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PARTICLE-TO-FLUID HEAT AND MASS TRANSFER IN PACKED BEDS OF FINE PARTICLES

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Abstract—By application of a simple model on heat or mass transfer between solids and flowing fluid in packed beds, experimental data of both Nusselt and Sherwood numbers reported in the previous literatures are interpreted theoretically in the range of low Péclet number, i.e. $Pe_p < 10$. Then it is suggested that channelling or local uneven contacting of fluids with solids is responsible for the further decrease of apparent heat- and mass-transfer coefficients in the above system.

NOMENCLATURE

a, surface area of particles per unit volume of bed [m²/m³];
b, radius of free surface [m];
C, concentration [mol/m³];
C_p, specific heat of gas at constant pressure [kcal/m³ degC];
D_p, particle diameter [m];
$\mathscr{D}_v$, mass diffusivity [m²/h];
f, fraction of dead water volume to total void volume;
G_0, mass flow rate of fluid per unit sectional area [kg/m² h];
h_p, particle-to-fluid heat-transfer coefficient based on total surface area of particles [kcal/m² h degC];
k_f, thermal conductivity of fluid [kcal/m h degC];
k_p, particle-to-fluid mass-transfer coefficient [m/h];
l, (1/2) × (average film thickness in channelling model) [m];
Q, total amount of heat transferred from particle to fluid per unit sectional area [kcal/m² h];
R, radius of particle [m];
s_v, surface area in unit stage per unit sectional area;
T, temperature [°C];
U_0, superficial fluid velocity based on sectional area of the empty tube [m/h];
V_ε, void volume in unit stage per unit sectional area [m];
Nu_p, Nusselt number, h_pD_p/k_f;
Pe_p, Péclet number, $D_pC_pG_0/k_f$ or $D_pU_0/\mathscr{D}_v$;
Re_p, Reynolds number, D_pG_0/μ;
Sh_p, Sherwood number, $k_pD_p/\mathscr{D}_v$.

Greek symbols

α, thermal diffusivity of fluid [m²/h];
γ, R/b in equation (2);
ε, void fraction;
ξ, ratio of average channelling length to particle diameter;
ρ, density of fluid [kg/m³];
θ, age of fluid defined by $Z/\{U_0/\varepsilon(1-f)\}$ [h];
ψ, dimensionless temperature or concentration;
ϕ_s, shape factor.

Subscripts

0, inlet fluid;
1, average of outlet fluid;
s, at the surface of solids.

INTRODUCTION

MUCH work has been done so far on heat or mass transfer between solids and flowing fluids in packed beds. Especially for the purpose of rational design of catalytic reactors, it has been recognized that the prediction of particle-to-fluid heat- or mass-transfer coefficients is important in the course of the estimation of local or overall conversions. Even though a number of experimental correlations have been reported in the recent literatures, they do not agree well with each other, particularly in the low Reynolds number range.

According to Ranz [1], heat-transfer coefficients between solids and flowing fluid in packed beds are given by

$$h_p D_p/k_f = 2{\cdot}0 + 0{\cdot}6\,(C_p\mu/k_f)^{\frac{1}{3}}\,(D_p G'/\mu)^{\frac{1}{2}} \qquad (1)$$

where G' is the effective mass flow rate of fluid and may be assumed as nine times superficial mass velocity based on the empty tube, namely $9{\cdot}0\,G_0$ in the case of packed beds. This equation can explain well the experimental data in turbulent flow range. Turbulence of fluid in void spaces may degrade effects of adjacent particles and boundary layers should develop on surfaces of packed particles in a similar manner to the case on an isolated particle. In the range of low Reynolds numbers, however, boundary conditions describing heat or mass transfer in flowing media are quite different, as Cornish [2] has suggested, from those for an isolated particle system. Therefore, the theoretical value, 2, for a single sphere in a stagnant medium does not have physical meaning any more.

