

INFLUENCE OF TURBULENCE AND VISCOSITIES  
ON THE KINETICS OF DROP BREAKING

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ABSTRACT

The proposed model assumes breaking to be caused by instant pulsation with sufficiently high amplitude. The critical amplitude is estimated taking into account capillary pressure inside the drop and viscous shear tension on its surface. The kinetics of drop size decrease is described using the distribution of pulsations and statistics of residence time of emulsion in the most turbulized area of flow.

The model was verified in vessels of four sizes with 17 agitators; relation of dispersed-to-continuous phase viscosities varied by approximately 600 times. Maximal turbulent dissipation rate values varied in the range of 10-10,000 W/kg.

The influence of viscosities was found to increase with an increase in turbulence intensity. The minimal mean drop diameter is obtained when emulsifying liquids with a definite viscosities ratio. It increases even if this ratio is changed as a result of a decrease in one of the viscosities.

The model allows to predict drop size distribution and to optimize emulsifying devices.

1. INTRODUCTION

Break-up of droplets of liquid in a turbulized flow of second phase is the main physical phenomenon controlling

the efficiency of emulsification process. Data on size distribution of drops formed under different flow conditions are essential for many industrial applications. This phenomenon is also of great scientific importance, as it is a direct reflection of certain features of turbulence. Experimental studies of drop break-up process have presumably started in 1949 [1] and since then have been constantly carried out in many laboratories in various experimental conditions. Those studies are based on the theoretical model of drop breaking in turbulent flow developed by Kolmogorov [2,3]. The model which is formulated in a generally accepted form in references [4,5] considers deformation and break up of drops of diameter  $d$  to be caused by turbulent pulsations of linear scale  $\lambda \approx d$ . For "Kolmogorov" range of linear scales

$$L \gg \lambda \gg \lambda_0 \approx \left(\frac{\nu}{\varepsilon}\right)^{1/4} \quad (1)$$

the stability condition for a drop in the flow with energy dissipation value  $\varepsilon$  is formulated out of the requirement of equilibrium of pulsation of pressure, corresponding to the mean square velocity component

$$\overline{v_m'^2} = \sqrt{\varepsilon_m^{2/3} \lambda^{2/3}} \quad (2)$$

and the capillary pressure, so that

$$\frac{\rho \overline{v_m'^2}}{2} \geq \frac{4\sigma}{d} \quad (3)$$

Thus, the diameter of the largest stable drops is determined as

$$d \approx c \varepsilon^{0.4} \left(\frac{\sigma}{\rho}\right)^{0.6}, \quad (4)$$

where  $C$  equals approximately 3 [6]. The value in this equation is usually estimated as the average value and is calculated as power consumption per unit of mass of the emulsion under treatment.

The very first comparisons of the model with experimental results have shown sufficiently good qualitative correspondence as regards the influence of main hydrodynamical parameters, such as mean flow velocity in tubes or rotational frequency of agitator, on the mean drop diameter [1,5,7,8]. However, more thorough studies of the process of drop formation carried out on various kinds of equipment, mostly in mixing vessels [6,9], have revealed certain serious shortcomings of the model. It has been found, in particular, that mean diameters of the drops formed in different mixing conditions are much smaller than those predicted by Eq.4. The experimental value of the constant  $C$  in this equation depends on the type and the size of the stirrer and does not usually exceed 0,15 [6, 10]. Moreover, experimental data have shown that the dependence between the mean drop size and turbulent energy dissipation rate is not always described by the relation  $d \sim \varepsilon^{-1/4}$  (Eq. 4). For instance, in case of drop breaking in turbulent jet with energy dissipation  $\varepsilon \approx 10^6$  W/kg, this dependence was found to be better described by the relation  $d \sim \varepsilon^{-0,25}$  [11].

Experimental research and practical experience have shown also, that drop sizes in the emulsions formed in the same flow conditions depend not only on the surface tension as predicted by the model, but on viscosities of phases as well. The most natural explanation of the influence of viscosities, based on the assumption that the drop size is less than Kolmogorov's microscale  $\lambda_0$ , proved to be incorrect. Calculations based on local energy dissipation values showed that even with dissipation rates being as high as  $10^6$  W/kg, mean diameters of drops

are 2-10 times greater than  $\lambda_0$  [11]. These results are in good correspondence with theoretical evaluations of Levich [12], which showed that Kolmogorov's assumptions concerning break-up of drops under viscous shear in turbulent flow are valid only for linear scales of the order of 10 and less, and cannot be applied to the actual range of breaking conditions. It has been found also that the increase in the viscosities of phases does not always prevent deformation. The influence is much more complex, and cases have been reported [13-16], when the increase in viscosity of one of the phases results either in increase or in decrease in the mean drop size. Such character of dependence does not match the theoretical results obtained during the last years [17-19] and needs additional investigation.

Finally, there is another factor not taken into account by Eq. 4 and other existing theoretical models, namely, the time factor. The Kolmogorov's model considers drop size  $d$  - diameter of the largest drops, stable in the turbulent flow with dissipation rate  $\mathcal{E}$  - as a constant value independent of residence time of emulsion in the breaking area. Obvious limitations imposed by such an approximation were noted by Kolmogorov himself in the work referred to above. The experiments described in a more recent work [13] have shown that the mean size of drops decreases along with the process duration, and the decrease rate depends on the intensity of mixing. The prediction of the drop size as a function of the break-up conditions should thus take into account the process kinetics, and the theoretical model of the break-up phenomena has to be transformed to meet this requirement.

