ON FLOW DYNAMICS IN A JET-LOOP REACTOR

Andrzej KMIEC, Lilla SZCZEPANIAK and Sahar S. H. ABDUL-LATIF

Institute of Chemical Engineering and Heating Equipment,
Wroclaw Technical University, 50-373 Wroclav, Poland

Received October 1, 1993
Accepted August 8, 1994

Results of experimental investigation of liquid flow velocity distribution, gas hold-up, and backmixing in the liquid phase in a jet-loop reactor with chambers of square cross-section are presented. Liquid velocity profiles in the draft tubes have been described applying the boundary layer theory. Aeration rates have been found to be comparable to those reported earlier for the reactor with chambers of circular shape as well as those obtained for reactors with other types of injectors. The axial dispersion coefficients in liquid phase have been found to depend very strongly on the air velocity in the draft tubes.

Gas–liquid contactors with ejector-type gas distributors have been recommended for chemical processes (Cramers et al.[1], Malone[2]) as well as for waste water treatment (Wachsmann et al.[3], Kmiec et al.[4], Yenkie et al.[5]). Such a reactor consists of a tank, an external forced liquid circulation loop, and an ejector located at the top or the bottom of the tank and directed downward or upward. Several authors investigated the hydrodynamics and the mass transfer of loop-Venturi reactors (Zahradnik et al.[6], Dutta et al.[7], Cramers et al.[8]). Tebel and Zehner[9] showed that in the case of the jet-loop reactor the whole injector–reactor system has particular characteristics regarding the dynamics of gas–liquid two-phase flow. Their analysis, however, concerned the jet-loop reactor with chambers of circular shape. The aim of the present paper is to give the results of experimental investigation of liquid velocity distribution, gas hold-up, and axial dispersion in the liquid phase in a jet-loop reactor with two chambers of square cross-section. Such a configuration seems to be justified for scale-up reasons in industrial applications of this type of reactors.

EXPERIMENTAL

Apparatus

The experimental set-up is shown in Fig. 1. The liquid stored in tank 1 is pumped by means of a centrifugal pump 2 through a rotameter 4 and injectors 5 to chambers with draft tubes 6 of the reactor. From the reactor, the aerated liquid flows back to the tank 1 by means of a syphon pipe under pressure of the liquid recirculating inside the reaction chamber 7. The reactor composed of two chambers 7 of square cross-section 0.12 × 0.12 m and 0.6 m height has the total volume of 17.28 dm^3. The draft tubes also of prism shape (0.07 × 0.07 × 0.5 m) are fixed in the axes of reactor chambers at a distance of 75 mm from the cover and 25 mm from the lower edge of the reactor. Figure 2 shows the design of injectors with the outlet nozzle of circular shape of 8 mm diameter and the suction tube for air of 4 mm inside and 6 mm outside diameters.

The injectors 5 (Fig. 1) sucking in atmospheric air through the vertical inner sucking pipe (see Fig. 2) accelerate the gas–liquid two-phase stream, which has a shape of circular jet, further on in the draft chamber of the reactor. The flow rate of liquid and hence the linear liquid flow velocity in the narrowest annular cross-sectional area of the nozzle (shown in Fig. 2) are controlled by means of valves Z_4, Z_5 (see Fig. 1). The aeration rate of stream inside the reactor also depends on the degree of their opening since the sucking pipe of the injector is not controlled.

Methods of Measurements

The liquid flow velocity in cross-section of the draft tube was measured by means of a Prandtl tube connected with a U-tube filled with carbon tetrachloride. The Prandtl tube was movable in the horizontal direction with the help of two fixing screws placed on the wall of the reactor at the distances of 105 and 230 mm from the end of the nozzle. The measurements were carried out for four

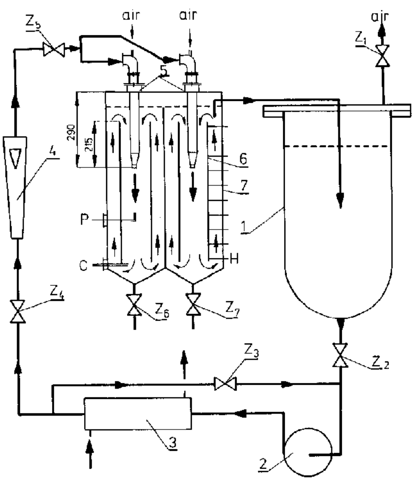

FIG. 1
Schematic diagram of the experimental set-up: 1 tank, 2 pump, 3 heat exchanger, 4 rotameter, 5 injectors, 6 draft tube, 7 reactor; C probe of conductometer, H pressure taps, P Prandtl tube, Z_1 – Z_7 valves

