The resulting generalizing relationship applicable to critical overload allows us to determine the values of the angular rotational velocity separating the flow regimes for the coolant in a heat tube.

NOTATION

ω_t, angular rotational velocity of the tube; ω_l, angular velocity of liquid motion; R_r, radius of tube rotation; g, free-fall acceleration; ρ, density; μ and ν, coefficients of dynamic and kinematic viscosity; h, stream depth; $\eta = \omega_t^2 R_r/g$, overload; ω_{cr}, critical angular velocity; ω_l^A and ω_l^B, effective angular velocities of liquid motion for sub- and supercritical frequencies of rotation; A and B, coefficients to account for nonuniformity of liquid motion at sub- and supercritical frequencies of rotation; R_t, tube radius; $R = R_r/R_t$, dimensionless radius of rotation; $\eta_{cr} = \omega_{cr}^2 R_r/g$, critical overload; $Re = \omega_t R_t h/\nu$, Reynolds number; $H = h/R_t$, dimensionless stream depth; β, angle through which normal to the stream surface is deflected; Δ, portion of the tube surface flushed by the stream; ψ, angle of immersion.

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MODEL OF HEAT TRANSFER IN THE LIQUID PHASE DURING AXIAL AND TWISTED TURBULENT MOTION OF LIQUID AND GAS FILMS IN SHORT CHANNELS

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UDC 66.063.2

Based on the model of a diffusion boundary layer in conjunction with a given law governing the attenuation of turbulent pulsations in a viscous sublayer, we have derived an equation to calculate the volumetric coefficients of mass transfer based on the experimentally determined hydraulic resistance in a two-phase system.

We know that the transfer of mass in a liquid film is significantly intensified in direct axial, and particularly in twisted, motion of phases in a contact tube. A large number of studies, involving both axial [1-8] and twisted motions [6-10], has been devoted to the study of the hydraulic features and the modeling of mass transfer in dispersed-circular flows. The mathematical models of mass transfer in these studies, as a rule, contain empirical coefficients which require correction as the conditions of phase interaction change, and for this we must use the experimental data related to mass transfer.

Model of Mass Transfer in a Turbulent Film. Let us examine the direct (ascending or descending) turbulent motion of a liquid and gas, when the tangential stress $\tau_{g\text{-}l}$ at the boundary of phase separation considerably exceeds the stress $\tau_w = \rho_l g\delta$ at the wall, the latter generated exclusively by gravitational forces ($\tau_{g\text{-}l} >> \tau_w$). Such a regime is achieved in tubular contact devices at high gas-stream velocities of $W_g > 12\text{-}15$ m/sec.

To describe the processes of transfer in a turbulent film, we will adopt the model of the diffusion boundary layer [11-13], according to which the basic resistance to the transfer of mass is concentrated in the viscous sublayer:

$$\frac{1}{\beta} = \int_0^{\delta_1} \frac{dy}{D + D_t}, \quad (1)$$

where D_t is the coefficient of turbulent diffusion in the viscous sublayer, defined by the law governing the attenuation of turbulent pulsations.

Kirov Chemical Engineering Institute, Kazan'. Translated from Inzhenerno-Fizicheskii Zhurnal, Vol. 60, No. 3, pp. 372-379, March, 1991. Original article submitted April 25, 1990.

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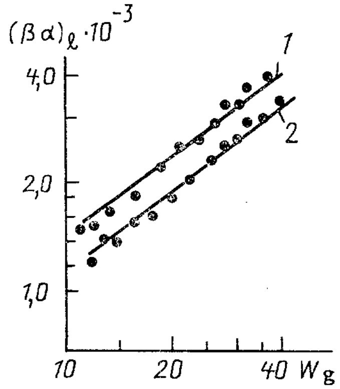

Fig. 1

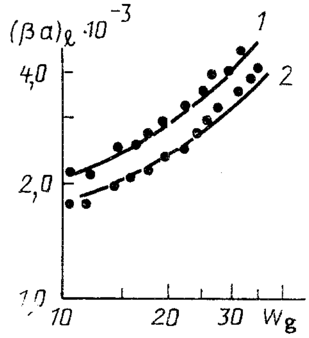

Fig. 2

Fig. 1. Ascending axial motion of phases in tube l = 0.15 m, d = 0.0168 m: 1) q = 2.43; 2) 1.85 $m^3/(m \cdot h)$. The curves represent calculations based on Eq. (15); the points represent experimental data [8]. The process of oxygen desorption out of water by air. $(\beta a)_l$, 1/h; W_g, m/sec.

