

## Wall-to-liquid mass transfer in fluidized beds and vertical transport of inert particles

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**Abstract:** Mass transfer coefficients in single phase flow, liquid fluidized beds and vertical hydraulic transport of spherical inert particles were studied experimentally using 40 mm and 25.4 mm diameter columns. The mass transfer data were obtained by studying the transfer of benzoic acid from a tube segment to water using the dissolution method. In all runs, the mass transfer rates were determined in the presence of spherical glass particles 1.2, 1.94 and 2.98 mm in diameter. The influence of different parameters, such as liquid velocity, particles size and voids on mass transfer in fluidized beds and hydraulic transport are presented. The data for mass transfer in all the investigated systems are shown using the Sherwood number ( $Sh$ ) and mass transfer factor – Colburn factor ( $j_D$ ) – as a function of Reynolds number ( $Re$ ) for the particles and for the column. The data for mass transfer in particulate fluidized beds and for vertical hydraulic transport of spherical particles were correlated by treating the flowing fluid–particle mixture as a pseudo fluid by introducing a modified mixture Reynolds number ( $Re_m$ ). A new correlation for the mass transfer factor in fluidized beds and in vertical hydraulic transport is proposed.

**Keywords:** mass transfer, fluidized bed, vertical transport, pseudofluid.

### INTRODUCTION

Research of mass transfer in liquid–solid systems is very important for equipment design for many applications. Due to industrial use, especially following the fast development of bioprocesses and water cleaning processes, a better knowledge of these systems becomes increasingly important. An industrial application of liquid–solid systems requires the determination of transfer characteristics, especially mass transfer. Mass transfer in fluidized beds has been widely investigated in terms of particle–fluid mass transfer by dissolution, by electrochemical and by ion-exchange methods.<sup>1–7</sup> Some of the results of mass transfer in fluidi-

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zed beds have been obtained as the transfer between an immersed surface and the liquid.<sup>8,9</sup> Wall-to-bed mass transfer has been intensively studied for several years because of its theoretical and practical importance.<sup>10,11</sup> In contrast to fluidized beds data, there are no published data on mass transfer in vertical and horizontal hydrotransport of particles.

This paper presents the experimental data for wall-to-liquid mass transfer in coarse particle-liquid systems: fluidized beds and vertical transport. A correlation for the mass transfer coefficient was developed on the bases of the Colburn  $j_D$  factor.

#### EXPERIMENTAL

The rates of mass transfer between a fluid and the wall in fluidized beds and vertical transport of spherical inert particles were studied. A schematic diagram of the experimental systems is shown in Fig. 1.