Certain steps in this direction have been described in an earlier article [20] which deals with breaking of large "non-viscous" drops. The purpose of the present article is to apply this approach to breaking of drops under the influence of both surface tension and viscosity of phases.

The experimental part of the research was carried out in mixing vessels. This kind of equipment was chosen because it allows for the independent variation of energy dissipation rates and residence time values by several orders of magnitude. It is essential also that these values can be calculated on the basis of the data on agitator and vessel geometry, using the available methods of hydrodynamical calculations [6].

## 2. THEORY

Let us consider break-up of a drop of diameter  $d_0$  and formation of two drops of diameter  $d_0/\sqrt{2}$  as a result of the first drop's deformation into an ellipsoid with the smaller radius  $r \leq d_0/2\sqrt{2}$ , caused by a random turbulent pulsation of pressure. The characteristic linear scale of this pulsation should be close to the length of the larger axis of the ellipsoid at the final stage of its existence before it breaks up. It has been evaluated as  $\ell \approx 2.17d_0$  [20]. The pulsation is deforming a portion of the media of the scale  $\ell$  containing the drop and the neighboring continuous phase. If the size of the drop does not exceed the microscale of turbulence too much, and the relative velocity of the drop and the neighboring liquid before the pulsation occurs is equal to zero, then the less viscous of the two liquids acquires higher acceleration at the initial phase of pulsation. When the pulsational pressure reaches its maximum, a difference in velocities of phases will also arise, resulting in a different effective velocity gradients in the liquid inside and outside the drop. As a result, an additional (in respect to the capillary pressure) "braking" stress will occur on the side of a more viscous phase, which can be approximately evaluated as

$$\Delta P_1 = \left( \frac{dV}{dr} \right)_k M_k. \tag{5a}$$

The index  $k$  denotes the phase with higher viscosity. This braking stress is reduced due to the effect of a similar stress on the other side of the drop surface:

$$\Delta P_2 = \left( \frac{dV}{dr} \right)_l M_l. \tag{5b}$$

In non-steady conditions these tangential stresses should not be of equal value. Eq.(1), which reflects the equilibrium of the drop deforming under the effect of a turbulent pulsation, thus acquires an additional member which depends on velocities of both phases:

$$\rho_c \frac{V'^2}{2} \approx 4 \sqrt{2} \frac{\sigma}{d_0} + S \left[ \left( \frac{dV}{dr} \right)_l M_l - \left( \frac{dV}{dr} \right)_c M_c \right]. \tag{6}$$

§ In this equation is a ratio of the surface area of the ellipsoid to its cross-sectional area.

For a simplified analysis, linear approximation of the velocity gradients can be used, while the initial drop diameter  $d_0$  and pulsational velocity  $V'$  are assumed as characteristic length and velocity. Transforming Eq. (6) in accordance with this assumption, we obtain:

$$\frac{V'^2}{2} = 4 \sqrt{2} \frac{\sigma}{\rho_c d_0} - \frac{V'}{d_0} M, \tag{7}$$

where

$$M = \left| \frac{\rho_d}{\rho_c} \chi_d - \chi_c \right|. \tag{8}$$

$\chi_d$  and  $\chi_c$  in Eq.(8) are coefficients of approximate expressions of mean velocity gradients inside and outside

the drop. An estimated value of these coefficients is about 2, while the exact values are to be determined experimentally.

Thus, the break-up of a drop of a diameter  $d_0$  occurs under the action of a turbulent pulsation with a velocity of linear scale  $\lambda \approx 2.17 d_0$ , whenever the amplitude exceeds a critical value set by Eq. (7). Relative frequencies of turbulent pulsations with different amplitudes are usually assumed to follow the Gaussian distribution. Evaluating the mean frequency of pulsations of scale  $\lambda$  according to the "law of 2/3", we may estimate the frequency of drop-breaking pulsations as a function of relative amplitude  $u = V'/U_m$ :

$$\nu_0 \approx \frac{0.6 \varepsilon^{1/3}}{d_0^{2/3}} \int_0^\infty \frac{\exp(-u^2)}{u} du, \tag{9}$$

where  $U_0$  - minimal value of  $U$ , satisfying the conditions of breaking - is obtained by solving Eq.(7) and using Eq. (2):

$$U_0(d_0) = \frac{0.775}{\varepsilon^{1/3} d_0} \left( \frac{M}{d_0} + \sqrt{\left( \frac{M}{d_0} \right)^2 + \frac{10 \sigma}{\rho_c d_0}} \right). \tag{10}$$

Occurrence of turbulent pulsations is usually considered to correspond to the Poisson process. In this case, the probability for a given drop to undergo the action of a pulsation of the amplitude  $U \geq U_0$  at least once during its residence time in breaking area  $\tau$  is determined as

$$P(\tau) = 1 - \exp(-\tau \nu_0). \tag{11}$$

On the other hand, in the conditions of intense local turbulent mixing usually existing in the breaking area,

the volume part of the emulsion residing in this area during the period between  $\tau$  and  $\tau+d\tau$  can be assumed to be

$$dP(\tau/\tau_0) = \exp\left(-\frac{\tau}{\tau_0}\right) d\left(\frac{\tau}{\tau_0}\right), \quad (12)$$

where  $\tau_0$  is the mean value of residence time.

