viscosity depending on the volume concentration of the suspended particles and the ratio of the viscosity of the particle material to that of the dispersion medium.

Since the situation considered in this article is very little different from the Einstein case, the coefficient 15 in Eq. (22) holds for $\Phi < 0.02$. The noticeable decrease of cross viscosity resulting from the addition of a dispersed phase to a viscoelastic liquid is well known and used in practice.

For $\mu_3 = 0$ the equations of state (21)-(23) give the classical results of the mechanics of dilute suspensions of spherical particles with a dispersion medium which is a Newtonian liquid.

It follows from the equations of state obtained that the addition of a dispersed phase with a small concentration to a Reiner–Rivlin liquid leads to a decrease in the cross viscosity, i.e., to a decrease in its non-Newtonian properties. Actually, μ_{3ef} can be written in the form

$$\mu_{3ef} = \mu_3[1 - \nu(\sigma)\Phi],$$

where $\nu(\sigma)$ lies between 0.6 and 15. The maximum value of ν corresponds to solid particles and the minimum value to gas bubbles.

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TURBULENT FLOW OF CONCENTRATED UNSTABLE EMULSIONS IN PIPES

V. F. Medvedev

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Unstable emulsions occur in a number of important technological processes, such as liquid extraction in oil refining and intrapipe demulsification in petroleum production. The hydrodynamic behavior of unstable emulsions differs from that of single-phase liquids in the damping of turbulent fluctuations of the dispersion medium by drops of the dispersed phase which are larger than the internal scale of turbulent fluctuations [1]. The turbulent flow of dilute unstable emulsions is described in [2].

When the content β of the dispersed phase of the emulsion lies in the range of $0.524 \leqslant \beta \leqslant 0.741$ (for $\beta = 0.741$ the phases of an unstable emulsion are inverted) the drops are closely packed, and shearing the emulsion requires an additional stress to deform them [3]:

$$\tau_0 = (0.195\beta - 0.102)\sigma/d, \quad 0.524 \leqslant \beta \leqslant 0.741$$

where σ is the interfacial tension, and d is the diameter of the drops of the emulsion. Thus, a concentrated unstable emulsion conforms to the Bingham model [4], and the equation of motion of concentrated emulsions in a pipe can be written in the form

$$\begin{gathered}(\mu_e + \mu_{ee})du/dy = \tau - \tau_0, \quad \tau_0 < \tau < \tau_w, \\ du/dy = 0, \quad \tau \leqslant \tau_0\end{gathered} \tag{1}$$

where u and τ are, respectively, the velocity and shear stress at a distance y from the wall, τ_w is the wall shear stress, and μ_e and μ_{ee} are the dynamic and eddy viscosities of the emulsion. It is shown in [5] that the dynamic viscosity of concentrated unstable emulsions can be determined in accordance with [6] as $\mu_e = \mu_1(1 - \beta)^{-2.5}$, where μ_1 is the dynamic viscosity of the dispersion medium.

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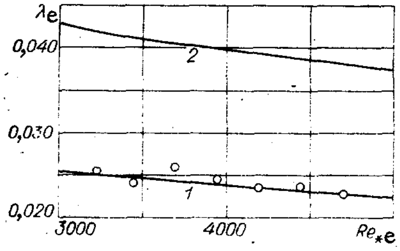

Fig. 1

We extend the hypothesis [7, 8] on the scale for fluctuating velocities in the flow of Newtonian fluids to the turbulent flow of concentrated emulsions which exhibit non-Newtonian properties; i.e., we assume that the scale for fluctuating velocities in developed turbulence is the dynamic velocity corresponding to the shear stress at a given radius r: $v_{*ye} = v_{*ye}(1 - y/r)^{1/2}$, where $v_{*e} = \sqrt{\tau_w/\rho_e}$ is the dynamic velocity. The density of the emulsion ρ_e is given by the sum $\rho_e = \rho_1(1 - \beta) + \rho_2\beta$, where ρ_1 and ρ_2 are, respectively, the densities of the dispersion medium and the dispersed phase. Then the eddy viscosity is found as in [2], taking account of the damping of turbulent fluctuations of the dispersion medium by drops of the dispersed phase.

By performing the integration of Eq. (1) in the same order as in [2], we obtain expressions for the drag coefficient of concentrated unstable emulsions λ_e. The are approximated to within 3% by the expressions

$$\lambda_e = 64/\mathrm{Re}_{*e}, \quad \mathrm{Re}_{*e} \leqslant 2320,$$
$$\lambda_e = \frac{0.3164}{(1+1{,}125\beta)\,\mathrm{Re}_{*e}^{0.25}}, \quad 2800 < \mathrm{Re}_{*e} < 10^5, \tag{2}$$

where $(1 + 1.125\beta)$ takes account of the effect of the damping of turbulence; Re_{*e} is the Reynolds number

$$\mathrm{Re}_{*e} = \frac{wD\rho_e}{\mu_e\left(1+\frac{\gamma P}{6}\right)},$$

where

$$\gamma = \begin{cases} 0, & 0 < \beta < 0.524, \\ 1, & 0.524 \leqslant \beta \leqslant 0.741; \end{cases}$$

$P = \tau_0 D/\mu_e w$ is the plasticity number, and w is the average flow velocity. Equations (2) describe the flow of unstable emulsions in pipes over the whole range of variation of the dispersed phase content.

