

# EFFECT OF VISCOSITY ON DROP DISPERSION IN MIXING EQUIPMENT

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A mathematical model is presented for breaking up drops during mechanical agitation in the turbulent regime in the absence of coalescence. The effect of the viscosity of the agitated phases is considered. An expression is obtained for calculating the diameter of the drops. The calculated drop diameters are compared with experimental values.

Processes which require agitating mutually insoluble liquids, in particular for extracting liquid or preparing stable emulsions, are characterized by a relatively wide change in the viscosity of the phases. Even in early experiments on agitating liquid-liquid systems, one of the questions studied was the effect of viscosity on the dimensions of the drops which were formed. Experimental results were very contradictory. Thus, for agitation in a turbulent mixture [1], an equation was obtained

$$a \sim (\mu_d / \mu_c)^{0.2},$$

from which it follows that a higher viscosity of the dispersed phase leads to a lower drop dimension, while a higher viscosity of the continuous phase to a higher drop dimension. A similar behavior of the continuous-phase viscosity was observed in [2], while in [3], a reverse relationship was obtained experimentally:

$$a \sim \mu_c^{0.45} / \mu_d^{0.27}.$$

Later work showed that the observed discrepancies are not a result of experimental errors and reflect the real effect of viscosity on the drop dimensions. As data [4-7] show, increasing the viscosity can either increase or decrease in the drop dimensions, depending on the viscosity of the second phase. In particular, this is illustrated by results from [8], where the equation

$$\frac{d_{32}}{d_n} = 0.095 K_N^{-2/5} (1 + 2.5 \varphi^{1/2}) \left( \frac{\mu_d}{\mu_c} \right)^\alpha \left( \frac{\mu_c}{\mu_d} \right)^\beta We^{-1/5} \quad (1)$$

was obtained. For  $\mu_d / \mu_c > 1$ , the exponents in (1) take the values  $\alpha = 0.2$  and  $\beta = 0$ ; for  $\mu_d / \mu_c < 1$ , the values  $\alpha = 0$  and  $\beta = 0.125$ ; the smallest drop diameter corresponds to cases where the phase viscosities are close in value.

This complex character of the effect of the phase viscosity is difficult to explain, and until recently has not been interpreted theoretically. The usual concepts on the mechanism of drop breakup in equipment with mixers were formulated basically [9, 10] on the basis of a model presented by Kolmogorov [11]. The assumptions of this model are well known. From the viewpoint of the viscosity effect, the model presumes the existence of two separate cases, which correspond to the conditions  $d_d \gg \lambda_0$  and  $d_d \ll \lambda_0$ , where the scale  $\lambda_0$  characterizes the upper bound of the dissipative scales:

$$\lambda_0 \sim (\nu^3 / \epsilon)^{1/4}.$$

In the first case the breakup of the drop is viewed as the result of the deforming effects from one of the inertial forces; in the second, only from viscous tangential stresses. Here in both cases, the only factor which prevents deformation is taken to be the capillary pressure. Later Levich [12] showed that the second of these cases cannot be realized in practice, because the condition of equilibrium between the viscous shear forces and the capillary pressure for  $d_d < \lambda_0$  can be fulfilled only in the range  $d_d \leq 10^{-7}$  m. In the case  $d_d >$

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$\lambda_0$ , analysis based on the model in [11] and the use of the "2/3" law leads [9, 10] to the equation

$$d_d \approx \sigma^{1/2} / (\rho^{1/2} \epsilon^{1/2}),$$

which generally does not include the viscosity.

Experimental results, and also estimates of local values of energy dissipation [13], however, show that the effect of viscosity appears for drop dimensions which exceed the intrinsic turbulence scale (in particular by a factor of 2-10, according to [14]). In order to explain the effect of viscosity on the scales of motion, for which viscous dissipation has not yet been characterized, several theoretical models [15-17] have been proposed, in which flow within the drop as it deforms is viewed as viscous, and the friction within the drop is viewed as a factor which prevents deformation and breakup. At the same time, experimental results, especially from (1) indicate that increasing the viscosity of one of the phases sometimes leads to a decrease in drop dimensions; consequently, under certain conditions the viscous interaction between the drops and the medium is a factor which aids breakup.