Assuming  $\tau_0$  to be equal to the mean period of drop breaking pulsations:

$$\tau_0 = \frac{1}{\dot{V}_0}, \quad (13)$$

where  $\dot{V}_0$  is described by Eq.(9), we estimate the part of drops of a diameter  $d$  undergoing break-up as

$$P(\tau_0) = \int_0^{\infty} e^{-\tau/\tau_0} (1 - e^{-\tau/\tau_0}) d\left(\frac{\tau}{\tau_0}\right) = 0.5.$$

If drop size distribution is Gaussian, as reported, for instance, in [13], the value of  $d$  must be close to the mean "Sauter's" value of drop diameter:

$$d = \frac{6\gamma}{f_s}, \quad (14)$$

where  $\gamma$  is a volume fraction of the disperse phase, and  $f_s$  - specific surface of drops per unit of volume of emulsion.

Remaining in the framework of a simplified monodisperse model, we may consider the  $\tau_0$  value from the Eqs (9) and (13) as time necessary to break half of the drops of diameter  $d_0$  into drops of diameter  $d_0/\sqrt{2}$ . According to (14), it can be expressed in a differential form:

$$\frac{d}{d\tau} d_0 \approx -0.115 d_0 \gamma_0. \quad (15)$$

and thus

$$\tau = - \int_{d_0}^{d_0} \frac{14.5 dz}{z^{1/3} \varepsilon^{1/3} \sqrt{\frac{2}{\pi}} \int_0^{\infty} \exp\left(-\frac{u^2}{2}\right) du} \quad (16)$$

This equation, together with Eqs. (9) and (10) describes the kinetics of the breaking process as a function of local turbulent dissipation rate and properties of phases.

It can be seen that in case of low viscosities of both phases and relatively big drop diameters, the values of parameter  $U$  practically do not depend on the viscosities. For such set of conditions Eqs. (9) and (16) may be used without any additional experimental constants. As for cases of higher viscosities or smaller drops, the parameter  $M$  must be taken into account. This parameter includes two constants -  $\chi_4$  and  $\chi_2$  which have to be determined on the basis of experimental results.

### 3. EXPERIMENTAL EQUIPMENT AND METHODS

Experimental research on drop break-up in non-coalescing conditions was carried out in glass and stainless steel mixing vessels of 4 sizes (volumes 4,5; 8; 16 and 250 litres; diameters from 17 to 60 cm). All vessels were equipped with 4 baffles, the baffles' width being equal to 0,1 of the vessel diameter. The height of the emulsion was 0,9 - 1,2 of the vessel diameter. Mixing was provided by turbine and paddle agitators with 2-6 blades, the blades height being 0,1 - 0,8 of the agitator

diameter. The ratio of vessel-to-agitator diameters varied in the range 1,7 - 8, and rotational speed of agitators - between 150 and 2500 RPM.

For any individual set of experimental conditions (vessel, agitator, pairs of phases), the lower limit of speed was selected to meet the requirement of fast disappearance of the oil layer, and the upper limit was maintained below the level of air insertion into the emulsion. Experiments were carried out with 6 pairs of immiscible liquids. Continuous phases were formed using water or water-glycerine solutions with viscosities from  $10^{-6}$  to  $3,87 \cdot 10^{-3} \text{ m}^2 \text{ s}^{-1}$ . Kerosene -  $\text{CCl}_4$  and transformer oil -  $\text{CCl}_4$  solutions with viscosities  $(2,5 \div 44,0) \cdot 10^{-3} \text{ m}^2 \text{ s}^{-1}$  served as a dispersed phase. Concentrations of solutions were selected in such a way as to obtain the difference in densities of phases of less than  $50 \text{ kg/m}^3$ , the oil always being the heaviest of the phases. The interfacial surface tension in the experiments ranged from 0,01 to  $0,052 \text{ N/m}$ . To prevent coalescence, several surface active agents were used, including gelatine and sodium oleate. Selection of SAA was carried out for each individual pair of phases. For that purpose the measurements of drop size were provided using from 1% to 4% volume concentration of emulsion. The choice of SAA was assumed to be correct if the mean drop size in emulsion was proved to be independent on the volume fraction of the disperse phase (Fig.1). To have an additional guarantee of the absence of coalescence, the dispersed phase concentration during the major part of the experiments was maintained as low as 2%. Measurements of drop sizes were carried out using microphotographic technique. Each mean drop size value was estimated on the basis of measurements of 250-400 individual drop sizes. The overall range of drop sizes was 1,5 -  $600 \mu\text{m}$ , mean (Sauter's) values ranged from 15 to  $250 \mu\text{m}$ . The duration of mixing in the experiments varied from 5 min to 6 hours.

