HYDRODYNAMIC CHARACTERISTICS OF GAS-LIQUID BEDS IN CONTACTORS WITH EJECTOR-TYPE GAS DISTRIBUTORS

Jindřich ZAHRADNÍK, František KAŠTÁNEK, Jan KRATOCHVÍL and Milan RYLEK

Institute of Chemical Process Fundamentals, Czechoslovak Academy of Sciences, 165 02 Prague 6 - Suchdol

Received December 11st, 1981

The effect of ejector-nozzle geometry on gas holdup and on the rate of interfacial mass transfer characterized by values of k_La was studied in a tower reactor with ejector-type gas distributor. It has been established that both gas holdup and k_La values are, in contactors of this type, unambiguously determined by the rate of energy dissipation in the place of gas–liquid dispersion formation *i.e.* in the ejector. No effect of nozzle type and geometry was observed on the character of dependences of gas holdup and k_La on the energy dissipation rate and consequently on the values of coefficients of empirical exponential-type relations used for experimental data correlation.

Tower reactors with ejector-type gas distributors have been recently often recommended for chemical and biotechnological processes in cases when the interfacial mass transfer is the rate-controlling step of the process[1]. It has been established that the ejector-type distributors utilizing the kinetic energy of the liquid jet for dispersion of gas throughput can ensure high intensity of interfacial contact in two-(three)-phase beds and thus bring a sufficiently large amount of gaseous reactant into contact with the liquid phase. Unlike in the case of other gas distributing devices commonly used in bubble-column reactors as *e.g.* perforated plates, sintered glass or metal plates, the decisive part of energy has to be in this case supplied for the circulation of the liquid phase through an ejector while gas can be either supplied to the ejector under pressure (forced supply) or can be sucked into the ejector due to pressure decrease in its expansion (suction) chamber (free suction). Despite the attention which has been lately paid to the ejector gas distributors, works aiming at the establishing of a general basis for ejector-distributor design and scaling--up can be found only scarcely up to now[2], most of the papers devoted to this topic being just application reports. As a result, there are still many questions left open concerning the factors determining the efficiency of these gas distributing devices and the effect of these factors on the character of gas–liquid beds formed in ejector-distributor contactors. At the present time, there are no sufficiently general relations available for the dependence of characteristic parameters of gas–liquid beds on ejector construction parameters. At the same time, it can be envisioned that in reactors with these gas distributing devices the character of gas–liquid (multiphase) bed can be influenced by the distributor construction much more significantly than *e.g.* in reactors with perforated (sintered) plate gas distributors and useful general relations between ejector construction parameters and gas–liquid bed characteristics would be thus of vital importance.

The analysis of information available in literature and our own previous experience with ejector gas distributors[3] made us believe that any *a priori* analysis aimed at theoretical derivation of quantitative relations between decisive construction parameters of ejectors and characteristics of gas–liquid beds represents a hardly feasible task at the present state of knowledge. Further progress towards a quantitative description of gas–liquid bed formation in contactors with ejector gas distributors has to be therefore based upon experimental evidence on the effect of individual factors governing the ejector performance on the properties of resulting gas–liquid bed.

It can be generally assumed that the character of gas–liquid bed formed by an ejector distributor depends on flow-rates of both phases and on their physico-chemical properties, on the construction parameters of the ejector and on the reactor geometry (*i.e.* on the height and diameter of the gas–liquid bed). As for the ejector, its performance (dispersing efficiency) can be expected to be determined (in the case of common Venturi-tube type ejector) by the type of liquid nozzle and by the shape and cross-sectional area of its outlet orifice, by the size and shape of the chamber where the gas is introduced into the ejector (suction chamber in the case of free suction) and by the size and geometry of the diffuser (mixing chamber), *i.e.* by its outlet cross-sectional area, total length and cone angle.