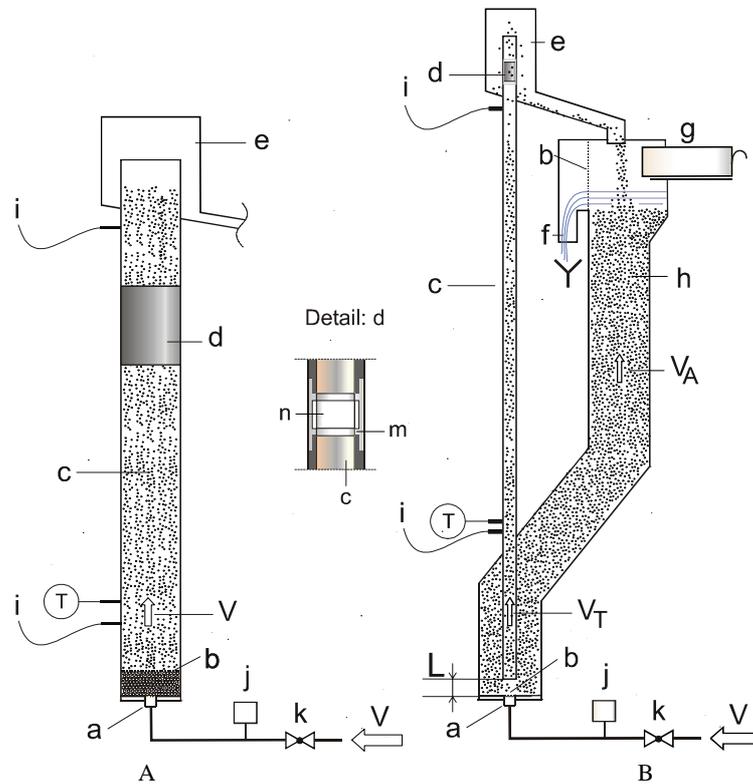


Fig. 1. Schematic diagram of the experimental system: A) fluidization setup; B) vertical transport setup (a - inlet nozzle, 20 mm ID; b - screen; c - fluidizing column 40 mm ID, transport tube 25.4 mm ID; d - segment prepared with benzoic acid; e - overflow; f - water overflow; g - box for water and particle flowrate measurements; h - modified spouted bed 70×70 mm<sup>2</sup> in cross-section; i - pressure taps; j - rotameter; k - valve; L - distance of 20 mm; T - thermocouple; V - inlet flowrate; V<sub>T</sub> - tube flow; V<sub>A</sub> - annular flow. Detail d: m - segment, n - melt of benzoic acid).

The experiments were conducted using 40 mm (fluidized beds, Fig. 1A) and 25 mm (fluidized beds and vertical transport, Fig. 1B) diameter columns. The fluidizing and transporting fluid was water. The experiments were conducted using inert glass spheres 1.20, 1.94 and 2.98 mm in diameter. The relevant properties of the fluid and particles are summarized in Table I.

TABLE I. Particle and fluid characteristics

Particles (glass spheres)				
$d_p$ / mm	$\rho_p$ / kg m <sup>-3</sup>	$\epsilon_{mf}$	$U_{mf}$ / m s <sup>-1</sup>	$U_t$ / m s <sup>-1(14)</sup>
1.2	2641	0.39	0.013	0.19
1.94	2507	0.42	0.026	0.29
2.98	2509	0.46	0.044	0.37
Fluid (water at 25 °C)				
$\rho_f$ / kg m <sup>-3</sup>	$\mu$ / Pa s	$D$ / m <sup>2</sup> s <sup>-1</sup>	$c^*$ / kg m <sup>-3</sup>	$Sc$
977	$0.893 \times 10^{-3}$	$9.24 \times 10^{-10}$	3.3554	969

In the vertical transport experiments, the fluid and particle flowrates were measured using a specially designed box (g), which allows all of the flow (fluid and particles) to be collected, separated and weighed (Fig. 1B). Normally, the particles recirculate and the suspension overflows at (e), while water overflows at (f). When the fluid and particle flowrates were to be measured, the box (g) was moved to the left to collect the entire flow for a short period of time (10 s to 1 min). The water was then separated from the particles. The particles were dried and weighed and the volume of water was recorded.

The range of experimental conditions is summarized in Table II.

TABLE II. Range of experimental conditions

$d_p$ / mm	1.2	1.94	2.98
$U/U_{mf}$ (Fluidized beds)	1.66–6.67	1.56–4.54	1.45–3.2
$\epsilon$ (Fluidized beds)	0.49–0.83	0.48–0.72	0.53–0.71
$U/U_t$ (Transport)	0.475–4.904	0.405–4.724	0.267–3.884
$W_p$ / kg m <sup>-2</sup> s <sup>-1</sup>	9.7–218.8	15.2–318.6	2.5–372.4
$G_p/G_f$	0.108–0.253	0.113–0.474	0.026–0.331
$\epsilon$ (Transport)	0.827–0.923	0.729–0.897	0.712–0.864

The dissolution of benzoic acid from a tube segment was followed. The tube segment was coated with benzoic acid by immersing an empty ring on a wire support into molten benzoic acid. By repeating the procedure, a compact layer of benzoic acid was formed on the ring segment of the column. The mass transfer coefficient was calculated from the equation:

$$k_c = \frac{\Delta m}{A t \Delta c} = \frac{\Delta m}{(D_c \pi h) t \Delta c} \quad (1)$$

The transferred mass,  $\Delta m$ , was determined by measuring the weight loss of benzoic acid. The mass transfer area  $A$  was calculated from the mean value of the diameter of the tube segment before and after dissolution. Since the weight loss of benzoic acid  $\Delta m$  is small, the bulk concentration was negligible; hence the equilibrium concentration  $c^*$  was taken as a driving force  $\Delta c$ .