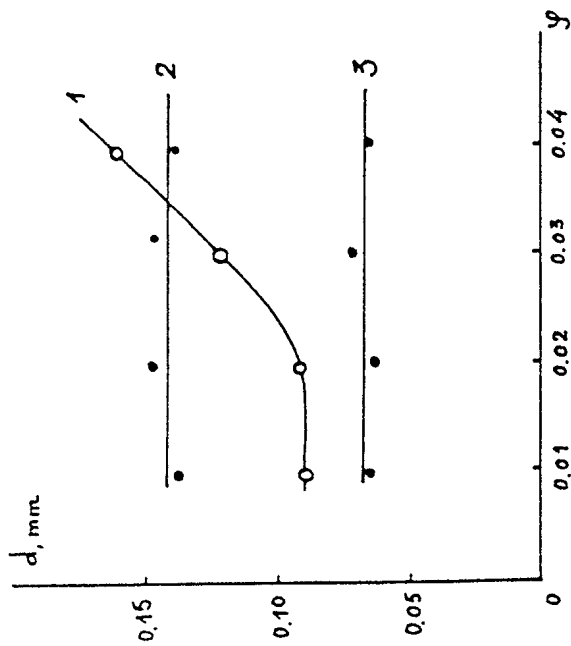


Fig. 1. Influence of volume fraction of disperse phase on mean drop diameter in the presence of various SAA. 1, 2 - the coalescence is absent; 3 - the coalescence is present.

It was shown in [20] that the size of drops formed in mixing vessels in the absence of coalescence is controlled by maximal local energy dissipation rate in the blades area, and by residence time of the emulsion in this area. These values for each set of conditions were evaluated using the formulae [20]:

$$\varepsilon = v_0^3 / l_m, \quad (17),$$

$$\tau_m = \theta \sqrt{V_m / V} \quad (18)$$

and

$$V_m = 0,5 z_0 h_0^2 l_0, \quad (19),$$

where  $Z_0, h_0$  and  $l_0$  - the number, height and length of the agitator blades, respectively,  $V$  and  $V_m$  - volumes of the vessel and of the area with dissipation rate  $\mathcal{E}$ ,  $\tau$  and  $\theta$  - mean residence time of emulsion in area  $V_m$  and overall mixing time,  $l_m$  - linear scale of area [20]. The value  $V_0$  - the difference of tangential velocities of blade and liquid - is calculated on the basis of mixing hydrodynamics as shown in [6].

#### 4. RESULTS

Fig. 2 shows typical curves of the decrease in the mean drop size. It is seen, that the decrease rate is very high during the first few minutes of mixing, then it goes down, and in 15-30 min the drops practically reach their final sizes. During the next 3-6 hours of mixing the mean drop size decreases only by 3-10%. Another specific feature of the very first stage of breaking consists in the fast change in the character of the drop size distribution. When the drops are relatively large, breaking occurs not only in a highly turbulized area near the mixer blades, but also in the bulk of the agitated flow, where energy dissipation rate is 100-1000 times lower. In these conditions the distribution is very wide, and sometimes two maximums are observed. However, this stage of the process is short, and when it is terminated, the distribution becomes more or less Gaussian, and its maximum is close to the mean (Sauter's) diameter. These results are shown in Fig. 3. The curve in the graph shows the results of 16 experiments carried out in different mixing conditions.

In the majority of the experiments only the "final" sizes of drops were measured. These final drop sizes proved to be independent of the type and size of agitator

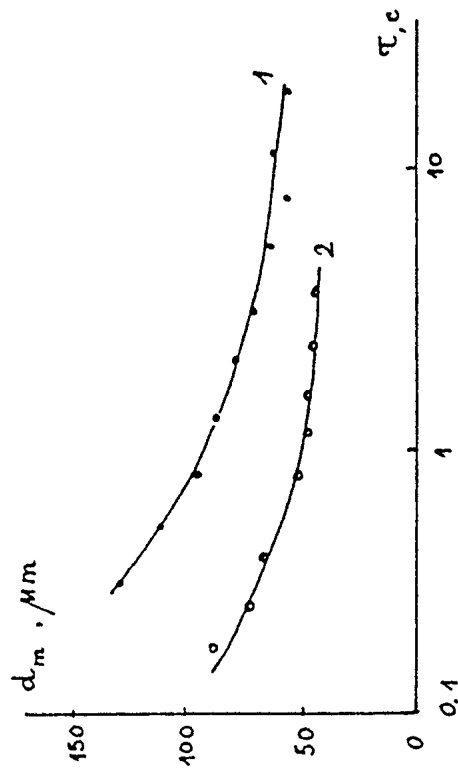


Fig. 2. Mean drop diameter vs. mean residence time in the breaking area. System "kerosene +  $C_{10}H_{14}$  - water".  
1 - turbine agitator,  $\mathcal{E} = 150 \text{ W} \cdot \text{kg}^{-1}$ .

2 - disk agitator,  $\mathcal{E} = 650 \text{ W} \cdot \text{kg}^{-1}$ .  
Continuous lines - theoretical calculations.