One of the possible mechanisms for such interaction can be related to the fact that deformation of the drop, as a discrete event, is a case of transient flow, and the motion of the phases is accelerated in the initial period. Starting from the moment when turbulent pulsations arise, the liquid which is within the drop and which directly connects it with the continuous-phase layer is accelerated by the action of a pulsating pressure. In the case where the scale of the drop exceeds the internal turbulence scale by less than an order of magnitude, it can be assumed that the less viscous of the two liquids will move with greater acceleration and, at the moment when the pressure difference reaches its maximum instantaneous value of  $\rho(v')^2/2$ , a velocity gradient will exist at the contact boundary between the drop and the continuous phase. Then an "accelerating" tangential stress

$$\tau_i = \left( \frac{dv'}{dr} \right)_i \mu_i,$$

will be applied on the side of the less viscous phase, and a "decelerating" stress

$$\tau_h = \left( \frac{dv'}{dr} \right)_h \mu_h,$$

will be applied on the side of the more viscous phase. Here  $(dv'/dr)_{l,h}$  is the velocity gradient at the boundary from the side of the less and more viscous phases. During the acceleration, the forces created by these tangential stresses do not equilibrate with each other, and their inequality can aid or prevent deformation. Here the stability condition of the drop can be represented in a unique form as

$$\frac{\rho(v')^2}{2} + \theta \mu_i \left( \frac{dv'}{dr} \right)_i \geq \frac{4\sigma}{d} + \theta \mu_h \left( \frac{dv'}{dr} \right)_h, \quad (2)$$

where  $\theta$  is the ratio of the lateral surfaces of the deformed drop to its cross sectional area. Using a linear approximation for the velocity gradients within the drop:

$$\frac{dv'}{dr} \approx \alpha_d \frac{v'}{d_d}$$

and from outside it:

$$\frac{dv'}{dr} \approx \alpha_c \frac{v'}{d_d},$$

where  $\alpha_d$  and  $\alpha_c$  are constants, and assuming that the transverse diameter of the deformed drop is proportional to the diameter of the initial drop, we obtain

$$\frac{4\sigma}{d} + \frac{v'}{d_0} (\gamma_c \mu_c - \gamma_d \mu_d) \leq \frac{\rho(v')^2}{2}. \quad (3)$$

by transforming Eq. (2) for the case where the viscosity of the dispersed phase is higher than that of the continuous phase ( $\mu_c > \mu_d$ ). Similarly, for the case  $\mu_d > \mu_c$ :

$$\frac{4\sigma}{d} \frac{v'}{d_0} (\gamma_d \mu_d - \gamma_c \mu_c) \leq \frac{\rho(v')^2}{2}. \quad (4)$$

In these equations  $\gamma_d$  and  $\gamma_c$  are coefficients which reflect the different velocity distributions near the contact surface from the side of the dispersed and continuous phases. Further analysis can be done in analogy to [13]. In the case of dividing the initial drop into two drops of half the volume, we obtain  $d = d_0/(2)^{1/3}$ , and, by solving (3) and (4) for  $v'$ , we determine the pulsation velocity sufficient for breaking up a drop of diameter  $d_0$ :

$$u_{d_0} = \frac{v'}{v_{cp}} \geq \frac{0.775}{\epsilon_m^{1/4} d_0^{1/4}} \left( \frac{M}{d_0} + \sqrt{(M/d_0)^2 + 10\sigma/(\rho_c d_0)} \right), \quad (5)$$

where

$$M = |(\rho_d/\rho_c) \gamma_d v_d - \gamma_c v_c|. \quad (6)$$

In Eq. (5),  $v_{av} = \epsilon^{1/m} \lambda^{1/3}$  is the root mean square pulsation velocity with a scale of  $\lambda = 2.17d_0$  (the scale of the major axis of an ellipsoid which is formed during the deformation of the drop is  $d_0$  [13]);  $\epsilon_m \approx \omega_0^3 r_0^3 (1 - v_0)^3 / L_b$  is the maximum local value of the energy dissipation in the zone of the mixer blades [13];  $L_b$  is the maximum characteristic dimension of the zone with energy dissipation  $\epsilon_m$  ( $L_b = h_b/2$  for  $h_b/2 > l_b$  and  $L_b = l_b$  for  $h_b/2 < l_b$ );  $v_0 = v_0/(\omega_0 r_0)$ ;  $v_0$  (m/sec) is the value of the circular component of the velocity at the radius of the end of the mixer blade  $r_0$ .

