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SIMILARITY OF THE TRANSFER PROCESSES IN INHOMOGENEOUS FLUIDIZED BEDS

Yu. S. Teplitskii

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Based on Fr and Ar numbers, which are the generalized characteristics of discrete and continuous phases, an effective method of modeling the transfer processes in gas fluidized beds is developed within the framework of similarity theory.

The development of calculation methods for transfer processes in inhomogeneous fluidized beds (FB) suffers from the absence of a sufficiently rigorous theory for the fluidization state of a granular medium [1, 2]. In connection with this, different empirical and semiempirical methods are widely used in investigations. Of particular value are methods of similarity theory [3] that make it possible, on the one hand, to conduct experimental studies in a rational way and, on the other hand, to obtain sufficiently broad generalizations of available experimental data that may be used in engineering practice.

In the present work, a method of similarity theory is described based on the use of the Fr and Ar numbers, $\mathrm{Fr} = (u - u_{mf})^2/gH_{mf}$, $\mathrm{Ar} = gd^3(\rho_s/\rho_f - 1)/\nu_f^2$, and permitting the modeling of different transfer processes in an inhomogeneous FB.

As is known, the main distinctive feature of such a system is the availability of two phases, namely, discrete (gas bubbles) and emulsion (particles with a gas between them) phases. This specifies the character of the transfer processes in an inhomogeneous FB in many respects, with each phase making its contribution to total transfer in each particular case. It is a priori evident that the corresponding FB characteristic will be expressed as a function of the characteristics of the discrete and emulsion phases. The goal of similarity theory in this case is to find, on the basis of analysis of physical regularities, a system with the minimum quantity of dimensionless numbers characterizing both phases and allowing one to model the real transfer processes in inhomogeneous FBs.

Bubble Phase. High values of the transfer characteristics of a FB (heat-mass transfer and diffusion coefficients, thermal conductivity) are primarily caused by gas bubbles which play the role of a peculiar "pneumatic" mixer creating large-scale circulation loops of the granular material and the gas in the bed. A characteristic dimension of these loops is the bed height (H or H_{mf}). The basic equation describing the gas flow distribution in FB phases (an equation of the two-phase theory) is

$$u = mu_{mf}(1 - \varepsilon_b) + u_b\varepsilon_b + nu_{mf}\varepsilon_b\,, \tag{1}$$

where mu_{mf} is the gas velocity in the emulsion phase, and nu_{mf} is the velocity of the straight-through (relative to the bubble) gas flow. Introducing $G_b = u_b\varepsilon_b$, which is the density of the gas flow in the form of bubbles (a "visible" bubble flow), we arrive at

$$G_b = u - u_{mf}(m(1 - \varepsilon_b) + n\varepsilon_b)\,. \tag{2}$$

The gas distribution in the phases may be described, as experience shows [4, 5], by simplified variants of Eq. (2) with an accuracy acceptable for practice:

a) for fine particles ($d < 1$ mm) by

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$$m = n = 1\ , \quad G_b = u - u_{mf} \tag{3}$$

which is the equation of the "ideal" two-phase theory developed by Toomey and Johnstone [4],

b) for coarse particles ($d \geq 1$ mm) by

$$m = 1\ , \quad G_b = u - u_{mf}\,(1 + (n - 1)\,\varepsilon_b) \tag{4}$$

which is the equation first used in [5]. The quantity n, characterizing the straight-through gas flow through the bubble, changes from 3 to 8 [5] and is a rather complicated function of the excess gas velocity and bed height. Eliminating ε_b from (4) we write this equation as

$$G_b = (u - u_{mf})\, u_b/(u_b + (n - 1)\, u_{mf})\ . \tag{4'}$$

From formulas (3), (4') we draw the following important conclusion. The excess gas velocity $u-u_{mf}$ may be chosen as the velocity scale characteristic of a bubble phase. Thus, we have two scales typical for the bubble phase, namely, H_{mf} and $u-u_{mf}$. These quantities should be supplemented with D_a (which influences the coalescence of gas bubbles), g (bubble rise occurs in a gravity field), and the current coordinate h.