The drag coefficient λ_e was determined experimentally for the turbulent flow of a concentrated unstable emulsion of transformer oil in water in a pipe 39.4 mm in diameter at a temperature of 16 ± 1°C. Unstable emulstions were obtained by the turbulent mixing of the liquids in a pipeline into which they were fed by forcing them out of tanks by compressed air.

Figure 1 compares the measured values of λ_e with a graph of Eqs. (2) (line 1) for an emulsion with a dispersed phase content $\beta = 0.6$; line 2 shows the variation of λ_e for pure liquid. It is clear that the drag coefficient is appreciably smaller for the flow of unstable emulsions than for a pure liquid as a result of the damping of turbulent fluctuations of the dispersion medium by drops of the dispersed phase.

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NATURAL CONVECTION AND HEAT TRANSFER IN POROUS INTERLAYERS BETWEEN HORIZONTAL COAXIAL CYLINDERS

V. A. Brailovskaya, G. B. Petrazhitskii, and V. I. Polezhaev

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INTRODUCTION

It is known that in finely dispersed porous materials with communicating pores filled with a liquid or gas, large-scale (with respect to the pore sizes) natural convection arises under specific conditions, which can have a significant effect on the heat-insulating properties of these materials. Investigations of the average characteristics of the heat transfer through plane horizontal and vertical layers of porous material and a comparison of them with experimental data are performed in [1-4]. The effect of convection on heat transfer in porous annular interlayers, which are elements of many engineering constructions (heat insulation of the volume contents of pipes, cables, and so on), is numerically investigated in this paper. Heat transfer in the annular interlayers of compression electric furnaces has been investigated in [5, 6]. Investigations have been carried out in [7, 8] for homogeneous annular interlayers filled with a liquid or gas.

§1. An annular interlayer of finely dispersed isotropic porous material is formed by two horixontal coaxial cylinders on whose outer and inner surface are maintained the constant temperatures T_2 and T_1, respectively.

In order to calculate the flow field and heat transfer the convection equations are used in the Boussinesq approximation, and the surface-friction force is replaced by the equivalent volume drag force in accordance with Darcy's law [9]. For the steady-state convection mode this system has the form

$$\begin{gathered}\mu\mathbf{v}/k = -\nabla p + \rho \mathbf{g}\beta\Delta T,\\ \operatorname{div}\mathbf{v} = 0,\quad \rho c_p(\mathbf{v}\nabla)T = \lambda^*\nabla^2 T,\end{gathered} \tag{1.1}$$

where ρ is the density, β is the volume expansion coefficient, μ is the dynamic viscosity coefficient, $\mathbf{v}$ is the velocity, c_p is the specific heat of the gas or liquid filling the pores, λ^* is the thermal conductivity of the porous medium without convection taken into account, p is the pressure difference from the static value, T is the mean temperature of the medium, ΔT is the difference between the local and some characteristic temperature, and k is the permeability coefficient of the porous medium.

Determining, as usual, the stream function ψ by the relationships $u = \partial\psi/\partial y$, and $v = -\partial\psi/\partial x$ (u and v are the components of the velocity $\mathbf{v}$ on the x and y axes) and eliminating the pressure from the equations of motion (1.1), we write down in the polar coordinate system in dimensionless form the system of equations for the stream function and the temperature Θ

$$\begin{gathered}\frac{\partial^2\psi}{\partial r^2}+\frac{1}{r}\frac{\partial\psi}{\partial r}+\frac{1}{r^2}\frac{\partial^2\psi}{\partial\varphi^2} = -\mathrm{Ra}^*\left(\frac{\partial\Theta}{\partial r}\cos\varphi-\frac{1}{r}\frac{\partial\Theta}{\partial\varphi}\sin\varphi\right),\\ \frac{\partial^2\Theta}{\partial r^2}+\frac{1}{r}\frac{\partial\Theta}{\partial r}+\frac{1}{r^2}\frac{\partial^2\Theta}{\partial\varphi^2} = \frac{1}{r}\left(\frac{\partial\psi}{\partial\varphi}\frac{\partial\Theta}{\partial r}-\frac{\partial\psi}{\partial r}\frac{\partial\Theta}{\partial\varphi}\right),\end{gathered} \tag{1.2}$$

where $\mathrm{Ra}^* = g\beta\delta k\rho^2 c_p\Delta T/\mu\lambda^*$ is the Rayleigh filtration number, which is the analogue of the Rayleigh criterion for a porous medium.

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