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HEAT EXCHANGE ALONG THE INITIAL SEGMENT IN A HEAT EXCHANGER WITH A HELICAL FLOW

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UDC 536.24

The characteristics of local heat transfer along the initial segment and along the segment with stabilized air flow in longitudinal flow past a bundle of coiled oval tubes are established

The characteristics of heat transfer in heat exchangers with longitudinal flow past a bundle of coiled oval tubes were examined in a number of papers [1-4], wherein it was shown that the observed increase in heat transfer is explained by the properties of helical flows in channels with a complicated shape. An equation was proposed in [2, 3] that describes the heat-transfer process in a stabilized turbulent flow in the range of numbers $Fr_m = 232-2440$:

$$\overline{Nu} = 0.023 Re^{0.8} Pr^{0.4} (1 + 3.6 Fr_m^{-0.357}) (T_w/T_f)^{-0.55}, \tag{1}$$

where

$$Re = u_{mm} d_e \rho/\mu, \tag{2}$$

$$Fr_m = S^2/dd_e, \tag{3}$$

which differs from the equation for circular pipes [5]

$$Nu = 0.023 Re^{0.8} Pr^{0.4} (T_w/T_f)^{-0.55} \tag{4}$$

by a factor that depends on the number Fr_m. For Fr_m numbers less than 100, the heat-transfer coefficient increases to a larger extent than follows from Eq. (1). Thus, for values of the number $Fr_m = 64$, heat transfer in a bundle of coiled tubes can be described by the function [1]

$$\overline{Nu} = 0.0521 Re^{0.8} Pr^{0.4} (T_w/T_f)^{-0.55}. \tag{5}$$

However, Eqs. (1) and (5) describe only the locally averaged heat transfer, since along the segment with stabilized flow, the experimental data are observed to separate within a range of approximately ±15%, as noted in [1-3], which must be taken into account in the heat-transfer law. In addition, it is necessary to establish the law governing the change in the heat-transfer coefficient along the initial segment of the flow. This paper is concerned with solving these problems.

The heat transfer was investigated using a generally accepted technique on experimental setup described in [1] with air as the heat transfer agent. The heat exchangers consisted of 37 coiled tubes of length 500 and 750 mm and the porosity for the heat-transfer agent was $m = 0.527-0.544$. The tubes were made of Kh18N10T steel. The tubes were heated by passing an alternating electric current through them from a OSU-100 transformer, controlled by AOMK-180 autotransformer. Chromel–Alumel thermocouples, welded to the interior of the central tube in the bundle at five sections along its length, were used to measure the temperature of the walls of the tubes. The densely packed lattice of the bundle had an ordered structure with the coiled tubes touching one another at the output part of the bundle along the long axis of the oval tube profile. The flow of the heat-transfer agent entered the bundle axisymmetrically. The measuring system permitted determining the Nusselt number with a limiting relative error of ± 7% with the following range of parameters: $S/d = 6.2-34$; $Fr_m = 64-2440$; $Re = 2 \cdot 10^3 - 4 \cdot 10^4$; $T_w/T_f = 1.0-1.73$; and $x/d_e = 3.75-103$.

The nature of the change in the heat-transfer coefficient along the length of the bundle of coiled tubes can be seen in Fig. 1, where the results of the experimental investigation of heat transfer in the bundle with a number $Fr_m = 924$ is shown as the functional relation:

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$$\mathrm{Nu}\,(T_w/T_f)^{0.55} = f\,(\mathrm{Re},\ x/d_e), \tag{6}$$

where the number $\mathrm{Nu} = \dfrac{q_0 d_e}{(T_w - T_{mm.f})\,\lambda}$.