We may assume that any characteristic (B) of the bubble phase (bubble velocity, frequency, bed expansion, etc.) is a function of the above parameters:

$$B = f\,(H_{mf}\,,\ u - u_{mf}\,,\ D_a\,,\ g\,,\ h)\ . \tag{5}$$

Using the π-theorem of the dimensional theory [3], we write the dimensionless analog of (5) instead of the latter:

$$B' = f\,(\mathrm{Fr}\,,\ H_{mf}/D_a\,,\ h/H_{mf})\ , \tag{6}$$

where $\mathrm{Fr} = (u-u_{mf})^2/gH_{mf}$ is the Froude number*, characterizing the ratio of the kinetic energy of gas bubbles to the potential energy of their rise.

Emulsion Phase. One of the fundamental propositions of the two-phase theory of the FB, confirmed repeatedly in practice, reads that the state of the emulsion phase is close to the onset of fluidization. Consequently, in interparticle gaps the characteristic space and velocity scales are d and u_{mf}** $u_{mf} \geq 0.6$ m/sec.. Moreover, the governing parameters of the emulsion phase include c_f, c_s, ρ_f, ρ_s, ν_f, D_f, λ_f***. For instance, for characterizing the emulsion phase (the coefficients of interphase heat and mass transfer, of emulsion-to-surface heat transfer, and so on) we can write

$$E = \varphi\,(d\,,\ u_{mf}\,,\ c_f\,,\ c_s\,,\ \rho_f\,,\ \rho_s\,,\ \nu_f\,,\ D_f\,,\ \lambda_f)\ . \tag{7}$$

Applying the π-theorem of the dimensional theory to (7) yields

$$E' = \varphi\,(\mathrm{Re}_{mf}\,,\ \mathrm{Pr}\,,\ \mathrm{Sc}\,,\ c_s/c_f\,,\ \rho_s/\rho_f)\ . \tag{8}$$

Using the Todes formula [10]

$$\mathrm{Re}_{mf} = \mathrm{Ar}/(1400 + 5.22\,\sqrt{\mathrm{Ar}})\ . \tag{9}$$

we can write instead of (8):

$$E' = \varphi\,(\mathrm{Ar}\,,\ \mathrm{Pr}\,,\ \mathrm{Sc}\,,\ c_s/c_f\,,\ \rho_s/\rho_f)\ . \tag{10}$$

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* The Froude numbers Fr and $\mathrm{Fr}_h = (u - u_{mf})^2/gh$ were first used in [6-8].

** In [9], experiments with a two-dimensional bed (61 × 2 cm) of particles witch d= 1.52 mm have shown some expansion (by 7-8 %) of the emulsion phase and its subsequent decrease to almost zero at $u - u_{mf} \geq 0.6$ m/sec.

*** The influence of λ_s is negligible due to convection of particles.

Remembering that the Pr and Sc numbers for gases change in a rather narrow range, we can simplify Eq. (10) to the form

$$E' = \varphi\,(\mathrm{Ar}\,,\; c_s/c_f,\; \rho_s/\rho_f)\,. \tag{10'}$$

Noteworthy here is the radical difference between the linear scales in (6) and (10). In (6), the scale is the height of the entire granular bed, and in (10) it is the diameter of a single grain.

Since the FB consists of two phases, its transfer characteristic in the general case is determined by superposition of (6) and (10′):

$$(FB)' = \psi\,(\mathrm{Fr}\,,\; \mathrm{Ar}\,,\; H_{mf}/D_a\,,\; h/H_{mf},\; c_s/c_f,\; \rho_s/\rho_f)\,. \tag{11}$$

The number of the determining complexes and simplexes in some cases may be significantly smaller (see below).

We consider now some examples to illustrate the approach developed.

