INTERFACIAL AREA IN BUBBLE TYPE PLATE REACTOR*

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Received November 6th, 1973

Interfacial areas were determined experimentally by the sulphite method in bubbled beds of liquids in columns 0·087, 0·152 and 0·294 m in diameter. Sieve plates with circular holes $d_0 = 1{\cdot}5-6$ mm, $F_0 = 3-6\%$ were used as distributors. The experimental data were correlated by a simple relation $a \sim (1 - e)$ based on the momentum and force balance in the heterogeneous gas–liquid bed.

Interfacial area is one of important hydrodynamic parameters which is necessary for calculation of the mass transfer rate. Here a simple relation suitable for correlation of interfacial areas has been derived. The interfacial area has been measured by the sulphite method which has already been discussed in detail earlier[1]. As distributors, sieve plates with circular holes were used. The interfacial area was not affected by the type of plate employed. As we were mostly concerned with bubbled beds of multi-stage column-type reactors a measurement was made only in a narrow interval of clear liquid heights 0·4 – 0·6 m. In this range no significant effect of the height of clear liquid on interfacial area has been observed.

THEORETICAL

A general relation for interfacial area of a heterogeneous bubble-type bed can be obtained by application of the momentum and force balance on an infinitesimal volume of the bubbled bed. The resulting balance relation is transformed by introduction of the term for friction forces by which the bubbling gas acts on liquid through the interfacial area. But application of such a relation for practical purposes is dependent on the possibility to express the friction factor between the gas and liquid and the dependence of gas porosity on space coordinates situated into the heterogeneous bed. This method of derivation of the relation for interfacial area in the gas–liquid bed formed on plates of separation columns which are characterized by small liquid holdups has been applied by Kolář[2]. In this case it is possible to assume that the hydrodynamic properties in the bed are homogeneous in the horizontal plane and are only a function of the vertical distance from the plate. This assumption significantly simplifies the situation as the relation for dependence of porosity on vertical distance from the plate can be applied which has been theoretically derived earlier[3]. Under

* Part VI in the series Gas–Liquid Reactors; Part V: This Journal *39*, 1672 (1974).

the following simplifying assumptions: *1*) negligible friction forces and surface tension acting on the physical interface of the system (walls, surface of plate); *2*) negligible momentum of liquid and volume forces acting on the gas; *3*) existence of homogeneous turbulence in the balanced infinitesimal volume; *4*) negligible linear liquid velocity as compared to that of gas; *5*) the relation $(hg\varrho_L - \Delta p_m)$ is negligible with respect to $e\,\Delta p_m$; *6*) an exponential approximation is introduced for φ_0 (porosity in the plate plane) the relation has been derived[2]

$$A \sim [\varphi_0/(1 - \varphi_0)]^{0.6}\,[(\Delta p_m h \varrho_g)/v\mu_g)]^{0.5}\,. \qquad (1)$$

The formally simple relation (*1*) seems suitable for correlation of the interfacial area of bubble-type systems which are characteristic by large liquid holdups. The crudest approximation which is made by application of relation (*1*) is the primary consideration on constancy of hydrodynamic parameters in the horizontal plane. In our recent publications[4,5] oriented to the mathematical model of porosity of the bubbled system and the effect of scaling up on porosity we have demonstrated the existence of circulation loops which of course means heterogeneity of hydrodynamic parameters in the horizontal direction. The force and momentum balances should be made in this case at least for a section including heterogeneous regions (ascending and descending streams) with differing hydrodynamic characters, however, volumes of these regions and relative velocities in both regions should be known which is rather complex[5].

Also assumption *1*, according to which the forces of surface tension are negligible, should be accepted with some reservations as for bubbled beds linear gas velocities are relatively small (up to two orders of magnitude). But taking the terms for surface tension into account would mean a significant complication in the derivation and therefore we assume that it will be more feasible to take into consideration the possible effect of surface tension on the quantity φ_0 in relation (*1*) and to express this effect by the empirically expressed dependence of φ_0. The parameter φ_0 is obviously dependent in general both on the geometrical arrangement of the system and on the flow rates of both phases and their physico-chemical properties. Actually it can only be said that φ_0 increases with increasing gas flow rate[2]. In principle, the parameter φ_0 can be measured and correlated empirically but for bubbled beds, due to low porosities in the vicinity of the plate, the measurement will be affected by considerable error. We assume in our considerations that the functional dependence for φ_0 in relation (*1*) must fulfill the condition

$$v \to 0\,, \quad \varphi_0 \to 0\,; \quad v \to \infty\,, \quad \varphi_0 \to 1\,. \qquad (2)$$

