11

Hydrodynamics and Mass Transfer in Heterogeneous Systems

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1. Introduction

Research of transport phenomena in liquid – particles systems, in past years, had a more theoretical then practical importance (Coudrec, 1985; Lee et al., 1997; Schmidt et al., 1999). For industrial use, especially with fast development of bio and water cleaning processes, better knowing of these systems become more important. An industrial application of these systems requires determination of transfer characteristics, especially mass transfer. Frequently mass transfer is the rate determining step in the whole process. However, in the real systems, it is not always easy to differentiate the limitation due to the mass transfers from that due to the hydrodynamic.

Mass transfer in liquid-solid packed and fluidized beds has been widely investigated in terms of particle–fluid mass transfer by dissolution, by electrochemical and by ion-exchange methods (Damronglerd et al. 1975; Koloini et al. 1976; Dwivedi & Upadhyay, 1977; Chun & Couderc, 1980; Kumar & Upadhyay, 1981; Rahman & Streat, 1981; Yutani et al. 1987). Some of the results of mass transfer in packed and fluidized beds have been obtained as the transfer between an immersed surface and the liquid (Riba et al. 1979; Bošković et al. 1994; Bošković-Vragolović, et al., 1996&2005). Liquid fluidization of particulate solids has a history which predates the now more commonly applied gas fluidization. The broad range of operations to which liquid fluidization has found applications are: classification of particles by size and density, a special case being sink-and-float separation by density; backwashing of granular filters and washing of soils; crystal growth; leaching and washing; adsorption and ion exchange; electrolysis with both inert and electrically conducting fluidized particles; liquid-fluidized bed heat exchangers and thermal energy storage; and bioreactors. Fluidized-bed bioreactors, which have received much attention during the past thirty years, are usually characterized by the catalytic use of enzymes or microbial cells that are immobilized by attachment, entrapment, encapsulation or self-aggregation. The most common application of such bioreactors is probably in wastewater treatment and, as in the case of the other operations mentioned above, liquid fluidization must in each case be weighed against competing schemes for achieving the same objective before it is adopted commercially (Epstein, 2003). In contrast to fluidized beds data, there are no published data on mass transfer in vertical and horizontal hydrotransport of particles.

Liquid–solid packed and fluidized beds have usually been investigated separately (Kumar & Upadhyay, 1981; Yutani et al. 1987; Comiti & Renaud, 1991; Schmidt et al., 1999), but many authors have noticed the similarity between the two systems (Dwivedi & Upadhyay, 1977).

2. Background of heterogeneous systems

There are a number of different types of systems designed for fluid-solid heterogeneous operations. Two-phase flow systems of fluid (gas or liquid) and particles, the way of realizing contact between the phases, namely how the introduction of fluid in the bed of particles can be (Fig. 1):
systems with a fixed bed of particles;
- packed bed (a),

systems with a moving bed of particles;
- Fluidized bed (b),
- spouted bed and spout-fluid beds (c),
- spouted bed and spout-fluid beds with draft tube (d),
- vertical two-phase flow systems (fluid-particle), i.e. transport systems (e).

Packed beds are an essential part of chemical engineering equipment. A packed bed is a column filled with a support material, in which one or more fluids flow (Fig. 1, a). In chemical processing, a packed bed is a hollow tube, pipe, or other vessel that is filled with a packing material. The packing can be randomly filled with small objects like Raschig rings or else it can be a specifically designed structured packing. Differently shaped packing materials have different surface areas and void space between the packing. Both of these factors affect packing performance. The purpose of a packed bed is typically to improve contact between two phases in a chemical or similar process. Packed beds are widely used in industry to contact two or more fluid phases at relatively low pressure drops. For process design purposes, it is essential that pressure drop be estimated for its proper operation. They are readily used in industry for catalytic reactions, combustion, gas absorption, distillation, drying, and separation processes.
A fluidized bed is a packed bed through which fluid flows at such a high velocity that the bed is loosened and the particle-fluid mixture behaves as though it is a fluid (Fig. 1, b). Thus, when a bed of particles is fluidized, the entire bed can be transported like a fluid, if desired. Both gas and liquid flows can be used to fluidize a bed of particles. In fluidized beds, the contact of the solid particles with the fluidization medium (a gas or a liquid) is greatly enhanced when compared to packed beds. The most common reason for fluidizing a bed is to obtain vigorous agitation of the solids in contact with the fluid, leading to excellent contact of the solid and the fluid and the solid and the wall. Fluidized beds are widely used in industry for mixing solid particles with gases or liquids. In most industrial applications, a fluidized bed consists of a vertically-oriented column filled with granular material, and a fluid (gas or liquid) is flow upward through a distributor at the bottom of the bed.
Fluidized bed technology has become a common method to increase heat and mass transfer in physical and chemical operations. Although a great amount of studies have been performed to characterize the physical properties of these systems, only a few satisfying models have been derived. This is especially true for liquid-solid fluidized beds, which usually play a minor role in practical applications. Furthermore, studies dealing with gas-solid or gas-liquid-solid fluidized beds may not be applied to liquid-solid fluidized beds, because of the very different physical behaviour of these systems (Schmidt et al., 1999).

