THE EFFECT OF THE COLUMN DIAMETER ON THE HYDRODYNAMIC PARAMETERS OF BUBBLE COLUMN WITH A PLATE AS A GAS DISTRIBUTOR

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Using the concept of one-loop circulation of liquid in a bubble column at low flow rates of gas, criterions have been derived analogously to the mixed vessels enabling the effect of the column diameter to be assessed. These criteria relate the rate of energy dissipation in the liquid to the mass transfer rate. The proposed model has supported theoretically our earlier experimental findings regarding the negligible effect of column diameter on $k_L a$ in the range of column diameters between 150 and 1 000 mm.

The liquid in a bubble column is not isotropic and it has been known that principal hydrodynamic parameters (porosity e and interfacial area a) markedly vary in the volume of the bubbling layer. The gas passing the column of liquid causes rippling of the liquid level which in return induces local dynamic changes of the hydrostatic pressure above the distributor. The gas thus passes the column only through the spots of minimum instantaneous hydrostatic pressure. Spatial inhomogeneity of individual zones above the gas distributor and its incessant time variations induce macroflows (circulation) within the column of liquid. The bulk flow of gas is then entrained into the space where these macroflows exist where, as a consequence, porosity and interfacial area are markedly higher than outside the volume of the bulk flow. Partial decay of macro-vortices near the liquid level is then accompanied by significant aeration of liquid in this region. Our experience shows that in bubble columns with a perforated plate of usual geometry ($\varphi > 0{\cdot}005$) serving as a gas distributor, the character of the macroflow is not random. Instead, depending on the velocity of gas, the liquid may circulate predominantly in a single loop (a single macro-vortex directing from a certain place on the gas distributor toward the column wall where it rises to the liquid level to return back toward the distributor) or in two loops (the macro-vortex rises in region near the axis of the column toward the liquid level where it splits and turns toward the walls to descend back to the distributor). Smaller and smaller vortices superimpose on the macro-vortices up to the smallest vortices which contribute to mass transfer in the liquid. The character of the macroflows is affected by the geometry of the column while the energy manifestations of these macroflows affect the turbulent field in the liquid and the structure of the gas liquid mixture. The coefficient of mass transfer, the interfacial area and porosity thus may depen also on the scale of the column.

The papers dealing with the effect of the scale on mass transfer in bubble columns are scarce, at least as far as the available literature is concerned. Akita and Yoshoda[1] have measured the coefficient of mass transfer in three columns of the following diameters: 150, 300 and 600 mm and found that this coefficient slightly increases with the column diameter (approximately with

0·17 power). In view of the complexity of the underlying problem and the lack of information it is believed that even a simplified approach to the problem of the effect of column diameter can be of considerable interest and practical utility.

In this paper we have confined ourselves to the assessment of the effect of energy conditions within the column on the value of the volume mass transfer coefficient $k_L a$ at low gas velocities. We have started from the concept of the characteristic macro-circulations of the gas–liquid mixture within the bubble column and their relation to the value of the rate of energy dissipation. The dependence of the magnitude of the specific interfacial area and the coefficient of mass transfer has been studied earlier. The presented work has an experimental support in the measurement of k_L, a or $k_L a$ in columns ranging in diameter between 90 and 1 000 mm with a maximum height of clear liquid 1 200 mm.

THEORETICAL

Assumptions: *1*) Constant contact time of the phases. (This means that we have a priori ruled out the condition of geometric similarity). *2*) In region of low superficial gas velocities ($v_g = 0-3$ cm/s) and if the gas distributor is a perforated plate, the mixing takes place by one-loop circulation[2]. *3*) The principal macro-vortices "trace" the diameter of the column; it is assumed, in accord with our earlier measurements[2], that for v_g = const. the cross-sectional area of the ascending macro-flow is proportional to the column cross section. In other words, these two characteristic dimensions are directly proportional

$$d^2 \sim D^2 . \tag{1}$$

4) Analogously to the mechanically mixed vessels we define an energy criterion, W, as

$$W \equiv P/\varrho_1 u_1^3 d^2 , \tag{2}$$

where P is the power exerted by the stream of gas in the liquid through which the liquid is forced into the circulatory motion. The modified Euler number is defined as

$$\mathrm{Eu} = \Delta p/\varrho_1 u_1^2 , \tag{3}$$

where u_1 is the mean velocity of liquid in the macro-vortex of the circulation loop.
5) The overall rate of energy dissipation in the aerated layer, E, and the quantity P are directly proportional

$$E \sim P , \tag{4}$$

while for the aerated layer we may write[3]

$$E = \dot{V}\Delta p \quad (\mathrm{kg\ m^2/s^3}) \tag{5}$$

where $\dot{V}$ is the volume flow rate of gas and Δp is the pressure drop of gas in the heterogeneous mixture.

