HEAT TRANSFER OF TOLUENE AND BENZENE DURING LAMINAR FLOW AND AT SUPERCRITICAL PRESSURES

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Experimental data was used to determine the lengths of pipe after which free convection disturbs the character of change in the heat-transfer coefficient of aromatic hydrocarbons (toluene and benzene).

Laws were established earlier for the change in wall temperature along a pipe and its dependence on the heat flux with a laminar flow of toluene and benzene and supercritical pressures [1-3].

Free convection may significantly affect heat transfer in the case of laminar flow and variable fluid properties. For this reason, the changes in wall temperature and heat-transfer coefficient may be different at supercritical pressures than under normal conditions. Test data on the heat transfer of toluene and benzene at supercritical pressures and with large temperature gradients between the wall and fluid bear this out [1-3, 4]. Tests on the heat transfer of toluene and benzene during laminar flow and at $p > p_{cr}$ were conducted in an open circulation loop in a stainless steel pipe with an inside diameter $d \approx 3$ mm and a heated section of length $l/d \leqslant 75$. The heated section was preceded by a hydraulic stabilization section with a length $l_{h.s.} = 0.06Re(d)$.

Heat transfer was studied during ascent and descent of the fluid in a vertical pipe in the following ranges of the regime parameters: for toluene $p/p_{cr} = 1.06$-3.07, $T_q/T_{cr} = 0.49$-1.05, $T_w/T_{cr} = 0.55$-1.56, $q = (0.3$-$4.0)\cdot 10^5$ W/m^2; for benzene $p/p_{cr} = 1.21$-2.63, $T_q/T_{cr} = 0.52$-1.12, $T_w/T_{cr} = 0.54$-1.55, $q = (0.10$-$4.5)\cdot 10^5$ W/m^2.

It is apparent from the graphs of the change in the heat-transfer coefficient along the pipe for the ascending flow of toluene and benzene that at low wall and fluid temperatures ($\Delta t < 100$°C) the criterion Nu decreases monotonically in the initial part of the pipe, as during normal convective heat transfer (curve 1, Fig. 1a and b), and the test data on the local value of the heat-transfer coefficient is described by Petukhov's equation [5]

$$\mathrm{Nu}^{e} = \mathrm{Nu}_0 \left(\frac{\mu_w}{\mu_q} \right)^{-0.14}, \tag{1}$$

where

$$\mathrm{Nu}_0 = 1.03 \left(\frac{1}{\mathrm{Pe}} \frac{x}{d} \right)^{-1/3}. \tag{2}$$

It can be seen from Fig. 2 that during ascending and descending flow the test data on heat transfer for the hydrocarbons is roughly the same in the case of low values of Δt. Thus, for laminar flow and supercritical pressures of the hydrocarbons with small temperature differences between the wall and liquid, heat transfer can be calculated from equations obtained for subcritical pressures, regardless of the direction of flow. In such calculations it is sufficient to allow for the change in the viscosity across the flow, since the effect of free convection on the character of flow and heat transfer is negligible ($Gr/Re < 50$ with descending flow and $Gr/Re < 100$ with ascending flow in the vertical pipe).

As the heat flux increases (with p = const, t_q^{in} = const, and ρu = const), wall temperature increases, there is a corresponding increase in Δt, Gr, and Gr/Re, and under the in-

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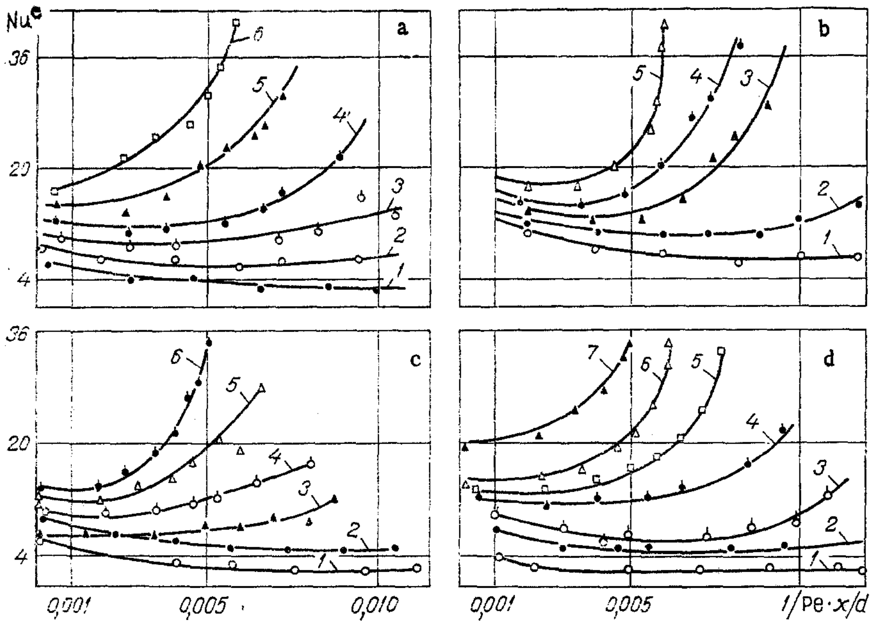