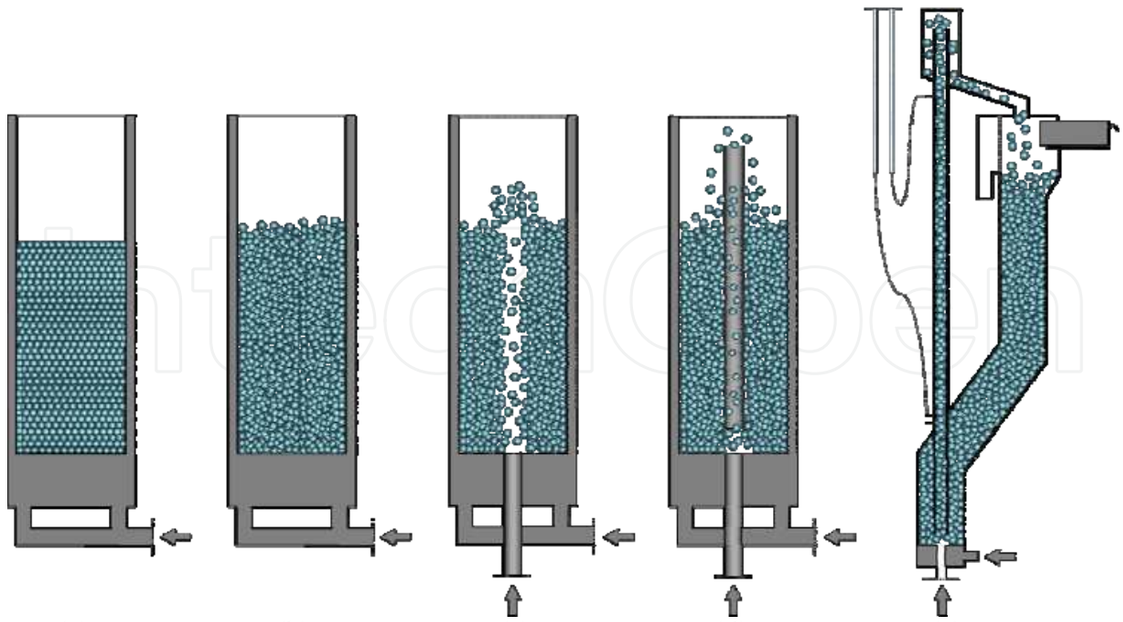

Fig. 1. Schematic diagram of heterogeneous systems:

a. Packed bed,
b. Fluidized bed,
c. Spout/Spout-fluid bed,
d. Spout/Spout-fluid bed with draft tube,
e. Vertical two-phase flow system (transport system).

A spouted bed can be formed in a vertical column where the fluid jet blows vertically upwards along the centre line of the centre line of the column thus forming a spout in which there is a fast moving fluid and entrained particle (Fig. 1, c). The remainder of the bed, the annulus between the spout and the column wall, is densely packed with particles moving slowly downwards and radially inwards. Fluid percolates through these particles from the spout.

The spouted bed technique, originally developed by Mathur & Gishler, 1955, has been successfully applied to a variety of diffusional, thermal, mechanical and other processes. Various modifications of the classical spouted bed have been proposed to improve its operability and provide better agreement between bed characteristics and process requirements. By introducing external annular fluid flow through the annulus bottom, spout-fluid bed can be formed (Nagarkatti & Chatterjee, 1974; Vuković et al., 1984; Sutanto et al., 1985). The draft tube in these systems is essentially a pneumatic or hydraulic riser which is relatively short even in the industrial scale units.

The addition, a draft tube in a spout and spout-fluid bed changes some of basic their characteristics (Fig. 1, d). The most important of these are:

- there is no bed height limitation imposed by a maximum spout and spout-fluid bed height.
- the minimum spout and spout-fluid (or circulating) velocity is lower in a bed with a draft tube.
- the draft tube forces all of the particles to travel through the entire annulus section surrounding the draft tube before re-entering the spout in the entry section thereby narrowing the particle residence time distribution in the annulus.

The spout and spout-fluid bed with draft tube is a very flexible fluid-particle contacting system because the fluid flowrates through the draft tube and annulus, and the solids circulation rate can be very easily controlled and adjusted in accordance with process needs.
As a kind of high-performance reactor for fluid-solid particle reactions, the spouted bed technology is applied to a variety of chemical processes such as dryness, prilling, coating, gasification, combustion, and pyrolysis, etc. During the past decades, numerous modified spouted bed designs have been developed to overcome some of the limitations of the conventional spouted beds, to accommodate the diverse properties of the materials handled and enhance the operability, heat and mass transfer characteristics, and gas-solid or fluid-solid contacting efficiency, for example, multiple spouted beds, spouted fluidized beds, conical spouted beds, draft tube spouted beds, pulsed spouted beds, and rotating spouted beds et al. In addition, many experimental and theoretical studies, which aimed at grasping the more useful flow characteristics of the spouted beds, have also been performed (Hu et al., 2008).
The vertical two-phase fluid-solids flow (Fig. 1, e), is a modified of draft tube spout and spout-fluid bed. In these systems the fluid jet penetrates upward along the central line, forming a spout and entraining the particles up the spout. The tube was mounted in a modified top closed spouted bed in order to obtain non-fluctuating controlled flow of particles. The annulus between the spout and the hopper wall is a densely packed bed with particles moving slowly downwards and radially inward toward the spout. These systems have complex hydrodynamics and the magnitude of the solids flowrate through the transport tube is a major consideration in all applications (Garić, et al., 1995&1996; Garić-Grulović, et al., 2005). All these systems, except packed and classical fluidized bed (Fig. 1), contain elements of a vertical two-phase fluid-particle flow. Design and practical realization of any of these systems that contain elements of a vertical two-phase flow, requires knowledge of the laws that connect the fluid velocity, particle velocity, voidage and pressure gradient as heat and mass transfer coefficients. The vertical two-phase flow systems are important because of their desirable characteristics for use in chemical and biochemical reactors. Unlike many studies of heat and mass transfer in the vertical flow of gas-particle, that include a wide area of application (Matsumoto et al., 1978; Duduković et al., 1996; Mansoori et al., 2002), in the literature are much less data for momentum, heat and mass transfer in systems with the vertical flow of liquid-coarse particles.

3. Investigation in heterogeneous systems

Wall-to-fluid mass transfer in packed beds, fluidized beds and hydraulic transport of spherical glass particles and in single-phase flow regime has been studied experimentally using adsorption and dissolution method. Experiments were performed using spherical glass particles from 1.2 to 3.06 mm in diameter with water in a 25.4 and 40mm I.D. column (Bošković-Vragolović, et al., 2007 & 2009; Garić-Grulović, et al., 2009).
The adsorption method was introduced as a relatively simple mass transfer measurement technique, applicable for liquid flow investigations (Mitrović, et al., 1989). The adsorption method is based on the dynamic adsorption of an organic dye onto a surface covered with a thin layer of a porous adsorbent. In our experiments, very diluted solution of methylene blue ($c_0 = 2.5 \cdot 10^{-3}$ g/dm^3) was used as a fluid in the presence of inert glass particles. The foils of silica gel were used as adsorbent. ("Merck", DC-Alufolien Kie-selgel). Concentration profiles of methylene blue were measured in the flow of water through investigated beds, i.e. packed and fluidized beds (Bošković-Vragolović, et al., 2009). Colour intensity of the surface was determinate by "Sigma Scan Pro" software. Besides basic usage, for