6) Proportionality is assumed between the criteria W and E provided that the flow rate criterion, F, is held constant

$$F = \dot{V}/u_1 d^2 \,. \tag{6}$$

or if

$$v_g/u_1 = \text{constant}\,. \tag{7}$$

It is assumed further that the systems are hydrodynamically similar if the criteria (2) or (3) and simultaneously (7) hold. Moreover, with regard to mass transfer and possible chemical reaction we introduce conditions *1* (the condition of equal residence times of the phases). Designating the compared model and reality by subscripts 1, 2 we obtain from the above assumptions the following set of equations

$$H_1 e_1/v_{g1} = H_2 e_2/v_{g2}\,, \tag{a}$$

$$H_1(1 - e_1)/w_1 = H_2(1 - e_2)/w_2\,, \tag{b}$$

$$H_1(1 - e_1)/u_{11}^2 = H_2(1 - e_2)/u_{12}^2\,, \tag{c}$$

$$v_{g1}/u_{11} = v_{g2}/u_{12}\,. \tag{d}$$

H_1 and H_2 are the heights of the layers in the two compared columns. From the condition of constant flow rate criterion it is apparent that for the case $v_{g1} = v_{g2}$ we must have that

$$u_{11} = u_{12}\,, \tag{7a}$$

in order that both systems may be similar from the view point of our considerations.

From Eq. (6) it further follows

$$\dot{V} \sim u_1 d^2$$

and substituting into (5)

$$E \sim u_1 d^2\,\Delta p\,,$$

or for bubble columns

$$E = u_1 d^2 h \varrho_1 g\,. \tag{8}$$

For the overall rate of energy dissipation related to a unit mass of liquid, E_w, we may write

$$E_w \sim u_1 d^2/A\,,$$

where A is the area of column cross section. Because $A \sim d^2$ (see assumption *1*) then

$$E_w \sim u_1\,. \tag{9}$$

For v_g = const. the rate of energy dissipation, E_w, does not depend, according to Eq. (*7a*), on the column diameter.

The parameter which reflects the transfer of mass in the liquid is the quantity e_w expressing the degree of dissipation in the liquid only and constitutes a smaller part of the overall rate of energy dissipation E_w. It has been derived that[4]

$$e_w \sim E_w\,. \tag{9a}$$

From (*7a*), (*9*) and (*9a*) it follows that

$$e_w \sim u_1 \tag{9b}$$

and further the conclusion that for v_g = constant, e_w should not depend on column diameter.

For the specific interfacial area, a, we have already derived the relation[5]

$$a \sim e_w^{0.4} e\,. \tag{10}$$

A similar relation for the mass transfer coefficient has been also found earlier[4]

$$k_L \sim e_w^{0.25}\,. \tag{11}$$

The volume mass transfer coefficient $k_L a$ is obtained from the combination of Eqs (*10*) and (*11*)

$$k_L a \sim e_w^{0.65} e\,. \tag{12}$$

According to our model the volume coefficient of mass transfer $k_L a$ thus should be affected by the column diameter only implicitly (for a given gas velocity) through the implicit dependence of porosity on the column diameter.

RESULTS AND DISCUSSION

For the assessment of the validity of the found theoretical conclusion regarding the degree of the effect of column diameter on the mass transfer coefficient it is necessary to confront these findings with experiments.

The dependence of the volume mass transfer coefficient on column diameter. The volume mass transfer coefficient was determined[6] from the rate of absorption of oxygen into in advance de-oxygenated water in 150, 300 and 1 000 mm column with the height of the layers up to 1 200 mm. It was safely established that in the range of gas velocities $v_g = 0-3$ cm/s, k_La does not depend on column diameter. (The geometry of the distributor: opening diameter $d_0 = 0{\cdot}16$ cm, relative free area of the plate $\varphi = 0{\cdot}005$.)

Dependence of the velocity of the ascending stream u_1 on column diameter. The values of the velocities of the ascending stream have been published in two papers[7,8]. In both cases it was established that this quantity is practically independent of column diameter for v_g = constant. This experimental finding appears to actually confirm our model concept leading to Eq. (*7a*).

The dependence of porosity on column diameter. According to our experience as well as that of other authors (see *e.g.* the review of Pata[9]) this dependence in the region of gas velocities under consideration is not believed to be significant.

Character of circulation of liquid. In the examined region of gas velocities ($v_g = 0-3$ cm/s) the columns with the mentioned type of gas distributor display safely the single-loop circulation starting from 300 mm in diameter column (up to 1 200 mm height of liquid investigated). The 150 mm column displayed a primary loop of the single-loop circulation just above the distributing plate but the loop did not reach as far as the liquid level. Instead, the stream forming the loop returned back toward the distributor still well within the layer. Above this loop there existed other S-shaped circulation loops. This course of circulation, which is typical for high values of H/D, does not influence our considerations because the condition *3* and Eq. (*1*) remain in effect.

