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HEAT-TRANSFER EQUATION FOR LAMINAR FLOW OF AROMATIC HYDROCARBONS AT SUPERCRITICAL PRESSURES

F. I. Kalbaliev, F. K. Babaev,
and G. I. Isaev

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Experimental data on local heat-transfer coefficients for toluene and benzene ascending and descending in a vertical pipe is generalized in the form of criterional equations.

Determining the wall temperature of apparatus operating at supercritical pressures is one of the main problems of convective heat transfer, the solution of which has been the goal of numerous works on turbulent liquid flow. Heat transfer has been studied for laminar flow and supercritical pressure only for the aromatic hydrocarbons toluene and benzene [1-9] and polymethylphenylsiloxane liquid [10].

Tests on heat transfer involving aromatic hydrocarbons were conducted in a closed circuit on an experimental tube made of stainless steel with an inside diameter $d = 3$ mm, wall thickness $\delta = 0.5$ mm, and length of heated section $l = 200-220$ mm. The tests were conducted in the following parameter ranges:

for toluene

$$p/p_{\text{cr}} = 1.06 - 3.07,\ T_l/T_{\text{cr}} = 0.49 - 1.05,\ T_{\text{w}}/T_{\text{cr}} = 0.55 - 1.56,$$
$$q = (0.31 - 3.90)\cdot 10^5\ \text{W/m}^2,\ \text{Re} = 375 - 4200;$$

for benzene

$$p/p_{\text{cr}} = 1.21 - 2.63,\ T_l/T_{\text{cr}} = 0.52 - 1.12,\ T_{\text{w}}/T_{\text{cr}} = 0.44 - 1.55,$$
$$q = (0.12 - 4.5)\cdot 10^5\ \text{W/m}^2,\ \text{Re} = 320 - 5300.$$

Located ahead of the heated section of the tube is a hydrodynamic stabilization section of length $l_{h.s} = 0.06d\ \text{Re}$. The length of the initial, thermal section, at $p > p_{cr}$, $T_l \ll T_m$, and $T_w \ll T_m$, is roughly $(1/\text{Pe})(x/d) = 0.01$. At $p > p_{cr}$, $T_l < T_m$, and $T_w \gtrsim T_m$, the tempera-

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M. Azizbekov Azerbaidzhan Institute of Petroleum and Chemistry. Translated from Inzhenerno-Fizicheskii Zhurnal, Vol. 44, No. 3, pp. 379-383, March, 1983. Original article submitted July 6, 1981.

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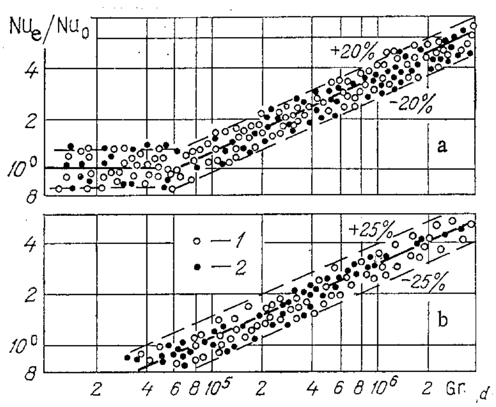

Fig. 1. Heat transfer in the laminar regime and at supercritical pressure for toluene and benzene at $T_l < T_m \leqslant T_w$ (*a* corresponds to the case of ascent; b — descent): 1) toluene; 2) benzene.