determination of local and average mass transfer coefficients between fluid and solid surface, adsorption method is also, suitable for fluid flow visualization.
The rates of the dye adsorption are supposed to be diffusion controlled. The quantity of the dye transferred during a fixed time period is a function of diffusion boundary layer conditions. The rate of adsorption decreases from high values at the beginning to a constant value for investigated dye concentration. This constant value is the actual mass transfer rate through the completely formed boundary layer. For high concentration of the dye the concentration at the film's surface is $c_0 \neq 0$ and the adsorption process is controlled by mixed kinetics. Also, for high mass transfer rates mixed kinetics occur at lower concentration.
Adsorption from a much diluted solution and far from equilibrium conditions is a very fast process. If the exposure time is short (less than 10 min) the mass transfer rate depends only on the diffusion through the boundary layer. The concentration of adsorbate just above the adsorbent's surface is $c_i = 0$.
Mass flux is,

$$N_A = k\left(c_0 - c_i\right) = k\, c_0 \tag{1}$$

and for exposure time ,

$$N_A = \frac{c_p}{t} \tag{2}$$

where c_p, is the surface concentration of organic dye on adsorbent layer, c_0 is the bulk concentration.
If the induction period can be short (in the case of thin boundary layers) a simplified expression to calculate mass transfer coefficient is obtained (Končar-Djurdjević, 1953):

$$k = \frac{c_p}{t\, c_0} \tag{3}$$

For this method, value of the surface concentration c_p, is necessary for data quantification. This method is used since 1953, but determination of this parameter, c_p which now can be done easily, using suitable software (Sigma Scan Pro), gives absorption method new actuality.
The dissolution method is based on mass transfer from tube segment of benzoic acid to water. The investigated systems (packed bed, fluidized bed and the transport tube in hydraulic transport) are equipped with a tube segment prepared from benzoic acid (Bošković-Vragolović, et al., 2007; Garić-Grulović, et al., 2001&2009).
The mass transfer coefficient is calculated from the equation:

$$k_c = \frac{\Delta m}{A\, t\, \Delta c} = \frac{\Delta m}{(D_t \pi h)\, t\, \Delta c} \tag{4}$$

The transferred mass, Δm, is determined by measuring the weight loss of benzoic acid. The mass transfer area A is calculated from the mean value of tube segment diameter after and before dissolution. Since the weight loss of benzoic acid Δm is small, bulk concentration is negligible so equilibrium concentration c^* was taken as a driving force Δc. The benzoic acid solubility and diffusivity in water were taken from the literature (Kumar, et al., 1978; Kumar & Upadhyay, 1980).

In each run the average fluid/water temperature was recorded and corresponding values of the diffusion coefficients, fluid viscosity, fluid density and equilibrium solubility were considered in calculations. The water temperature was ranged from 14 to 16°C. Important properties of particles and bed are summarized in table 1.

Particles (glass spheres)					
d_p (mm)	1.2	1.94	2.06	2.98	3.04
ρ_p (kg/m^3)	2641	2507	2461	2509	2465
ε_{mF}	0.392	0.422	0.424	0.462	0.447
U_{mF} (m/s)	0.0129	0.0255	0.0268	0.0435	0.0409
U_t (m/s)	0.208	0.299	0.3025	0.371	0.3680

Table 1. Particle characteristics

4. Hydrodynamics and mass transfer

4.1 Hydrodynamics in heterogeneous systems

For design of heterogeneous systems, used for the mass transfer is important to know hydrodynamics characteristics of systems (i.e. liquid velocity, particle velocity and voidage). Investigation of hydrodynamics in liquid fluidized beds are covered by a large number of research papers, where are presented: the minimum fluidization velocity, the expansion of fluidized bed, the fluid-particle friction coefficient, the drag coefficient of bed and movement of particles (Makkawi & Wright, 2003; Yang & Renken, 2003).

For liquid-particle systems, the two most important characteristics are: for any liquid velocity of the fluidized bed voidage is uniform and the bed voidage is increasing function. Many authors have noted that on the diagram U = f (ε) in the logarithmic scale is present a change of slope for ε≈0.85. Both Garside & Al Dibouni 1977, and Riba & Couderc 1977, have observed a change in the slope of their data on such a plot at ε≈0.85 and suggested correlations for low ε (ε < 0.85) and high ε (ε> 0.85) systems. Carlos & Richardson, 1968, and Kmiec, 1978, concluded, analyzing the movement of labelled particles in fluidized beds, that probably there is a change in the mechanism of momentum transfer at voidage ε≈0.85.

Our investigations in liquid fluidized beds showed agreement with mentioned literature results (Fig. 2).

Our investigation have shown that in the vertical liquid-particle flow, at voidage ε≈0.85 there is also change in mechanism of momentum transfer, which represents the transition from turbulent to parallel flow of particles (Fig 3&4).

In the hydraulic transport of coarse particles, the two characteristic flow regimes are defined as:

a. Turbulent flow, where the particles move vertically but with some radial movement. This regime is characteristic of the lower fluid and particle velocities and appears very much like the inverse settling of a particle suspension. At the same time, the flowing suspension is very much like a particulate fluidized bed, where the whole fluidized suspension flows relative to the tube walls.
b. Parallel flow, where the particles move vertically along a parallel streamline. This regime is characteristic of the higher fluid and particle velocities.

The 2D-photographic illustration of these two regimes is shown in Fig. 3, for glass particles, 5 mm in diameter. The water was coloured with methylene blue. Particle tracks were recorded using a strong light behind the column and an exposure of 0.5 s.