The frequency of pulsations of scale  $\lambda$  which have an amplitude  $v' \geq v_d'$  is [13]

$$v_{d_0} \approx \frac{0.6 \epsilon_m^{1/4}}{d_0^{1/4}} P(u), \quad P(u) = 1 - (2/\sqrt{\pi}) \int_0^u \exp(-x^2/2) dx, \quad (7)$$

and the average duration of agitation, which corresponds to breaking up a drop of diameter  $d_0$  in an agitator of given geometry is

$$\tau_{d_0} \approx \frac{d_0^{3/4}}{\epsilon_m^{1/4} P(u)}. \quad (8)$$

In (8),  $V_m$  is the volume of the zone near the mixer blade at which the energy dissipation is  $\epsilon_m$ . For turbine and blade mixers [13]

$$V_m = 0.5 z_b h_b^2 l_b, \quad (9)$$

for rotary mixers

$$V_m = z_b (0.13 l_b)^3. \quad (10)$$

By comparing (5) with the expression [13],

$$u = \frac{2.45}{\epsilon_m^{1/4} d_0^{1/4}} \left( \frac{\sigma}{\rho_c} \right)^{1/2}, \quad (11)$$

it is easy to see that Eq. (11) represents a special case corresponding to the condition

$$M \ll 10\sigma/(\rho_d d_0).$$

Measurements [13] have been conducted on the dimensions of drops in an emulsion stabilized by an emulsifying agent; agitation was carried out in equipment with reflecting baffles 0.17 and 0.3 m in diameter with turbine mixers and double-bladed mixers for drop dimensions from 40 to 200  $\mu\text{m}$ . In order to verify the application of our results and also to determine the constants  $\gamma_c$  and  $\gamma_d$  in (6) independently, the range of the experimental investigations was greatly expanded. Drop dimensions were measured for dispersed solutions of kerosene +  $\text{CCl}_4$  and transformer oil +  $\text{CCl}_4$  in water and in a water solution of glycerin. The local values of the dissipation energy reached 5000 W/kg (as opposed to 500 W/kg in [13]). The drop diameter  $d_{3,2}$  was varied from 400 to 15  $\mu\text{m}$ . The physical properties of the phases are shown in Table 1. The experiments were conducted in equipment with four reflecting baffles; the characteristics of the equipment and the mixers are shown in Tables 2 and 3. The drop diameters were measured by removing a sample from the vessel and photographing it or else by counting the drops directly under a microscope. In order to prevent coalescence, the emulsifying agents for each experimental system were blended individually. In order to measure the drop diameters for four different values of the emulsion concentration, the criterion for the absence of coalescence was taken to be the independence of the drop diameter



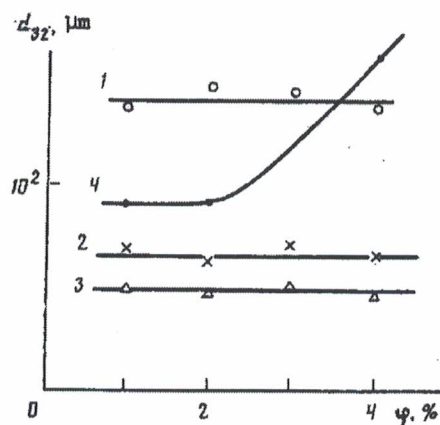


Fig. 1. Effect of emulsion concentration on the drop dimensions with the use of various surface-active agents ( $D = 0.17$  m; turbine mixer;  $d_M = 0.05$  m;  $n = 600$  rpm; mixing time 2400 sec). (see table below.)

Curve number	Continuous phase	Dispersed phase	Surface-active agent
1	Water	Transformer oil + $CCl_4$	Sodium oleate
2	Water	Kerosene + $CCl_4$	SEK
3	Water solution of glycerin	Transformer oil + $CCl_4$	Gelatin
4	Water	Transformer oil + $CCl_4$	Syntanol