The benzoic acid solubility and diffusivity were taken from the literature.<sup>12,13</sup> In each run, the average fluid temperature was recorded and the corresponding values of the diffusion coefficients, fluid viscosity, fluid density and equilibrium solubility were considered in the calculations.

## RESULTS AND DISCUSSION

The relationships between the particle Sherwood number and the particle Reynolds number for different particle diameters in fluidized beds and vertical transport are presented in Fig. 2. From this Figure, the influence of liquid velocity on mass transfer can be seen. The data are separated into three groups with the particle diameter as the critical parameter. The experimental results show that with increasing liquid velocity (or Reynolds number), the mass transfer coefficient (or Sherwood number) slightly decreases in fluidized beds.

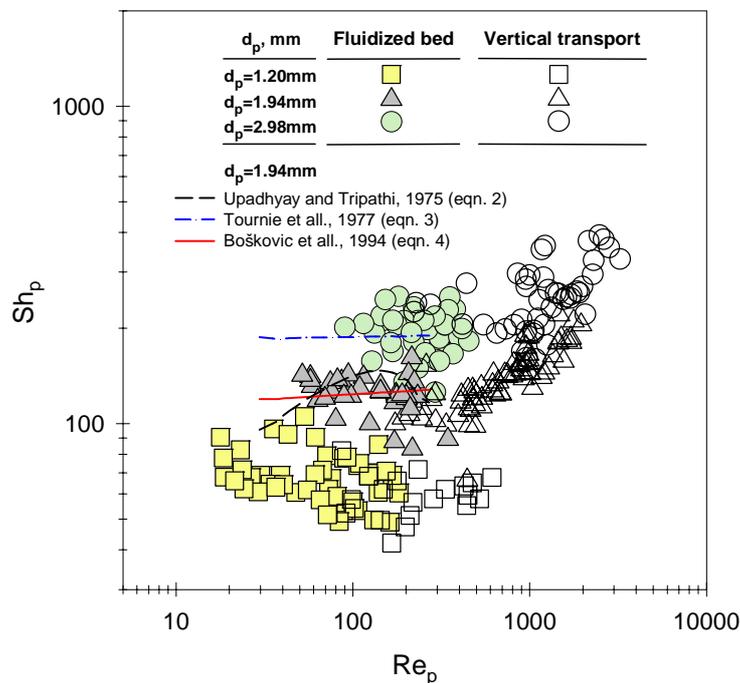


Fig. 2. The relationship between the Sherwood number and Reynolds number for the particles.

The influence of liquid velocity in fluidized beds is connected with the influence of the bed concentration (bed voidage). There are two modes of influence: increasing the liquid velocity increases the mass transfer, but increasing the liquid velocity decreases the particle concentration. Increasing velocity and concentration have a positive influence on mass transfer. Due to both effects, the diffusion boundary layer becomes thinner and mass transfer increases. Finally, as can be seen in Fig. 2, the influence of particle concentration is slightly greater than the influence of liquid velocity.

With increasing liquid velocity, the Sherwood number or mass transfer coefficient is constant for low transport velocities, but for higher transport velocities, the mass transfer increases. This implies that because of the low particle concen-

tration in a transport column, the influence of particles on the diffusivity boundary layer is smaller.