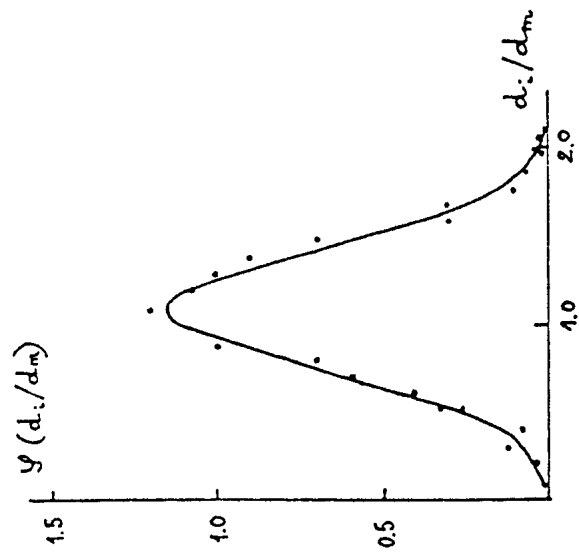


Fig. 3. Distribution of drop sizes

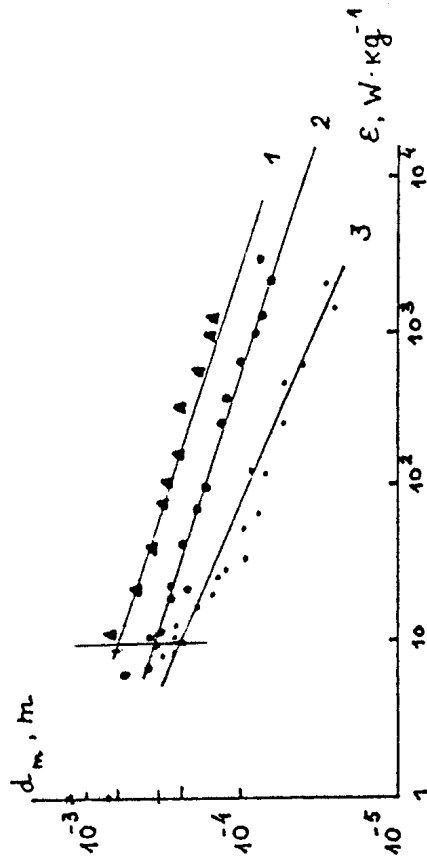


Fig. 4. Dependence of final mean drop diameter on the turbulent energy dissipation rate.

1 - system "kerosene +  $\text{CCl}_4$  - glycerine + water",  $\nu_c = 3,87 \cdot 10^{-5} \text{ m}^2 \text{ s}^{-1}$ ,  $\rho_c = 1090 \text{ kg} \cdot \text{m}^{-3}$ ,  $\nu_d = 2,5 \cdot 10^{-6} \text{ m}^2 \text{ s}^{-1}$ ,  $\rho_d = 1100 \text{ kg} \cdot \text{m}^{-3}$ ,  $\phi = 0,052 \text{ N} \cdot \text{m}^{-1}$ .

2 - system "transformer oil +  $\text{CCl}_4$  - water"  $\nu_d = 4,41 \cdot 10^{-5} \text{ m}^2 \text{ s}^{-1}$ ,  $\rho_d = 1050 \text{ kg} \cdot \text{m}^{-3}$ ,  $\phi = 0,02 \text{ N} \cdot \text{m}^{-1}$ .

3 - system "kerosene +  $\text{CCl}_4$  - water"  $\nu_d = 3 \cdot 10^{-6} \text{ m}^2 \text{ s}^{-1}$ ,  $\rho_d = 1010 \text{ kg} \cdot \text{m}^{-3}$ ,  $\phi = 0,01 \text{ N} \cdot \text{m}^{-1}$ .

and on the vessel volume, provided that the maximum energy dissipation values  $\mathcal{E}$  are the same, and residence times are sufficiently long (Fig. 4). Some of the essential results which reflect the influence of energy dissipation rate and viscosities of phases, are shown in Figs 4 and 5. It can be seen that the experimental curve  $\log d = f(\log \mathcal{E})$  has a different slope for different pairs of liquids. For the least viscous one (kerosene +  $\text{CCl}_4$  - water) it is practically equal to -0,4, whereas for such systems as kerosene +  $\text{CCl}_4$  - glycerine + water or transformer oil +  $\text{CCl}_4$  - water it is closer to -0,3.

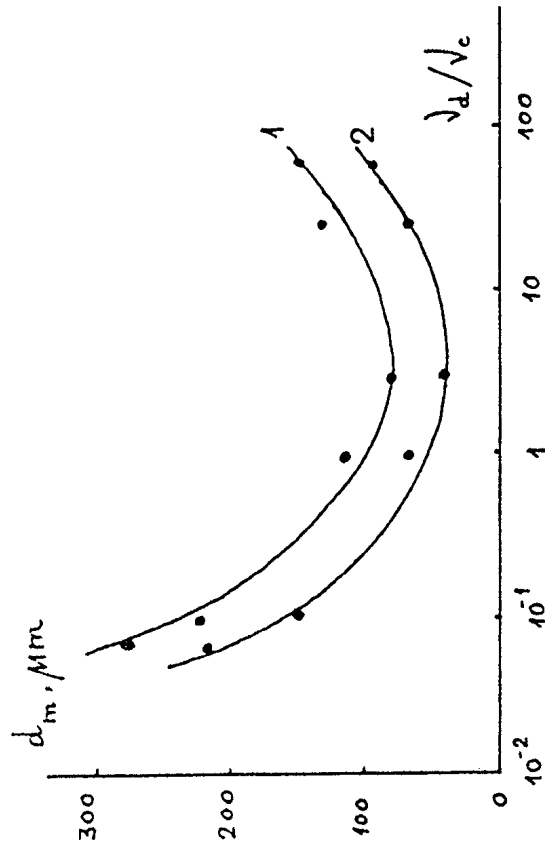


Fig. 5. Mean drop size vs. viscosity ratio of phases  
1 -  $\mathcal{E} = 116 \text{ W} \cdot \text{kg}^{-1}$ , 2 -  $\mathcal{E} = 475 \text{ W} \cdot \text{kg}^{-1}$ .