Fig. 1. Change in the Nusselt number along the pipe: ascending flow: a) toluene; p = 9.6; ρu = 87; $q \cdot 10^{-5}$: 1) 0.14; 2) 0.47; 3) 0.84; 4) 1.37; 5) 1.54; 6) 2.10; b) benzene; p = 6.0 MPa; ρu = 85 kg/m^2·sec; $q \cdot 10^{-5}$ W/m^2: 1) 0.28; 2) 0.70; 3) 1.27; 4) 1.81; 5) 2.4; descending flow: c) toluene; p = 5.0; ρu = 107; $q \cdot 10^{-5}$: 1) 0.02; 2) 0.17; 3) 0.74; 4) 0.94; 5) 1.33; 6) 1.70; d) benzene; p = 9.0 MPa, ρu = 95 kg/m^2·sec; $q \cdot 10^{-5}$ W/m^2: 1) 0.07; 2) 0.39; 3) 1.14; 4) 1.39; 5) 1.56; 6) 2.10; 7) 3.15.

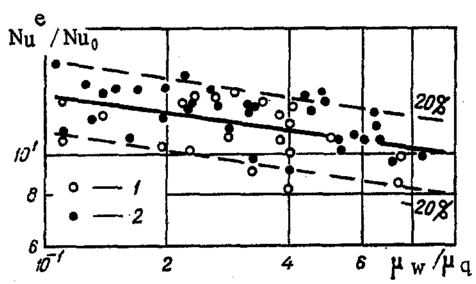

Fig. 2

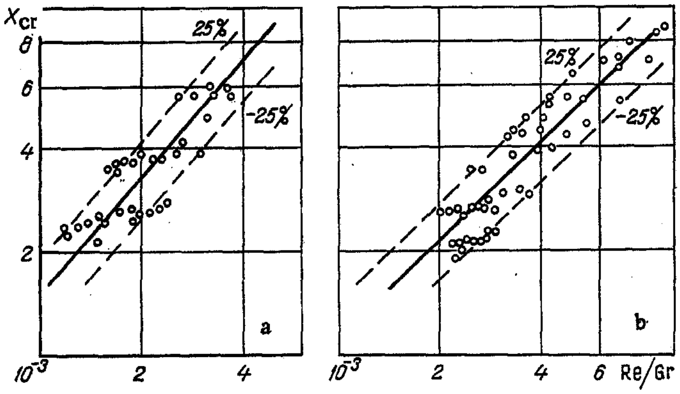

Fig. 3

Fig. 2. Heat transfer of toluene and benzene with laminar flow and small temperature gradients between the wall and fluid: 1) ascending; 2) descending flow.

Fig. 3. Dependence of X_{cr} on Re/Gr: a) ascending; b) descending flow.

fluence of free convection the character of the change in the heat-transfer coefficient along the pipe begins to differ from the previous case (curves 3 and 4, Fig. 1a and b). The effect of free convection on the heat-transfer rate begins in the outlet part of the pipe and shifts to the initial part of the pipe as heat flux increases. A discontinuity is seen on the curve $\mathrm{Nu}=f(\frac{1}{\mathrm{Pe}}\frac{x}{d})$ in this case. In such tests, normal convective heat transfer occurs in the initial part of the pipe, this heat transfer corresponding to a viscous regime of laminar flow. The flow in the final part of the pipe corresponds to a viscous-gravitational regime of convective heat transfer. The heat-transfer coefficient is highest on the pipe length where the transition occurs from one regime to another: the criterion Nu decreases in the initial part of the pipe to a certain value, and subsequently increases. A further increase in heat flux leads to a situation whereby the heat-transfer coefficient increases along the entire length of the pipe and the initial stabilization section disappears (curve 6, Fig. 1a).

Analysis of the data on the heat transfer of toluene and benzene with laminar flow and supercritical pressures shows that during both ascending and descending flow and with large temperature differences between the wall and fluid, free convection causes the heat-transfer coefficient to increase rather than decrease in the initial part of the pipe. In the final part of the pipe, the values of the criterion Nu become several times greater than in the initial part of the pipe.

During descending flow of toluene and benzene, we obtained the same character of change in the heat-transfer coefficient as during ascending flow of these hydrocarbons in a vertical pipe (Fig. 1c and d). The only difference is that during descending flow the effect of free convection on heat-transfer rate is seen at low values of Δt ($\Delta t > 60°C$) and, accordingly, at low values of Gr/Re (Gr/Re > 70). Thus, given the same value of Gr/Re, the pipe length on which free convection will have an effect is greater for an ascending flow than for a descending flow. This is apparent from Fig. 3. For both ascending and descending flows of toluene and benzene at high values of Gr/Re, the viscous-gravitational regime of convective heat transfer is seen along the entire pipe length and is accompanied by an increase in the heat-transfer coefficient along the pipe.