Comparison of the data for fluidized bed ( $d_p = 1.94$  mm) with several literature correlations<sup>9,15,16</sup> show significant difference between our data and the available correlations. As shown, there are significant differences among the systems investigated. Upadhyay and Tripathi<sup>15</sup> proposed the following correlation for mass transfer in packed and fluidized beds (Fig. 2):

$$Sh_p = 1.6218Re''^{0.5553} Sc^{1/3} \text{ for } Re'' > 20 \quad (2)$$

where  $Re'' = Re_p/(1-\varepsilon)$ , in the following range of variables:  $0.01 < Re'' < 12000$ ,  $572 < Sc < 70000$  and  $0.268 < \varepsilon < 0.9653$ . The data calculated using Eq. (2) show a maximum which was not confirmed by our experimental data.

Also, Tournie *et al.*<sup>16</sup> gave the following correlation for mass transfer particle–fluid in a fluidized bed:

$$Sh_p = 0.253Re_p^{0.004} Ga^{0.319} Mv^{0.299} Sc^{0.4} \quad (3)$$

where  $Mv = (\rho_p - \rho_f)/\rho_f$ . Eq. (3) is recommended in the following range of variables:  $1.6 < Re_p < 1320$ ,  $2470 < Ga < 442106$ ,  $0.27 < Mv < 1.14$ ,  $305 < Sc < 1595$  (also shown in Fig. 2). The data show significant difference between this correlation and our experimental values.

Mass transfers in a liquid fluidized bed, were correlated by Bošković *et al.*, by the Equation:<sup>9</sup>

$$Sh_p = 0.261Re_p^{0.03} Ga^{0.324} Sc^{1/3} \quad (4)$$

in the following range of variables:  $Re_p = 15-400$ ,  $Sc = 1361-1932$ . The predicted values using correlation (4) for mass transfer between fluid and an immersed sphere in fluidized beds of spherical inert particles are in good agreement with the experimental data.

The influence of particle diameter on mass transfer can be seen in Fig. 3, which presents data for the Sherwood number as a function of the Reynolds number for the column in fluidized beds and vertical transport. The data for mass transfer for single phase flow are shown in the same plot for comparison. As can be seen, there is no influence of particle diameter on mass transfer in systems with particles. The presence of particles improves mass transfer for lower fluid velocities, *i.e.*, Reynolds number, when the particle concentration in the flowing mixture is generally higher. A decrease in the liquid velocity ( $Re < 10000$ ), increases the mass transfer for fluidized bed relative to the values in single phase flow by a factor of about 10. For higher Reynolds number (approximately  $Re > 10000$ ), there is no significant difference in mass transfer between single phase flow and hydraulic transport. For a high liquid velocity, the concentration of particles is small and their influence on the mass transfer coefficient is negligible.

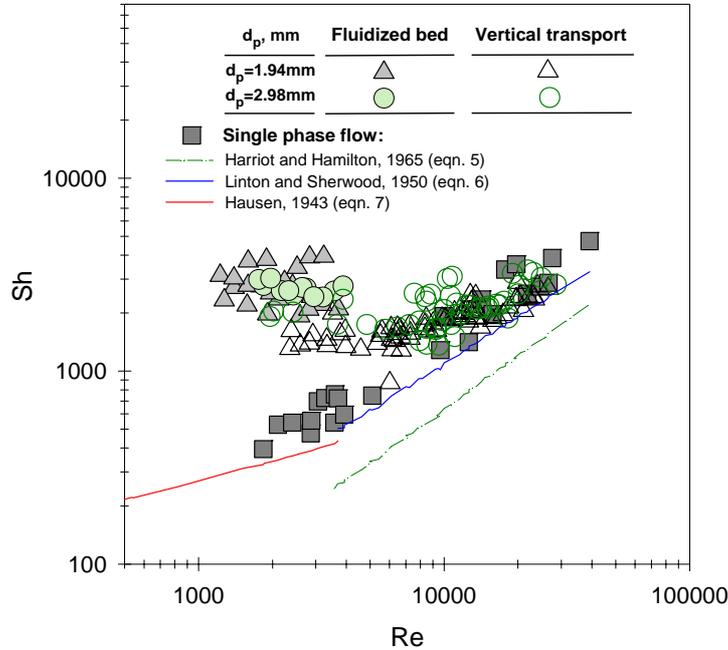


Fig. 3. The relationship between the Sherwood number and Reynolds number for the column.