Fig. 5 presents the results obtained during emulsifying different pairs of liquids in the same mixing conditions. It is easy to see that the drop size depends not on absolute values of viscosities, but rather on their ratio. The increase in the viscosity of the more viscous phase results in the increase in the drop size, while the increase in the viscosity of the less viscous phase leads to the opposite effect. It should be noted that the results reported above are in good correspondence with the data obtained by other researchers. For instance, the empirical equation proposed in reference [16] for calculating mean drop diameter in mixing vessels contain the expression  $(M_d/M_c)^{\alpha} \cdot (M_c/M_d)^{\beta}$ ,

where  $\alpha = 0.2$  and  $\beta = 0$  for  $M_d > M_c$ , whereas for  $M_c > M_d$  values  $\alpha = 0$  and  $\beta = 0,125$  were obtained.



The kinetic curves presented in Fig. 2 are also very similar to those obtained in [13].

### 5. DISCUSSION

As it follows from the equations of the model presented above, the mean time value  $\bar{\tau}_0$  corresponding to breaking of drops of a diameter  $d_0$  may be expressed as a function of the parameter  $U_0$ , so that

$$\frac{\tau_0 \varepsilon^{1/3}}{d_0^{2/3}} = 1.67 / P(U_0), \quad (20)$$

where

$$P(U_0) = \sqrt{\frac{2}{\pi}} \int_{U_0}^{\infty} \exp\left(-\frac{u^2}{2}\right) du. \quad (21)$$

The graph in Fig. 1 shows that this function is gently sloping when  $U$  values are low, and becomes very steep when  $U$  values are equal or greater than 2. If the dissipation rate is constant, the time needed for each next act of breaking corresponding to a 1.26 decrease in the drop size, is much longer than the duration of the previous step of breaking. For values of  $U = 4$ , the mean breaking time considerably exceeds the duration of all the previous breaking steps. Thus, the drop diameter corresponding to  $U=4$  can practically be considered the final diameter, and the time needed to arrive at this value can be determined from Eqs. (9,13), without reference to the initial drop size.

These results may explain the qualitative character of the kinetics of emulsifying process observed experimentally. As for the quantitative description of the kinetics, it may be noted that for cases when

$$\frac{M}{d_0} \ll \sqrt{\frac{10\varepsilon}{\rho d_0}}, \quad (22)$$

the  $U$  value determined by Eq.(10) does not depend on any experimental constant. Consequently, for low-viscous systems, such as kerosene +  $CCl_4$  - water, the above equations may be used for a direct quantitative description of the process in experimentally observed range of drop sizes. The comparison of the kinetic curves  $d_0 = f(\tau)$  with experimental ones calculated on the basis of Eqs. (9,13), proved to be satisfactory (Fig. 2). The final values of the mean drop diameter correspond to the  $U$  values in the range 3.8 - 4.5. According to Eq. (9), this corresponds to the average experimental error of determination of the mean drop size of (+ 5%±10%). Thus, as it follows from Eq. (10), for non-viscous conditions described by (22), the final drop size may be estimated as

$$d_0 \approx 4 \varepsilon^{-0.4} \left(\frac{\varepsilon}{\rho}\right)^{0.6}. \quad (23)$$

Data presented in Fig.4 shows fairly good correlation between this equation and the results of the experiments with a pair of two low-viscous phases (curve 1).

When the viscosity of one of the phases is relatively high, experimental values of  $d$  were found to deviate from the curves predicted by simplified expression (23) (curves 2 and 3), and the effect of the viscosities is to be taken into account by using the expression of the parameter  $U$  in its complete form (Eq.10). To make it possible, the numerical values  $\chi_1$  and  $\chi_2$  were estimated, using the results of experiments with systems kerosene +  $CCl_4$  - glycerine + water ( $\nu_d/\nu_c = 0.065$ ) and transformer oil +  $CCl_4$  - water ( $\nu_d/\nu_c = 4.41$ ). It should be noted that values  $\chi_1$  and  $\chi_2$  were determined independently, because in the first case the  $M$  value is influenced only by the viscosity of the continuous phase, and in the second case - only by the viscosity of the