The difference in the character of change in the heat-transfer coefficient with different values of Gr/Re complicates the calculation of convective heat transfer along the entire pipe when the hydrocarbons are at supercritical pressures. When a viscous regime exists in the initial part of the pipe and a viscous-gravitational regime of convective heat transfer prevails in the final part, it is necessary to have two equations to calculate the heat-transfer coefficient. In such cases, heat transfer in the initial part of the pipe can be calculated from Eq. (1), while heat transfer in the final part can be calculated from equations which take into account the effect of free convection — such as the equations in [4]

$$\mathrm{Nu}_{q,d} = 0.0033\,\mathrm{Re}_{q,d}^{0.50}\,\mathrm{Pr}_{q}^{0.43}\,\mathrm{Gr}_{q,d}^{0.435}\left(\frac{d}{x}\right)^{0.40} \quad (3)$$

for ascending flow of the hydrocarbons and

$$\mathrm{Nu}_{q,d} = 0.0033\,\mathrm{Re}_{q,d}^{0.50}\,\mathrm{Pr}_{q}^{0.43}\,\mathrm{Gr}_{q,d}^{0.415}\left(\frac{d}{x}\right)^{0.40} \quad (4)$$

for descending flow.

However, such a method of calculation requires the establishment of boundaries of applicability for these equations. It is therefore necessary to determine the pipe length $X = \frac{1}{\mathrm{Pe}}\frac{x}{d}$, after which free convection affects heat transfer. Graphs of the function $\mathrm{Nu} = f\left(\frac{1}{\mathrm{Pe}}\frac{x}{d}\right)$ are used to determine the pipe length X_{cr} over which the criterion Nu has the lowest value. This data is analyzed in the form of the function $X_{cr} = f(\mathrm{Re/Gr})$ for ascending and descending flows (Fig. 3).

The following equations were obtained to determine X_{cr} on the basis of the test data on the heat transfer of toluene and benzene with laminar flow and supercritical pressures.

$$X_{cr} = 2.9\,(\mathrm{Re/Gr})^{1.10} \quad (5)$$

for ascending flow

$$X_{cr} = 0.78\,(\mathrm{Re/Gr})^{0.95} \quad (6)$$

for descending flow.

Equations (5) and (6) were obtained for laminar flow of toluene and benzene in a vertical 3-mm-diameter pipe of length $l/d \geq 75$ at a mass velocity ρu = 80-110 kg/m^2·sec, a fluid temperature at the pipe inlet of 15-25°C, and a wall-fluid temperature gradient Δt = 100-300°C for ascending flow and Δt = 70-250°C for descending flow. The Reynolds number changed from 350 to 2200 along the pipe and, accordingly, Gr/Re = 250-850 for the ascending flow and Gr/Re = 80-500 for descending flow.

NOTATION

p, pressure; T, t, temperature; q, heat flux; $\Delta t = t_w - t_q$, temperature difference; μ, ν, absolute and kinematic viscosities; ρ, density; u, velocity; ρu, mass velocity; x, distance from pipe inlet; d, pipe diameter; $Nu = \alpha d/\lambda_q$; $Re = \rho u d/\mu_q$; $Pr = \mu_q c_{p_q}/\lambda_q$; $Pe = Re\,Pr$; $Gr = \frac{\beta_q g d^3}{\nu^2}\Delta t$ respectively, the Nusselt, Reynolds, Prandtl, Peclet, and Grashof numbers; X, corrected pipe length. Indices: w, wall; q, liquid; cr, critical; e, experimental.

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INFLUENCE OF A MAGNETIC FIELD ON THE CONTACT WETTING ANGLE IN BUBBLE BOILING

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It is experimentally shown that an external magnetic field applied to a bubble boiling region improves wetting of the surface heating the boiling liquid.

Investigations of recent years have shown that one possible way to improve heat transfer in bubble boiling is to apply an electric or a magnetic field to the boiling region. Here one finds appreciable changes in the heat-transfer coefficient and the critical heat flux [1-3], in the density of vapor forming centers, separation size of the bubbles, and their separation frequency [3-5]. All of this indicates that if one knows the mechanism of the action of the field on the boiling, one can use an electric field as a means of controlling heat and mass transfer. However, in spite of the practical value of these effects, the mechanism of the influence of the field on bubble boiling has not yet been elucidated conclusively. It is suggested that the external field alters the structure of the heated boundary layer or of the double electric field. In [1] the intensifying action of the electric field on the heat transfer is explained by the generation of electroconvective fluxes, and according to the research data of [6] the electric field alters the contact wetting angle. Further investigations are needed to pinpoint the mechanism of the field action.

It has been established earlier that the contact wetting angle strongly influences the heat-transfer coefficient [7], the critical heat flux [8], the density of vapor forming centers and their stability, the separation dimensions of vapor bubbles, and the mechanism and frequency of their separation [9, 10]. The present author attempts to explain the influence of a magnetic field on the contact wetting angle and to evaluate the relative changes in the microparameters D_0, Θ under the action of this field.

The experimental equipment has been described in [5]. The boiler chamber is equipped with windows for motion pictures and visual observation of the boiling process, as well as an external heater to maintain the saturation temperature. Inside the chamber are mounted a heat-transfer surface, a device to create and control the magnetic field, and a system of thermocouples to measure temperatures. The magnetic field intensity varied over a range in-

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