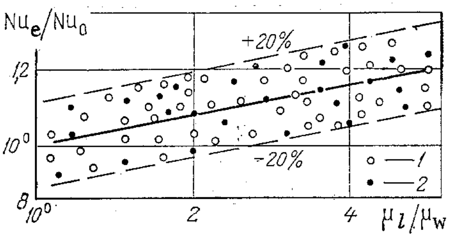

Fig. 2. Heat transfer of aromatic hydrocarbons with laminar flow regime and $p > p_{cr}$, $T_l < T_w < T_m$. Same notation as in Fig. 1.

ture of the wall along the tube changes nonmonotonically. Thus, the heat-transfer coefficient on the thermal section is different from the normal value. The change in the coefficient in this section and, accordingly, the length of the section at low Reynolds numbers and $p > p_{cr}$, $T_l < T_m$, and $T_w \geqslant T_m$ depends on the physical properties of the test liquids, the effect of free motion on heat transfer, and other factors.

The study of heat transfer involving toluene and benzene showed that the laws of convective heat transfer are different under laminar flow and supercritical pressures than at subcritical state parameters. In particular, heat transfer in laminar flow with $p > p_{cr}$ and $T_l \leqslant T_m \leqslant T_w$ is accompanied by pulsations in the liquid pressure and wall temperature. The wall temperature distribution along the tube is wavelike in character. The process is accompanied by improved heat transfer. Under these conditions, the heat transfer rate is affected significantly by the free motion of the liquid arising from mechanical changes in the density of the investigated hydrocarbons across and along the flow. The effect of the latter depends on the direction of motion of the liquid, as was confirmed by heat-transfer data obtained in tests with toluene and benzene rising and falling in a vertical pipe.

Figure 1 presents data on heat transfer in the nonisothermal flow of the hydrocarbons in the pipe. It follows from the figure that the effect of free motion on heat transfer rate is negligible in the case of ascent and $\mathrm{Gr}_{l,d} < 6 \cdot 10^4$ (corresponding to the conditions $T_l < T_w < T_m$).

Experimental data on local heat-transfer coefficients at $p > p_{cr}$ and $\mathrm{Gr}_{l,d} < 6 \cdot 10^4$ obtained for laminar flow of toluene and benzene in a vertical pipe of diameter d = 3 mm is described by an ordinary equation of convective heat transfer. For example:

$$\mathrm{Nu}_{l,d} = \mathrm{Nu}_0 \left(\frac{\mu_l}{\mu_w}\right)^{0.15}, \qquad (1)$$

where Nu_0 is determined from the equation in [11]:

$$\mathrm{Nu}_0 = 0.33\,\mathrm{Re}_{l,d}^{0.50}\,\mathrm{Pr}_l^{0.43}\left(\frac{d}{x}\right)^{0.40}. \tag{2}$$

Comparison of the calculated and experimental values of the heat-transfer coefficient, in Fig. 2, shows that Eq. (2) describes the test data to within ±20% up to $\mu_l/\mu_w = 5$. The heat transfer rate for rising aromatic hydrocarbons and $\mathrm{Gr}_{l,d} > 6\cdot 10^4$ (corresponding to $p > p_{cr}$, $T_l \leqslant T_m \leqslant T_w$) depends to a considerable degree on the effect of the free motion (see Fig. 1a).

Under the above conditions, heat transfer is accompanied by pulsations of liquid pressure and wall temperature. These effects are connected with a marked change in the physical properties of the test liquids in the transverse and longitudinal directions of the flow and the effect of this change on free motion. In individual tests, the physical properties changed in the flow cross section in the following ranges up to a wall temperature of 400°C: $\rho_l/\rho_w = 4\text{–}6$, $\lambda_l/\lambda_w = 1.2\text{–}1.5$. At a wall temperature above 420°C, the toluene and benzene undergo thermal degradation. Local heat-transfer coefficients of the hydrocarbons in the initial thermal section are described by the following equation up to a wall temperature of 420°C