It has been the aim of our work to study (in a specific contactor and for a specific two-phase system) the effect of selected ejector construction parameters on the character of gas–liquid bed formed in the contactor and on the intensity of interfacial mass transport in the bed. The study was limited to the ejector with free suction of air. In such a case the total energy supplied to the system is only that needed for pumping of the liquid circulated through the ejector. The rate of gas suction and consequently the volumetric gas flow rate in the contactor are then determined by the rate of liquid circulation and by ejector parameters and cannot thus been independently adjusted. From the ejector construction parameters stated above, nozzle type and geometry were chosen as independent variables considering the decisive effect of these factors on the pressure drop across the ejector and consequently on its sucking capacity. Gas holdup ratio (bubble bed porosity), ε_G, and volumetric liquid-side mass transfer coefficient, k_La were chosen to characterize quality of gas dispersion and intensity of interfacial mass transport in resulting gas–liquid beds.

EXPERIMENTAL

The schematic sketch of experiment set-up is given in Fig. 1. Glass-wall column reactor (I.D. 0·292 m) was used for experiments and the ejector distributor was mounted by a flange to the column bottom. Common type of Venturi-tube ejector was used drawn schematically in Fig. 2. Length of diffuser was 0·43 m, its crosssectional area at the outlet 0·02 m^2 and the ratio of diffuser outlet to inlet diameter was 2·5. Eight nozzles of thre different types (*a*, *b*, *c*) were used (Fig. 3) all of them having circular cross-section. Six nozzles of the type *a* were used with outlet orifice diameters ranging between 6 and 11 mm, orifice diameters of the *b*- and *c*-type nozzles were 6 and 8 mm respectively. Experiments were carried out in air–water system at zero absolute throughput of the liquid phase (semibatch arrangement). The ratio of clear liquid height to the column diameter (H_0/D_K) equaled five in all experiments. The pump EPK-70 (Sigma, Lutín) was used for liquid circulation, the liquid flow rate through the ejector was regulated by by-pass adjustment keeping the pump discharge constant. The circulation flow rates ranged for individual nozzles between $0{\cdot}5 . 10^{-3}$ and $2{\cdot}0 . 10^{-3}$ m^3 s^{-1}, corresponding gas suction rates were $0{\cdot}28 . 10^{-3}$

to $5 \cdot 04 . 10^{-3}$ m^3 s^{-1}, gas superficial velocities thus being in the region from 0·004 to 0·075 m . . s^{-1}. For all the nozzles used the values of characteristic gas–liquid bed parameters ε_G and k_La were determined as functions of superficial gas velocity in the reactor *i.e.* of the liquid circulation flow rate. Values of ejector pressure drop were also determined for the liquid flow rates and ejector nozzles used.

The dynamic method was applied for k_La measurements based upon monitoring of the unsteady oxygen absorption into previously deoxygenized water in the bed *i.e.* on the evaluation of a system response to an input step change nitrogen–air. The Clark-type fast response oxygen electrode was used for measurement of dissolved oxygen concentration. Complete mixing of the liquid phase in bubble bed was assumed for the probe response evaluation, the dynamics of the oxygen electrode was described by the two-regional model proposed by Linek and Beneš[4]. Gas holdup values were determined by the bed expansion method *i.e.* from the difference of clear liquid height and the overall height of aerated bed. The ejector pressure drop was calculated as a difference of the pump discharge pressure and the bubble-bed pressure drop.

RESULTS AND DISCUSSION

In Figs 4 and 5 experimental data of gas holdup and k_La obtained for all the nozzles studied are plotted against the superficial gas velocity, w_G. Graphs clearly prove the existence of unambiguous dependences of gas holdup and k_La on w_G. Both these

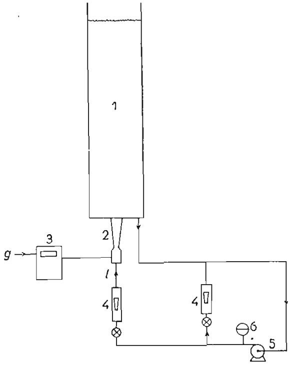

FIG. 1

Experiment set-up. 1 Glass-wall column, 2 Venturi-tube ejector, 3 gas flow meter, 4 rotameters, 5 liquid pump, 6 manometer

