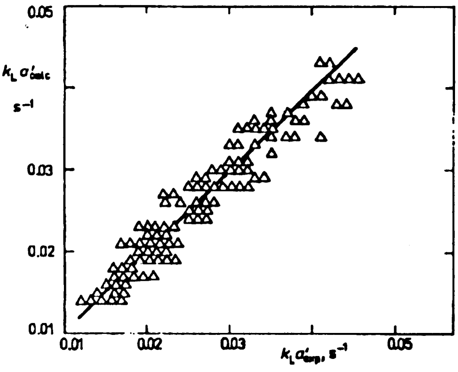

FIG. 5

Comparison of experimental and calculated values $k_L a'$ from Eq. (*11*) presumption can be accepted is seen from Figs 6 and 7 where the dependence of $k_La' = f(u_G)$ is plotted for three different models of the flow of liquid under otherwise identical conditions. The mass transfer coefficients in liquid for plug flow and ideal mixing of the liquid were obtained by usual procedure from an identical data set using the logarithmic mean of terminal driving forces. From Figs 6 and 7 it is seen that the coefficients k_La' whose calculation presumed the dispersion flow differ from the mass transfer coefficients for both the extreme cases, the difference between the coefficients being more distinct at higher flow rates of gas. From the comparison given it follows that the calculation procedure based on the presumption of ideal mixing of liquid on a plate with mobile packing is not fully justified.

CONCLUSIONS

The mass transfer was followed in a liquid on the plate of an apparatus with mobile packing. From the experimental findings obtained from desorption of CO_2 from water by action of air stream, the mass transfer coefficients were estimated (in a product with specific interfacial area) whose calculation presumed an axial dispersion in the liquid on the plate. The equation used for the calculation was obtained by integration of the differential equation describing the dependence of the concentration of the component desorbed on the height of layer at the steady state. The results are presented graphically and in the form of empirical correlation.

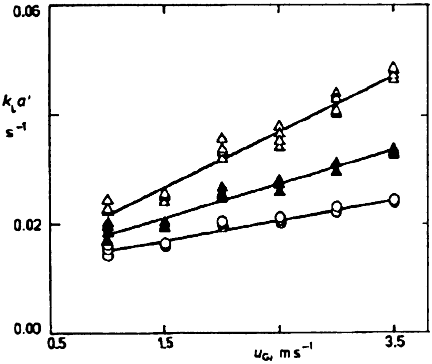

FIG. 6

Comparison of mass transfer coefficients for various models of flow of liquid at $u_L = 8.93 . 10^{-3}$ m s^{-1} and $h_0 = 21 . 10^{-3}$ m: ○ plug flow, ▲ dispersion flow, Δ ideal mixing

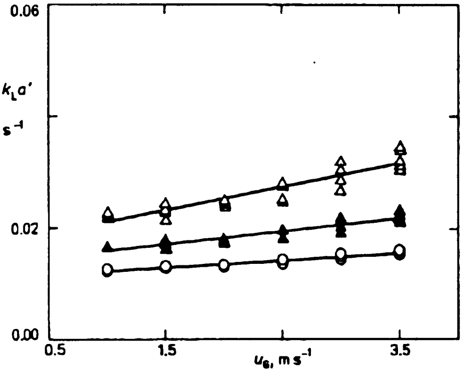

FIG. 7

Comparison of mass transfer coefficients for various models of flow of liquid at $u_L = 5.36 . 10^{-3}$ m s^{-1} and $h_0 = 32 . 10^{-3}$ m: ○ plug flow, ▲ dispersion flow, Δ ideal mixing

SYMBOLS

a	specific interfacial area (referred to volume of hold-up liquid), m^{-1}
a'	specific interfacial area (referred to volume of contact space), m^{-1}
c	molar concentration, kmol m^{-3}
D_L	dispersion coefficient in liquid, $m^2\,s^{-1}$
D_1	constant, $m^{-(p_1+p_2+p_3)}\,s^{p_1+p_2-1}$
H	Henry's constant, kmol m^{-3} Pa^{-1}
h_0	static bed height, m
h_d	height of expanded packing, m
k_L	mass transfer coefficient, m s^{-1}
M	dimensionless group of physical quantities, Eq. (*3*)
n	number of measurements
Pe	diffusion Péclet number
p	partial pressure, Pa
p_i	i = 1, 2, 3; exponents in Eq. (*10*)
R	gas constant, Pa m^3 $kmol^{-1}$ K^{-1}
T	absolute temperature, K
u	velocity, m s^{-1}
$\bar{u}_L$	mean flow rate of liquid in the layer, m s^{-1}
V	volume, m^3
$\dot{V}$	volumetric flow rate, $m^3\,s^{-1}$
Z	dimensionless length coordinate
z	length coordinate, m
z_a	height of contact space, m
$\bar{\delta}$	mean relative quadratic error, %
ε_L	dimensionless hold-up of liquid
τ_m	mean residence time of liquid on the plate, s

Subscripts

A	referring to component A
calc	calculated value
exp	experimental value
G	referring to gas
i	referring to liquid–gas interface
L	referring to liquid

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