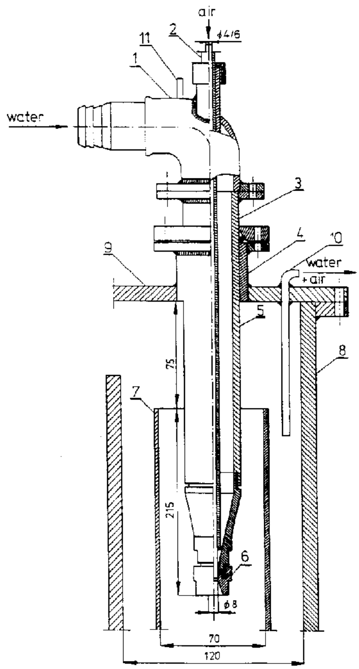

FIG. 2
Design of injector: 1 elbow, 2 sucking tube, 3 extension tube, 4 fixing tube, 5 nozzle, 6 nozzle diffuser, 7 draft tube, 8 reactor, 9 cover, 10 syphon pipe, 11 overpressure tap

flow modes: liquid flow without aeration (one-phase flow; the sucking tubes of the injectors were closed), liquid flow with aeration (gas–liquid two-phase flow), and the two flows with addition of ion exchanger particles (d_p = 0.7 mm, ρ_p = 2 460 kg m^{-3}) amounting 3% of the total volume of the reactor. Such particles may be added as adsorbents in the case of using the reactor for waste water treatment (Kmiec et al.[4]).

Series of measurements of local liquid flow velocity distributions were carried out for the various liquid flow modes through the injector. The local streams of liquid flowing into the opening of the Prandtl tube at a given flow rate are compared in Eqs (*1*) and (*2*) for the one-phase and two-phase flows, respectively.

$$V_l = A_l w_l \qquad (1)$$

$$V_{la} = A_{1a} w_{la} (1 - \varepsilon_g) \qquad (2)$$

Assuming that these values are equal and that $A_l = A_{la}$, we obtain the following formula

$$\varepsilon_g = 1 - (w_l / w_{la}) \; . \qquad (3)$$

The value of ε_g calculated from Eq. (*3*) is the local value of the gas hold-up at the point of measurement of liquid velocity. Only the axial liquid velocities measured at the distance of 105 mm from the injector outlet were used in the analysis. The accuracy of measurement of liquid linear velocities w_l and w_{la} was satisfactory for our purposes in this case.

Beside that, the air hold-up ε_g was estimated for the gas–liquid two-phase flow from hydrostatic pressure gradients along the reactor

$$-\Delta P / \Delta H = \boldsymbol{g} (\varepsilon_g \rho_g + \varepsilon_l \rho_l) \; , \qquad (4)$$

where

$$P = P_a + \boldsymbol{g} H_l \rho_l (1 - \varepsilon_g) \; . \qquad (5)$$

Here P_a is the pressure above the liquid inside the reactor, and H_l is the height of the liquid layer for a given pressure tap. The values of hydrostatic pressure P were measured along the wall of the draft tube by means of U-tubes connected with the pressure taps, the air pressure P_a was measured with a separate U-tube filled with water. Knowing the pressure gradients (dP/dH), the value of ε_g along the height can be found from Eq. (*4*).

The power input was calculated from the formula used earlier by Bhutada and Pangarkar[10]

$$N = \Delta P_i V_l \; , \qquad (6)$$

where ΔP_i is the pressure drop over the injector which was estimated from the pressure values measured at the inlet and the outlet of the injector (overpressure and suction pressure, respectively; see Fig. 2). The overpressure at the inlet was measured with one U-tube (filled with mercury or some other liquid and connected to tap 11), the suction pressure with another U-tube (filled with water) connected directly to the sucking tube of the injector for the time of measuring the suction pressure only.