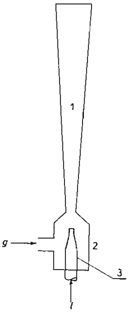

FIG. 2

Venturi-tube ejector. 1 Diffuser, 2 suction chamber, 3 nozzle dependences were linear in the whole region of experimental gas flow rates and independent on the nozzle type and geometry. Understandably however these factors did influence the dependence of gas suction rate $\dot{V}_G$ on the liquid circulation

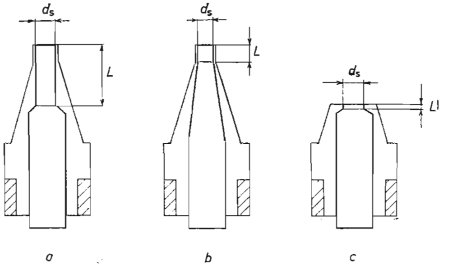

FIG. 3

Ejector nozzles. Type *a* $L = 25$ mm, $d_s = 6-11$ mm; type *b* $L = 7$ mm, $d_s = 6$ mm; type *c* $L = 1.5$ mm, $d_s = 8$ mm

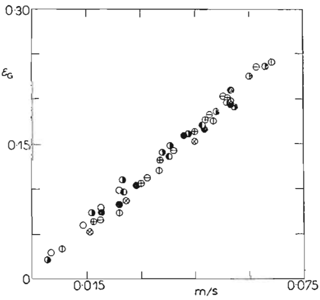

FIG. 4

Gas holdup as a function of superficial gas velocity. Nozzle *a*: ○ $d_s = 11$ mm, ⊗ $d_s = 10$ mm, ● $d_s = 9$ mm, ◐ $d_s = 8$ mm, ◑ $d_s = 7$ mm, ⊕ $d_s = 6$ mm; nozzle *b*: ⦶ $d_s = 6$ mm; nozzle *c*: ⊖ $d_s = 8$ mm

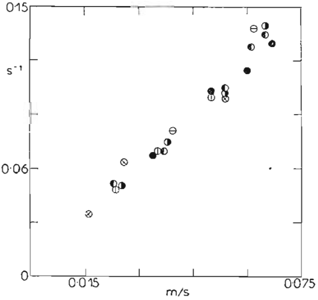

FIG. 5

Volumetric liquid-side mass transfer coefficient as a function of superficial gas velocity. Nozzle *a*: ⊗ $d_s = 10$ mm, ● $d_s = 9$ mm, ◐ $d_s = 8$ mm, ◑ $d_s = 7$ mm; nozzle *b*: ⦶ $d_s = 6$ mm; nozzle *c*: ⊖ $d_s = 8$ mm flow rate, Q_L (Fig. 6), as the suction capacity of an ejector depends on the ejector (nozzle) pressure drop, ΔP_e which is a function of the liquid flow rate in nozzle orifice and of the nozzle construction.

In Fig. 7 pressure drop characteristics of all the nozzles studied are compared. Graphs demonstrate direct proportionality of the ejector pressure drop, ΔP_e, on the correlation term $w_{Ls}^2 \varrho_L/2$ obtained by rearrangement of the expression for the loss height in the Bernouilly equation, the proportionality coeficients (slopes of the graphs) having physical meaning of nozzle pressure drop coefficients, ξ. Values of these coefficients varied with the nozzle orifice cross-section (with the ratio of nozzle cross-sectional area to that of liquid supply piping) and with the character of cross-section reduction. It may seem surprising that higher values of ΔP_e (and consequently of ξ) were observed for the *c*-type nozzle in comparison with data for the equal diameter *a* type nozzle. Such a result corresponds however well with the dependence of the pressure drop coefficient K of the McAllister, McGinnis, Plank equation for dry sieve-tray pressure drop on the ratio of plate thickness to plate hole diameter, T/d_0, presented by Kaštánek and Rylek[5]. This dependence has a minimum for $T/d_0 = 2{\cdot}5$. Assuming that the nozzle can be considered as a limiting case of a per-