Gas Bubble Dimensions. We use the three most verified formulas. In order to avoid the effect of the input device, we consider the case of gas distribution in pores:

$$D_h = 0.42\,(u - u_{mf})^{2/5}\,h^{4/5}/g^{1/5} \quad [11]\,, \tag{12}$$

$$D_h = 0.78\,(u - u_{mf})^{1/2}\,h^{3/4}/g^{1/4} \quad [12]\,, \tag{13}$$

$$D_h = 1.3\,(u - u_{mf})^{2/3}\,h^{2/3}/g^{1/3} \quad [7]\,. \tag{14}$$

In dimensionless form, expressions (12) to (14) are as follows:

$$D_h/h = 0.42\;\mathrm{Fr}\,d^{1/5}\,(h/H_{mf})^{-1/5}\,, \tag{12'}$$

$$D_h/h = 0.78\;\mathrm{Fr}^{1/4}\,(h/H_{mf})^{-1/4}\,, \tag{13'}$$

$$D_h/h = 1.3\;\mathrm{Fr}^{1/3}\,(h/H_{mf})^{-1/3}\,. \tag{14'}$$

These formulas may be further simplified by using the local Fr_h number. Then, for instance, Eq. (14) acquires the form

$$D_h/h = 1.3\;\mathrm{Fr}_h^{1/3}\,. \tag{14''}$$

Figure 1 compares the experimental D_h values [5, 13-18, 20, 21] and those calculated by Eqs. (12′)-(14′). As is seen, formula (14′) is best for describing the experimental data in the entire Pr_h range.

Figure 2 shows the results of processing the data on gas bubble size measurements in the beds under pressure using the procedure proposed by A. I. Podberezskii and V. A. Rybchinskii. The generalized dependence turns out to be very close to the Powe formula (13′):

$$D_h/h = 0.8\;\mathrm{Fr}_h^{0.28}\,. \tag{15}$$

Bed Expansion. This phenomenon is associated entirely with the character of motion of gas bubbles in a bed and, consequently, we determine the bed expansion by a formula of type (6). Indeed, processing of numerous experimental data [19, 22] has resulted in the following simple dependences:

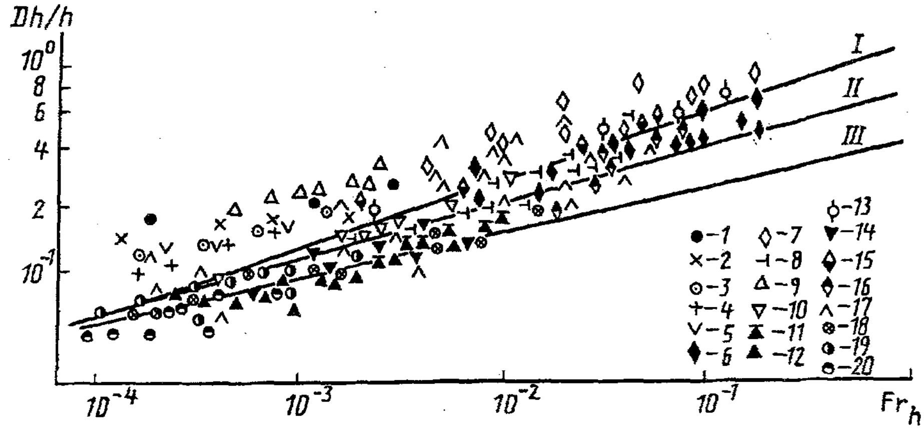

Fig. 1. Comparison of experimental D_h values calculated by different formulas: 1-5) [13]; 6,7) [14] ($D_a = 0.3$ m); 8) [14] ($D_a = 0.7$ m); 9) [15]; 10) [16]; 11) [17]; 12) [18]; 13) [14] (0.4 × 0.25 m column cross-section); 14) [18]; 15) [5]; 16) [20]; 17) [21]; 18 to 20) [18]; lines I, II, III calculation by Eqs. (14′), (13′), (12′), respectively.