Pfeffer [3] combined "free surface model" introduced by Happel [4] with "thin boundary-layer solution" [5]; he showed a simple relationship among heat- or mass-transfer coefficients, Péclet number and void fraction as follows:

$$\left.\begin{aligned} h_p D_p/k_f &= 1{\cdot}20\left[\frac{1-(1-\varepsilon)^{\frac{5}{3}}}{W}\right]^{\frac{1}{3}} (D_p C_p G_0/k_f)^{\frac{1}{3}} \\ W &= 2 - 3\gamma + 5\gamma^5 - 2\gamma^6 \\ \gamma &= R/b = (1-\varepsilon)^{\frac{1}{3}}\,. \end{aligned}\right\} \qquad (2)$$

This equation is applicable to the high Péclet number–low Reynolds number region, as he assumed the existence of viscous flow and a thin thermal or diffusional boundary layer on the surface of a particle.

It is worth remarking on the proposal of Zabrodsky [6] on a heat-transfer mechanism in a fluidized bed. He tried to interpret experimental values of Nusselt number by means of a model with "micro-breaks", that is, local by-passing of gas due to non-uniformity of fluidization.

CHANNELLING MODEL

When the flow rate of fluid is small in void spaces of packed beds, mutual interactions of neighboring particles should be taken into account for the transport between particles and flowing fluid. Pfeffer *et al.* [3, 7] interpreted these phenomena by means of "free surface model". This model consists of a cell made of two concentric spherical surfaces, the inner representing the surface of a particle itself and the outer the boundary of a fluid envelope. They assumed the outside surface of each cell to be frictionless and to be kept at a constant temperature or concentration. The latter assumption is valid only at sufficiently high Péclet numbers, and should be replaced by the adiabatic condition (impermeable surface) if this model is to be used outwith the range $Pe_p \gg 1$.

The void spaces are so narrow in actual packed beds that the fluid path may be approximated by a film bounded by two parallel planes indicating surfaces of solids. Surface area of the parallel planes corresponds to that of solids, and the volume of the fluid film corresponds to that of void spaces. Suppose that some groups of particles are packed closely to form aggregates while others are packed loosely. This distribution of voidage in beds will result in selective flowing of fluid (channelling) in the loose section in a bed of fine particles.

Flow structure in an actual packed bed may be simplified as illustrated in Fig. 1, where f is defined to be the portion of dead water volume

to total void volume and then ξD_p the average axial length of unit channelling.
Postulates:

(1) In the stagnant block, the temperature or concentration of the fluid is in equilibrium with that of the interspersed solids;

(2) Transfer of heat or mass can be described in two-dimensional co-ordinates and fluid flow in channel is assumed to be plug flow;

(3) Heat or mass transfer is restricted to molecular diffusion in the direction perpendicular to flow.

In the model introduced by the above

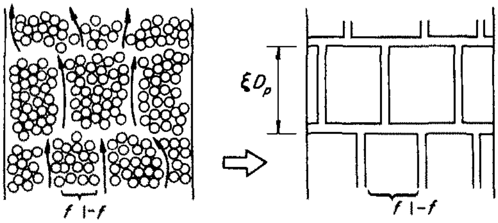


FIG. 1. The concept of channelling model.

simplifications, surface area of particles contained in one unit stage of channelling, ξD_p per unit sectional area is

$$s_v = \xi D_p . a \tag{3}$$

where a represents surface area of particles per unit volume of packed beds, which is

$$a = \frac{6(1-\varepsilon)}{\phi_s D_p}$$

ϕ_s: shape factor.

Fluid volume in one unit stage per unit sectional area is

$$V_\varepsilon = \varepsilon . \xi D_p . \tag{4}$$

Average thickness of the fluid film in this model bed $2l$ is calculated as

$$2l = V_\varepsilon / \tfrac{1}{2} s_v \tag{5}$$

where s_v corresponds to the area of both surfaces of the film. Equations (3), (4) and (5) give

$$l = \frac{\varepsilon \phi_s}{6(1-\varepsilon)} D_p \cdot \tag{6}$$

The average residence time of fluid in one unit stage is

$$\Delta\theta_{av} = \int_0^\infty E(\Delta\theta) . \Delta\theta \, d(\Delta\theta) = \xi D_p / \{U_0/\varepsilon(1-f)\} \tag{7}$$

where $E(\Delta\theta)$ represents residence time distribution of fluid in one unit stage. Now the average value of dimensionless temperature or concentration of the fluid passing one unit stage in the model bed ψ_1 is

$$\psi_1 = \frac{T_1 - T_s}{T_0 - T_s} \quad \text{or} \quad \frac{C_1 - C_s^*}{C_0 - C_s^*}$$

where T_0, T_1 represent temperatures of inlet and outlet fluid respectively, while T_s means equilibrium temperature with solids. Furthermore, C_0, C_1 and C_s^* represent corresponding concentration respectively.