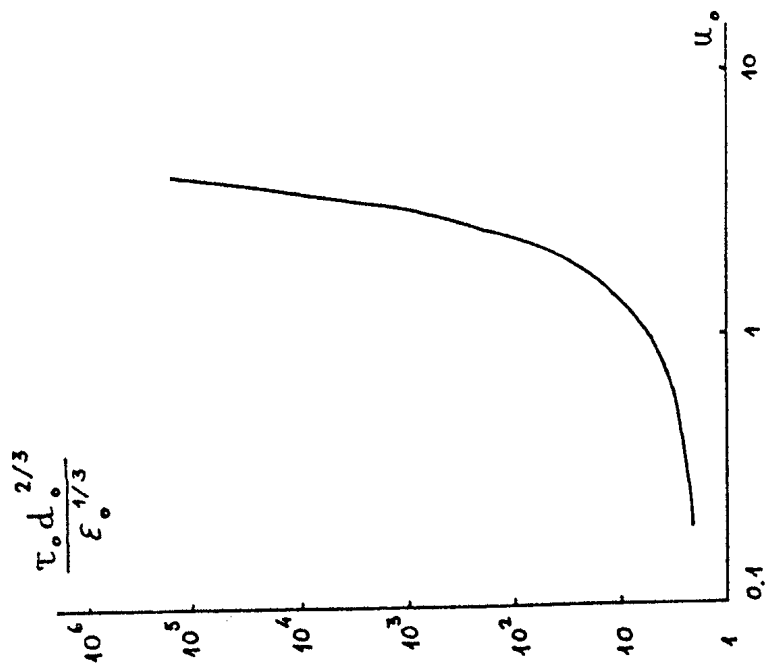


Fig. 6. Mean time of the breaking act vs.  $U_0$  value.

dispersed phase. Fig. 6 shows that the correlation between theoretical curves calculated using Eq. (10), and the experimental results for different systems is quite satisfactory.

The final drop diameter determined by these equations may be assumed to be the size of the largest drops stable in turbulent flow with a given dissipation rate. In this case, the Kolmogorov's model provides a correct description of stability conditions only in the range of highly "non-viscous" linear scales as determined by (22). However, it should be applied using the real, and not average, local values of turbulent dissipation rates, and

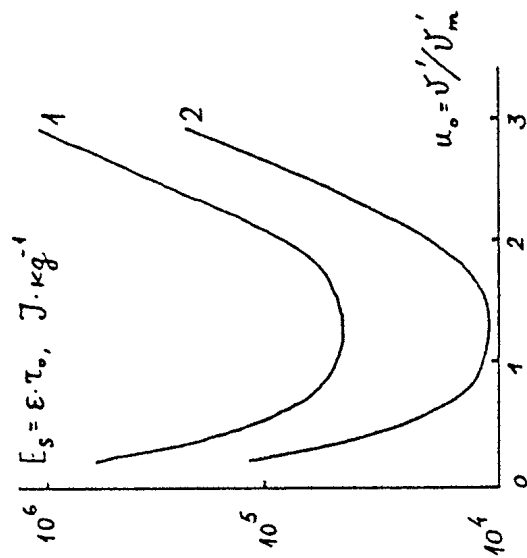


Fig. 7. Specific energy consumption for an individual act of break-up.

$$M = 3 \cdot 10^{-5} \text{ m}^2 \text{ s}^{-1}, \quad G/\rho_c = 10^{-5} \text{ m}^2 \text{ s}^{-2}, \\ 1 - d_0 = 0.5 \text{ Mm}, \quad 2 - d_0 = 1 \text{ Mm}.$$

based on the maximal probable amplitudes of turbulent pulsations of velocity rather than on their mean square values.

This is a simplified approach to be used mainly for analyzing drop-breaking processes in mass transfer equipment, where relatively large drops are formed, and viscosities of both liquid phases are low. For ordinary emulsifying processes and equipment, the influence of viscosities should be taken into account, and Eq. (10) should be used in its complete form.

The equations above allow also for drawing some conclusions concerning the practical side of emulsifying processes. As it follows from the graph in Fig. 7, an individual act of breaking can be carried out in certain optimal conditions corresponding to  $U \approx 1.25$ . The values

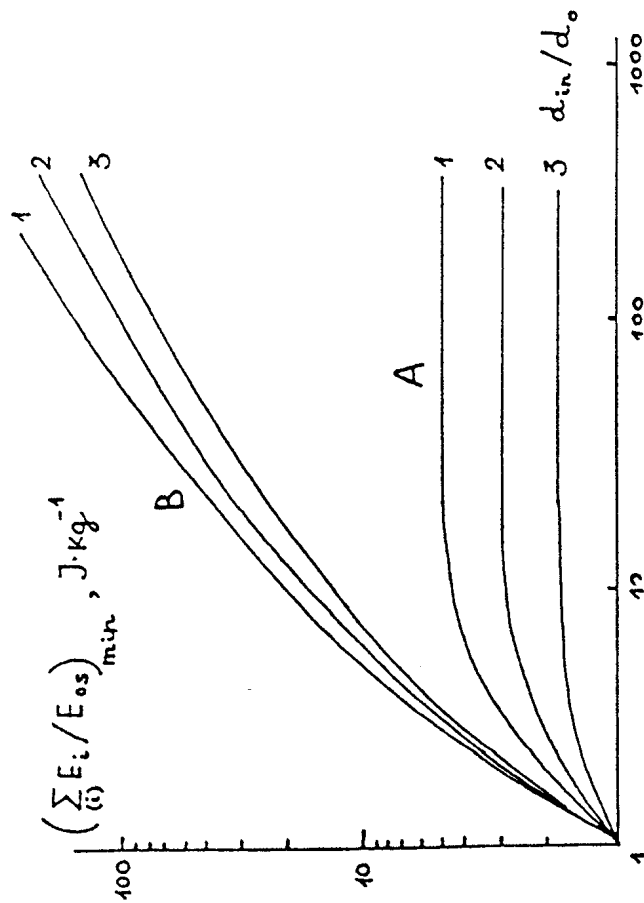


Fig. 8. Specific energy consumption for the decrease in mean drop size from  $d$  to  $d_0$ .