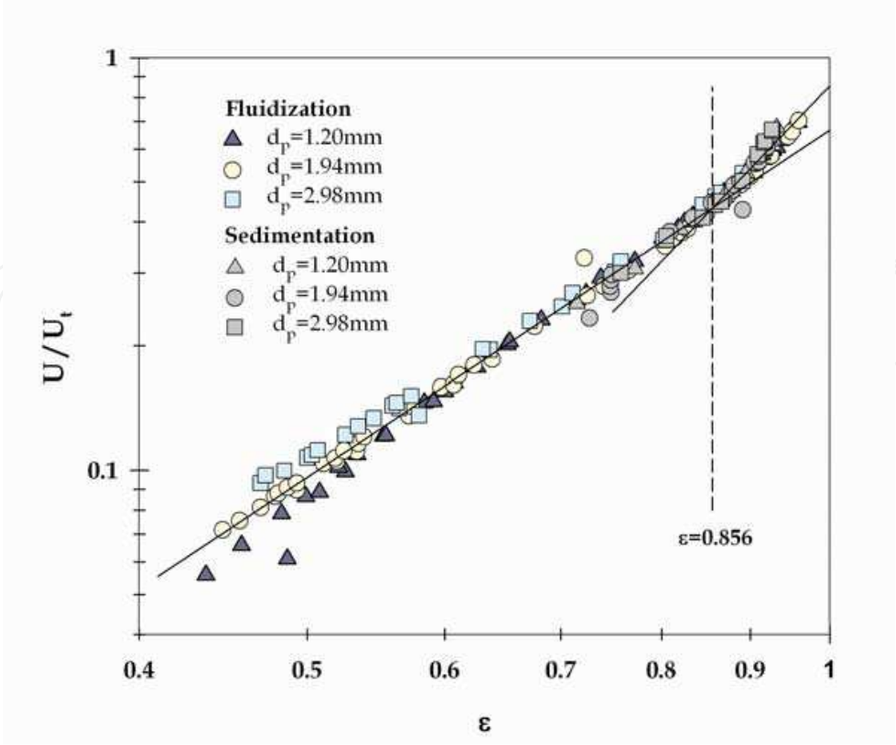

Fig. 2. The relationship between velocity and voidage for water-fluidized and sedime-nting beds of spherical glass particles (D_t = 40mm)

(A) Turbulent flow (B) Parallel flow

Fig. 3. Illustration of turbulent (A) and parallel (B) particle flow. Twodimensional column 60x8 mm, glass spheres 5 mm in diameter, water coloured by methylene blue

In our previous work (Grbavčić et al., 1992; Garić-Grulović et al., 2004&2008), we found that the choking criterion for vertical pneumatic transport lines of Day et al., 1990, is also valid as a criterion for regime designation in the hydraulic transport of coarse particles. Day et al., 1990, studied choking phenomena in vertical gas–solids flow. Using the steady state one-dimensional suspension momentum equation, with constant fluid properties including frictional effects between the bed and the wall (Leung, 1980, Grbavčić et al., 1992), could be expressed as,

$$-\frac{dP}{dz} = (\rho_p - \rho_f)g(1-\varepsilon) + F_w + \gamma\frac{d\varepsilon}{dz} \tag{5}$$

where

$$\gamma = \rho_p v^2 - \rho_f u^2 \tag{6}$$

They proposed γ=0 at the inlet of the transport tube as their choking criterion. This criterion together with a semi-theoretical relationship for the slip velocity at the inlet leads to an equation for predicting the choking velocity that agrees quite well with all available experimental data for vertical gas-particle flow. Fig. 4 shows the relationship between dimensionless parameter γ^* and transport line voidage, where:

$$\gamma^* = \frac{\rho_p v^2 - \rho_f u^2}{\rho_f U_t^2} \tag{7}$$

As can be seen, from Fig. 4, the critical voidage for regime transition in vertical transport is about $\varepsilon \approx 0.85$. The value of voidage $\varepsilon \approx 0.85$ is shown as a typical value in particulately fluidized and sedimenting beds (Fig. 2). Since in the turbulent flow regime in vertical transport, the frequency of particle collisions with the tube wall is much higher, it is reasonable to expect higher mass transfer rates in this flow regime.

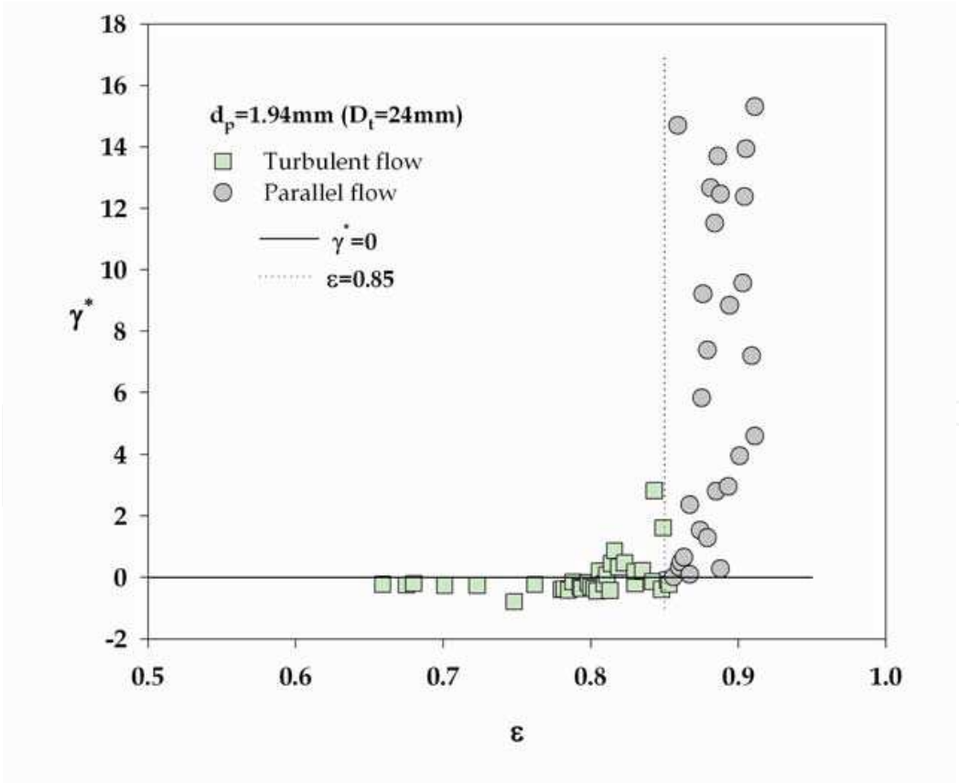

Fig. 4. Relationship. γ^* vs. ε for hydraulic transport (Grbavčić et al., 1992; Garić-Grulović et al., 2004&2008)

4.2 Mass transfer in heterogeneous systems

Mass transfer in heterogeneous systems is influenced by basic hydrodynamics parameters: fluid velocity, bed voidage and particles concentration. Fig. 5, shows relationship among mass transfer coefficient, liquid velocity and voidage in all investigated systems. Increasing liquid velocity increases mass transfer in packed bed. In fluidized bed increasing liquid velocity leads to expansion of bed (ε> 0.7) i.e. reduction of the concentration of particles in the bed, which contributes to reducing the mass transfer coefficient. In hydraulic transport, increasing liquid velocity and slightly changes of voidage have positive influence on mass transfer. Finally, as can be seen in Fig. 5, the influence of particle concentration for all system is slightly greater then the influence of liquid velocity.