The axial dispersion coefficients were measured by means of a conductometer whose probe was fixed at the lower end of the draft tube. A sample of 10 ml aqueous sodium chloride was used as a tracer. The tracer was injected just under one of the injectors into the liquid stream (jet). The electric signal obtained from the conductometer was recorded continuously with a recorder. In this way a plot of concentration vs time dependence was obtained where the concentration c is given in arbitrary units depending on the amplification rate of the signal. The time dependence of the outlet concentration at the end of the draft tube (with the liquid stream containing the NaCl tracer) was analyzed as a function of residence time by the method of Van der Laan[11] where the Peclet number is given by Eq. (*7*) in which the second moment M_2 is the variance of the concentration dispersion.

$$M_2 = (2/Pe^2)[\exp(-Pe) + Pe - 1] \quad (7)$$

The root of Eq. (*7*), i.e. the value of *Pe*, was found by the Newton method. The axial dispersion coefficient depends on the mean liquid velocity in the draft tube (w_e), the active length of the draft tube (L = 0.285 m), and the Peclet number (*Pe*) as follows

$$D_l = w_e L/Pe \, . \quad (8)$$

RESULTS AND DISCUSSION

Two series of measurements of local liquid flow velocity distribution were carried out in the cross-sections of the draft chamber, viz. at the distances of 105 and 230 mm from the end of the injector. The local water flow velocities measured inside the draft chambers at the distance of 105 mm from the injector outlet are shown in Fig. 3.

In general, the velocity distributions can be analyzed by applying Schlichting's[12] boundary layer theory for (circular) jet flows. In our case the lowest value of the nozzle *Re* number (= $w_{cd}d_{cd}\rho_l/\mu_l$) is 6 070, i.e. it exceeds the critical value of Re_{cr} = 1 000 (see Abramovich[13]) at which a jet flow becomes turbulent. The radial velocity distributions in turbulent jet flows are described by the following generalized curve:

$$w/w_{max} = 1/(1 + 0.25\eta^2)^2 \, , \quad (9)$$

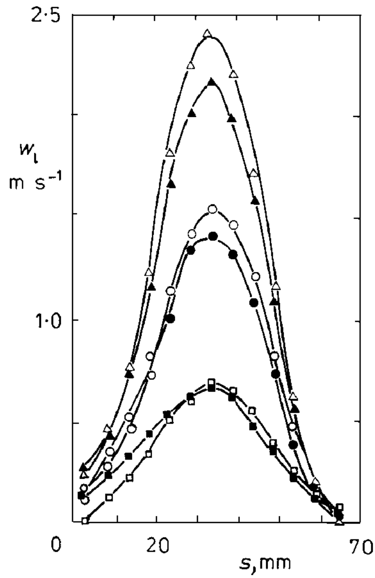

FIG. 3
Local liquid flow velocity distributions in the draft chamber at the distance of 105 mm from the injector outlet: ■●▲ without aeration, ❐❍Δ with aeration, ❐■ V = 0.167 dm^3 s^{-1}, ❍● V_l = 0.33 dm^3 s^{-1} Δ▲V_l = 0.5 dm^3 s^{-1}

where

$$\eta = 1.287\ r/r_{1/2}\ . \qquad (10)$$

The above function is valid if $x/d_{cd} > 6$, where d_{cd} is the diameter of the nozzle outlet, and $r_{1/2}$ is the (radial) distance from the axis at which $w = w_{max}/2$. In our case, an equivalent diameter of the nozzle outlet has to be used, i.e. $d_{cd} = d_{cd2} - d_{cd1}$.

The generalized velocity distributions measured for the flow rate $V_l = 0.167$ dm^3 s^{-1} at the two distances (x_1 = 105 mm, x_2 = 230 mm) are given in Fig. 4. As seen in Fig. 4, both the one-phase and the two-phase jet flows can be described by Eq. (*9*). It should be stressed that the theoretical description was developed for circular jets and the experimental data have been gathered for the square geometry.

The dependences of the liquid flow velocity in the axis of the draft chamber on the driving jet velocity are given in Fig. 5. It can be seen that the liquid velocity inside the draft chamber, w_{max}, is proportional to the driving jet velocity, w_{cd}, in the case of aerating action of injector as well as without it, whereas in the case of two-phase flow the linear velocity is slightly higher. Thus the ratio of the circulating liquid velocity in the

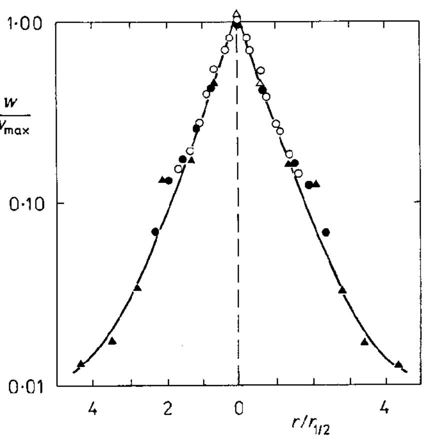

FIG. 4
Generalized velocity distributions for single (●▲) and two-phase flow (❍Δ) for V_l = 0.167 dm^3 s^{-1}: ❍● distance x_1 = 105 mm, Δ▲ distance x_2 = 230 mm from the injection outlet