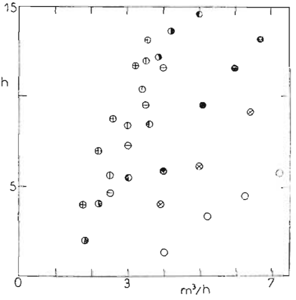

FIG. 6

Rate of gas suction as a function of circulating liquid throughput. Nozzle *a*: ○ $d_s = 11$ mm, ⊗ $d_s = 10$ mm, ● $d_s = 9$ mm, ◐ $d_s = 8$ mm, ◑ $d_s = 7$ mm, ⊕ $d_s = 6$ mm; nozzle *b*: ⦶ $d_s = 6$ mm; nozzle *c*: ⊖ $d_s = 8$ mm

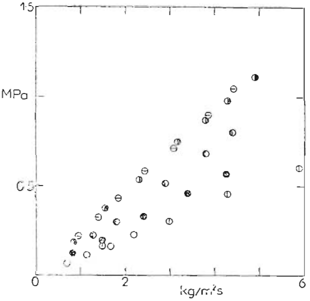

FIG. 7

Ejector pressure drop as a function of correlation term ($w_{Ls}^2 \varrho_L/2$). Nozzle *a*: ○ $d_s = 11$ mm, ● $d_s = 9$ mm, ◐ $d_s = 8$ mm, ◑ $d_s = 6$ mm; nozzle *b*: ⦶ $d_s = 6$ mm; nozzle *c*: ⊖ $d_s = 8$ mm forrated plate (a single orifice plate), the nozzle parameter L/d_s (Fig. 3) can be considered as analogous to T/d_0 for sieve trays. It can be then seen that the value K (and consequently that of ΔP) corresponding in the Kaštánek and Rylek's graph[5] to the ratio $L/d_s = 0{\cdot}2$ (nozzle c) is significantly higher than that for $L/d_s = 3$ (nozzle a).

It has been shown in our recent paper[6] devoted to the comparison of sieve-tray and ejector-type gas distributors that significantly higher values of gas holdup and k_La can be observed in ejector-distributor contactors compared with data obtained at equal gas flow rates in beds formed on sieve-trays. Results of such comparison which have been proved also by our present data shown in Figs 4 and 5 suggest that application of ejector gas distributors enables better gas utilization in the reactor It can be however seen from Fig. 8 that the dependence of k_La on gas holdup ratio is almost identical for both types of gas distributors concerned in the whole region of superficial gas velocities examined in our present work ($w_G = 0{\cdot}075\ \mathrm{ms}^{-1}$) and the experimental data can be reasonably well fitted by a single exponential function

$$k_La = A\varepsilon_G^y\,. \tag{1}$$

The least squares method yielded values $A = 0{\cdot}590\ \mathrm{s}^{-1}$, $y = 1{\cdot}015$ independent on geometry of nozzles and sieve trays for which the k_La and ε_G data were obtained Apparently this experimental evidence seems to suggest that it was solely the increase of gas holdup and consequently of the specific interfacial area, a, which was responsible for the higher values of k_La observed in gas–liquid beds created by an ejector gas distributor in comparison with sieve-tray distributor contactors. Values of the liquid-side mass transfer coefficient, k_L, were on the other hand apparently more or less identical for both types of distributors despite significantly different

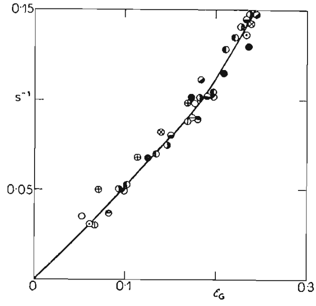

FIG. 8

Dependence of k_La on gas holdup ratio. Ejector distributor. Nozzle a: ○ $d_s = 10$ mm, ● $d_s = 9$ mm, ◐ $d_s = 8$ mm, ◑ $d_s = 7$ mm; nozzle b: ⊕ $d_s = 6$ mm; nozzle c: ⊗ $d_s = 8$ mm. Sieve-tray distributor: ⦶ $\varphi = 0{\cdot}2\%$, ⊖ $\varphi = 0{\cdot}5\%$, ⊙ $\varphi = 1{\cdot}0\%$, $d_0 = 3$ mm; ◓ $\varphi = 0{\cdot}2\%$, ◒ $\varphi = 0{\cdot}5\%$, ◕ $\varphi = 1{\cdot}0\%$, $d_0 = 1{\cdot}6$ mm; ◔ $\varphi = 0{\cdot}2\%$, $d_0 = 1{\cdot}0$ mm. Data calculated from Eq. (1) for $A = 0{\cdot}59$, $y = 1{\cdot}015$