For φ_0 a simple dependence on the gas flow rate is introduced

$$\varphi_0 = 1 - [1/(1 + Kv^n)]\,, \qquad (3)$$

which satisfies the condition (2) and thus also the assumption of the dependence of φ_0 on the gas flow rate. With regard to large liquid holdups, which are for bubble-type reactors characteristic, we do not take into account (unlike for bubbled beds with low gas holdups, see Červenka[6]) the dependence of φ_0 on the clear liquid height. Then for $a' = A/H$ (m^2/m^3 of mixture) by substituting Eq. (3) into Eq. (1) the relation is obtained

$$a' = (Kv^n)^{0.6} (\Delta p_m h \varrho_g / v \mu_g)^{0.5} (1/H) . \tag{1a}$$

For bubbled beds it can be written[6]

$$\Delta p_m \sim h \varrho_L g . \tag{4}$$

Then

$$a' \sim K^{0.6} v^{(0.6n-0.5)} (1 - e) (\varrho_L \varrho_g / \mu_g)^{0.5} \tag{5}$$

Thus for the given system

$$a' = C v^{(0.6n-0.5)} (1 - e) . \tag{6}$$

Constant C must be determined from measurements of interfacial areas, exponent n can be determined from results of measurements of the dependence of porosity on distance from the plate. With respect to the probable dependence of constant K (relation (3)) on the reactor diameter it is necessary to assume that, in general, constant C is also dependent on the reactor size. Relation (6) represents the extremely simple relation for correlations of interfacial area and is in agreement with the expected[2] simple exponential function for v and Δp_m or e.

Relation for the dissipated energy in the heterogeneous gas–liquid bed can be derived theoretically from the Bernoulli equation[7]. Under the simplifying assumption of negligible free plate area occupied by the liquid (this assumption can be accepted for a plate through which no liquid flows) the energy dissipated in the bed[8] is

$$L = V(\Delta p - \Delta p_{C+t}) - W(\Delta p - \Delta p_{C+t}) + \\ + W(\varrho_g/z) V^2 (1 - \varphi^2)/\varphi^2 + (\gamma_L W - \gamma_g V) H . \tag{7}$$

For batch reactors ($W = 0$) of the bubble type ($V \rightarrow 0$) the second and fourth right hand side terms in Eq. (7) can be neglected and we can write

$$L = V(\Delta p - \Delta p_{C+t}) . \tag{8}$$

In the preceding study[9] we discussed in detail the significance of individual pressure terms and the possibility of substitution

$$(\Delta p - \Delta p_{C+t}) \sim \Delta p_m \tag{9}$$

for conditions on a plate without downcomers with liquid flow (relatively small liquid holdup, high gas velocities). For conditions of a heterogeneous bubbled bed with large liquid holdups the approximation (*9*) is justified even more and thus for the dissipated energy in the heterogeneous bed it can be written

$$L = V\Delta p_m = vS\,\Delta p_m\,. \qquad (10)$$

For the energy dissipated in a unit of bed volume $l = L/AH$ (kg/m s^{-3}) it holds

$$l = v(\Delta p_m/H)\,. \qquad (10a)$$

We assume that the energy dissipated in the heterogeneous bed above the plate is partially utilized to the formation or deformation of the interfacial area, partially for mixing of fluids in both phases *i.e.* that the interfacial area related to a unit volume per unit of time is proportional to the energy dissipated while the proportionality constant is dependent on physico-chemical behaviour of the mixture and obviously on the size of the unit as well due to a differing mixing mechanism and bed structure in columns of different diameters. Thus we can write

$$a' = C'l \qquad (11)$$

and by use of Eq. (*4*)

$$a' = C'v(1 - e)\,. \qquad (12)$$

In general the value of exponents with v and Δp_m in Eq. (*10*) could differ from one, but in the preceding studies[9] it proved useful in a similar derivation (mass transfer coefficient) when their values were equal to one. From the formal agreement of the both derived Eqs (*6*) and (*12*), it can be expected that the exponent n in Eq. (*3*) should equal to 2·5.