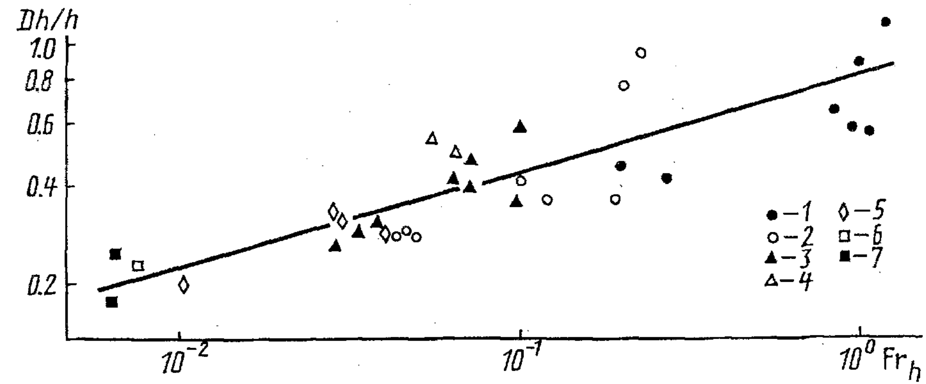

Fig. 2. Correlation of vertical size data for gas bubbles in millet bed under pressure: 1 to 7) $P = 0.1$; 0.6; 1.1; 1.6; 2.1; 2.6; Eq. (15); column cross-section 0.1 × 0.015 m.

for fine particles:

$$H/H_{mf} - 1 = 0.7\ \mathrm{Fr}^{1/3}\ (H_{mf}/D_a)^{1/2}\ , \tag{16}$$

for coarse particles:

$$H/H_{mf} - 1 = 0.54\ \mathrm{Fr}^{0.54}\ . \tag{17}$$

Solid Phase Mixing. As is known [10], the high values of thermal conductivity and diffusion in the FB are attributable, first of all, to powerful convective particle flows in the FB caused by gas bubbles passing through the FB. The mixing intensity of the particles depends on the both phases thus permiting use of an equation of type (11) to correlate the experimental data. In [23, 24], for calculation of the vertical K_z and horizontal $D_{x,y}$ diffusion coefficients of particles, the following relations are obtained:

$$K_z/(u - u_{mf})\ H_{mf} = 0.1\ \mathrm{Re}_{mf}^{-0.4}\ , \tag{18}$$

$$D_{x,y}/(u - u_{mf})\, H_{mf} = 0.013\, (D_a/H_{mf})^{1/2}\, \mathrm{Fr}^{-0.15}\,. \tag{19}$$

Relations (18) and (19) correlate the large number of experimental data, and their correctness has been confirmed many times (e.g., formula (19) is included in the handbook [25]).

Mass Transfer Coefficient between a Bubble and an Emulsion Phase. Below we give a typical example of using the developed approach for the rational generalization of experimental data.

In [26], for the surface coefficient of mass transfer between a single bubble and an emulsion phase (Ar > 500), the following expression is obtained:

$$\mathrm{Sh} = 6.7 \cdot 10^{-4}\, (\mathrm{Pe}_b\, \mathrm{Re}_b)^{1/2}\, \mathrm{Ar}^{1/2}\,. \tag{20}$$

Using the obvious relationship between the surface and volume mass transfer coefficients

$$\beta = 6\varepsilon_b\, k/\langle D_b\rangle\,, \tag{21}$$

based on formula (20), the relations of the "ideal" two-phase theory (3), and the preliminarily averaged dependence (14′), we arrive at

$$\beta H_{mf}/(u - u_{mf}) = k_0\, \mathrm{Fr}^{-1/3}\, \mathrm{Ar}^{0.5}\,. \tag{22}$$

Processing of the literature data [26-29] on β values by the least-square method enabled us to refine the dependence of this coefficient on Fr and Ar in (22) and resulted in the following simple formula:

$$\beta H_{mf}/(u - u_{mf}) = 0.21\, \mathrm{Fr}^{-0.13}\, \mathrm{Ar}^{0.14}\,, \tag{23}$$

that describes the experimental data with a root-mean square deviation of 18% (Fig. 3).

FB-to-Surface Heat Transfer. The most physically well-based heat-transfer model, accounting for two phases of the system, is the packet model [30] in accordance with which

$$\alpha_{cc} = (1 - \varepsilon_b)\, \alpha_e + \alpha_{\mathrm{conv}}\,. \tag{24}$$

Due to different reasons (of which the main one is the entrance of the bubble frequency, a priori unknown, into the expression for α_c), the packet model has not been sufficiently developed to obtain calculation relations. A more effective and simple model is the one-phase model treating the FB core as an inhomogeneous infiltrated granular medium with effective porosity

$$\varepsilon = 1 - (1 - \varepsilon_b)\,(1 - \varepsilon_{mf})\,. \tag{25}$$

The coefficient α_{cc}, within the framework of this model, is given as the sum of the conductive and convective components:

$$\alpha_{cc} = \alpha_{\mathrm{cond}} + \alpha_{\mathrm{conv}}\,, \tag{26}$$

By analogy with the heat transfer for a pure gas, α_{conv} is determined in terms of Re and Rr, and α_{cond} in terms of the mean particle concentration $1-\varepsilon$, Ar number, and simplexes c_s/c_f, ρ_s/ρ_f.