Fig. 2. Ascending twisted motion of gas and liquid in a tube l = 0.15 m, d = 0.0168 m, $S_{tur.s}$ = 0.06 m: 1) q = 2.43; 2) 1.85 $m^3(m \cdot h)$. The curves represent calculations based on Eq. (15); the points represent experiment [8].

In [14, 15] the function $D_t(y)$ is assumed to be in the form of the exponential function:

$$D_t = u_* \delta_1 (y/\delta_1)^n , \tag{2}$$

in which the value of the exponent n depends on the hydrodynamic conditions of phases interaction. The value of n establishes the relationship between the coefficients of mass transfer and the coefficients of molecular diffusion D.

When $y = \delta_D$, $D_t = D$ and expression (2) has the form

$$D = u_* \delta_1 (\delta_D / \delta_1)^n \tag{3}$$

or

$$D/\delta_D = u_* (\delta_D / \delta_1)^{n-1}, \tag{4}$$

where $D/\delta_D \approx \beta$ is the coefficient of mass transfer.

Using the familiar relationship for the gas—liquid system $\delta_D/\delta_1 \approx Sc^{-1/2}$, from Eq. (4) we obtain the relationship between the coefficients

$$\beta \sim D^{\frac{n-1}{2}} . \tag{5}$$

In gas (steam)—liquid systems (film, surge, and similar equipment) the function $\beta \sim D^{1/2}$ has been established experimentally [6, 16, 17]. In this case, from relationship (5) we have n = 2.

We find the coefficient of mass transfer for the case in which n = 2 from expressions (1) and (2):

$$\frac{1}{\beta} = \delta_1 \int_0^1 \frac{d(y/\delta_1)}{D + u_* \delta_1 (y/\delta_1)^2} = \frac{1}{u_*} \int_0^1 \frac{d(y/\delta_1)}{D/u_* \delta_1 + (y/\delta_1)^2} =$$

$$= \frac{1}{u_*} \int_0^1 \frac{d\psi}{(\sqrt{a})^2 + \psi^2} = \frac{1}{u_*} \frac{1}{\sqrt{a}} \operatorname{arctg} \left(\psi \sqrt{\frac{1}{a}} \right) \Big|_0^1 ,$$

where $\psi = y/\delta_1$, $a = D/u_* \delta_1 = (R_1 Sc)^{-1}$. As a result, we obtain

$$\beta = \frac{u_*}{\operatorname{arctg}\sqrt{R_1 \mathrm{Sc}}\sqrt{R_1 \mathrm{Sc}}} = \frac{1}{\operatorname{arctg}\sqrt{R_1 \mathrm{Sc}}} \sqrt{\frac{\tau_{g-\ell}}{\rho_\ell R_1 \mathrm{Sc}}}. \tag{6}$$

The parameter $R_1 = u_*\delta_1/\nu$ in the turbulent motion of a single-phase flow over a solid surface has the value $R_{10} = 11.6$ from [18]. In a gas—liquid system with a movable surface of phase separation, as well as in the presence of hydrodynamic perturbations, the boundary-layer parameter R_1 can assume other values.

We will express the parameter R_1 in the gas—liquid system in terms of the familiar values of $R_{10} = 11.6$. For this we will write the relationship

$$\frac{R_{10}}{R_1} = \frac{(u_*\delta_1)_0}{u_*\delta_1}. \tag{7}$$

We will find the value of δ_{10}/δ_1 from the expressions for the determination of the coefficients of friction:

$$C_{f0} = \frac{2\tau_0}{\rho u_{\infty 0}^2}, \quad C_f = \frac{2\tau_{g-\ell}}{\rho u_\infty^2}. \tag{8}$$

We will adopt the familiar assumption to the effect that the momentum flux across the boundary layer is a constant quantity:

$$\nu\rho \frac{\partial u}{\partial y} = \tau = \mathrm{const}.$$

On this basis, the tangential stresses in expressions (8), with slight error, can be presented in the following form:

$$\tau_0 = \rho\nu\left(\frac{u_1}{\delta_1}\right)_0, \quad \tau_{g-\ell} = \rho\nu\frac{u_1}{\delta_1}, \tag{9}$$

where u_1 is the velocity of the liquid at the boundary of the viscous sublayer.

Given an identical hydrodynamic regime of single-phase flow motion at the solid wall and the flow of the liquid film $u_{\infty 0} = u_\infty$ and $u_{10} = u_1$. Then, from Eqs. (7)-(9) we obtain the value of R_{1l} in the boundary layer of the film, interacting with the gas flow:

$$R_{1\ell} = 11{,}6\sqrt{C_{f0}/C_{f\ell}} = 11{,}6\sqrt{\tau_0/\tau_{g-\ell}} \tag{10}$$

It follows from expression (10) that the correction of the parameter R_{1l} is achieved by altering the momentum flux τ_{g-l} in the film in comparison to the momentum flux τ_0 in the boundary layer of the single-phase flow.