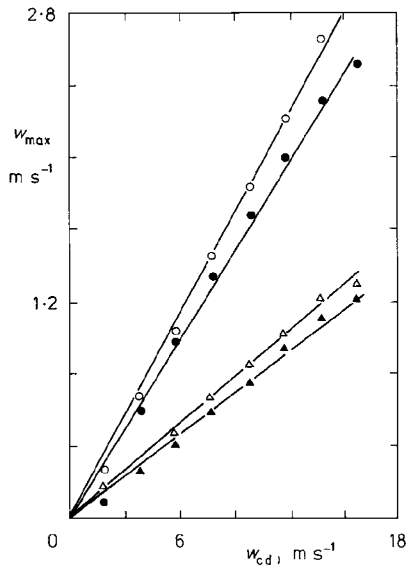

FIG. 5
Dependences of the water flow velocity in the axis of the draft chamber (maximum velocity) on the driving jet velocity: ❍● distance x_1 = 105 mm, Δ▲ distance x_2 = 230 mm from the injector outlet; ●▲ without aeration, ❍Δ with aeration

draft chamber, w_e, i.e. the mean velocity in the draft chamber, and the driving velocity, w_{cd}, should have constant values for a given type of flow.

Applying the principle of momentum conservation in the system of the driving jet stream and the flow stream inside the draft chamber (as suggested by Tebel and Zehner[9]) the following equation is proposed for our geometry:

$$(\Pi/4)\ (d_{cd2}^2 - d_{cd1}^2)\ w_{cd}\rho_l w_{cd} = \xi_I a_t^2\ (w_e^2/2)\ \rho_l\ , \tag{11}$$

wherefrom it can be obtained on rearrangement:

$$w_e/w_{cd} = [\Pi(d_{cd2}^2 - d_{cd1}^2)/\xi_I 2a_t^2]^{1/2}\ , \tag{12}$$

where ξ_I is the turbulent flow friction factor for the reactor.
When the ratio of the velocities is expressed in terms of volumetric flow rates

$$w_e/w_{cd} = (V_e/V_{cd})[(\Pi/4)(d_{cd2}^2 - d_{cd1}^2)/a_t^2] \tag{13}$$

the following relationship is obtained

$$V_e/V_{cd} = [8a_t^2/\Pi\xi_I(d_{cd2}^2 - d_{cd1}^2)]^{1/2}\ . \tag{14}$$

The volumetric flow rate V_e is calculated as a sum of the products of local liquid velocities and areas of cross-sectional quadratic increments related to the point of measurements.

An experimental verification of the above formula is presented in Fig. 6 where the volumetric flow ratio values are shown to be nearly constant – about 10 on average – for a three-phase gas–liquid system with 3 vol.% of solid ion exchanger particles. Com-

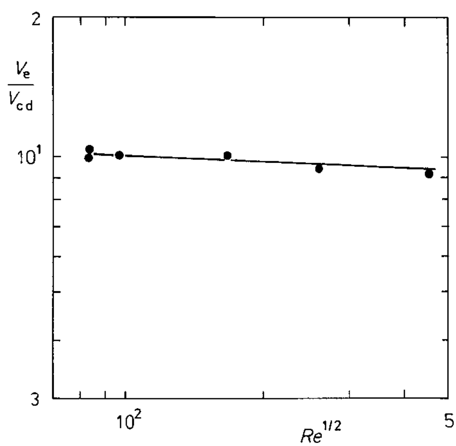

FIG. 6
Volumetric flow ratio dependence on $Re^{1/2}$

parable values were obtained by Tebel et al.[14] for water and the same type of reactor which differed in geometrical dimensions and shape of chambers. In the case of non-Newtonian liquid, Tebel et al.[14] reported much lower values of this ratio. Assuming the volumetric flow ratio equal to 10.0, the flow resistance factor obtained from Eq. (*14*) is 4.5.