$$\mathrm{Nu}_{\mathrm{cond}} = \theta\,(1 - \varepsilon\,,\ \mathrm{Ar}\,,\ c_s/c_f\,,\ \rho_s/\rho_f)\,. \tag{27}$$

One of the recent and most universal formulas for Nu_{cc} is [31]

$$\mathrm{Nu}_{cc} = 0.85\,(1 - \varepsilon)^{2/3}\, \mathrm{Ar}^{0.1}\, (c_s/c_f)^{0.24}\, (\rho_s/\rho_f)^{0.14} + 0.046\, \mathrm{Re}\, \mathrm{Pr}\, (1 - \varepsilon)^{2/3}/\varepsilon\,. \tag{28}$$

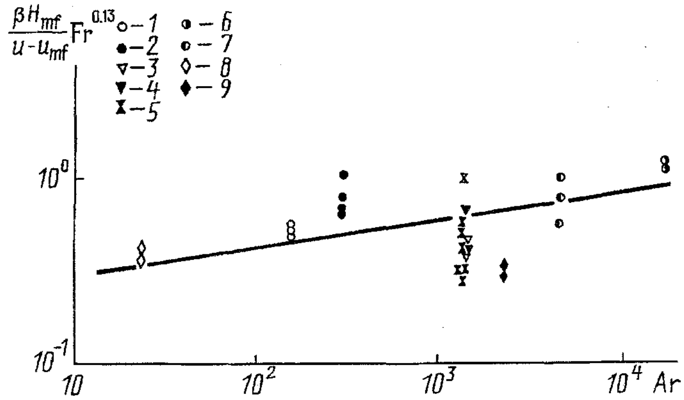

Fig. 3. Correlation of experimental data on mass transfer between discrete and emulsion phases: 1) [29]; 2 to 4) [28]; 5) [26]; 6 to 9) [27]; solid line, by Eq. (23).

With account of the equality $\varepsilon_b = 1 - H_{mf}/H$ and formulas (16), (17), and (25), we can easily introduce the Fr number and simplex H_{mf}/D_a into the expression for Nu_{cc}.

Heat Transfer Coefficient in the FB Freeboard Region Heat transfer in this region is determined [10] by the concentration of particles ejected from the bed by gas bubbles. The velocity of the latter determines the speed of the escaping particles and determines their concentration distribution over the height of the freeboard region [10]:

$$\rho = \rho_{su} \exp\left(-2g(h - H_{mf})/w_{su}^2\right). \tag{29}$$

Assuming $w_{su} \approx (v_b)_{su} = 0.711 \cdot \sqrt{g(D_b)_{su}}$, G. I. Pal'chenok and A. Hassan [32] have established, on the basis of (29), a simple relationship between ρ and $(D_b)_{su}$:

$$\rho = \rho_{su}\left(-2g(h - H_{mf})/(D_b)_{su}\right). \tag{30}$$

Using (12) and (14), the authors of [32] have derived simple relations for the determination of ρ:

for fine particles:

$$\rho/\rho_{su} = \exp\left(-1.2\,(h - H_{mf})\,\mathrm{Fr}^{-1/3}/H_{mf}\right), \tag{31}$$

for coarse particles:

$$\rho/\rho_{su} = \exp\left(-2.4\,(h - H_{mf})\,\mathrm{Fr}^{-1/5}/H_{mf}\right). \tag{32}$$