The coefficient of friction in the motion of the single-phase flow in a tube is equal to [18]:

$$C_{f0} = \frac{\lambda_0}{4} = \frac{0{,}0791}{\mathrm{Re}^{0,25}}, \quad \mathrm{Re} = \frac{u_\infty d}{\nu}. \tag{11}$$

The velocity at the outside boundary of the turbulent boundary layer of the film, with the exception of a small segment of hydrodynamic stabilization, is close to the average velocity of the fluid in the interphase surface, i.e., $u_\infty = u_b$. An expression was derived in [15, 19] on the basis of a hydrodynamic analogy between momentum transfer and the transfer of mass for purposes of calculating u_b in the developed turbulent boundary layer separating the gas from the liquid:

$$u_b = \sqrt{\frac{\tau_{g-\ell}}{\rho_\ell}} R_{1\ell}\,(\pi/2 + \sqrt{R_{1\ell}}) \tag{12}$$

Consequently, the value of R_{1l} can be determined from a solution of Eqs. (10)-(12). The calculations carried out for the axial and twisted motions of the phases in the tubes demonstrate that the parameter R_{1l} in the boundary layer of the film is smaller than $R_{10} = 11.6$ and in the majority of cases amounts to $R_{1l} = 6$-8, while the average value of u_b ranges within 1.1-1.3 of the average liquid velocity in the turbulent film.

In the mass-transfer model (6), (10)-(12) the decisive parameter is the tangential frictional stress τ_{g-l}, whose reliability of determination is significantly dependent on the accuracy with which the coefficient of mass transfer is calculated. We presently have no methods on hand for an exact calculation of τ_{g-l} in direct axial flows of gas and liquid, and this applies even more particularly to twisted dispersed-circular flows. The value of τ_{g-l} can be calculated on the basis of the well-known coefficient of resistance [1, 4, 6] or from the results obtained in the measurement of the pressure difference ΔP in the dispersed-circular flow. We will illustrate

TABLE 1. Ascending Axial Motion in a Tube with d = 0.0168 m, l = 0.2 m. The Oxygen Desorption Process by Air Out of Water

w_t, m/sec	q, $m^3/(m \cdot h)$	ΔP, Pa	Calc. of $(\beta a)_\ell$ from (15), 1/h	Experiment [7] $(\beta a)_\ell$, 1/h	Divergence, %
14,6	2,398	1324,3	1384,5	1366	+1,0
18,6	2,398	1912,9	1716,5	1567	+8,7
25,1	2,398	2403,4	1974,9	1935	+2,0
28,8	2,398	2992,0	2234,5	2049	+8,3
32,1	2,398	3482,5	2434,5	2370	+2,6
34,9	2,398	4071,1	2654,5	2545	+4,1
37,6	2,398	4267,3	2729,8	2700	+1,1
40,1	2,398	4708,8	2883,3	2886	—0,1
44,6	2,398	5297,4	3080,9	3211	—4,2
15,1	1,355	931,9	1170,0	915	+21,7
15,1	1,852	1128,1	1285,0	1194	+7,1
15,1	2,398	1324,3	1388,7	1416	—2,0
15,1	2,99	1520,5	1483,9	1797	—21,0
39,2	1,355	3139,2	2339,4	1921	+17,8
38,2	1,852	3727,0	2552,0	2408	+5,6
38,5	2,398	4708,8	2879,0	2918	—1,4

examples of calculating the coefficients of mass transfer on the basis of an experimental investigation of ΔP in cylindrical channels under conditions of axial and twisted phase motion.

Let us write the condition for the balance of forces acting on the gas flow in projection onto the vertical axis in a contact tube:

$$\Delta P_{tu} S_g = \tau_{g-\ell} F \cos\theta. \tag{13}$$

From this we find the tangential stress

$$\tau_{g-\ell} = \frac{\Delta P_{tu} S_g}{F \cos\theta}, \tag{14}$$

where ΔP_b represents the drop in pressure across the two-phase system, ascribed to the friction generated by the gas and the liquid. On the basis of Eqs. (6) and (14) we obtain the formula with which we can calculate the volumetric coefficient of mass transfer:

$$(\beta a)_\ell = \frac{\beta_\ell F}{V} = \frac{4}{\pi d^2 l \operatorname{arctg} \sqrt{R_1\ {}_\ell \mathrm{Sc}_\ell}} \sqrt{\frac{\Delta P_{tu} S_g F}{\rho_\ell R_1\ {}_\ell \mathrm{Sc}_\ell \cos\theta}}. \tag{15}$$

Since considerable difficulties are encountered in an exact calculation of the interphase surface F in the case of direct liquid film and gas motion in the tube, we will assume its value to be equal to the geometric surface, without consideration of wave formation, i.e., $F = \pi(d - 2\delta)l$. This assumption, in the case of high liquid flow rates and high gas velocities may sometimes lead to an underestimated value (of up to 20-30%) of the volumetric coefficient of mass transfer (15).