A comparison of several literature correlations<sup>17-19</sup> of mass transfer for single phase flow are shown in Fig. 3.

The Harriot and Hamilton<sup>17</sup> Equation for a wide range of values of Reynolds and Schmidt numbers ( $2000 < Re < 70000$  and  $1000 < Sc < 2260$ ),

$$Sh = 0.0096Re^{0.913}Sc^{0.346} \quad (5)$$

indicates the adverse ratio  $L_{t,A}/D_t$ , give a value of 0.73 in the present investigation. The predicted values from correlation (5) and the present experimental data show a significant difference.

For specific systems (concentration boundary layer is not formed, small ratio  $L_{t,A}/D_t$ ), the Equation is:<sup>18</sup>

$$Sh = 0.023Re^{0.8}Sc^{1/3} \left( 1 + \left( \frac{D_t}{L_{t,A}} \right)^{0.7} \right) \quad (6)$$

The Linton and Sherwood, Eq. (6), for turbulent flow ( $Re > 10000$ ), gives better results than Eq. (5).

An Equation for the average heat transfer coefficient for the entrance region and the latter part of the tube in laminar flow was proposed by Hausen,<sup>19</sup> which when applied to mass transfer is:

$$Sh_{lam} = 3.66 + \frac{0.0668 Re Sc \left( \frac{D_t}{L_{t,A}} \right)}{1 + 0.4 \left( Re Sc \left( \frac{D_t}{L_{t,A}} \right)^{1/2} \right)^{2/3}} \left( \frac{\mu}{\mu_{zid}} \right)^{0.14} \quad (7)$$

The data calculated using Eq. (7) are in good agreement with the experimental data.

The experimental data of wall-to-liquid mass transfer in fluidized beds and hydraulic transport are presented using the Colburn mass transfer factor  $j_D$ . The mass transfer factor as a function of Reynolds number for particles is shown in Fig. 4 and as a function of Reynolds number for the column in Fig. 5. From the definition of the mass transfer factor:

$$j_D = \frac{Sh}{Re Sc^{1/3}} = \frac{k_c}{U} Sc^{2/3} \quad (8)$$

and the previous conclusions from Figs. 4 and 5, it is clearly seen that with the increasing fluid velocity, the mass transfer factor decreases. The same trend is observed for  $Re > 10000$ .

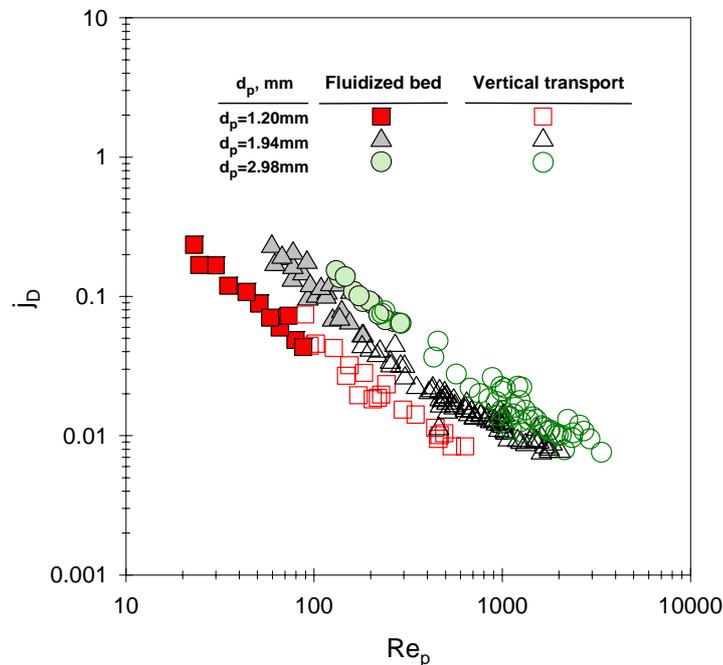


Fig. 4. Variation of the mass transfer factor with Reynolds number for the particles.