It follows from Fig. 1 that the experimental data on heat transfer for different x/d_e are distributed in an equidistant manner, i.e., the Re number has practically no effect on the relation of the Nu number to the quantity x/d_e for turbulent and transient flow regions. The observed separation of the experimental data on heat transfer for different x/d_e in a ±15% range of $\overline{\mathrm{Nu}}$ can be explained by the difference in the flow conditions for the heat-transfer agent past different regions of the coiled tube, where thermocouples are placed, depending on the mutual position of the coiled tubes in the bundle [6, 7]. For the mutual positions of the tubes in the bundle examined, the thermocouples could be located on the tube walls according to the maximum size of the oval either in the region of the throughput channels of the bundle or at the point at which neighboring tubes touch or in intermediate regions of flow past the tubes [6, 7]. Since the values of the local velocity of flow past the tubes in the regions indicated are different [6, 7], while in [2, 3] the flow-determining factor was taken as the mean mass velocity in the cross section of the bundle examined, with the technique used to analyze the experimental data, it must be expected that the variations in the Nu number must be periodic along the bundle relative to the value $\overline{\mathrm{Nu}}$, determined by Eqs. (1) or (5). Thus, at a distance $x/d_e = 20$ from the bundle inlet with $\mathrm{Fr}_m = 9.24$ (Fig. 1), the coefficient of heat transfer is less than at a distance $x/d_e = 36.2$, while at distances $x/d_e = 52.5$ and 59.3, it is lower than at $x/d_e = 20$. If the origin for the longitudinal coordinate l is placed in the outlet section of the bundle, while the positive direction for the coordinate l is taken as the direction upwards along the flow, then, introducing the relative coordinate l/S, it is possible to generalize the experimental data on local heat transfer for bundles of coiled tubes with different Fr_m numbers using the following function:

$$\mathrm{Nu}/\overline{\mathrm{Nu}} = 1 + 0.15\cos\frac{2\pi l}{S}. \tag{7}$$

The spread in the experimental data observed in Fig. 2 relative to the relation (7) can be explained by the effect of the tolerances in step size of the coiled tubing and the porosity of the bundle with respect to the heat-transfer agent. The closer the real porosity of the bundle to the porosity of the bundle with densely packed tubes, the closer is the correspondence between the experimental data and the function (7). The functional relation (7) is valid only for the ordered packing of tubes examined, although the range of variation of the heat-transfer coefficient, equal to ±15%, is practically independent of the nature of the packing of the tube bundle. Then, local heat transfer with stabilized turbulent flow in bundles of coiled tubes can be described by the equation

$$\mathrm{Nu} = 0.023\left(1 + 0.15\cos\frac{2\pi l}{S}\right)\mathrm{Re}^{0.8}\,\mathrm{Pr}^{0.4}\,(1 + 3.6\mathrm{Fr}_m^{-0.357})\,(T_w/T_f)^{-0.55}. \tag{8}$$

The data on heat transfer, in accordance with [2, 3], in bundles of coiled tubes can also be represented in a more convenient form, when the effective thickness of the near wall layer δ, representing the integral geometric characteristic of the bundle, is used as the characteristic size:

$$\delta = 0.5\,(1 + 3.6\mathrm{Fr}_m^{-0.357})^{-4}\,d_e \tag{9}$$

and the average temperature of the near-wall layer is used as the characteristic temperature (Fig. 3). In this case, the locally average heat transfer with stabilized turbulent flow is described by the heat-transfer law

$$\overline{\mathrm{Nu}}_{\delta m} = 0.020\,\mathrm{Re}_{\delta m}^{0.8}\,\mathrm{Pr}_m^{0.4}, \tag{10}$$

where

$$\overline{\mathrm{Nu}}_{\delta m} = q_0\delta/\lambda_m\,(T_w - T_{mm.f}), \tag{11}$$

$$\mathrm{Re}_{\delta m} = \frac{\rho_m\delta\int_F u\,dF}{\mu_m F}, \tag{12}$$

ρ_m, μ_m, and λ_m are the density, viscosity, and thermal conductivity, determined at the average temperature over the thickness of the near-wall layer $T_m = (T_w + T_f)/2$. The heat-transfer law (10) is analogous to the equation for calculating heat transfer in circular tubes using the tube radius as the characteristic size [8]. Local heat transfer with stabilized turbulent flow in bundles of coiled tubes, taking into account the characteris-