TABLE 1. Properties of Continuous and Dispersed Phases

System number	Solid phase			Dispersed phase			$\sigma$ , N/m	Emulsifying agent	
	composition	$\rho_c$ , kg/m <sup>3</sup>	$\nu_c \cdot 10^6$ , m <sup>2</sup> /sec	composition	$\rho_d$ , kg/m <sup>3</sup>	$\nu_d \cdot 10^6$ , m <sup>2</sup> /sec		design- nation	concentration, g/dm <sup>3</sup>
1	Water	1000	1	Kerosene + $CCl_4$	1000	3	0,01	SEK	0,5
2				Transformer oil + $CCl_4$	1050	21,3	0,016	Sodium oleate	1,0
3				Transformer oil + $CCl_4$	1000	44,1	0,02		
4	Water solution of glycerin	1090	25,4	Kerosene + $CCl_4$	1100	2,5	0,046	Gelatin	1,0
5		1100	29,4	Transformer oil + $CCl_4$	1090	30,3	0,052		
6		11200	38,7	Kerosene + $CCl_4$	1100	2,5	0,052		

on the concentration of the dispersed phase (Fig. 1). Information on the emulsifying agents is shown in Table 1. Most of the measurements were made for volume concentrations of the dispersed phase equal to 2%.

It has been shown [13, 18] that, if there is no coalescence, the drop dimensions in the dispersed phase do not reach the limiting (equilibrium) value; however rate of decrease in the dimensions rapidly slows with time. In this regard, we choose an agitation time from 30 to 60 min, depending on the dimensions of the equipment and the mixers, such that doubling it did not decrease  $d_{32}$  by more than 5%.

In order to verify the nature of the drop dimensions as a function of the phase viscosities and to determine the values of  $\gamma_c$  and  $\gamma_d$ , the experiments were designed to change the ratio of the viscosities in the dispersed and continuous phases by a factor of 600, while maintaining the equipment and mixer geometry and the number of rotations. The results of the measurements (Fig. 2) confirm the conclusion that the nature of the effect of each of the phases on the drop dimensions is complex. As follows from Fig. 2, increasing the

TABLE 2. Equipment Parameters

$V \cdot 10^3, m^3$	$D, m$	$H, m$	Type of bottom
4.5	0.17	0.20	Flat
8	0.29	0.124	
16		0.24	
250	0.6	0.9	Elliptical

TABLE 3. Mixer Type and Dimensions

Nos. (Fig. 4)	Type of mixer	$d_m, M$	$z_b$	$h_b, m$	$l_b, m$	$\beta_b, \text{deg}$	$n, \text{rpm}$
1	Turbine	0.05	6	0.01	0.0125	90	500...1500
2		0.075		0.015	0.0185		500
3		0.1		0.02	0.025		150...550
4				0.03			150...350
5				0.04			150...250
6		0.125		0.025	0.031		700...900
7		0.16		0.062	0.04		125
8	Double-bladed	0.075	2	0.0075	0.03		800
9		0.1		0.01	0.04		300...1000
10				0.025			150...500
11		0.125		0.0125	0.05		300
12		0.16		0.016	0.064		200
13		0.18		0.018	0.072		200...800
14	Three-bladed	0.09	3	0.018	0.03	45	350...600
15	Six-bladed	0.15	6	0.015	0.06	90	100...600
16	Rotary	0.05	18	0.0045 *	0.009 *		800...1000
17		0.08		0.0072 *	0.014 *		600...800

\*For rotary mixers  $h_b = 0.09d_M$  and  $l_b = \pi d_M / 18$ .

viscosity of the dispersed phase for a low-viscosity continuous phase (Table 1, systems 1-3) leads to an increase in drop dimensions, while for a high-viscosity continuous phase (systems 4 and 5) to a decrease in drop dimensions. From the graph it can also be seen that the function  $d_{32} = f(v_d/v_c)$  is asymmetric. The minimum is displaced from  $v_d/v_c = 1$  on the abscissa and is found in a range of values\*  $v_d/v_c = 3-4$ , which indicates some difference in the effect of the viscosity of the continuous and dispersed phases. The nature of the function is in complete agreement with the structure (5). It also follows from Eq. (5) that for fixed physical properties of the phase properties ( $M = \text{const}$  and  $\sigma/\rho = \text{const}$ ), the viscosity effect should increase as the drop diameter decreases; that is as the maximum value of the energy dissipation  $\epsilon_m$  increases. Analysis of experimental functions of the drop diameter versus the local energy dissipation (Fig. 3) leads to the same conclusion. The observed increased effect of the viscosity of the more viscous phase as  $\epsilon_m$  is increased is in complete agreement with the faster increase of the term  $(M/d_0)^2$  compared to  $\sigma/(\rho d_0)$  in (5) as the drop diameter is decreased.