macro-scale flow patterns observable visually in beds formed by ejector- or sieve-tray distributors respectively. Indeed such considerations comply well both with the theory of interfacial mass transfer[7] and with experimental evidence demonstrating that only increase of micro-scale turbulence can contribute to the increase of intensity of interfacial contact[8].

It has been proved experimentally[7,9] that in bubble beds formed by sieve-trays the interfacial area and consequently the intensity of interfacial mass transport depend on the total rate of energy dissipation in both phases defined in this case as $E_{db} = \Delta P_v \dot{V}_G$ or related to the unit mass of the bed $e_{db} = \Delta P_v \dot{V}_G / V_L \varrho_L$. If an ejector distributor is used for gas dispersion it can be assumed analogically that the intensity of interfacial mass transport depends on the rate of energy dissipation in the place of two-phase dispersion formation (*i.e.* in the ejector) which is determined by the rate of liquid circulation and by the ejector (nozzle) pressure drop

$$E_{de} = \Delta P_e Q_L \tag{2}$$

or related to a bed mass unit

$$e_{de} = \Delta P_e Q_L / V_L \varrho_L \,. \tag{3}$$

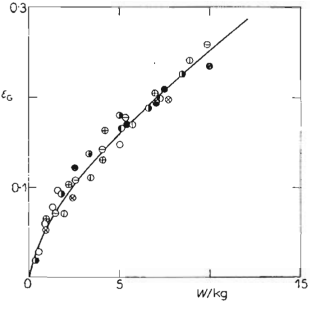

FIG. 9

Gas holdup as a function of the rate of energy dissipation. Nozzle *a*: ○ $d_s = 11$ mm, ⊗ $d_s = 10$ mm, ● $d_s = 9$ mm, ◐ $d_s = 8$ mm, ◑ $d_s = 7$ mm, ⊕ $d_s = 6$ mm; nozzle *b*: ⦶ $d_s = 6$ mm; nozzle *c*: ⊖ $d_s = 8$ mm. Data calculated from Eq. (*5*) for $a_1 = 0{\cdot}05$, $b_1 = 0{\cdot}69$

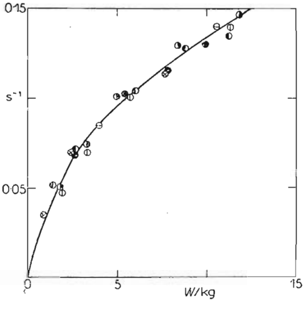

FIG. 10

Dependence of $k_L a$ on the rate of energy dissipation. Nozzle *a*: ⊗ $d_s = 10$ mm, ● $d_s = 9$ mm, ◐ $d_s = 8$ mm, ◑ $d_s = 7$ mm; nozzle *b*: ⦶ $d_s = 6$ mm; nozzle *c*: ⊖ $d_s = 8$ mm. Data calculated from Eq. (*6*) for $a_2 = 0{\cdot}04$, $b_2 = 0{\cdot}54$

In Figs 9 and 10 data of gas holdup and $k_L a$ are plotted against the specific rate of energy dissipation in ejector, e_{de}. Both graphs clearly demonstrate that in spite of significantly different pressure drop characteristics of individual nozzles (see Fig. 7), the values of gas holdup and $k_L a$ were dependent solely on the specific rate of energy dissipation in ejector, e_{de}, which can be expressed after substitution for ΔP_e, Q_L and V_L by relation

$$e_{de} = \xi(\pi/8)\, w_{Ls}^3 d_s^2 \,. \tag{4}$$

Experimental dependences ε_G *vs* e_{de} and $k_L a$ *vs* e_{de} presented in Figs 9 and 10 thus prove validity of our assumption on the decisive role of the rate of energy dissipation in the ejector on the intensity of interfacial contact in gas–liquid beds formed in contactors with ejector gas distributors. Both dependences were well fitted by exponential type relations