Introducing the ratio of liquid velocities for one-phase and two-phase flows from Fig. 5 into Eq. (*3*), we obtain the air hold-up. Figure 7 compares our results calculated in this way and by the method of measurement of hydrostatic pressure gradients applying Eq. (*4*) with the data by Wachsmann et al.[3] and those by Bhutada and Pangarkar[10]. It can be seen that the air hold-up in the draft chamber of the reactor is about 10% at the power input of about 2 kW m^{-3}. This value is close to the result reported by Wachsmann et al.[3] for this type of reactor. The data by Wachsmann et al.[3] shown in Fig. 7 were read from their paper for the ratio of the draft tube and the reactor diameters corresponding to ours which equals 0.51. Nevertheless, our construction of the chambers differs from that of Wachsmann et al.[3] in the configuration, which was circular in cross-section.

The results of investigation of axial dispersion are shown in Figs 8 and 9 as functions of dispersion coefficient vs liquid volumetric flow rate through the injector and vs the mean air superficial velocity in the draft tube, respectively. As seen from Fig. 8 the dispersion coefficient increases strongly with increasing liquid volumetric flow rate in the injector. The results obtained for the jet-loop reactor in this work (line *1* in Fig. 9) are much lower as compared with the data for the other types of reactors. At low gas velocity (and low liquid velocity as well) they are considerably lower than those for the

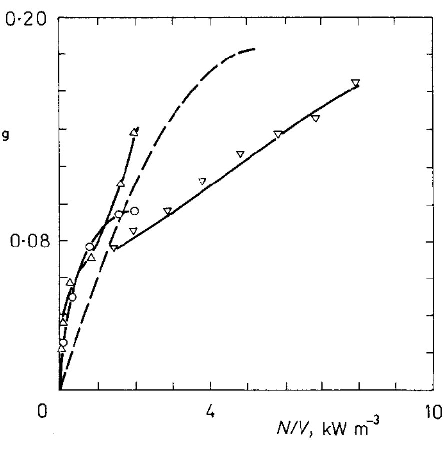

FIG. 7

Comparison of air hold-up: ❍ local measurements of this work (Eq. (*3*)), Δ pressure gradient measurements of this work (Eq. (*4*)), – – – results by Wachsmann et al.[3], ∇ data by Bhutada and Pangarkar[10]

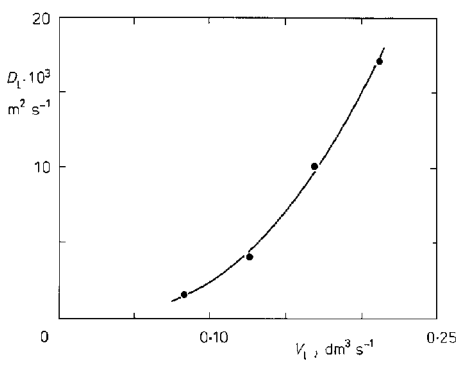

FIG. 8

Axial dispersion coefficient as a function of the liquid volumetric flow rate through the injector bubble column obtained from the correlation by Joshi[15] (line *2*), then they go up closer to the values reported by Verlaan et al.[16] for the airlift reactor. This dependence is related with the flow behaviour in the reactor: while at a low velocity the flow rate is rather moderate, at higher rates of flow it is very turbulent. At the same time, the air hold-up increases considerably too, which brings in additional turbulence into the liquid flow.

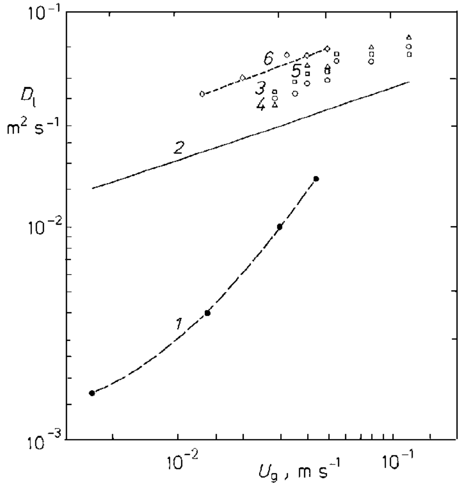

FIG. 9
Dependences of axial dispersion coefficient on superficial air velocity for different systems: *1*(●) the present paper, *2* correlation by Joshi[15] for bubble column, *3*(❒), *4*(❍), *5*(Δ) data by Varlaan et al.[16] for airlift reactors with increasing opening of valve and various flow rates, *6*(✧) data by El-Temtamy et al.[17] for three-phase system with particles d_{p_1} = 0.45 mm and liquid velocity w = 3.25 cm s^{-1}

CONCLUSIONS

1) Liquid flow velocities in the chamber of the jet-loop reactors depend on the geometry of the injector–reactor system. Their profiles in cross-sections of the draft tubes can be described according to the theory of Schlichting[12] for single phase, circular jet streams.