Total amount of heat exchanged between particles and fluid in one unit of channel is

$$Q = C_p \rho U_0 (T_1 - T_0) \cdot \tag{8}$$

Particle-to-fluid heat-transfer coefficient based on the total surface area of packed particles, h_p, is defined by

$$Q = h_p s_v (T_s - T_0) \cdot \tag{9}$$

Combination of equations (8) and (9) gives h_p as follows:

$$h_p = \frac{C_p \rho U_0 (T_1 - T_0)}{s_v (T_s - T_0)} = \frac{C_p \rho U_0 \phi_s}{6(1-\varepsilon)\xi}(1 - \psi_1) \cdot \tag{10}$$

Therefore, the Nusselt number is

$$Nu_p = \frac{h_p D_p}{k_f} = \frac{\phi_s (1 - \psi_1)}{6(1-\varepsilon)\xi} \cdot \frac{D_p C_p G_0}{k_f} \cdot \tag{11}$$

The basic equation for determining ψ_1 is the familiar two-dimensional diffusion equation in a co-ordinate system shown in Fig. 2.

$$U \frac{\partial T}{\partial z} = \alpha \frac{\partial^2 T}{\partial x^2} \tag{12}$$

where

$$U = U_0/\varepsilon(1-f)\cdot$$

Boundary conditions are

$$T = T_s: \quad x = \pm l, \qquad z > 0$$

$$T = T_0: \quad z = 0.$$

Equation (12) reduces to one dimensional

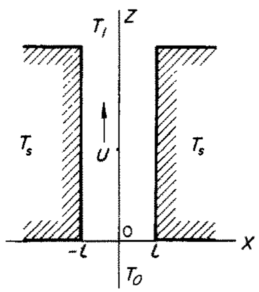


FIG. 2. Co-ordinate system.

unsteady-state equation by substituting

$$\theta = z/\{U_0/\varepsilon(1-f)\}$$

and

$$\psi = (T - T_s)/(T_0 - T_s),$$

to obtain

$$\frac{\partial \psi}{\partial \theta} = \alpha \frac{\partial^2 \psi}{\partial x^2} \tag{13}$$

$$\psi = 0: \quad x = \pm l, \qquad \theta > 0$$

$$\psi = 1: \quad \theta = 0.$$

The solution of equation (13) is as follows [8]:

$$\psi = \frac{4}{\pi} \sum_{n=0}^{\infty} \frac{(-1)^n}{2n+1} \times \exp\left[-\alpha(2n+1)^2 \pi^2 \Delta\theta/4l^2\right] \times \cos\frac{(2n+1)\pi x}{2l} \tag{14}$$

from which dimensionless temperature of outlet fluid is given by

$$\psi_1 = \frac{1}{2l}\int_0^\infty \mathrm{d}(\Delta\theta) \int_{-l}^{l} \psi \,.\, E(\Delta\theta)\,\mathrm{d}x. \tag{15}$$

Two typical cases of residence time distributions are considered here.

(I) Residence time of flowing fluid in each channel is equal to the average residence time, that is

$$E(\Delta\theta) = \delta(\Delta\theta - \Delta\theta_{av})$$

δ: Dirac's delta function.

(II) Residence time distribution of fluid in one stage is equal to that of an ideal mixing cell, that is

$$E(\Delta\theta) = \frac{1}{\Delta\theta_{av}} \,.\, \exp -\left(\frac{\Delta\theta}{\Delta\theta_{av}}\right)$$

ψ_1 is calculated for both cases.