Axial Phase Motion. The total expenditures of energy in the motion of a two-phase dispersed-circular flow through a cylindrical channel are made up of the energy expended on the acceleration of the liquid film from the initial velocity to the average velocity, on the separation of the liquid drops by the gas flow, on the transport of the drops within the channels, and on the friction of the gas against the interphase surface of the film [6]:

$$\Delta P = \Delta P_{acc} + \Delta P_{sep} + \Delta P_{en} + \Delta P_{tu}. \tag{16}$$

It was demonstrated in [6] that the energy expended on the acceleration of the liquid film, i.e., ΔP_{acc}, amounts to no more than 0.25-0.5% of ΔP, while the energy expended on the separation of the droplets ΔP_{sep} from the surface of the film does not exceed 2-3%. Calculation of the pressure losses ΔP_{en} on the transport of liquid drops, without consideration of the slippage rate, thus leading to overestimated results, amounts to 15-25% of the total pressure difference [6]. At the same time, on the basis of the data from [20], this component of the pressure difference ranges within 6-11% of ΔP. As a result, we can draw the conclusion that the pressure difference ascribed to the friction of the gas and the liquid ranges within the limit

$$\Delta P_{tu} = (0{,}8 - 0{,}95)\, \Delta P.$$

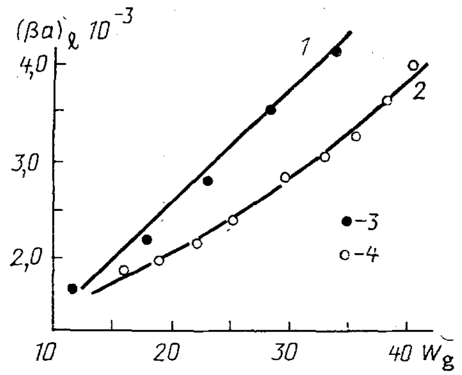

Fig. 3. Heat-release coefficients as a function of gas velocity in the twisted motion of phases with varied turbulator pitch in the tube: l = 0.2 m, d = 0.0168 m, q = 2.39 $m^3/(m\cdot h)$; 1, 3) $S_{tur.s}$ = 0.062 m; 2, 4) 0.118 m; 1, 2) calculation based on Eq. (15); 3, 4) experiment [7].

TABLE 2. Descending Twisted Motion in a Tube with d = 0.0168 m, l = 0.15 m

W_g, m/sec	q, $m^3/(m\cdot h)$	ΔP, Pa	Calc. of $(\beta a)_\ell$, 1/h	Exp. [8] $(\beta a)_\ell$, 1/h	Divergence, %
		$S_{tur.s}$=0,06m			
25,4	1,55	1960	2325	1682	+27,6
25,4	2,14	2338	2536	2115	+16,6
25,4	2,71	2770	2762	2700	+2,2
25,4	3,32	3093	2916	3380	—15,9
36,2	1,55	3921	3411	2500	+26,7
36,2	2,14	4440	3628	3240	+10,6
36,2	2,71	4931	3823	3840	—0,4
36,2	3,32	6010	4233	4780	—12,9
		$S_{tur.s}$=0,12 m			
20,6	1,55	1000	1453	1137	+21,7
20,6	2,14	1343	1668	1550	+8,1
20,6	2,71	1667	1882	2040	—8,4
20,6	3,32	2026	2081	2485	—19,4

With direct axial phase motion through the tube the parameters θ and S_g in Eq. (15) have the following values:

$$S_g = \pi (d - 2\delta)^2/4, \quad \theta = 0° .$$

Results from the calculation of $(\beta a)_l$ in accordance with Eq. (15) and comparison against the experimental data from [7, 8] in the case of an ascending axial motion are shown in Table 1 and in Fig. 1. The thickness of the film was determined from empirical expressions found in [1, 6].

Twisted Phase Motion. Let us examine the helical motion of a liquid and gas film in a tube with a strip turbulator.