$$\varepsilon_G = a_1 e_{de}^{b_1} \,, \tag{5}$$

$$k_L a = a_2 e_{de}^{b_2} \,. \tag{6}$$

Values of coefficients $a_1 = 0{\cdot}05$; $b_1 = 0{\cdot}69$; $a_2 = 0{\cdot}04$; $b_2 = 0{\cdot}54$ calculated from our experimental data by the least squares method were independent on the nozzle type and geometry. In Fig. 11 the total rate of energy dissipation corresponding to a unit of gas volume supplied per a time unit into the bed $\Delta P_e Q_L / \dot{V}_G$ is plotted against the volumetric gas flow rate $\dot{V}_G$. The graph shows that with increasing rate of gas supply to the bed the amount of energy dissipated per a unit of the gas flow rate also increases *i.e.* that the specific amount of energy supplied (consumed)

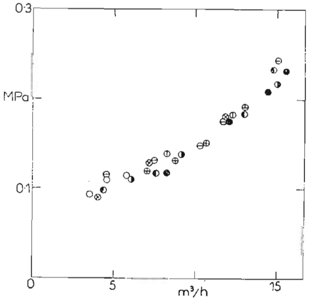

FIG. 11

Energy consumed for dispersion of a gas volume unit as a function of gas throughput. Nozzle *a*: ○ $d_s = 11$ mm, ⊗ $d_s = 10$ mm, ● $d_s = 9$ mm, ◐ $d_s = 8$ mm, ◑ $d_s = 7$ mm, ⊕ $d_s = 6$ mm; nozzle *b*: ⦶ $d_s = 6$ mm; nozzle *c*: ⊖ $d_s = 8$ mm for dispersion of a unit volume of gas is higher at higher gas throughputs. This observed increase of specific energy consumption can be apparently viewed as a characteristic feature of ejectors with free suction of gas distinguishing them significantly *e.g.* from the various types of plate distributors for which the energy supplied for dispersion of a unit of gas volume is constant (independent on gas flow rate) and is determined by the bed pressure drop $\Delta P_v = H_0 \varrho_L g$ *i.e.* by the clear liquid height and by liquid phase density. It can be assumed that better gas utilization in contactors with ejector gas distributors (compared with sieve-tray distributor contactors) discussed above can be related to these differences in the energy amount supplied for dispersion of a gas volume unit and in its respective changes with gas flow rate. On the other hand it can be envisioned that the increase of specific energy consumption (related to a gas volume unit) with increasing gas-flow rate observed in ejector-distributor contactors affects negatively the energetic performance of these contactors.

It can thus be concluded that the results of our experiments have proved that the character of gas–liquid bed and the intensity of interfacial contact in the bed are even in contactors with ejector gas distributors unambiguously determined by the rate of energy dissipation in the place of dispersion formation *i.e.* in this case in the ejector nozzle. The nozzle type and geometry determine the rate of energy dissipation for a specific liquid phase throughput they however do not influence relations between the rate of energy dissipation and hydrodynamic characteristics of resulting gas–liquid beds. In a specific ejector-distributor contactor, a single specific value of the energy dissipation rate (related to a unit of bed mass) corresponds to each demanded value of $k_L a$ (representing the intensity of interfacial contact) independently on the relative contribution of individual variables ΔP_e, Q_L, V_L, ϱ_L determining the rate of energy dissipation.

