USE OF MIXING APPARATUS FOR HIGHLY CONCENTRATED SUSPENSIONS

V. M. Barabash, L. N. Braginskii, and E. G. Kozlova* UDC 66.063.8

A random mechanism has been suggested for the rising of solid particles from the bottom of an apparatus when mixing suspensions. The equations derived reflect the effect of the concentration of solid particles on the hydraulic resistance of the apparatus body and the mixing intensity on the degree of uniform distribution of the solid particles in the mixed body. The results of theoretical analysis were compared with the test results in a wide range of solid phase concentrations.

Agitators are most extensively used for mixing suspensions of various cencentrations, the apparatus capacity going up to several thousands of cubic meters.

Mixing in such apparatus requires the use of expensive metal-bearing and power-consuming, often unique, drives with high torques or multiple drive systems. Improving the techno-economic indexes of such apparatus by improving their design accuracy and reliability is therefore important.

Methods of designing and selecting the apparatus and mixing devices [1, 2] are based on approximate theoretical calculations [3], ignoring any significant influence of the solid particles on flow characteristics in two-phase systems. Such methods are extended to suspensions with low solids (up to 10% by vol.). Tests were made on the effect of solids concentration on the hydrodynamic flow characteristics, hydraulic resistance coefficient, and also on the rise of particles from the apparatus bottom.

Data on suspension flow in tubes and channels [4, 5] point out that the solid particles present in the flow influence the hydraulic resistance. To verify this in apparatus with agitators, measurements were made of the peripheral velocity and torque profiles when mixing suspensions with solid phase up to 40% by vol. The tests were made in cylindrical apparatus of diameter 0.2 to 0.8 m without any fixed internal components, with the height of the working medium in the apparatus close to its diameter. Mixing was done by standard uncovered turbines and two- and three-blade agitators. Water was the continuous phase while sand of grain size $2.5 \cdot 10^{-4}$ m and glass globules of diameter $8 \cdot 10^{-4}$ m represented the discontinuous phase.

In apparatus with agitators without baffle plates, most commonly used for mixing suspensions, hydraulic resistance coefficient of the body c_f is determined assuming equilibrium between torque M_t applied to the agitator and the force moment of hydraulic resistance on the wall and bottom of the apparatus:

$$c_f = \frac{4M_t}{\pi \gamma \rho_s \omega_0^2 r_0^5 \bar{R}^3 \bar{v}_m^{\,2}},$$

where ρ_s is the suspension density, kg/m³; ω_0) angular rotation velocity of the agitator, rad/sec; r_0) agitator radius, m; $\gamma = 4\bar{H}/R + 1$; H) height of filling of the apparatus, m; R) radius of the apparatus, m; $\bar{R} = R/r_0$; $\bar{v}_m = v_m/(\omega_0 r_0)$; and v_m) mean peripheral flow velocity, m/sec. This equation has been used and v_m determined from the results of measuring the peripheral velocity profile:

$$v_m = \frac{1}{R}\int_0^R v(r)\,dr.$$

*Deceased.

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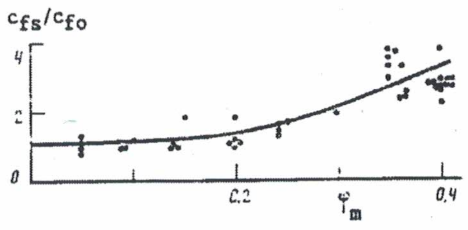

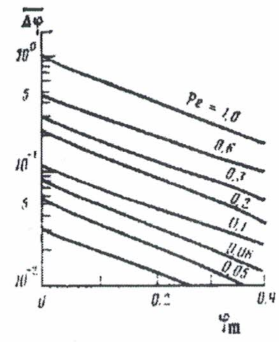

Fig. 1 Fig. 2

Fig. 1. Dependence of the apparatus body resistance coefficient on the volume concentration of the solid particles in the suspension.

Fig. 2. Dependence of the relative degree of heterogeneity of solid phase distribution along the height of the apparatus on the mean volume concentration and Peclet number.

Figure 1 gives the test results as a dependence of the resistance coefficient to the suspension flow c_{fs} on volume concentration φ. The vertical axis represents the values of c_{fs}/c_{f0}, where c_{f0} is the resistance coefficient in the absence of solid particles in the flow ($\varphi = 0$). The c_{f0} values obtained in this work agree with the values calculated using the equation of [2]

$$c_{f0}=0.095\mathrm{Re}^{-0.25},$$

where $\mathrm{Re} = \rho v_m R/\mu$ is the Reynolds number; and ρ and μ) density, kg/m³, and dynamic viscosity of the continuous phase, Pa·sec, respectively.

The results of measurements also showed that the solid particles in the test concentration ranges do not significantly influence the radial velocity distribution pattern. The method of [1, 2] can thus be used to calculate the peripheral velocity profile in the case of homogeneous media with allowance for the varation of the body resistance coefficient c_f.

Information on the effect of solid phase concentration on the hydrodynamic structure of the suspension flow helps analyze the rising of particles from the bottom and their distribution in the effective volume when mixing highly concentrated suspensions based on the local flow characteristics.