The pressure difference across a twisted dispersed-circular flow of gas and liquid, in addition to the above-noted resistances (16), includes the hydraulic resistance against the twisting of the gas flow:

$$\Delta P = \Delta P_{acc} + \Delta P_{sep} + \Delta P_{en} + \Delta P_{tu} + \Delta P_{tw} . \tag{17}$$

With helical phase motion at a gas velocity of W_g < 30 m/sec we observe no liquid drop separation from the surface of the film [8]. The centrifugal force generated in the rotation of the gas—liquid flow hinders the separation and entrainment of the liquid. If the velocity of the gas is W_g > 30 m/sec, then we observe separation of liquid particles from the surface of the film and this, however, under the action of centrifugal force, leads once again to compression of these liquid particles to the film [8]. Consequently, the pressure loss components ΔP_{acc}, ΔP_{cr}, and ΔP_b are negligibly small in comparison with ΔP and their values need not be taken into consideration.

The drop in pressure generated by the twisting of the gas flow and by friction against the surface of the strip turbulator can be found from the following expression [21]:

$$\Delta P_{tw} = \lambda_{tur} \frac{l}{d} \frac{\rho_g W_g^2}{2}, \tag{18}$$

where λ_{tur} is the coefficient of hydraulic resistance which is calculated from the empirical equations found in [21].

The pressure difference ascribed to the friction of the gas and liquid is equal to $\Delta P_{tu} = \Delta P - \Delta P_{tw}$.

For a tube with a strip turbulator the parameters S_g and θ in Eq. (15) have the value

$$S_g = \frac{\pi}{4}(d - 2\delta)^2 - (b\delta)_{tur.s}$$
$$\theta = \operatorname{arctg}\frac{\pi d}{S_{tur.s}}. \tag{19}$$

Equation (15) with the parameters in (19) has been tested for contact devices with descending and ascending phase motion, with different turbulator spacing. The thickness of the liquid film was calculated from empirical expressions in [9]. The results from the calculations of the volumetric coefficients of mass transfer in the case of oxygen desorption by air out of water and comparison against experimental data [7, 8] are shown in Table 2 and in Figs. 2 and 3.

CONCLUSION

A detailed description of the transfer processes within the liquid film leads to the relationship between the coefficients of mass transfer and the length and amplitude of the wave at the interphase surface [2, 3]. However, precise measurement of these parameters encounters considerable difficulties, particularly in the twisted motion of the phases in the contact tube.

The developed mass-transfer model (6), (10)-(12) links the kinetic coefficients with the tangential stress, which can be expressed with sufficient accuracy in terms of the pressure drop across the two-phase system. The relationship between the coefficients of mass transfer and the pressure difference has been established experimentally by numerous researchers [5, 6, 22, 23].

Consequently, the significant intensification of mass transfer within the liquid film in direct high-speed turbulent motion of a dispersed-circular flow is achieved through a major increase in tangential stress at the interphase surface.

The model of mass transfer in a turbulent liquid film ($Re_f > 1600$) has been tested in the following range of changes in regime and structural parameters: $q = 1.35$-3.32 $m^3/(m \cdot h)$, $W_g = 14$-45 m/sec, $l/d < 20$. An absolute majority of results from the calculation of the volumetric coefficients in (15) is found to fall within the limits of experimental error (±20%).

Equations (6) and (15) can be recommended for design purposes in the planning and redesigning of contact equipment with direct turbulent liquid and gas film movements, and this is based exclusively on the hydraulic research.

NOTATION

$b_{tur.s}$, width of turbulator strip, m; D, coefficient of molecular diffusion, m^2/sec; d, tube diameter, m; F, phase separation surface, m^2; l, tube length, m; q, liquid flow rate referred to tube perimeter, $m^3/(m \cdot h)$; S_t, area of free tube cross section, m^2; $S_{tur.s}$, spacing of strip turbulator, m; u_*, dynamic friction velocity, m/sec; u_b, liquid velocity at the interphase surface, m/sec; V, working volume of contact device, m^3; W_g, velocity of gas in total tube cross section, m/sec; y, transverse coordinate of the boundary layer, m; β, coefficient of mass transfer, m/sec; θ, angle of liquid film movement; δ, film thickness, m; δ_1, thickness if viscous sublayer, m; δ_D, thickness of diffusion sublayer, m; $\delta_{tur.s}$, thickness of turbulator strip, m; ν, coefficient kinematic viscosity, m^2/sec; ρ, density, kg/m^3; τ, tangential frictional stress, $kg/(m \cdot sec^2)$. Complexes: $Sc = \nu/D$, the Schmidt number; $Re_f = 4g/\nu$, the Reynolds number for the film. Subscripts: g, gas phase; l, liquid phase; o, single-phase flow; tur.s, strip turbulator.

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