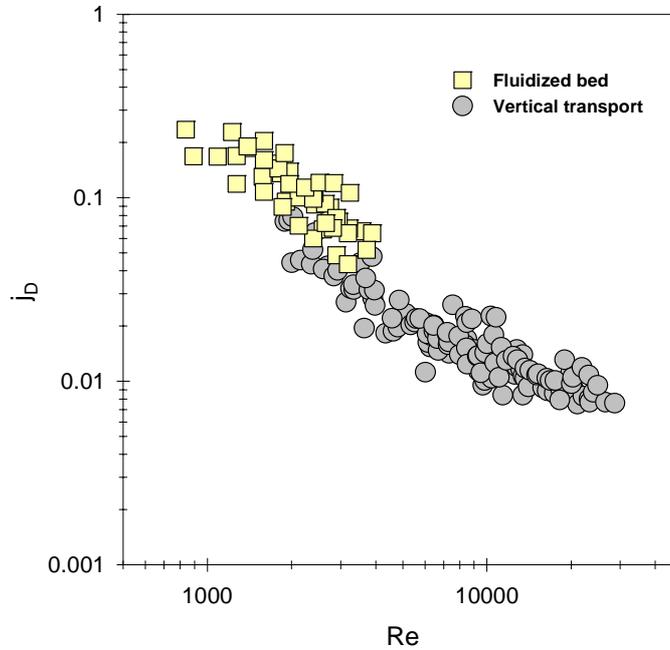


Fig. 5. Variation of the mass transfer factor with Reynolds number for the column.

As can be seen in Fig. 5, there is a significant difference between the experimental data for mass transfer in fluidized beds and vertical transport. As a result of this, the whole system was treated as a pseudo fluid or suspension with a mean density:

$$\rho_m = \varepsilon \rho_f + (1 - \varepsilon) \rho_p \quad (9)$$

An effective mixture viscosity:<sup>20</sup>

$$\mu_m = \mu \exp\left(\frac{5(1 - \varepsilon)}{3\varepsilon}\right) \quad (10)$$

and a mean mixture superficial velocity:

$$U_m = \frac{G_f}{\rho_f A_t} + \frac{G_p}{\rho_p A_t} \quad (11)$$

For the suspension mixture, the Reynolds number is:

$$Re_m = \frac{U_m D_c \rho_m}{\mu_m} \quad (12)$$

Also, the fluidized bed and the vertical transport data overlap for  $2000 \leq Re_p \leq 4000$  because the concentrations of the particles in these systems are nearly the same.

The results of the mass transfer factor as a function of the Reynolds number of the mixture are presented in Fig. 6. There is a unique correlation for the mass

transfer data for fluidized beds and vertical transport. These data are correlated by the Equation:

$$j_D = 45 Re_m^{-0.88} \quad (13)$$

with an absolute deviation of 15.7 % and a relative deviation of -2.0 %.