When studying the rising of solid particles from the bottom, the mechanism assumed that their vertical transport in the bottom zone was caused by pressure pulsations generated by the impulses of longitudinal component of velocity $v_{0\ell}'$. The analysis was made for the viscous sublayer zone (close to the bottom) since the particle transport in this zone is considerably less intense than in the main apparatus due to the attenuation of turbulence.

When there is no cohesion between the particles at the bottom, from the condition of gravity equilibrium per unit surface

$$F_g = d_p(\rho_p-\rho)g\varphi_b$$

and the lifting force per unit surface

$$F_\ell=\rho(v_{cr}')^2/2$$

we derive the equation for the critical value of pulsating velocity v_{cr}' ensuring the rising of the solid particles:

$$v_{cr}'=\left[\frac{2(\rho_p-\rho)}{\rho}gd_p\varphi_b\right]^{1/2}.$$

where d_p is the particle diameter, m; ρ_p) particle density, kg/m³; g) gravity acceleration, m/sec²; and φ_b) maximum volume concentration of solid phase at the apparatus bottom with the densely packed particles.

Let us write the equation for the balance of particle flow transported from the bottom in a unit time through a unit surface by turbulent pulsations of frequency ω_{cr} and settling to the bottom by gravity:

$$\varphi_b d_p \omega_{cr} = W_{sr} \varphi_{\delta_0},$$

where W_{sr} is the settling rate of the solid particles (allowing for obstructions), m/sec; and φ_{δ_0}) vol. concentration of solid particles at distance δ_0 from the sediment.

ω_{cr} may be expressed in terms of frequency corresponding to the pulsating velocity v_{op}':

$$\omega_{op} = v_{op}'/\delta_0.$$

and some function characterizing the proportion of the pulsations of the velocity scale of the viscous sublayer δ_0, whose amplitude is greater than or equal to v_{cr}':

$$\omega_{cr} = \omega_{op}\varphi(v_{cr}'/v_{op}'). \tag{1}$$

By assuming the normal law of distribution [7] of the pulsating velocities of a given scale, and also considering that the following condition is relevant at the boundary of the viscous sublayer according to [6]:

$$v_0'\delta_0/\nu_s^{\delta_0} \simeq 11.5, \tag{2}$$

Eq. (1) is rewritten as

$$\omega_{cr} = \frac{v_0' v_{op}'}{11.5\nu_s^{\delta_0}}[1 - 2\Phi(v_{cr}'/v_{op}')], \tag{3}$$

where ν_s is the suspension viscosity, m²/sec; and Φ) Laplace's function.

Taking into consideration the relation between the longitudinal and transverse components of pulsating velocity in the boundary layer [8]

$$v_{op}' = \beta v_0'.$$

where β is a coefficient describing the degree of homogeneity of flow turbulence, the value of the normal component of pulsating velocity ensuring the rising of solid particles from the bottom can be expressed from Eq. (3) as

$$v_0' = \left[\frac{(11.5/\beta)\,\omega_{cr}\nu_s^{\delta_0}}{1 - 2\Phi(v_{cr}'/(\beta v_0'))}\right]^{1/2}. \tag{4}$$

For changing over from the pulsating velocities to the averaged flow velocities, as a first approximation, the following equation may be assumed to apply at the boundary of the viscous sublayer:

$$v_0' = (\varepsilon_0^{\ell}\delta_0)^{1/3}, \tag{5}$$

where ε_0^{ℓ} is the local value of energy dissipation at the boundary of the viscous sublayer, W/kg.

By using Eq. (2), Eq. (5) may be rewritten as:

$$\varepsilon_0^{\ell} = 0.085(v_0')^4/\nu_s^{\delta_0}. \tag{6}$$

On the other hand, ε_0^{ℓ} may be determined in terms of the averaged flow characteristics:

$$\varepsilon_0^{\ell} = \nu_f(dv/dy)^2.$$

Having determined dv/dy by using the known changes of the averaged flow velocities near the solid surface [9]:

$$v/v^* = 8.71(v^*\delta_0/\nu_s^{\delta_0})^{1/7},$$

where $v^* = \sqrt{\tau/\rho_s^{\delta_0}}$ is the dynamic velocity; and $\rho_s^{\delta_0}$) suspension density at the viscous sublayer level, kg/m³, we find

$$\varepsilon_0^{\ell}=0.271\tau^2/((\rho_s^{\delta_0})^2 \nu_s^{\delta_0}). \tag{7}$$

The tangential stress τ in this equation is determined [2] as

$$\tau(r)=0.144\rho_s^{\delta_0}(\varepsilon_0^{\ell}\nu_s^{\delta_0})^{0.25}v(r), \tag{8}$$

where v(r) is the suspension flow rate at radius r, m/sec.

By balancing Eqs. (6) and (7) and also using Eqs. (4) and (8), we derive the equation for suspension flow rate ensuring the absence of sedimentation at radius r:

$$v(r)\geqslant 7.3v_0'. \tag{9}$$

The absence of the sedimentation of solid particles at the bottom is determined as:

$$v(R)\geqslant 7.3v_0',$$

where v(R) is the velocity of the peripheral flow of the suspension at radius R which is determined by using the equations of [2] for apparatus with a central agitator without any internal devices.