In agreement with our theoretical derivation it is thus confirmed that in the region of gas velocities dealt with in this work (for bubble columns this region encompasses the majority of the situations of practical interest) the column diameter does not affect the volume mass transfer coefficient. This is a rather important finding particularly from the viewpoint of modelling of bubble columns where mass transfer (even in the presence of chemical reaction) depends also on k_L, a or k_La. According to the available results it seems that for the transfer from a model to reality it suffices that the flow criterion F be constant.* The independence of k_La on the height of the layer has been confirmed by experiments in columns between 150 and 1 000 mm

* The condition of hydrodynamic similarity, particularly that of H = constant, in the model and reality usually encounters practical difficulties in the laboratory.

in diameter for heights 600 and 1200 mm. However, it will be still necessary to test the dependence of u_1 on the height H, or the ratio H/D for greater H and column diameters. Owing to the possible variations of porosity along the height of the column for tall columns it has to be reckoned with possible dependence of u_1 on the height of the layer and with the effect of the height of the layer on k_L and particularly a.

In accord with another paper[10], in which we studied the effect of column diameter on the rate of absorption accompanied by chemical reaction in columns 90, 150 and 300 mm, we can state that the least diameter of the column, which can still yield meaningful results as far as the modelling is concerned, is 150 mm.

For direct transfer of the results on the overall rate of absorption from the model an important role is that of the residence time of the phases. Also the character of the flow of liquid and gas in the model and reality must be known (*e.g.* the coefficient of backmixing, coefficients of turbulent diffusion in conjugation with a realistic model reflecting RTD).

The problem of scale-up for agitated gas–liquid vessels has been examined recently by Miller[11] who dealt with also non-agitated bubble reactors. The basic requirements for scale-up brought up by Miller are the geometric similarity of the reactors, equal effective power input and the condition that the volumetric gas flow rates through the individual sparger opening must be kept constant (this in order to form bubbles of the same size). If the experiments are carried out at constant v_g and φ (this conditions are usually met by experiments) the third condition of Miller is automatically satisfied. The condition of geometric similarity may be justified perhaps in case of mechanically stirred vessels with an impeller where the regime of mixing can be altered by different combinations of the dimensions of the impeller and the vessel. However, the geometrical similarity in columns not agitated mechanically at constant v_g significantly changes the residence time. This, of course, could markedly affect the value of the coefficient of mass transfer. In addition, in bubble columns the character of macromixing also significantly changes (the fact that is even visually well apparent); the second condition of Miller is then practically identical to our approach leading to Eq. (*12*).

The scale-up based on constant values of the rate of energy dissipation thus appears to be a reasonable compromise for the majority of practical situations encountered both in mechanically agitated columns[12] as well as in simple bubble columns.

It should be realized that the insubstantial effect of the column diameter in the range between 150 and 1000 mm should be rated relatively to the scatter of the published results for columns of equal diameter and different systems. The data of different authors on k_La measured with the same systems and equal column diameters differ[1,13]. More careful comparative studies are necessary for columns of different diameter and different height of the layer using a standardized methods aimed particularly at testing the effect of column diameter in region above 1000 mm. It is likely that systems apt to foaming, or systems whose structure is significantly different from that of those solutions which on bubbling behave like water, will provide results leading to different conclusions. This has been deduced from the analysis of some already published data[1,13,14] (see for instance the somewhat different trends of the course of porosity plotted *versus* column diameter for water and water solutions of sulphite aerated by air[14]). One of the possibilities already pursued[15] for a more closer assessment of the role of the scale in mass transfer offers a sensitive dynamic pick-up of the frequency and amplitude characteristics of pressure pulsations in the layer. This method permits characterization of the energy spectrum of turbulence by characteristic frequency and noise as functions of column diameter and mass transfer.

LIST OF SYMBOLS

a specific interfacial area
d_0 opening diameter of plate
d equivalent diameter of the area of cross section of the ascending stream of liquid in the circulation loop
D column diameter
e porosity of layer
E overall rate of energy dissipation in gas and liquid
E_w overall rate of energy dissipation related to unit mass of liquid
e_w rate of energy dissipation in liquid only related to unit mass of liquid
Eu Euler number, Eq. (*3*)
F flow rate criterion, Eq. (*6*)
g acceleration due to gravity
H height of layer
k_L mass transfer coefficient
k_La volume mass transfer coefficient
P power input
ΔP pressure drop in layer
u_l mean velocity of ascending stream of liquid in the circulation loop
v_g superficial velocity of gas
w superficial velocity of liquid
W energy criterion, Eq. (*2*)
$\dot{V}$ volume flow rate of gas
φ relative free area of the plate
ϱ_l density of liquid

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