The relation for φ_0 which will be further on applied is

$$\varphi_0 = 1 - 1/(1 + 6{\cdot}025\,.\,10^{-5}v^{2\cdot5})\,, \qquad (3a)$$

where v is in cm.

RESULTS AND DISCUSSION

In Fig. 1 is plotted the dependence of porosities on the vertical distance from the plate determined experimentally[10,11]. In spite of a considerable scatter of experimental data three characteristic zones appear with respect to the porosity distribution along the bed height: region above the plate corresponding approximately to 1/3 of the original clear liquid height (h = 0·6 m) in which the porosity is sharply increasing from the nearly zero value just above the plate to the about constant porosity in the second region, situated in the second third of the original height of clear liquid (porosity is a function of gas velocity), while finally in the upper region again a sharp increase in porosities takes place. These regions are also in agreement with the visual observation of the heterogeneous bed: the lower third is characterized by the presence of individual bubbles or their clusters, in the upper third the bed has the character of a foam visually similar to the homogeneous moving froth which under certain conditions also forms on plates with low liquid holdups. Experimental values of limiting porosities φ_0 (actually measured 2·5 cm apart from the plate) were zero in the region of linear gas velocities 0–9 cm/s and $\varphi_0 = 0{\cdot}03$ for velocities 13·7 cm/s. With regard to the character of experimental curves $e - x$ in the region near the plate, the shape of the curve cannot be satisfactorily extrapolated to zero values and we think that the applied experimental technique (measurement of the clear liquid height at the wall, where usually the density of the bed is higher than in the central part) is not the best one for a safe determination of φ_0. In Fig. 1 are on the ordinate axis denoted points of φ_0 calculated according to relation (*3*) with the

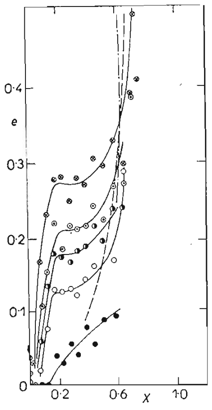

FIG. 1

Dependence of Porosity e on Distance from the Plate x(m) for the Column 150 mm in Diameter

● $v = 2$ cm/s, ○ $v = 5$ cm/s, ◑ $v = 7{\cdot}1$ cm/s, ⊙ $v = 9{\cdot}2$ cm/s, ⊗ $v = 13{\cdot}7$ cm/s. Points on the ordinate, the values of φ_0 according to Eq. (*3a*). ------ Correlation according to Eq. (*13*), φ_0 according to Eq. (*3a*); ·—·—· Correlation according to Eq. (*13*), φ_0 according to[6].

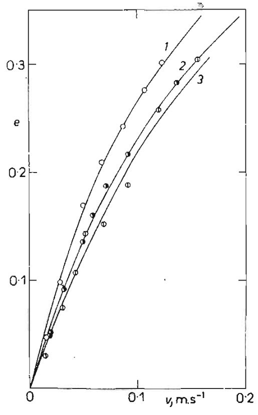

FIG. 2

Mean Porosity e of Bubbled Bed for Aqueous Solution of 0·4M-Na_2SO_3–O_2 in Dependence on v (m/s)

1 $D = 0{\cdot}294$ m, *2* $D = 0{\cdot}152$ m, *3* $D = 0{\cdot}087$ m.

exponent $n = 2{\cdot}5$. It is possible to expect that the calculated values of φ_0 will not differ much from actual values which could be determined by use of a more progressive experimental technique. Especially the agreement of experimental and calculated values of φ_0 for highest gas velocities which are plotted in Fig. 1 is worth noticing ($\varphi_{0,\mathrm{exp}} = 0{\cdot}031$, $\varphi_{0,\mathrm{calc}} = 0{\cdot}038$, according to Eq. (*3a*)).

