The effect of fine particle influence on numerical simulation of bidisperse fluidized bed

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Abstract. In this paper, we consider methods for choosing the effective diameter of solid particles for fluidized bed modeling using the Euler-Euler approach to describing a multiphase medium. Calculations are carried out for the modal average diameter $D_{mod}$ and average diameter of Sauter $D_{32}$. Studies have also been conducted for the bidisperse composition of particles, with separation into large and small fractions. The results of numerical simulation were compared with experimental data.

1. Introduction

Fluidized bed (gas-solid particles) devices are widely used in the petrochemical and oil refining industry and the gasification of coal and biomass [1-3]. Reactors are widespread due to the high efficiency of heat and mass transfer between components. Solid particles are a mixture of approximately spherical granules of different diameters, thus, constitutes a polydisperse discrete phase.

The particle size distribution plays an important role in the behavior of the fluidized bed. In some cases, the addition of small particles leads to the expansion of the layer and an increase in the porosity of the dense phase [4, 5]. However, in other cases, the addition of these particles leads to a decrease in solid-phase porosity [6, 7]. The authors [8] showed that the addition of small particles just slightly reduces the effective viscosity. Experimental results showing a decrease in bubble size using fast X-ray tomography were obtained in [9]. The effect of fine particles on fluidization was investigated in [10] by numerical simulation. It was shown that the addition of a large number of small particles suppresses the bubbles, and the device can achieve a stable quasistationary mode of operation.

One of the possible ways to obtain data is numerical simulation. Numerical calculations of the fluidization process are usually based on the Euler-Lagrange and Euler-Euler approaches. The solid phase is considered as discrete particles in the Euler-Lagrange approach (for example, [11, 12]). In this approximation, each particle is considered separately. This approach requires significant computational resources and laborious for industrial devices that work with hundreds of tons of particles. In the Eulerian-Eulerian approach, the carrier (gas, liquid) and the discrete (solid) phase are considered continuous. To take into account the interaction of particles in the fluidized bed, by analogy with the kinetic theory of gases, an equation was added describing the change in the kinetic energy of the particles due to their collisions (for example, [13]). This approach is less demanding in terms of computational resources and requires discretization of particle characteristics. Discretization can be carried out using a multi-group approach, the essence of which is to divide the domain of definition into groups of diameters and assign them to correspond some value of the cumulative distribution function that characterizes the fraction of the group in the particle layer, and the transport equations are solved for...
each group separately. A multi-group approach to simulate a fluidized bed was used in [14, 15]. In [16], it was shown that the method of taking into account polydispersity can affect the hydrodynamic characteristics of the flow obtained by numerical simulation.

Thus, a universal method for the discretization of the characteristics of polydisperse solids is required. The method should provide the optimal choice of the smallest number of particle fractions for the numerical calculation and reflect the effects associated with polydispersity.

2. Problem formulation

2.1. Effective diameter model

Let us consider several ways of choosing the number and size of discrete phases for numerical simulation using the Eulerian-Eulerian approach. The calculation results are verified with experimental data obtained at the laboratory facility. The experimental setup is a glass tube with an internal diameter of 2.2 cm. A porous material is at the bottom of the tube to ensure a uniform airflow. The process of fluidization was controlled on a high-speed camera, the bed height was estimated by frames.

In this paper, we consider discretization methods based on the choice of the effective diameter and the particle terminal velocity. The method essence is to obtain two effective diameters, the first for coarse particles forming a dense phase, and the second for fine particles carried away by the flow. As effective diameters, we consider two models: $D_{s}$ is Sauter mean diameter, $D_{mod}$ is modal diameter.

$$D_{s} = \frac{\int_{0}^{\infty} f(D)D^{3}dD}{\int_{0}^{\infty} f(D)D^{2}dD}, \quad D_{mod} = \arg \max_{D} \left(f(D)\right).$$

(1)

Terminal velocity

$$u_{t} = \left(\frac{4gD(\rho_{p} - \rho_{f})}{3\rho_{f}C_{D}}\right)^{0.5},$$

(2)

$$C_{D} = \frac{24}{Re_{f}} + \frac{6}{\sqrt{Re_{p}}} + 0.4, \quad Re_{f} = \frac{\rho_{f}u_{t} - u_{g}}{\mu_{f}} < 200000.$$

Equating the known gas velocity $u_{g}$ to the terminal velocity (1), it is necessary to obtain the diameter, relative to which the initial mixture of particles can be divided into the two groups described above. $D_{s}$ is the root of equation (3).

$$u_{g} = \left(\frac{4gD(\rho_{p} - \rho_{f})}{3\rho_{f}C_{D}}\right)^{0.5}.$$  

(3)

When under the action of the forces of gravity and resistance the particle of diameter $D_{s}$ is in equilibrium ($u_{g} = 0$), the particles with a diameter less than $D_{s}$ will be carried away by the stream, and particles with a large diameter tend down. This fact allows us to determine the effective diameters (4).

$$D_{s1} = \arg \max_{D} \left(f(D)\right), \quad D_{s2} = \frac{\int_{0}^{D_{s1}} f(D)D^{3}dD}{\int_{0}^{D_{s1}} f(D)D^{2}dD},$$

$$D_{mod1} = \frac{\int_{D_{s2}}^{\infty} f(D)D^{3}dD}{\int_{D_{s2}}^{\infty} f(D)D^{2}dD}, \quad D_{mod2} = \frac{\int_{D_{s1}}^{D_{s2}} f(D)D^{3}dD}{\int_{D_{s1}}^{D_{s2}} f(D)D^{2}dD}.$$

(4)

The effective diameters and volume fractions of the experimental data are presented in Table 2.
Table 1. The effective diameters and initial volume fractions.