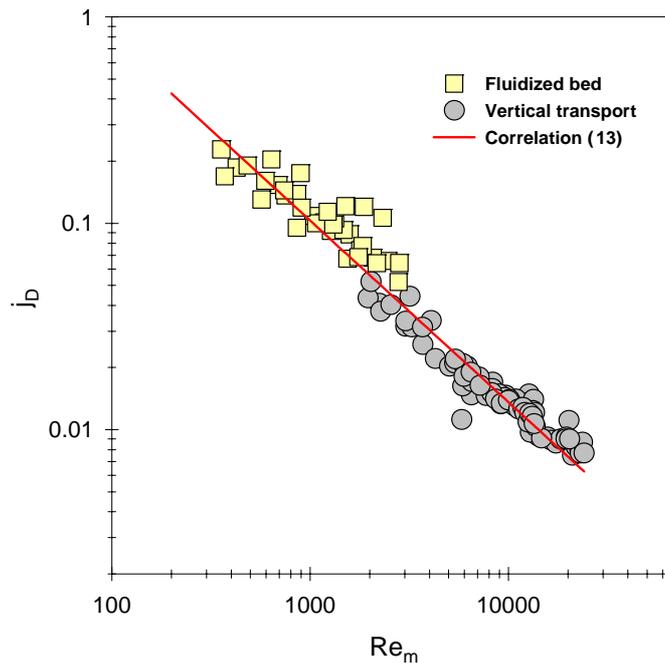


Fig. 6. Correlation between the mass transfer factor and the Reynolds number of the mixture.

#### CONCLUSIONS

Experiments were performed to determine the mass transfer coefficient between the fluid and the wall in liquid fluidized beds and vertical transport of inert particles. These experiments showed the following:

- With increasing liquid velocity, the mass transfer coefficient slightly decreases in fluidized beds. In vertical transport, the mass transfer coefficient is constant for low velocities and increases for higher velocities.
- The wall-to-liquid mass transfer in fluidized beds and in hydraulic transport is higher than in single phase flow for Reynolds number of the column lower than 10000. For higher Reynolds numbers, there is no significant difference between single phase flow and hydraulic transport.
- The mass transfer factor as a function of Reynolds number is independent of the particle size in fluidized beds and vertical transport.
- A new correlation for the mass transfer factor in particulate fluidized beds and vertical hydraulic transport is proposed.

## NOMENCLATURE

$A$	Mass transfer area, $m^2$
$c^*$	Equilibrium concentration, $kg\ m^{-3}$
$\Delta c$	Concentration difference (driving force), $kg\ m^{-3}$
$D$	Molecular diffusion coefficient, $m^2\ s^{-1}$
$D_c$	Column diameter, $m$
$d_p$	Particle diameter, $m$
$\dot{G}_f$	Fluid mass flowrate, $kg\ s^{-1}$
$\dot{G}_p$	Particle mass flowrate, $kg\ s^{-1}$
$h$	High of tube segment, $m$
$j_D$	Mass transfer factor, $(k_c/U)Sc^{2/3}$
$k_c$	Mass transfer coefficient, $m\ s^{-1}$
$\Delta m$	Total mass transferred, $kg$
$Re$	Reynolds number for column, $UD_c\rho_f/\mu$
$Re_m$	Mixture Reynolds number, $U_m D_c \rho_m / \mu_m$
$Re_p$	Reynolds number for particles, $U d_p \rho_f / \mu$
$Sc$	Schmidt number, $\mu / \rho_f D$
$Sh$	Sherwood number for column, $k_c D_c / D$
$Sh_p$	Sherwood number for particles, $k_c d_p / D$
$t$	Exposure time, $s$
$U$	Superficial fluid velocity, $m\ s^{-1}$
$U_t$	Terminal velocity, $m\ s^{-1}$
$U_m$	Superficial fluid mixture velocity, $m\ s^{-1}$
$U_{mf}$	Minimum fluidization velocity, $m\ s^{-1}$
$W_p$	Particle mass flux, $kg\ m^{-2}\ s^{-1}$

## Greek letters

$\varepsilon$	Voidage
$\varepsilon_{mf}$	Voidage at minimum fluidization
$\mu$	Viscosity of the fluid, $Pa\ s$
$\mu_m$	Viscosity of the fluid-particle mixture, $Pa\ s$
$\rho_f$	Fluid density, $kg\ m^{-3}$
$\rho_p$	Particle density, $kg\ m^{-3}$
$\rho_m$	Mean mixture density, $kg\ m^{-3}$