Fig. 6, shows relationships between mass transfer coefficient and superficial liquid velocity. The highest mass transfer coefficient was at minimum fluidization velocity because of high concentration of moving particles.

Constant movement of particles in the bed causes mixing of fluid near the wall reduces the thickness of the boundary layer and increases the mass transfer (Yutani et al., 1987; Schmidt et al., 1999). In addition, investigation of Schmidt et al., 1999, indicate existence of a maximum of Sherwood number Sh = f (k), at bed voidage about 0.8. Results of our investigations (Fig. 5&6), show existence of maximum of mass transfer coefficient at transition from packed to fluidized bed, but there is no maximum for fluidized bed. In addition, with increasing fluid velocity fluidized bed is expands (ε>0.8, Fig. 6&7).

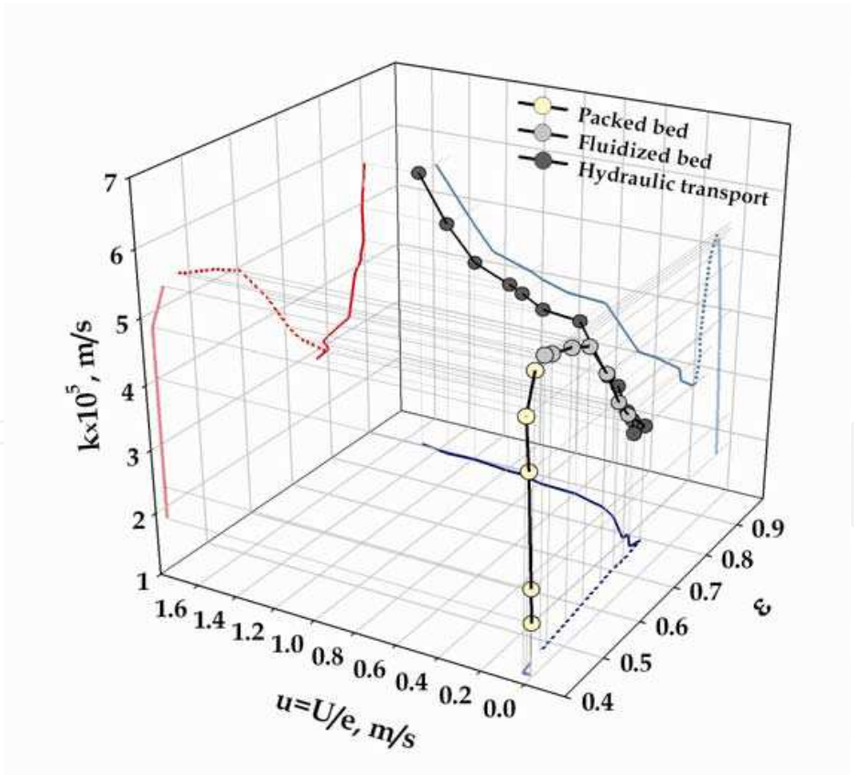

Fig. 5. The relationship between mass transfer coefficient, interstitial fluid velocity and voidage, for packed beds, fluidized beds and vertical transport (dissolution method; d*p*=1.94mm)

Chromatogram on Fig. 7, gives clear visualization of fluid flow around particles in the packed bed. Colour intensity is proportional to local mass transfer rates. Also, on Fig. 7, are shown chromatograms for fluidized beds at minimal velocity and for highly expanded fluidized bed. Uniform colour intensity could be observed in both cases, with higher intensity for minimal fluidized bed velocity.

Fig. 8, presents mass transfer coefficient as a function of superficial fluid velocity for packed beds, fluidized beds, vertical transport and single phase flow. With increasing liquid velocity in packed beds, mass transfer coefficient increases for all bed voidage. With increasing superficial liquid velocity in fluidized beds mass transfer coefficient slightly decreases, tends to the value of mass transfer coefficient for single phase flow. In the vertical transport for low transport velocities mass transfer coefficient is constant, and it is higher than mass transfer coel1icient for single liquid flow. For higher transport velocities mass transfer coefficient increases, but there no significant difference between transport and single liquid flow.

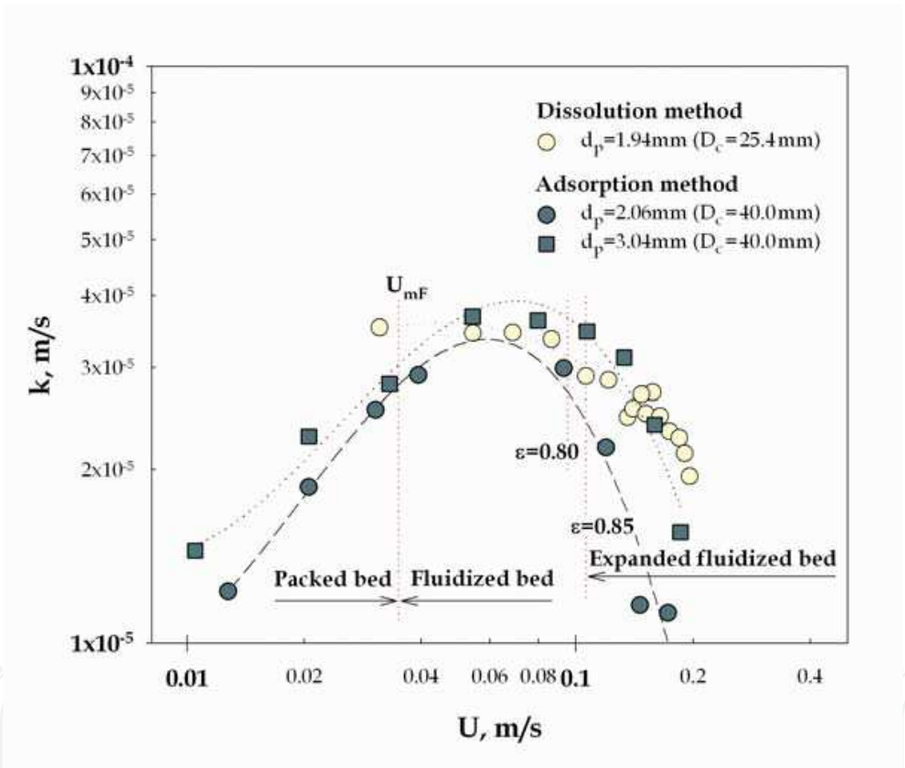

Fig. 6. Mass transfer coefficient vs. superficial fluid velocity for packed and fluidized beds

Fig. 9, present the relationship between particle Sherwood number and particle Reynolds number for different particle diameter in fluidized beds and vertical transport. The data are separated into tree groups with particle diameter. The experimental results show that with increasing Reynolds number (liquid velocity), Sherwood number (mass transfer coefficient), slightly decreases in fluidized beds.