| Velocity, m/s | Fine particle diameter, µm | Coarse particle diameter, µm | Fine particle volume fraction | Coarse particle volume fraction |
|--------------|---------------------------|-------------------------------|------------------------------|------------------------------|
|              | $D^{i}_{on}$              | $D^{i}_{off}$                 | $D^{j}_{on}$                 | $D^{j}_{off}$                 | $\alpha^{i}_{on}$ | $\alpha^{i}_{off}$ | $\alpha^{j}_{on}$ | $\alpha^{j}_{off}$ |
| 0.0716       | 30.5                      | 32.4                         | 66.9                        | 56.4                         | 0.005               | 0.411               | 0.014               | 0.402               |
| 0.0892       | 33.4                      | 35.6                         | 67.2                        | 56.4                         | 0.014               | 0.402               | 0.014               | 0.402               |
| 0.1088       | 36.4                      | 39.6                         | 67.7                        | 56.4                         | 0.033               | 0.383               | 0.033               | 0.383               |
| 0.1213       | 38.2                      | 42.0                         | 68.3                        | 56.4                         | 0.052               | 0.364               | 0.052               | 0.364               |

All tests were carried out for 30 g of particles, the initial volume fraction of solid particles in a fixed bed was $\alpha_{0} = 0.41639$.

2.2. Fluidized bed model

In the numerical simulation of the fluidized bed, a continuous Eulerian-Eulerian multiphase model was used, supplemented by the kinetic theory of gases to take into account collisions of solid particles. For each phase, the laws of conservation of mass, momentum are satisfied. In the study of this problem, the following equations were solved:

Conservation of mass

$$\frac{\partial \alpha_i \rho_i}{\partial t} + \nabla \cdot (\alpha_i \rho_i \vec{v}_i) = 0,$$

where $\alpha_i$ is volume fraction for $i$-th phase, $\rho_i$ is density, $\vec{v}_i$ is velocity.

Conservation of momentum

$$\frac{\partial \alpha_i \rho_i \vec{v}_i}{\partial t} + \nabla \cdot (\alpha_i \rho_i \vec{v}_i \vec{v}_i^T) = \alpha_i \rho_i \vec{v}_i \vec{v}_i + \alpha_i \rho_i \vec{g} + \sum_j \vec{R}_ij,$$

where $\vec{F}_i = \alpha_i \mu_i (\nabla \vec{v}_i + \nabla \vec{v}_i^T) + \alpha_i (\lambda_i - (2/3) \mu_i) \nabla \cdot \vec{v}_i \vec{I}$ is stress tensor, $\vec{v}_i$ is velocity, $\lambda_i, \mu_i$ are bulk and shear viscosities, $\vec{I}$ is unit tensor, $\vec{R}_ij = K_p \left( \vec{v}_i - \vec{v}_j \right)$ is interfacial interaction force.

The equation for the solid phase granule temperature is represented as

$$\frac{3}{2} \frac{\partial}{\partial t} \left[ \alpha_i \rho_i \Theta_i \right] + \nabla \cdot \left[ \alpha_i \rho_i \vec{v}_i \Theta_i \right] = \left( -p_i \vec{F}_i + \vec{F}_i \right) : \nabla \vec{v}_i + \nabla \cdot \left[ k_{\Theta_i} \nabla \Theta_i \right] - \gamma_{\Theta_i} + \phi_{\Theta_i},$$

where $k_{\Theta_i}$ is granule energy diffusion coefficient, $\Theta_i$ is solid phase granule temperature, $\gamma_{\Theta_i}$ is energy dissipation from particle collisions, $\phi_{\Theta_i}$ is energy exchange between solid and gas phases.

To close the equation system, semi-imperial models of interphase interaction were used. For the liquid (gaseous) phase ($i$-th) and solid granular phase ($j$-th), the interaction model was taken from [13]; in the case of two solid granular phases, the model [18] was used. The energy dissipation due to particle collisions was described by the model obtained in [19].

3. Results

The results of numerical simulation using the methods described for selecting an effective diameter for bidisperse particles were compared with experimental data, as well as with the results of calculations performed with one effective diameter for monodisperse particles. The distribution of the volume fraction of the catalyst at the height of the tube, obtained in the calculations, is shown in Figure 1.

It can be seen that in all calculations the height of the dense layer of particles is within the limits of the observed minima and maxima. From Figure 1 a and b, it can be seen that at low speeds the addition of fines practically does not affect the average heights of the dense layer obtained in the calculations, this is because the volume fraction of particles with a diameter less than $D^{i}_{on}$ is small.

The time-averaged layer heights obtained experimentally and numerically are shown in Figure 2. The average heights of the dense layer, obtained in calculations with two effective diameters $D^{i}_{on}$ and
$D^2_{32}$ (for small and coarse fractions, respectively), showed better agreement with experiment than the heights obtained from calculations with one effective diameter $D_{32}$. In the case when the effective diameters were chosen as a mode, it turned out that the heights obtained in the calculations with two effective diameters $D_{mod}^{1}$ and $D_{mod}^{2}$ turned out to be greater than the experimental average, moreover, the heights in the calculations with one effective diameter $D_{mod}$.

![Figure 1. The distribution of the volume fraction of particles at the height of the tube for four values of flow velocity.](image)

Figure 1. The distribution of the volume fraction of particles at the height of the tube for four values of flow velocity.
Figure 2. Time-averaged layer height as a function of flow velocity.

Based on this fact, for polydisperse particles and with a similar distribution function, we can recommend the following conditions for choosing effective diameters

\[ D^1_{\text{eff}} \leq D_1 \leq D^2_{\text{eff}}, \quad D^2_{\text{eff}} \leq D_2 \leq D^3_{\text{eff}} \]  

(8)

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