It can be further assumed that the total amount of energy dissipated in a time unit in the ejector can be divided into three terms due to the way of its utilization: the energy spent for gas suction (in the case of free suction gas supply) *i.e.* energy consumed on the account of gas supply pressure losses, energy directly spent for gas dispersion *i.e.* for bubble bed formation, and the unprofitably dissipated energy (energy losses). While it has been proved that the total rate of energy dissipation is determined by the nozzle type and geometry it can be envisioned that the efficiency of utilization of the dissipated energy (*i.e.* the ratio of the total energy directly utilized for gas dispersion) depends on other construction parameters of ejectors as *e.g.* on the suction orifice diameter, on the size and shape of ejector (suction) chamber, and on the shape and geometry of diffuser (mixing chamber). Validity of this assumption has been proved qualitatively by results of our preliminary experiments devoted to estimation of the effect of suction orifice diameter and diffuser construction on the quality of gas dispersion. Similarly, Zlokarnik's work[2] can be mentioned reporting a favourable effect of slot shaped nozzle and mixing chamber on the quality of interfacial contact in resulting bubble bed. Further experimental evidence is nevertheless still needed to obtain a base for quantitative estimation of the effect of construction parameters of ejector diffuser and suction chamber on the ejector performance and its energy effectiveness. Realization of experimental programs aimed at obtaining such data is therefore to be recommended. Results of an experimental study devoted to comparison of dispersion efficiency and energetic performance of ejector distributors with free gas suction and with forced gas supply would be also very useful for better understanding of ejector-distributors performance and thus for formulation of principles of ejector-distributors design and scaling-up.

Finally it would be apparently very useful to study in detail the effect of bubble-bed pressure drop (clear liquid height) on parameters of bubble bed (on the efficiency of gas-liquid dispersion formation) in a wide range of $\Delta P_v(H_0)$ including such conditions when the bed pressure drop $\Delta P_v = H_0 \varrho_L g$ is comparable with the ejector pressure drop, ΔP_e, *i.e.* when both constituents of the total circulation pump discharge pressure, $P_i = \Delta P_e + \Delta P_v$, which has to be considered in evaluation of the total energy effectiveness of ejector-distributor contactors[6], are of the same order. Results of such experiments would contribute valuably both to better understanding of principles of buble bed formation in contactors with ejector-type gas distributors and to the estimation of the effect of ejector-distributor reactors scaling-up on their energetic performance.

LIST OF SYMBOLS

a	specific interfacial area
a_1, a_2	correlation coefficients (Eq. (*5*)—(*6*))
A	correlation coefficient (Eq. (*1*))
b_1, b_2	correlation coefficients (Eq. (*5*)—(*6*))
d_0	plate-hole diameter
d_s	nozzle diameter
D_K	column diameter
e_{db}	rate of energy dissipation in a unit of bubble bed mass
e_{de}	rate of energy dissipation in an ejector related to a unit of bubble bed mass
E_{db}	total rate of energy dissipation in bubble bed
E_{de}	total rate of energy dissipation in an ejector
g	gravitational acceleration
H_0	clear-liquid height
k_L	liquid-side mass transfer coefficient
L	characteristic dimension of an ejector nozzle
P_i	pump discharge pressure
ΔP_e	ejector pressure drop
ΔP_v	bubble-bed pressure drop
Q_L	liquid throughput (rate of liquid flow through an ejector)
T	plate thickness
$\dot{V}_G$	volumetric gas flow rate

V_L volume of liquid in a bubble bed
w_G superficial gas velocity
w_{Ls} liquid velocity within a nozzle orifice
y correlation coefficient (Eq. (*1*))
ε_G gas holdup ratio
ξ pressure drop coefficient
ϱ_L liquid phase density

REFERENCES

1. Van Landeghem H.: Chem. Eng. Sci. *35*, 1912 (1980).
2. Zlokarnik M.: Chem. Eng. Sci. *34*, 1265 (1979).
3. Kaštánek F., Zahradník J., Rylek M., Čermák J., Kratochvíl J.: Int. Chem. Eng. *20*, 57 (1980).
4. Linek V., Beneš P.: Biotechnol. Bioeng. *19*, 741 (1977).
5. Kaštánek F., Rylek M.: Verfahrenstechnik *4*, 303 (1970).
6. Zahradník J., Kratochvíl J., Kaštánek F., Rylek M.: Chem. Eng. Commun., in press.
7. Kaštánek F.: This Journal *42*, 2491 (1977).
8. Zakrzewski W., Lippert J., Lübbert A., Schügerl K.: Chem.-Ing. Tech. *53*, 135 (1981).
9. Kaštánek F.: This Journal *41*, 3709 (1976).

Translated by M. Rylek.