When using Eqs. (4) and (9), information is necessary about $\nu_s^{\delta_0}$. This in turn calls for a solution to the problem of vertical distribution of the suspended solid particles relative to the mixing conditions, design features and geometric characteristics of the apparatus and the agitator, and also the properties of the working media. Let us examine Eq. (5) characterizing the diffusion transport of the phases through some horizontal surface of the apparatus with an agitator working in a periodic regime:

$$-W_p\varphi-D_p\frac{d\varphi}{dh}=0; \tag{10}$$

$$-W_{med}(1-\varphi)+D_{med}\frac{d\varphi}{dh}=0, \tag{11}$$

where W_p and W_{med} are the flow velocities of the particles and the medium relative to the apparatus wall, m/sec; and D_p and D_{med} the turbulent transport coefficients of the particles and the liquid medium, m²/sec.

By deriving from Eqs. (10) and (11) the velocities of the particles and the liquid medium assuming that the difference of these velocities is no different from the sedimentation rate, and assuming [2] as a first approximation that

$$D_p=D_{med}=D_t,$$

we find

$$W_{sr}=\frac{-D_t}{\varphi(1-\varphi)}\frac{d\varphi}{dh}. \tag{12}$$

When the volume concentration of the solid phase φ exceeds 5%, the sedimentation rate, strictly speaking, should be determined with allowance for the constraint of particle movement. For various materials, the sedimentation rate under these conditions according to [8] is:

$$W_{sr}=W_{sr}'(1-\varphi)^n. \tag{13}$$

The value of n depends on the physical and mechanical characteristics of the solid particles and the flow regimes. The values given in the literature are n = 2-7. For spherical particles in a turbulent suspension flow regime, n ≈ 2.5 has been recommended [8].

The free sedimentation rate of the particles at Re ≤ 1 is determined using Stokes' law [8]:

$$W_{sr}'=d_p^2 g(\rho_p-\rho)/(18\mu),$$

at Re > 1

$$W_{sr}' \simeq 1.15[d_p g(\rho_p-\rho)/\rho]^{1/2},$$

where $Re = ud_p/\nu$; u is the particle flow rate relative to the liquid, m/sec; and ν is the kinematic viscosity of the continuous medium, m²/sec.

By comparing Eqs. (12) and (13) at n = 2.5, we find

$$\frac{W_{sr}'dh}{D_t} = -\frac{d\varphi}{\varphi(1-\varphi)^{3.5}}. \qquad (14)$$

The integration of Eq. (14) with allowance for the boundary condition

$$\varphi(h)=\varphi_0 \quad \text{at} \quad h=0$$

gives equations for calculating the concentration distribution of solid particles along the height of the apparatus at given mean concentration φ_m:

$$\frac{W_{sr}'h}{D_t} + \ln\left|\frac{1+\sqrt{1-\varphi_0}}{1-\sqrt{1-\varphi_0}}\right| - \frac{2}{\sqrt{1-\varphi_0}}\left[1+\frac{1}{3(1-\varphi_0)}+\frac{1}{5(1-\varphi_0)^2}\right] =$$

$$= \ln\left|\frac{1+\sqrt{1-\varphi}}{1-\sqrt{1-\varphi}}\right| - \frac{2}{\sqrt{1-\varphi}}\left[1+\frac{1}{3(1-\varphi)}+\frac{1}{5(1-\varphi)^2}\right]; \qquad (15)$$

$$\frac{1}{H}\int_0^H \varphi(h)\,dh = \varphi_m.$$

To determine the relative degree of inhomogeneity of the distribution of the solid phase along the height of the apparatus, the following value is usually considered:

$$\overline{\Delta\varphi} = |\varphi(0)-\varphi(H)|/\varphi_m.$$

By calculating the coefficient of turbulent diffusion D_t according to [2] and also the concentration of solid particles at the bottom $\varphi(0)$ and in the near-surface zone $\varphi(H)$ by resolving Eq. (15), $\overline{\Delta\varphi}$ can be theoretically determined relative to the properties of the working media and the mixing conditions (Fig. 2).

Finding the degree of inhomogeneity $\overline{\Delta\varphi}$ helps find the volume concentration of the particles close to the bottom of the apparatus and hence the mixing conditions ensuring the absence of sedimentation.

A comparison of the theoretical and experimental results of peripheral velocities obtained in 20- to 250-liter capacity apparatus for mixing water–sand suspensions using various agitators of diameter 0.1 to 0.36 m established that the only unknown parameter of the model, i.e., coefficient β, can be adopted as 1.2.

Thus, the simultaneous resolution of Eqs. (4) and (9) helps calculate and select the required mixing intensity to prevent sedimentation.

The results of calculation by the proposed method show that, for mixing suspensions with concentration of 40% by vol., the power of the agitator drives in apparatus of 500-600 m³ capacity can be reduced by 2-5 times and the values of specific power from 0.6-0.8 to 0.1-0.2 kW/m³. The method developed has been included in the new edition of the guide [10] and is being used for designing agitators for large reservoirs.

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