The experimental results in Fig. 2 were used to determine the values of  $\gamma_d$  and  $\gamma_c$ . As shown in [13], during the agitation of phases with low viscosities, which correspond to the relationship  $M/d_0 \ll \sqrt{\sigma/(\rho d_0)}$ , the value of  $d_0$  computed for  $M = 0$  is very close to the measured value of  $d_{32}$ . By assuming that the value of  $d_0$  corresponds to  $d_{32}$  even for  $M \neq 0$ , the coefficient  $\gamma_d$  was determined by comparing calculated results from (5)-(11) with the results of experiments conducted with the system of transformer oil +CCl<sub>4</sub> and water (Table

\*We note that in [18] the smallest value of the average drop diameter corresponded to a ratio  $\mu_d/\mu_s = 3.7$ . These experiments were conducted in an apparatus with a diameter  $D = 150$  mm; a turbine mixer was used with  $d_M = 49$  mm and  $n = 400$  rpm. The agitation time was 60 min.



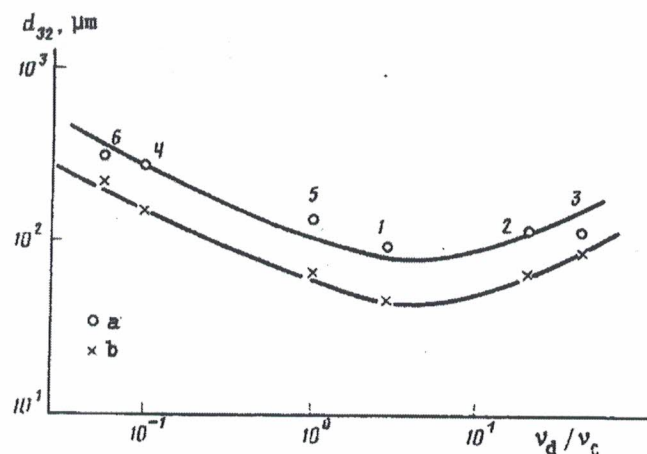


Fig. 2. Dependence of  $d_{32}$  on the phase viscosity ratio ( $D = 0.17$  m; turbine mixer,  $d_M = 0.05$  m); mixer rotation frequency; a) 500 rpm, b) 800 rpm; the point numbers correspond to the number column in Table 1; the lines are calculated from (5)-(12).

1, system 3), that is under the conditions  $v_d \gg v_c$ . The result is  $\gamma_d = 1.2$ . The value  $\gamma_c = 3$  was found in an analogous manner from experimental data with the system kerosene +  $\text{CCl}_4$  and a water solution of glycerin (Table 1, system 6), for which  $v_c \gg v_d$ . Thus, the expression for  $M$  in (5) takes the form

$$M = |1.2(\rho_d/\rho_c)v_d - 3v_c|. \quad (12)$$

Calculated drop diameters from Eqs. (5)-(11) and (12) are shown in Fig. 3 by solid lines. As follows from the curves, they are in complete agreement with the experimental data. In Fig. 4, the calculated values of  $d_{32}$  are compared with measured data for all the agitating mixers used in this investigation (Table 3). The figure also includes experimental results obtained for dispersion in the absence of coalescence [19, 20]. In [19], the emulsification was conducted in a vessel 0.05 m in diameter and mixed with a double-blade mixers with diameters 0.017, 0.025, and 0.033 m; in [20] with an apparatus diameter of 0.3 m and a turbine mixer of diameter 0.1 m. We note that the difference in phase densities in these investigations was 220-250  $\text{kg/m}^3$ , which greatly exceeds the corresponding values of the liquids used in our investigation. The correlation of experimental and calculated values of the drop diameter in Fig. 4, as well as the comparisons between calculations and experiments in Figs. 2 and 3, can be viewed as completely satisfactory. The root mean square deviation of the calculated drop diameters from the measured quantities  $d_{32}$  is  $\pm 0.15d_{32}$ .