Case (I),

$$\psi_1^{\mathrm{I}} = \sum_{n=0}^{\infty} \frac{8}{(2n+1)^2 \pi^2} \times \exp\left[-\alpha(2n+1)^2 \pi^2 \Delta\theta_{av}/4l^2\right]. \tag{16}$$

Case (II),

$$\psi_1^{\mathrm{II}} = \sum_{n=0}^{\infty} \frac{8}{(2n+1)^2 \pi^2} \times \frac{1}{1 + \alpha(2n+1)^2 \pi^2 \Delta\theta_{av}/4l^2}. \tag{17}$$

Plots of ψ_1 against $\alpha \,.\, \Delta\theta_{av}/l^2$ for both cases are shown in Fig. 3. The meaning of $\alpha \,.\, \Delta\theta_{av}/l^2$ appearing above is found as

$$\frac{\alpha \,.\, \Delta\theta_{av}}{l^2} = \frac{k_f}{C_p \rho} \cdot \frac{\xi D_p}{U_0/\varepsilon(1-f)} \cdot \left\{\frac{6(1-\varepsilon)}{\varepsilon \phi_s D_p}\right\}^2 = \frac{36(1-\varepsilon)^2 \xi(1-f)}{\varepsilon \phi_s^2} \cdot \frac{1}{Pe_p}. \tag{18}$$

If parameters in the above equations are taken as $\varepsilon = 0{\cdot}4$, $\phi_s = 1$, $f = 0$ and $\xi = 1$ (ideal case), $\alpha \,.\, \Delta\theta_{av}/l^2$ is larger than 3 in the region of $Pe_p < 10$, which leads to the results $(1 - \psi_1) \simeq 1$ in both cases.

In this range of parameters, equation (11) reduces to

$$Nu_p = \frac{h_p D_p}{k_f} = \frac{\phi_s}{6(1-\varepsilon)\,\xi} Pe_p. \quad (19)$$

It might be justified here that heat-transfer coefficient defined by equation (9) does not mean the film heat-transfer coefficient in the usual sense for the region of small values of Péclet numbers. It should be kept in mind that the residence time of fluid even in one unit stage with the length of one particle diameter is

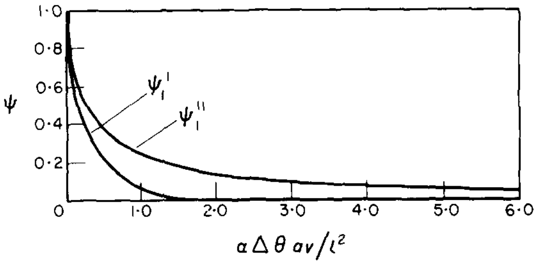


FIG. 3. Plots of ψ_1 vs. $\alpha\Delta\theta_{av}/l^2$ in equations (16) and (17).

longer than the time necessary for attainment of thermal equilibrium with solid particles.

Therefore the coefficient defined by equation (9) under flow condition of low Péclet number might be named the "apparent heat-transfer coefficient". In the ideal case mentioned above without dead water or channelling ($f = 0$, $\xi = 1$), apparent heat-transfer coefficient h_p should be expressed as $C_p G_0 \phi_s/6(1-\varepsilon)$ in the range of $Pe_p < 10$.

For Sherwood number, almost the same derivation can be applied in a bed uf fine particles, by replacing the temperature in the above derivations by the concentration of flowing fluid.

$$Sh_p = \frac{k_p D_p}{\mathscr{D}_v} = \frac{\phi_s}{6(1-\varepsilon)\,\xi} \,.\, Pe_p. \quad (20)$$

Even though almost all of the experimental data on heat and mass transfer in fixed or fluidized beds have been correlated in literatures by plotting j-factor versus Reynolds numbers, it is not justified physically to use this correlation in the region of low Péclet number, since existence of thermal or diffusion boundary layer has no significant meaning in the system any more.