With increasing liquid velocity the Sherwood number for vertical transport is constant for low transport velocities, but for higher transport velocities mass transfer increases. It implies that because of low particle concentration in transport column the influence of particles on diffusivity boundary layer is not significant for higher transport velocities.

Comparison of the data for fluidized bed (d_p=1.94mm) with several literature correlations (Upadhyay & Tripathi, 1975; Tournie et al., 1977; Bošković, et al., 1994), show significant difference between our data and the available correlations. The data calculated with it could be seen that are quit different.
The following correlation proposed by Upadhyay & Tripathi, 1975, for mass transfer in packed and fluidized bed,

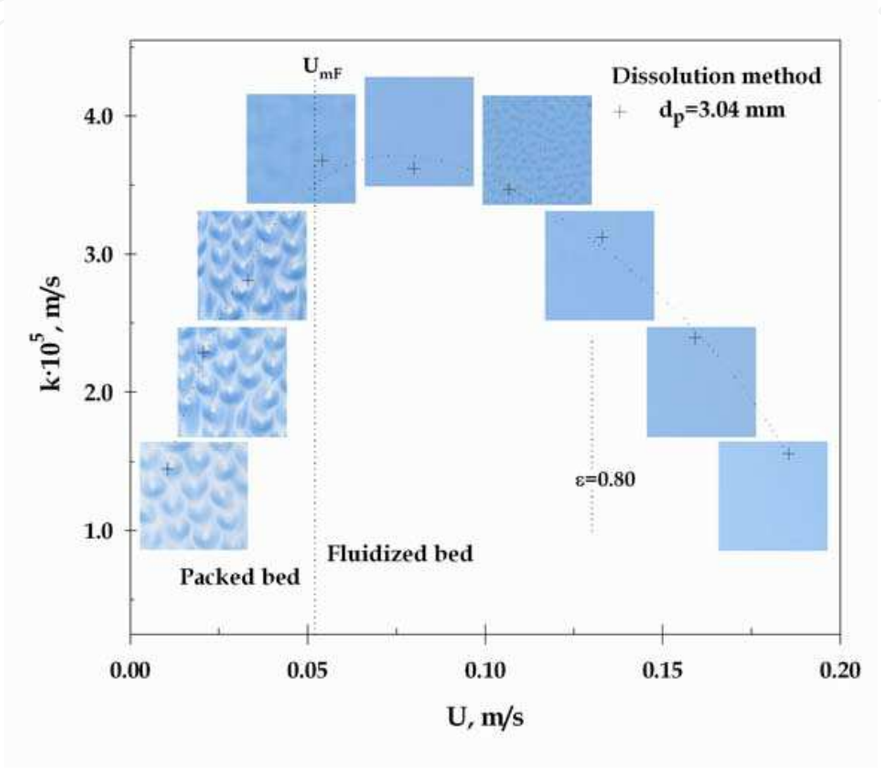

Fig. 7. Relationship between mass transfer coefficient and superficial liquid velocity in packed and fluidized beds

$$Sh_p = 3.8155 \cdot Re''^{0.2687} Sc^{1/3} \quad \text{for} \quad Re'' < 20 \tag{8}$$

$$Sh_p = 1.6218 \cdot Re''^{0.5553} Sc^{1/3} \quad \text{for} \quad Re'' > 20 \tag{9}$$

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 with correlation (eq. 9) show the maximum which has not been confirmed by our experimental data (Fig. 9).
Also, Tournie et al., 1977, have given the correlation for mass transfer particle-fluid in fluidized bed,

$$Sh_p = 0.253 \cdot Re_p^{0.004} Ga^{0.319} Mv^{0.299} Sc^{0.4} \tag{10}$$

where $Mv=(\rho_p-\rho_f)/\rho_f$, and the equation (10) is recommended in the following range of variables: $1.6<Re_p<1320$, $2470<Ga<4.42\cdot106$, $0.27<Mv<1.14$, $305<Sc<1595$. The data calculated

with Tournie et al., 1977, correlation for mass transfer particle-fluid in fluidized bed (eq. 10), show significant difference our experimental values (Fig. 9).

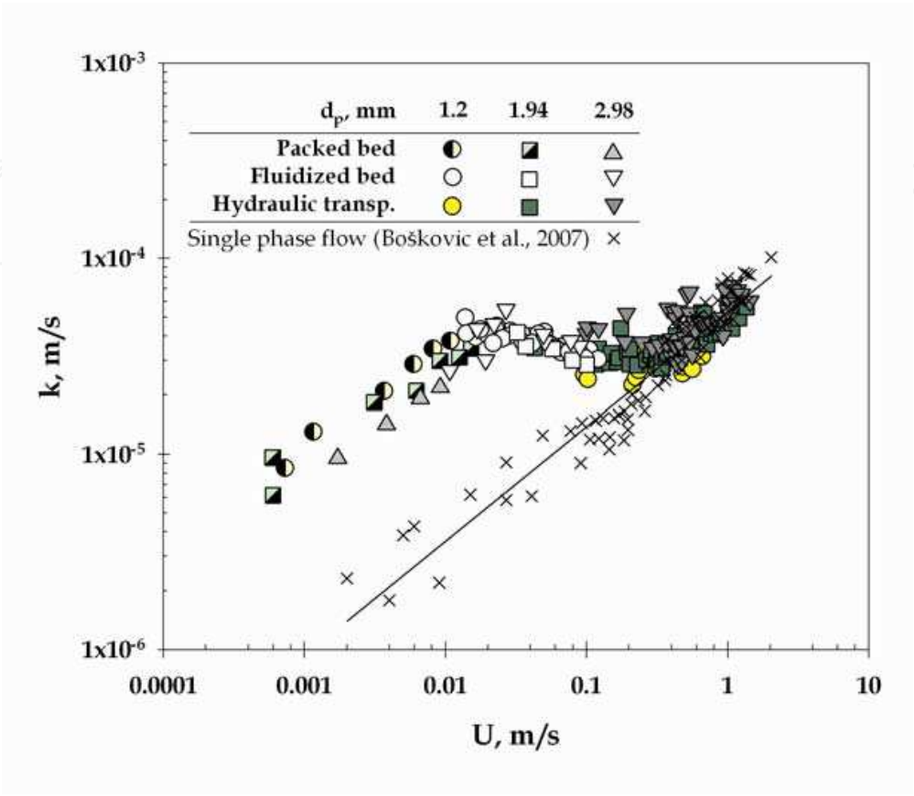

Fig. 8. Mass transfer coefficient vs. superficial fluid velocity for packed beds, fluidized beds, vertical transport and for single liquid flow by dissolution method