#### NOTATIONS

a) specific surface separating the phases,  $\text{m}^{-1}$ ;  $D$ ) equipment diameter, m;  $d_0$ ) initial drop diameter, m;  $d$ ) diameter of drop formed from breakup of initial drop, m;  $d_M$ ) mixer diameter, m;  $d_{32}$ ) average volume/surface diameter of the drop, m;  $H$ ) filling height of the equipment, m;  $h_b$ ) mixer blade width, m;  $K_N$ ) power coefficient;  $l_b$ ) mixer blade length, m;  $n$ ) rotation frequency of the mixer, rpm;  $r_0$ ) mixer radius, m;  $V$ ) equipment volume,  $\text{m}^3$ ;  $V_m$ ) volume of breakup zone,  $\text{m}^3$ ;  $v_0$ ) circular velocity of the liquid at radius  $r_0$ , m/sec;  $v'$ ) root mean square pulsation velocity, m/sec;  $v_d$ ) settling (floating) velocity of the drop, m/sec;  $z_b$ ) number of mixer blades;  $\omega_0$ ) angular velocity of mixer, rad/sec;  $\beta_b$ ) inclination angle of the mixer blades, deg.;  $\epsilon_m$ ) maximum local value of the energy dissipation,  $\text{m}^2/\text{sec}^3$ ;  $\lambda$ ) turbulence scale, m;  $\lambda_0$ ) internal turbulence scale, m;  $\mu_d$ ,  $\mu_c$ ) dynamic viscosity of the dispersed and continuous phases, Pa·sec;  $\mu_h$ ,  $\mu_l$ ) dynamic viscosity of the phase with greater/lesser viscosity, Pa·sec;  $v_d$ ,  $v_c$ ) kinematic viscosity of the dispersed and continuous phases,  $\text{m}^2/\text{sec}$ ;  $\rho_d$ ,  $\rho_c$ ) density of the dispersed and continuous phases,  $\text{kg/m}^3$ ;  $\phi$ ) volume fraction of the dispersed phase;  $\sigma$ ) interphase surface tension, N/m;  $We = v_d^2 d \rho / \sigma$ ) Weber number.

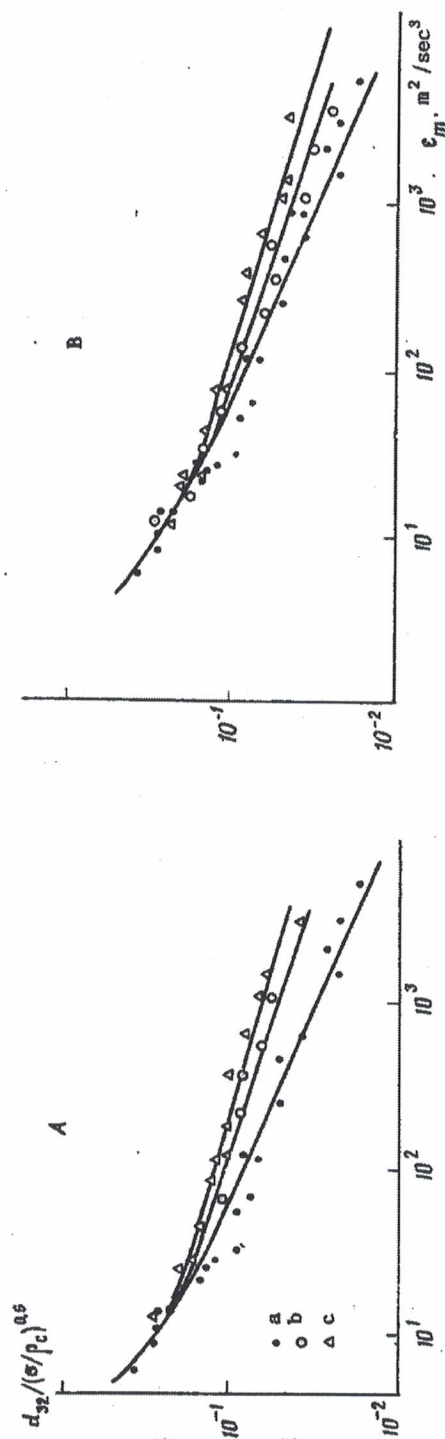


Fig. 3. Reduced drop diameter as a function of the local value of the energy dissipation, computed according to [13]: A)  $v_d = 3 \cdot 10^{-6} \text{ m}^2/\text{sec}$ , a)  $v_c = 1 \cdot 10^{-6} \text{ m}^2/\text{sec}$ , b)  $v_c = 25.1 \cdot 10^{-6} \text{ m}^2/\text{sec}$ , c)  $v_c = 38.7 \cdot 10^{-6} \text{ m}^2/\text{sec}$ ; B)  $v_c = 1 \cdot 10^{-6} \text{ m}^2/\text{sec}$ , a)  $v_d = 3 \cdot 10^{-6} \text{ m}^2/\text{sec}$ , b)  $v_d = 21.3 \cdot 10^{-6} \text{ m}^2/\text{sec}$ , c)  $v_d = 44.1 \cdot 10^{-6} \text{ m}^2/\text{sec}$ . Lines show calculations according to (5)-(12).