A - in area with  $\varepsilon = \text{const}$ .

B - under the optimal conditions ( $U; \approx 1,25$ )

$$G/\rho = 10^{-5} \text{ m}^2 \text{ s}^{-2},$$

$$1 - M=0, \quad 2 - M=10^{-5} \text{ m}^2 \text{ s}^{-1}, \quad 3 - M=10^{-4} \text{ m}^2 \text{ s}^{-1}.$$

of minimal specific power consumption per mass unit of emulsion are quickly increasing with the decrease in drop size. If the whole chain of consequent break-up acts starting with  $d$  and ending with  $d_0$ , is carried out in the optimal conditions, i.e., if for each 1-stage of the process  $U = 1,25$ , the overall power consumption value may be only 1,5-4 times higher than the optimal power consumption at the last break-up step (curve A in Fig.8). On the other hand, if the whole chain of breaking takes

place in the area with uniform dissipation rate, the minimal value of overall power consumption is some 5-20 times greater (curves B in Fig.8). Real emulsifying processes take place in the conditions corresponding to the upper rather than to the lower curve in the graph, and the proposed model provides possibilities for a significant improvement in this field.

## REFERENCES

1. Baranaev M.K., Teverovskiy E.N., Iregubova E.L., Reports of AN SSSR, 66, #5, 821, 1949.
2. Kolmogorov A.N., Ibid., 66, #5, 825, 1949.
3. Il'inze G.O., AICHE J., 1, 289, 1955.
4. Calderbank P.H., Trans. Instn. Chem. Engrs., 36, 443, 1958.
5. Pitterskiikh G.P., Valashek E.R., Khim. Prom., 1, 35, 1959.
6. Braginsky L.N., Begachev V.I., Barabash V.M., "Mixing in Liquids. Physical Fundamentals and Engineering Analyses Methods", Khimiya, Leningrad, 1984.
7. Pavlushenko S.S., Braginsky L.N., Journal of Appl. Chem., 36, # 8, 3122, 1963.
8. Nagata S., "Mixing. Principles and Applications", Halsted Press, New York, 1976.
9. Treybal S., "Liquid Extraction", McGraw Hill, New York, 1963.
10. Braginsky L.N., Begachev V.I., Theoretical Foundations of Chem. Eng., 3, #1, 103, 1969.
11. Mohr K.H., Chem. Techn., 36, #4, 157, 1984.
12. Levich V.G., "Physico-Chemical Hydrodynamics", Moscow, 1959.
13. Moro S., Process in Organic Coatings, #6, 271, 1978.

14. Bouguatotis B.S., Tohrnton J.D., Inst. Chem. Engrs., **26**, **43**, 1967.
15. Weinstein B., Treybal R.F., AIChE, **19**, **#2**, 304, 1973.
16. Nishikawa M., Mori F., Fujeda S., Journal of Chemical Engineering of Japan, **20**, **#1**, 82, 1987.
17. Arai K., Konno M., Matunaga J., Saito S., Journal of Chem. Eng., **10**, **#4**, 325, 1977.
18. Lagiseti I.S., Das P.K., Kumar R., Gandhi K.S., Chem. Eng. Sci., **41**, **#1**, 65, 1986.
19. Davies J.T., Chem. Eng. Sci., **42**, **#7**, 1671, 1987.
20. Braginsky L.N., Belevitskaya M.A., Theoretical Foundations of Chemical Technology, **24**, **#4**, 509, 1990.
21. Yaroshenko V.V., Braginsky L.N., Barabash V.M., Ibid., **21**, **#1**, 35, 1988.

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### Book Reviews

**Impure Science: Fraud, Compromise, and Political Influence in Scientific Research.** Robert Bell. Wiley: New York, pp. xvi+301. \$ 22.95.

First of all, this book was written by an economist, specifically, by a professor of economics. Second, the author *appears* to be an admirer of Congressman Dingell and his works. These comments are not offered in criticism; they do perhaps serve to suggest the direction of any bias which the reader may detect.

As the title suggests, Prof. Bell subsumes a large number of actions under the concept "fraud." These include hanky-panky in the award of grants by the National Science Foundation, abuse of peer review (using the review process to steal other's ideas), political motives in "big science" (i.e., patronage and the Supercollider), and, of course, *l'affaire* Baltimore. This is a considerable grab-bag.

Unfortunately, as far as I am concerned, the author does not distinguish between fraud *in* science and fraud *by* scientists. Fraud by scientists involves such absolutely human traits as stealing other's ideas, padding expense accounts, unfairly deprecating another's work, and so forth and (alas) so on. But (it may come as a suprise to some) scientists are human beings, and they will be guilty of what other human beings are guilty. The immoral items listed above are found in every group in society; if I had to guess, I would say somewhat less among scientists than among, say, attorneys. This ascribes no particular merit to scientists. Simply, the opportunities are fewer.

What I recognize as fraud *in* science is another matter entirely. This is *important*. By fraud *in* science I mean the deliberate falsification of data for purposes of gain or reputation. Of course, nothing is quite as black and white as the previous sentence suggests. For example,

1. Reporting the synthesis of compound X when you know that it is actually Y is one thing;
2. Reporting the synthesis of compound X in unqualified terms when you suspect it may not be X is another;
3. Just being plain wrong in your assertion out of ignorance or sloppy workmanship is another.