## ИЗВОД

ПРЕНОС МАСЕ ЗИД-ФЛУИД У ФЛУИДИЗОВАНИМ СЛОЈЕВИМА И ПРИ  
ВЕРТИКАЛНОМ ТРАНСПОРТУ ИНЕРТНИХ ЧЕСТИЦАНЕВЕНКА БОШКОВИЋ-ВРАГОЛОВИЋ<sup>1</sup>, РАДМИЛА ГАРИЋ-ГРУЛОВИЋ<sup>2</sup> И ЖЕЉКО ГРБАВЧИЋ<sup>1</sup><sup>1</sup>Технолошко-металуршки факултет, Катедра за хемијско инжењерство, Универзитет у Београду,  
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У овом раду је експериментално испитиван прелаз масе између зида и флуида при струјању флуида кроз колону и у присуству флуидизованих сферних инертних честица и честица у транспорту. За одређивање коефицијента прелаза масе коришћена је метода праћења растварања слабо растворне супстанце – бензоеве киселине – са дела зида колоне у води. Коришћене су колоне пречника 40 mm и 25,4 mm, а као инертне честице коришћене су стак-

лене сфере пречника 1,2; 1,94 и 2,98 mm. Приказани су утицаји различитих параметара као што су брзина флуида, величина честица и порозност слоја на коефицијент прелаза масе у флуидизованим слојевима и при вертикалном транспорту честица. Експериментални резултати су приказани у облику зависности Шервудовог броја и фактора преноса масе од Рејнолдсових бројева за колону и честице. Уведена је и величина псеудофлуид која третира двофазни систем течност–честице као суспензију одакле је проистекао модификовани Рејнолдсов број мешања. Увођењем ових величина дата је и завршна експериментална корелација, јединствена за флуидизоване слојеве и вертикални транспорт честица.

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#### REFERENCES

1. S. Damronglerd, J. P. Couderc, H. Angelino, *Trans. Inst. Chem. Engrs.* **53** (1975) 175
2. P. N. Dwivedi, S. N. Upadhyay, *Ind. Eng. Chem. Proc. Des. Dev.* **16** (1977) 157
3. S. Kumar, S. N. Upadhyay, *Ind. Eng. Chem. Fundam.* **20** (1981) 186
4. T. Chhun, J. P. Couderc, *Chem. Eng. Sci.* **35** (1980) 1707
5. T. Koloini, M. Sopčić, M. Žumer, *Chem. Eng. Sci.* **32** (1976) 637
6. K. Rahman, M. Streat, *Chem. Eng. Sci.* **36** (1981) 293
7. N. Yutani, N. Ototake, L. T. Fan, *Ind. Eng. Chem. Res.* **26** (1987) 343
8. J. P. Riba, R. Routie, J. P. Couderc, *Fluidization*, J. F. Davidson, D. L. Keairns, Eds. Cambridge, University Press, Cambridge, England, 1979, p. 157
9. N. Bošković, Ž. B. Grbavčić, D. V. Vuković, M. Marković–Grbavčić, *Powder Techn.* **79** (1994) 217
10. J. Lee, H. Chun, L. Shemilt, *J. Chem. Eng. Jap.* **30** (1997) 246
11. S. Schmidt, J. Buchs, C. Born, M. Biselli, *Chem. Eng. Sci.* **54** (1999) 829
12. S. Kumar, S. N. Upadhyay, *Ind. Eng. Chem. Fundam.* **19** (1980) 75
13. S. Kumar, S. N. Upadhyay, K. V. Mathur, *J. Chem. Eng. Data* **23** (1978) 139
14. D. Kunii, O. Levenspiel, *Fluidization Engineering*, Wiley, New York, 1969
15. S. N. Upadhyay, G. Tripathi, *J. Chem. Eng. Data* **20** (1975) 20
16. P. Tournie, C. Laguerie, J. P. Couderc, *Chem. Eng. Sci.* **32** (1977) 1259
17. P. Harriot, M. R. Hamiltin, *Chem. Eng. Sci.* **20** (1965) 1073
18. W. H. Linton, T. E. Sherwood, *Chem. Eng. Progr.* **46** (1950) 258
19. H. Hausen, *Allg. Wärmetechnik* **9** (1959) 75
20. E. Barnea, J. Mizrahi, *Chem. Eng. J.* **5** (1973) 171.