Assuming, as in the FB [(see (28)], that $(\alpha_{\mathrm{cond}})_{fb} \approx (1-\varepsilon)^{2/3} \approx \rho^{2/3}$, and using (31) and (32), the authors of [33] have obtained the following expressions for calculating the heat transfer characteristics in the freeboard region:

for fine particles:

$$A = \frac{(\alpha_{cc})_{fb} - (\alpha_{\mathrm{conv}})_{fb}}{\alpha_{cc} - (\alpha_{\mathrm{conv}})_{fb}} = \exp\left(-0.8\,(h - H_{mf})\,\mathrm{Fr}^{-1/3}/H_{mf}\right), \tag{33}$$

for coarse particles:

$$A = \exp\left(-1.6\,(h - H_{mf})\,\mathrm{Fr}^{-1/5}/H_{mf}\right). \tag{34}$$

Finally, on the basis of the simple phenomenological model in [34], we derive a formula not containing α_{cc}:

$$(\mathrm{Nu}_{cc})_{fb} - (\mathrm{Nu}_{\mathrm{conv}})_{fb} = 0.29\ \mathrm{Fr}^{0.66((h-H_{mf})/H_{mf})^{1.32}} \times \mathrm{Ar}^{0.27} \exp(-1.4\,(h - H_{mf})/H_{mf})\,. \tag{35}$$

Relations (12) to (19), (23), (28), and (31) to (35), whose series may be extended (see, e.g., [35]), confirm convincingly the concept of using the Fr and Ar numbers to generalize a wide spectrum of experimental data on the FB and its freeboard region. The Fr and Ar numbers represent, in essence, the generalized characteristics of discrete and continuous phases of the FB. Their simultaneous use makes it possible, in our opinion, to correctly and rationally account for one of the main FB features, namely, the existence of two phases. The formulas obtained on the basis of the Fr and Ar numbers are distinguished by their physical sense and simplicity, are verifiable within wide limits of experimental conditions, and are convenient for use in engineering practice.

In conclusion, we note that the ideas described may be easily extended to retarded FBs in which the presence of a fixed packing imposes additional restrictions on the motion of the gas bubbles. This gives rise, correspondingly, to the additional quantities characterizing the packing (ε_p, D_b, S_v, S_h, etc.) that appear in (5) [36, 37].

NOTATION

Ar $= gd^3(\rho_s/\rho_f - 1)/\nu_f^2$, Archimedes number; c, specific heat; D_f, gas diffusion coefficient; D_a, apparatus diameter; D_t, tube diameter; D_b, bubble diameter; $D_h = 0.7D_b$, vertical size of a bubble; d, particle diameter; $\mathrm{Fr}_h = (u-u_{mf})^2/gh$, local Froude number; $\mathrm{Fr} = (u-u_{mf})^2/gH_{mf}$, Froude number; g, free fall acceleration; H_{mf}, H, bed height at filtration rates u_{mf} and u, respectively; h, height above the gas distributor; k, mass transfer coefficient based on unit surface of a bubble; $\mathrm{Nu} = \alpha d/\lambda_f$, Nusselt number; p, pressure; $\mathrm{Pe}_b = v_b D_b/2D_f$, Peclet number; $\mathrm{Pr} = c_f\eta_f/\lambda_f$, Prandtl number; $\mathrm{Re} = ud/\nu_f$; $\mathrm{Re}_b = v_b D_b/2\nu_f$, $\mathrm{Re}_{mf} = u_{mf}d/\nu_f$, Reynolds numbers; $\mathrm{Sc} = \nu_f/D_f$, Schmidt number; $\mathrm{Sh} = kD_b/2D_f$, Sherwood number; S_v, S_h, vertical and horizontal pitches of the tubes; u_{mf}, u, rate at the onset of fluidization and the filtration rate, respectively; u_b, absolute velocity of a bubble; v_b, relative velocity of a bubble ($u_b = u - u_{mf} + v_b$); w_{su}, root-mean-square rate of ejection of particles out of the bed; α, heat transfer coefficient; β, discrete-to-emulsion phase mass transfer coefficient based on unit volume of the bed; ε_{mf}, bed porosity at u_{mf}; ε_b, bubble concentration; ε, mean porosity of the bed determined by (25); ε_p, packing porosity; η_f, dynamic gas viscosity; ν_f, kinematic gas viscosity; $\rho = \rho_s(1-\varepsilon)$, bed density; λ, thermal conductivity. Indices: b, bubble; cond, conductive; conv, convective; cc, conductive-convective; f, gas; e, emulsion of particles; su, fluidized bed surface; fb, freeboard region; s, particles; $\langle\ \rangle$, averaging over the entire bed.

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