2) The air hold-up in the reactor increases with increasing power input but depends also on the geometry of the injector–reactor system.

3) The axial dispersion coefficient in the liquid phase is found to strongly depend on the superficial air velocity. The results are valid for reactors with chambers of square cross-section and should be confronted with results obtained for other arrangements.

LIST OF SYMBOLS

a_t	flank size of draft chamber (= 0.07 m), m
A	cross-sectional area of stream, m^2
d_{cd}	nozzle outlet diameter (in our case $d_{cd} = d_{cd2} - d_{cd1}$), m

d_{cd1}	inside annular diameter of the nozzle outlet, m
d_{cd2}	outside annular diameter of the nozzle outlet, m
D_1	axial dispersion coefficient, $m^2\ s^{-1}$
$\boldsymbol{g}$	gravity acceleration, 9.81 $m\ s^{-2}$
H	height, m
H_1	height of liquor layer, m
L	active length of draft tube (= 0.285 m), m
N	power, kW
P	pressure, Pa
r	radial distance from jet axis, m
s	perpendicular distance from the wall of draft tube, m
U	superficial velocity of air in draft tube, $m\ s^{-1}$
V	volumetric flow rate, $m^3\ s^{-1}$
w	liquid flow velocity, $m\ s^{-1}$
w_{cd}	liquid velocity in the nozzle outlet of the injector, $m\ s^{-1}$
w_e	average liquid velocity calculated from local liquid velocities inside the draft tube, $m\ s^{-1}$
x	axial distance from the outlet in jet stream, m
ε_g	air hold-up, $m^3\ m^{-3}$
ρ	density, $kg\ m^{-3}$
μ	viscosity, Pa s
η	the function given by Eq. (*10*)
ξ	the flow resistance factor

Subscripts

cd	nozzle
e	mean
g	gas
l	liquid
la	liquid aerated
p	particle

Dimensionless numbers

$Pe = w_e L/D_1$

$Re = w_{cd} d_{cd} \rho_l/\mu_1$

REFERENCES

1. Cramers P. H. M. R., van Dierendonck L. L., Beenackers A. A. C. M.: Chem. Eng. Sci. *47*, 2251 (1992).
2. Malone R. J.: Chem. Eng. Prog. *76*, 6, 53 (1980).
3. Wachsmann W., Raebiger N., Vogelpohl A.: Chem.-Ing.-Tech. *57*, 346 (1985).
4. Kmiec A., Kucharski J., Szczepaniak L., Szymanowski W.: Inzynieria i Aparatura Chemiczna *1*, 3 (1989).
5. Yenkie M. K. N., Geissen S. U., Vogelpohl A.: Chem. Eng. J. *49*, B1 (1992).
6. Zahradnik J., Kastanek F., Kratochvil J., Rylek M.: Collect. Czech. Chem. Commun. *47*, 1939 (1982).
7. Dutta N. N., Raghavan K. V.: Chem. Eng. J. *36*, 111 (1987).

8. Cramers P. H. M. R., Beenackers A. A. C. M., van Dierendonck L. L.: Chem. Eng. Sci. *47*, 3557 (1992).
9. Tebel K. H., Zehner P.: Chem. Eng. Technol. *12*, 274 (1989).
10. Bhutada S. R., Pangarkar V. G.: Chem. Eng. Commun. *61*, 239 (1987).
11. Van der Laan E. Th.: Chem. Eng. Sci. *5*, 187 (1958).
12. Schlichting H.: *Boundary Layer Theory*. McGraw-Hill, New York 1960.
13. Abramovic G. N.: *The Theory of Turbulent Jets*. M. I. T. Press, Boston 1963.
14. Tebel K. H., Zehner P., Langer G., Mueller W.: Chem.-Ing.-Tech. *58*, 820 (1986).
15. Joshi J. B.: Trans. Inst. Chem. Eng. *58*, 155 (1980).
16. Verlaan P., Vos J. Ch., van't Riet K.: J. Chem. Technol. Biotechnol. *45*, 181 (1989).
17. El-Temtamy S. A., El-Sharnoubi Y. O., El-Halwagi M. M.: Chem. Eng. J. *18*, 151 (1979).

Translation revised by J. Panchartek.