According to Eq. (*3a*) for $v = 0{\cdot}45$ m/s the calculated value of $\varphi_0 = 0{\cdot}4$, for $v = 2$ m/s, $\varphi_0 = 0{\cdot}97$, whereas the experimental range found by Lívanský[12] for the given range of velocities was $\varphi_0 = 0{\cdot}4 - 0{\cdot}99$. So, it seams that the approximation (*3a*) can be considered sufficiently universal in a wide range of gas velocities.

To check the suitability of the proposed relations (*6*) and (*12*) and of the exponent $n = 2{\cdot}5$, an attempt has been made to express the experimental porosities by the relation proposed by Kolář[3]

$$x/h = [\varphi_H\varphi_0/(\varphi_H - \varphi_0)]\,\{(e - \varphi_0)/e\varphi_0 + \ln[(1 - \varphi_0)\,e/(1 - e)\,\varphi_0]\}\,, \quad (13)$$

which was experimentally verified for distillation plates[12] with the frequent simplification[2] $\varphi_H = 1$. The verification was made for the curve $e - x$ ($v = 13{\cdot}7$ cm/s) where, with regard to higher gas velocities, hydrodynamic conditions in the bed are close to those for distillation plates. For φ_0 relation (*3a*) was applied. In Fig. 1 the calculated dependence of porosities according to relation (*13*) is plotted as the dotted line. From this it can be seen that the calculated curve is close to the experimental one only in the regions at the top of the bed (the small deviation visible in the graph

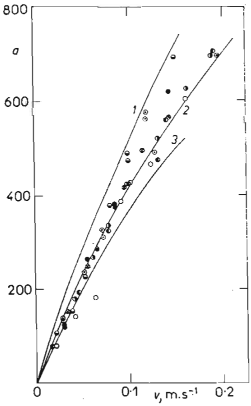

FIG. 3.

Dependence of Effective Interfacial Area, Related to Unit Bed Volume a' (m^{-1}) on Linear Gas Velocity v (*m*/s)

1 Correlation according to Eq. (*14*) for $D = 0{\cdot}294$, *2* correlation according to Eq. (*12*), *3* correlation according to Eq. (*14*) for $D = 0{\cdot}087$ m. ○ $D = 0{\cdot}087$ m, $\varphi = 3\%$, $d = 1{\cdot}6$ m, $h = 600$ mm; ◒ $D = 0{\cdot}294$ m, $\varphi = 3\%$, $d = 1{\cdot}6$ mm, $h = 600$ mm; ● $D = 0{\cdot}152$ m, $\varphi = 3\%$, $d = 1{\cdot}6$ mm, $h = 400$ mm; ⊗ $D = 0{\cdot}152$ m, $\varphi = 3\%$, $d = 1{\cdot}6$ mm, $h = 600$ mm; ⊙ $D = 0{\cdot}152$ m, $\varphi = 6\%$, $d = 1{\cdot}6$ mm, $h = 600$ mm; ◐ $D = 0{\cdot}152$ m, $\varphi = 3\%$, $d = 3$ mm, $h = 600$ mm; ◑ $D = 0{\cdot}152$ m, $\varphi = 6\%$, $d = 5$ mm, $h = 600$ mm.

can be caused *e.g.* by insufficient accuracy of experimental porosities used where the foam structure is similar to that of homogeneous moving froth. Relation (*13*) is not in agreement with the shape of the porosity dependence in the region in vicinity of plates and in the central region; this could have been expected because of the effect of forces of surface tension which were neglected in the derivation of the corresponding relation for porosity. Nevertheless, the agreement of the presented relation (*13*) in the region of homogeneous foam is very good which also proves that the made approximation (*3a*) was reasonable. In Fig. 1 is plotted the dependence of porosities according to relation (*13*) where the relation $\varphi_0 = 0{\cdot}824 \exp(-3{\cdot}6h^{1/2}v^{-1/4})$ was substituted[6] for φ_0. The agreement of experimental data with the calculated curve is not so good as with the use of our relation, obviously because the experimental relation[6] has been found as the correlation of results of heterogeneous beds with low liquid holdups and it does not fit the conditions in bubbled reactors (moreover, by use of the cited relation in Eq. (*1*) the dependence of the interfacial area on gas velocity has not been expressed satisfactorily).