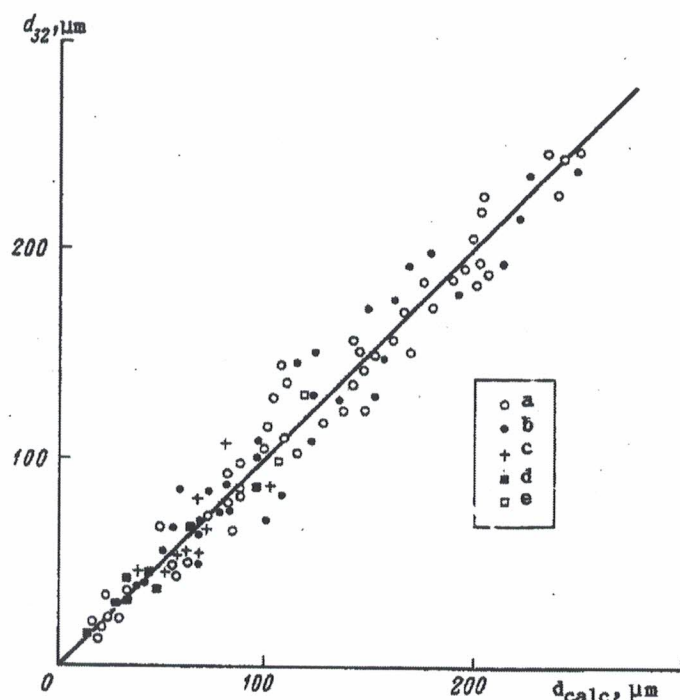


Fig. 4. Comparison of experimental and calculated drop diameters: a) turbine mixers (Table 3, Nos. 1-7); b) blade mixers (Table 3, Nos. 8-15); c) rotary mixers (Table 3, Nos. 16-17); d) blade mixers, experimental data from [20]; e) open turbine mixer, experimental data from [19].

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# COMPUTER-MODELING THE AERODYNAMICS OF GASEOUS SUSPENSIONS IN A VORTEX CHAMBER

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Nonstationary gas flow in a vortex chamber with side injection at high Reynolds numbers has been mathematically described. The effect of pulsations of the radial component of gas velocity on the flow of finely-dispersed particles has been studied.

Vortex chamber is the main element of swirling-flow equipment used for intensifying heat and mass transfer when drying finely-dispersed material and also of vortical and spiral-vortical dryers [1, 2]. Finely-dispersed material is injected into the spiral channels of spiral-vortical dryer in a stream of drying agent and later the gaseous suspension enters the vortical chamber (Fig. 1) consisting of an outer cylindrical shell, central cylindrical insert, and chamber bottoms. The radial component of the flow vector at the inlet can be regarded as much less than the tangential component. The gaseous suspension exits through an annular clearance at the chamber bottom.

A significant turbulization of gas flow greatly complicates the description of solid particles flow in the working zone of the vortex chamber while the velocity pulsations of the drying agent can exert much influence on the particle residence distribution in the working zone and the heat and mass transfer coefficients.

Aerodynamic modeling of vortex chamber with side-slit injection calls for several assumptions to simplify the problem. At any point in the working zone, the velocity vector can be assumed as parallel to the plane perpendicular to the chamber axis. Based on this assumption, gas flow can be regarded as planar. The effect of the molecular gas viscosity is ignored in the entire flow zone with the exception of a thin boundary layer.

The main equations for modeling the gas flow are the continuity equation

$$(\nabla, U)=0$$

and Eulerian flow equation

$$\frac{\partial U}{\partial t} + (U, \nabla)U = -\frac{1}{\rho} \nabla P,$$

where  $U$  is the velocity vector,  $\rho$  the density,  $P$  the pressure, and  $t$  the time.

Following the approach of [3] and the assumptions made, the problem involves finding a complex analytical function of velocity satisfying the required boundary conditions.