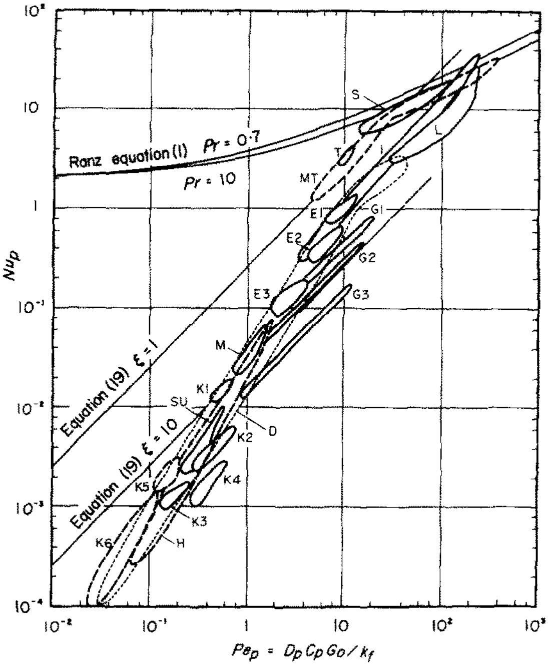
DISCUSSION

In Fig. 4, experimental data of Nu_p are compared with equation (19), where ξ is taken as 1·0 and 10 and $\phi_s = 1{\cdot}0$ and $\varepsilon = 0{\cdot}4$ are assumed. The experimental data are taken from a number of investigations on heat transfer in packed beds [9–20]. As indicated in the figure, equation (20) can explain the inclination of the most of all observations at low Péclet numbers, that is $Nu_p \propto Pe_p$.

Mass transfer data obtained previously by Bar-Ilan [21], Aerov [22], Resnick [23], and Thodos [24] are illustrated in Fig. 5 in comparison with equation (20). Bar-Ilan measured the rate of sublimation of naphthalene solids immersed in dummy particles, and the results coincided with the prediction proposed here. The published data on mass transfer at low Péclet numbers are very scanty and limited only to gas systems.

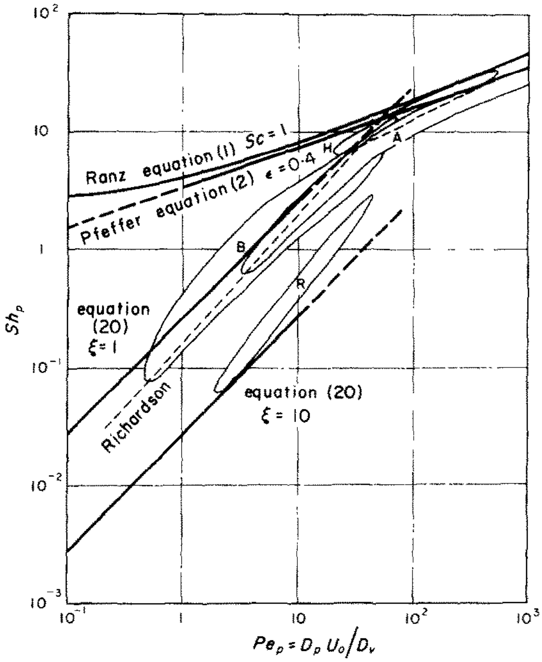
In fluidized beds, Richardson and his colleagues [25, 26] measured heat- and mass-transfer coefficients in shallow, particulately fluidized beds. Their mass-transfer correlations are also included in Fig. 5 (dotted lines). Their data can be interpreted as the ideal case of the present model without dead space or channelling, because the particles may be dispersed homogeneously in the shallow fluidized bed. Their results show rather stronger dependence on flow rate ($Sh_p \propto Re_p^{1\cdot18}$) than indicated in equation (20), which might be attributed to the increase of voidage with increasing flow rate of fluidizing media.

At sufficiently low Péclet numbers where $\alpha\,.\,\Delta\theta_{av}/l^2$ in equation (16) and (17) is large enough, apparent heat-transfer coefficient does not depend on dead water fraction f as deduced before. Therefore, channelling length factor ξ can be back estimated from figures by fitting with equation (19). Thus obtained ξ may be a function of particle size, flow rate, viscosity of fluid, packing structures, etc. However, because



KEY	Authors	Particle diameter (mm)	Fluid
L:	Löf and Hawley [9]	32·8	air
G1:	Grootenhuis [10]	0·39	air
G2:		0·21	
G3:		0·064	
E1:	Eichhorn and White [11]	0·66	air
E2:		0·52	
E3:		0·28	
S:	Satterfield and Resnick [12]	5·1	
K1:	Kunii and Smith [14]	1·02	air
K2:		0·57	
K3:		0·40	
K4:		0·11	
D:	Dannadieu [27]	0·13 ~ 1·1	air
I:	Kunii and Ito [15]	0·45~5·0	air
M:	Mimura [16]	3·7	air
SU:	Suzuki [17]	1·10	air
T:	Tokutomi [19]	1·0	air
K5:	Kunii and Smith [14]	1·02	water
K6:		0·57	
H·	Harada [18]	0·55~3·1	water
MT:	Mitsumori [20]	1·18 ~11·0	water

FIG. 4. Comparison of equation (19) with heat-transfer data.