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

This equation was obtained from the test data for a tube $d \leqslant 3$ mm at $\mathrm{Re}_{l,d} = 800\text{–}2300$, $\mathrm{Gr}_{l,d}/\mathrm{Re}_{l,d} = 75\text{–}1100$, and a temperature difference $\Delta T = 50\text{–}300$°C. When the hydrocarbons descend in the pipe, free motion affects heat transfer at $\mathrm{Gr}_{l,d} > 3\cdot 10^4$ (see Fig. 1b). The test data on local heat-transfer coefficients with descending toluene and benzene and $\mathrm{Gr}_{l,d} < 3\cdot 10^4$ is described by Eq. (1), while the following equation describes the data for $\mathrm{Gr}_{l,d} > 3\cdot 10^4$

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

This equation was obtained from the test data for $d \leqslant 3$ mm, $\mathrm{Re}_{l,d} = 700\text{–}2000$, $\mathrm{Gr}_{l,d}/\mathrm{Re}_{l,d} = 43\text{–}1500$, and $\Delta T = 50\text{–}300$°C and describes test data on heat transfer for toluene and benzene to within ±25%.

It is apparent from Fig. 1 that Eqs. (3) and (4) satisfactorily describe heat transfer by the aromatic hydrocarbons. It thus becomes unnecessary to introduce additional corrections into these equations in the form of physical property ratios calculated from the liquid and wall temperatures. It should be noted that heat transfer for turbulent flow of aromatic hydrocarbons is described well by the equation in [12]

$$\mathrm{Nu} = \mathrm{Nu}_0(\rho_l/\rho_w)^c, \tag{5}$$

in which the nonisothermal nature of the flow in the transverse direction is accounted for only by the ratio $(\rho_l/\rho_w)^c$. Evidently, the Gr number accounts for this effect to some degree in Eqs. (3) and (4). The quantity Nu_0 in Eq. (5) is determined from equations obtained for isothermal turbulent flow.

The lack of information on the thermophysical properties of liquid polymethylphenolsiloxane prevented us from checking the possibility of describing the experimental data in the work by Kaplan et al. [10] with Eq. (3).

NOTATION

p, pressure; T_w, T_l, temperature of wall and liquid, respectively; p_{cr}, T_{cr}, critical pressure and temperature; T_m, temperature corresponding to maximum heat capacity at $p > p_{cr}$; λ, thermal conductivity; C_p, heat capacity; α, heat-transfer coefficient; β, coefficient of cubical expansion; μ, ν, absolute and kinematic viscosities; q, heat flux; d, inside diameter of tube; x, length of section from tube inlet; $\mathrm{Nu} = \alpha d/\lambda$, $\mathrm{Re} = \rho u d/\mu$, $\mathrm{Pr} = \mu C_p/\lambda$, $\mathrm{Gr} = \beta g d^3 \Delta t/\nu^2$, Nusselt, Reynolds, Prandtl, and Grashof numbers, respectively.

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COOLING OF OVERHEATED SURFACES BY A FALLING LIQUID FILM

V. G. Ivanov, V. M. Kozlov,
and A. B. Musvik

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A method is presented for calculating the cooling of overheated surfaces by a falling liquid film. Comparison of the results with available empirical data shows that they agree well.

The problem of rewetting overheated surfaces under transient conditions with a moving wetting front has attracted the attention of researchers in recent years in connection with the problem of emergency cooling of the active zone of nuclear reactors [1, 2] and other equipment [3].

To keep the temperature of fuel elements in the core of a nuclear reactor below the tolerable level during emergency loss of coolant, reactor designers usually provide an emergency cooling system which injects water or floods the area with water from below.

When coolant is delivered to the surface of fuel elements, a water film is formed and flows down the element at a certain velocity.

The physical reality of the rewetting process is complicated, since the initial temperature of the fuel element wall is fairly high and it is not wetted by the liquid. Here, a so-called cooling front is formed. This front advances over the surface as it cools to the temperature at which the liquid begins to wet the surface. In the literature [4], this temperature is known as the Leidenfrost point. It was shown in [5, 6] that this point is a constant value at constant pressure. Thus, the problem of cooling a heated surface reduced to studying the laws of the process of advance of an isotherm corresponding to the Leidenfrost

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