The mass transfers in liquid fluidized bed, Bošković et al., 1994, are correlated by the equation,

$$Sh_p = 0.261 \cdot Re_p^{0.03} Ga^{0.324} Sc^{1/3} \tag{11}$$

in the following range of variables: Re_p=15÷400, Sc=1361÷1932. This equation is derived for mass transfer between fluid and an immersed sphere in fluidized beds of spherical inert particles. The predicted values using the correlation by Bošković, et al., 1994 (eq. 11), are in good agreement with our experimental data (Fig. 9).

Fig. 10, present the relationship between mass transfer factor and particle Reynolds number in fluidized beds obtained by adsorption and dissolution methods. Also this picture gives the comparison of our experimental data with several literature correlations (Chu et al., 1953; Fan et al., 1960; Upadhyay & Tripathi, 1975; Dwivedi & Upadhyay, 1977; Bošković, et al., 1994), expressed through the mass transfer factor (j_D), which is defined as:

$$j_D = \frac{Sh}{Sc^{1/3} Re} \tag{12}$$

The data are calculated with correlation proposed by Chu et al., 1953, as follows,

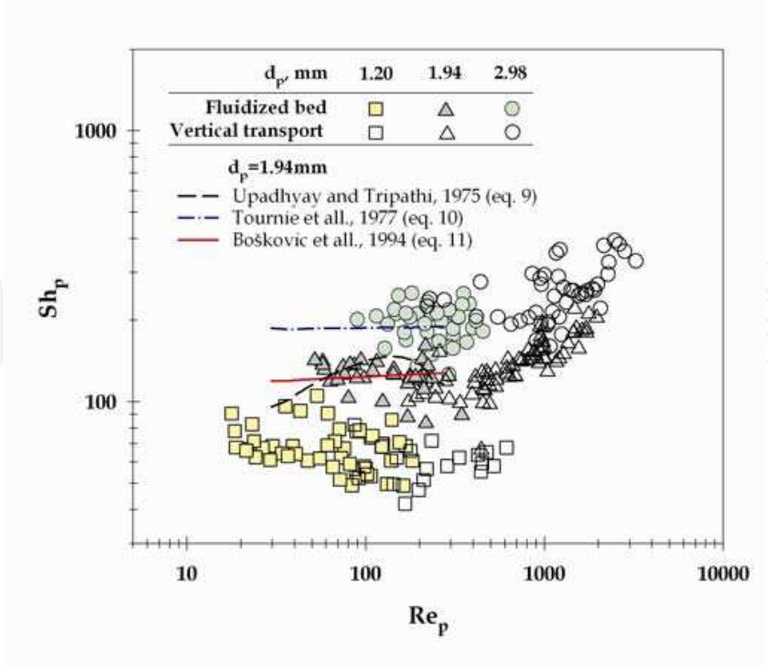

Fig. 9. Relationship between Sherwood number and Reynolds number for particles (dissolution method)

$$j_D = 5.7 \cdot Re^{"-1.22} \qquad \text{for} \qquad 1 < Re^{"} < 30 \tag{13}$$

$$j_D = 1.77 \cdot Re^{"-1.56} \qquad \text{for} \qquad 30 < Re^{"} < 1000 \tag{14}$$

where $Re^{"} = Re_p/(1-\varepsilon)$, show good agreement with our experimental data (Fig. 10).

Fan et al., 1960, on the basis of adsorption tests of phenol on activated carbon particles in liquid fluidized bed came to the following correlation,

$$j_D = 1.865 \cdot Re^{' 0.48} \tag{15}$$

where $Re^{'} = Re_p/\varepsilon$. The predicted values from the correlation (15) and our experimental data have shown significant difference (Fig. 10).

The data calculated with correlation proposed by Upadhyay and Tripathi, 1975, for mass transfer in packed and fluidized bed (from eq. 9&12),

$$j_D = 1.6218 \cdot Re^{"0.4447} \qquad \text{for} \qquad Re^{"} > 20 \text{)(eq. 9)} \tag{16}$$

where Re"=$Re_p/(1-\varepsilon)$, show good agreement with the our experimental data (Fig. 10).

Also Fig. 10, gives the comparison of our experimental data with correlations of Dwivedi & Upadhyay, 1977, who proposed the following correlations,

$$j_D \cdot \varepsilon = 1.1068 \cdot Re_p^{-0.72} \qquad \text{for} \qquad Re_p < 10 \tag{17}$$

$$j_D \cdot \varepsilon = 0.4548 \cdot Re_p^{-0.4069} \qquad \text{for} \qquad Re_p > 10 \tag{18}$$

based on the many different experimental data for mass transfer of liquid in packed and fluidized beds. As can be seen, our data show very good agreement with correlation for lower values of Reynolds number for particles in the fluidized beds.

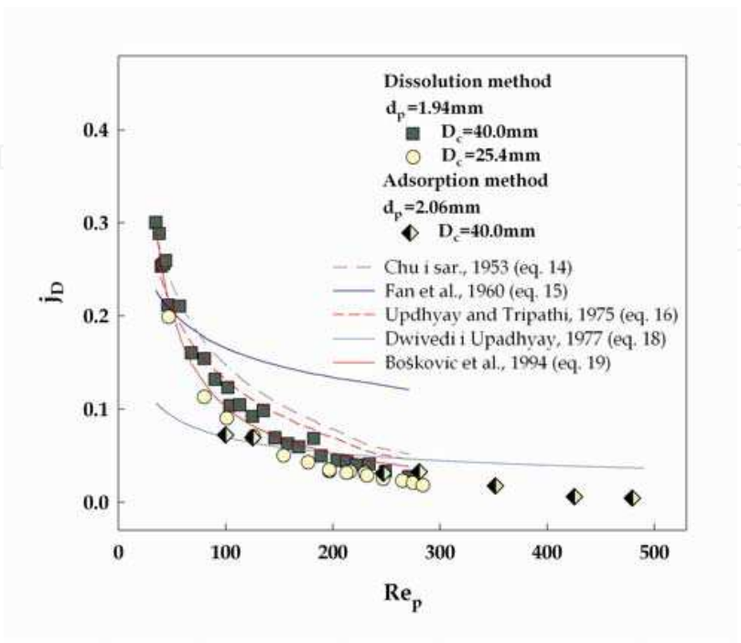

Fig. 10. Variation of mass transfer factor with Reynolds number for particles

And finally, the correlation of Bošković et al., 1994, has been derived for the mass transfer between fluid and an immersed sphere in fluidized beds of spherical inert particles (from eq. 11&12),