KEY	Authors
B:	Bar-Ilan and Resnick [21]
A:	Aerov and Umnik [22]
R:	Resnick and White [23]
H:	Hobson and Thodos [24]

FIG. 5. Comparison of equation (20) with mass-transfer data.

of the lack of experiments covering a wide range of flowing conditions or packing structures, a clear relationship between ξ and above factors cannot be found. Nevertheless, it can be noted by an inspection of the figures that ξ depends more on particle diameter than on flow rate. In Fig. 6, ξ are plotted against particle diameter, where ξ seems to be in inverse proportion to particle diameter. In the correlated data, Grootenhuis [10] measures heat-transfer coefficients between sintered metal and air flowing through it. It can be understood easily that void structures in sintered metal are different from those in packed beds. His results are cited only for comparison. Eichhorn [11] made the measurement of heat-transfer coefficients in DVB-particle–air system where particles were heated by electric induction and exchanged heat with flowing air. As they neglected the axial conduction and employed temperature difference between fluid and particles at the exit of the bed to calculate temperature driving force, their

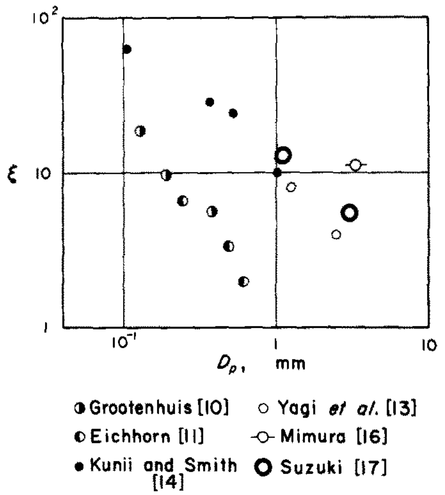


FIG. 6. Channelling length factor ξ against D_p

resultant Nu_p values may be somewhat higher than those defined by equation (9).

In actual packed beds, the existence of channelling may result in a different mechanism of axial mixing from the ordinary diffusion model, which should be investigated in the future.

CONCLUSION

The apparent anomaly of the previous experimental observations on heat or mass transfer between fluid and packed solids in the region of low Péclet number was interpreted by assuming a simple model taking channelling of fluid in the beds into consideration.

In the flow range of low Péclet number ($Pe_p < 10$), attainment of thermal or diffusional equilibrium between solids and flowing fluid within one-particle row is ascertained. The apparent particle-to-fluid heat- or mass-transfer coefficient is expressed in terms of heat capacity of fluid flow.

Channelling length factor was back estimated by assuming a simple equation, which increases with decreasing particle diameter.

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Résumé—En appliquant un modèle simple pour le transport de chaleur ou de masse entre les solides et le fluide en écoulement dans les lits fixes, les résultats expérimentaux présentés sous la forme de nombres de Nusselt et de Sherwood dans les travaux antérieurs sont interprétés théoriquement dans la gamme des faibles nombres de Peclet, c'est-à-dire $Pe_p < 10$. On suggère alors que des digitations ou des contacts non-uniformes localement entre le fluide et le solide sont responsables de la décroissance supplémentaire des coefficients apparents de transport de chaleur ou de masse dans le système ci-dessus.

Zusammenfassung—Durch Anwendung eines einfachen Modells auf den Wärme- und Stoffübergang zwischen Festkörpern und einem strömenden Fluid in Festbetten lassen sich Versuchswerte sowohl für Nusseltzahlen als auch für Sherwoodzahlen aus der kürzlich erschienenen Literatur im Bereich kleiner Pécletzahlen d.h. $Pe_p < 10$ theoretisch interpretieren. Es wird vermutet, dass Kanalbildungen oder örtlich ungleicher Kontakt mit dem Fluid für die Abnahme der scheinbaren Wärme- und Stoffübergangskoeffizienten verantwortlich sind.

Аннотация—В статье предложена простая модель тепло-и массообмена между твердыми частицами и потоком среды в плотном слое, с помощью которой обработаны известные в литературе данные о числах Нуссельта и Шервуда при низких числах Пекле ($Pe_p < 10$). Предполагается, что каналообразование или неравномерный контакт частицы со средой приводит к дальнейшему падению эффективных коэффициентов тепло-и массообмена в данных системах.