From the experimental interfacial area (Fig. 2) and porosities of the heterogeneous bed (the system aqueous solution of 0·4M-Na_2SO_3—O_2) (Fig. 3), the pairs (a, e) were taken for individual v and the average value of constant C' was calculated from relation (*6*) or (*12*). For the column diameter 152 mm, $C' = 54{\cdot}62$. The agreement of experimental and calculated interfacial areas is in the range $\pm 1{\cdot}9$ relative %. The proposed correlation fits the reality better than the relation by Sharma and Mashelkar (exponential dependence $v^{0.7}$) or that by Reith[10] though the quantity a cannot be calculated by our method without knowing the porosity (or holdup of the bubbled bed).

In one of our earlier studies[13] the relation between the porosity of the heterogeneous bubbled bed and interfacial area has been studied. From relations (*4*) and (*11*) of the cited paper *i.e.* from the relation according to Reith for interfacial area $a' = Kv/(2v + 0{\cdot}2)$ and from our relation for porosity $e = \alpha K^{7/15} v^{4/5}/(2v + 0{\cdot}2)^{7/15}$, an approximate relation results for ratios of porosities in the column of a given diameter to that of compared diameter which is proportional to the corresponding ratio of interfacial areas

$$(a_1/a_2) \sim (e_1/e_2)^2 \,. \qquad (14)$$

From Fig. 2 was determined the ratio of porosities in two columns of different diameters at the standard velocity $v = 0{\cdot}1$ m/s in the following manner: $(\bar{e}_{0.1})_{300} : (\bar{e}_{0.1})_{150} = 1{\cdot}124$; $(\bar{e}_{0.1})_{150}/(\bar{e}_{0.1})_{90} = 1{\cdot}068$. From relation (*14*) were then calculated the interfacial areas in columns 90 and 300 mm (curves *1*, *3* Fig. 3). The agreement of experimental and calculated interfacial areas for various reactor diameters is surprisingly good, especially when we realize that for derivation of Eq. (*14*) relation for interfacial area according to Reith was applied, which – as has already been said – is not the most accurate one (this relation was used as interfacial area was there given explicitely and was not a function of porosity). The mean deviation between mean experimental and calculated interfacial areas is less than relative 15%.

LIST OF SYMBOLS

A interfacial area related to unit area of distributing plate
a' interfacial area related to unit of volume of heterogeneous bed
C, C' constants dependent on the given system
D diameter of column
e porosity of bed
g gravitational acceleration
h clear liquid height
H height of bubbled bed
K constant (Eq. (3))
L energy dissipated in the bed
l energy dissipated in the foam per unit bed volume
Δp_m pressure drop in heterogeneous bed
Δ_p over-all pressure drop (bed + plate)
Δp_{C+t} pressure drop due to concentration and friction across the plate
S plate area
v linear gas velocity
V volumetric gas flow rate
W volumetric liquid flow rate
x distance from the plate
φ_H porosity at the reference plane $x = H$
φ_0 porosity at the plate plane, $x = 0$
φ free plate area
ϱ_G density of gas
ϱ_L density of liquid
μ_G viscosity of gas
ν_L specific gravity

REFERENCES

1. Nývlt V., Kaštánek F.: This Journal, in press.
2. Kolář V.: This Journal *35*, 3678 (1970).
3. Kolář V.: Chem. Eng. Sci. *24*, 1285 (1969).
4. Kaštánek F., Nývlt V., Rylek M.: This Journal, in press.
5. Zahradník J., Kaštánek F., Rylek M.: This Journal *39*, 1403 (1974).
6. Červenka J.: *Thesis*. Czechoslovak Academy of Sciences, Prague 1972.
7. Standart G.: Chem. Eng. Sci. *20*, 611 (1965).
8. Standart G., Rylek M.: This Journal *31*, 2159 (1966).
9. Kaštánek F., Rylek M.: This Journal *35*, 3368 (1970).
10. Reith T.: *Thesis*. Technische Hogeschool, Delft 1968.
11. Nývlt V.: *Thesis*. Czechoslovak Academy of Sciences, Prague 1973.
12. Lívanský K.: *Thesis*. Czechoslovak Academy of Sciences, Prague 1969.
13. Kaštánek F., Nývlt V., Rylek M.: This Journal *39*, 528 (1974).

Translated by M. Rylek.