$$j_D = 0.261 \cdot Ga^{0.324}\, Re_p^{-0.97} \tag{19}$$

shows excellent agreement for the mass transfer wall-to-fluid (Fig. 10).
Fig. 11, presents the mass transfer factor as a function of Reynolds number in packed, fluidized beds and hydraulic transport for different experimental techniques. It could be seen that there is no significant difference between mass transfer factors obtained by this two methods (Bošković, et al. 1994; Bošković-Vragolović, et al., 2007). Often used dissolution method is very reliable, and agreement of data shows that the adsorption method gives good results, also. Advantage of adsorption method is possibility to obtain local mass transfer coefficients.

5. Conclusion

The basic hydrodynamic parameters of packed bed, fluidized beds and vertical transport are analyzed.
The influence of different parameters as liquid velocity, particles size and voidage on mass transfer in packed beds, fluidized beds and hydraulic transport are presented. The data for mass transfer in all investigated systems are shown using Sherwood number (Sh) and mass transfer factor-Colburn factor (j_D) as a function of Reynolds number (Re) for particles and for column.

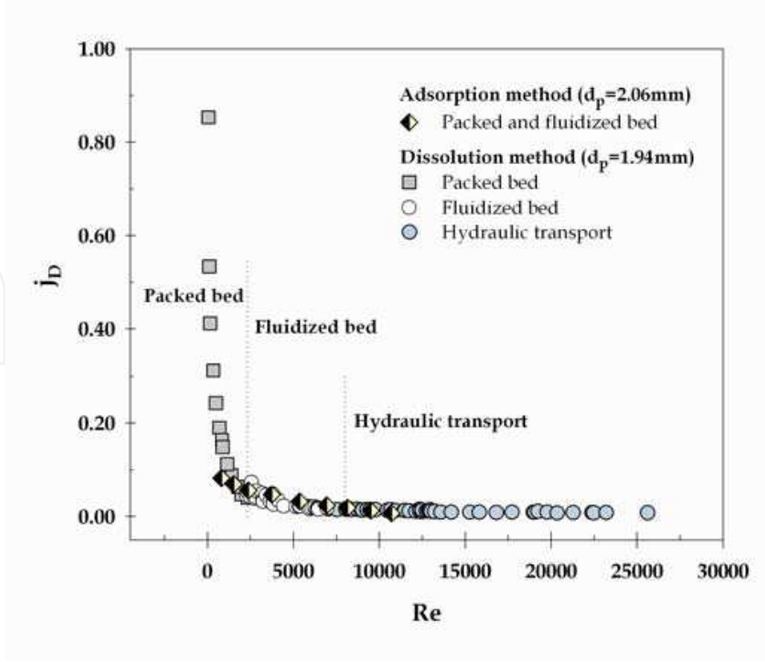

Fig. 11. The relationship between mass transfer factor and Reynolds number

With increasing liquid velocity in packed beds mass transfer coefficient increases while in fluidized beds mass transfer coefficient decrease. The highest mass transfer coefficient was at minimum fluidization velocity because of high concentration of moving particles.
In the vertical transport for low transport velocities mass transfer coefficient is constant, and it is higher than mass transfer coel1icient for single liquid flow. For higher transport velocities mass transfer coefficient increases, but there no significant difference between transport and single liquid flow.

6. Nomenclature

A	mass transfer area, m^2
c^*	equilibrium concentration, kg/m^3
c_i	concentration of adsorbate, kg/m^3
c_0	concentration of the bulk, kg/m^3
cp	surface concentration, kg/m^3
Δc	concentration difference (driving force), kg/m^3
D_{AB}	molecular diffusion coefficient, m^2/s
d_p	particle diameter, m
D_t	diameter of the column/transport tube, m
F_w	pressure gradient due to the suspension-wall friction, Pa/m
g	gravitational acceleration, m/s^2
Ga	Galileo number ($=(d_p^3\rho_f^2 g)/\mu^2$)
h	high of tube segment, m
j_D	mass transfer factor, $(k/U)\cdot Sc^{1/3}$
k	mass transfer coefficient, m/s
Δm	transferred mass, kg

Mv	density number (=$(\rho_p-\rho_f)/\rho_f$)
N_A	mass flux, kg/(m²s)
P	dynamic pressure, Pa
Re	Reynolds number for column (=$D_t\rho_f U/\mu$)
Re_p	Reynolds number for particles (=$d_p\rho_f U/\mu$)
Sc	Schmidt number (=$\mu/(\rho_f D_{AB})$)
Sh	Sherwood number for column (=kD_t/D_{AB})
Sh_p	Sherwood number for particles (=kd_p/D_{AB})
t	exposure time, s
u	mean interstitial fluid velocity in the transport tube (=U/ε), m/s
U	superficial fluid velocity in the spout and in the transport tube, m/s
U_{mF}	minimum fluidization velocity, m/s
U_t	particle terminal velocity in an infinite medium, m/s
v	radially averaged particle velocity in the transport tube, m/s
z	vertical distance coordinate, m
	Greek letters
γ	defined by Eq. (6)
γ^*	defined by Eq. (7)
ε	voidage
ε_{mF}	voidage at minimum fluidization
ρ_f	fluid density, kg/m³
ρ_p	particle density, kg/m³
μ	viscosity of the fluid, Ns/m²

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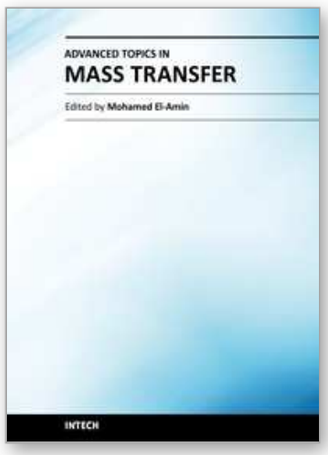

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