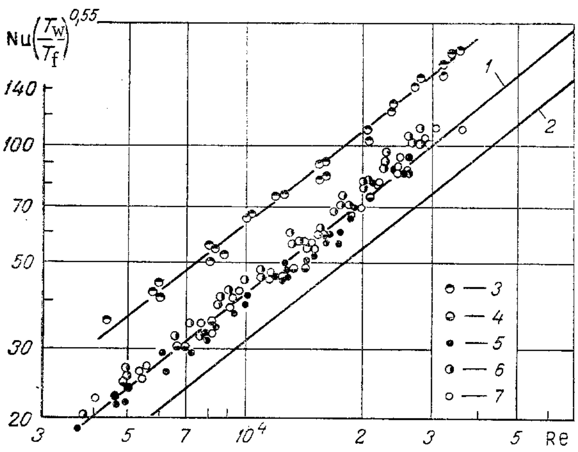

Fig. 1

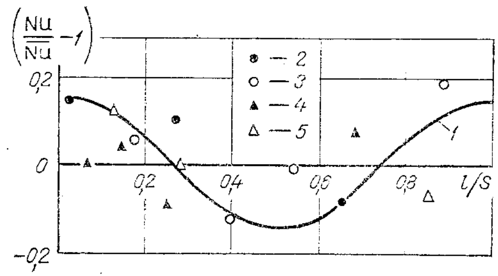

Fig. 2

Fig. 1. Experimental data on heat transfer in the space between the tubes in a heat exchanger with $Fr_m = 924$: 1) Eq. (1); 2) (4); 3-7) experimental data with $x/d_e = 3.75$; 59.3; 52.5; 36.2; and 20, respectively.

Fig. 2. Effect of mutual position of coiled tubes on the heat-transfer coefficient: 1) dependence (7); 2-5) experimental data with $Fr_m = 64$, 232, 924, and 1050, respectively.

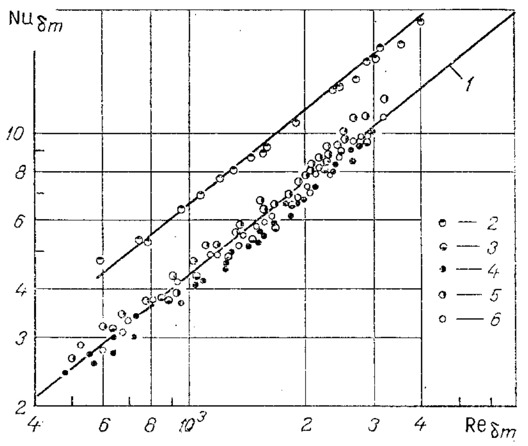

Fig. 3

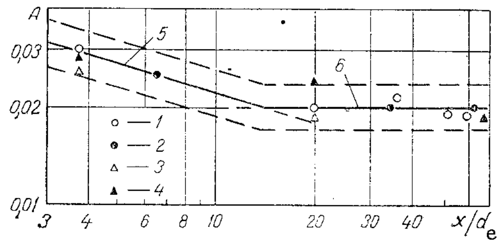

Fig. 4

Fig. 3. Generalizing dependence for heat transfer in bundles of coiled tubes: 1) heat-transfer law in bundles of coiled tubes: 1) heat-transfer law (10); 2-6) experimental data for $x/d_e = 3.75$, 59.3, 52.5, 36.2, and 20, respectively.

Fig. 4. Effect of the length of the initial segment on the local heat-transfer coefficient: 1-4) experimental data with $Fr_m = 924$, 2440, 1050, and 232, respectively; 5) dependence (15); 6) the heat-transfer law (10). $A = c_m = \frac{\overline{Nu}\,\delta_m}{Re_{\delta m}^{0.8}\,Pr_m^{0.4}}$

tics of the flow past separate segments of the tubes for the method proposed for analyzing the experimental data, can be calculated according the equation

$$Nu_{\delta m} = 0{,}020\left(1 + 0.15\cos\frac{2\pi l}{S}\right) Re_{\delta m}^{0.8}\,Pr_m^{0.4}. \tag{13}$$

The experimental data presented in Figs. 1 and 3 also indicate the higher heat-transfer coefficient on the initial segment of the bundle of coiled tubes compared to the heat-transfer coefficient on the segment with the stabilized flow. In analyzing the experimental data in the form $Nu_{\delta m} = Nu(Re_{\delta m}, Pr_m, x/d_e)$, the effect of the initial segment on the heat-transfer coefficient can be taken into account by introducing into Eq. (10), instead of a constant factor, the expression

$$c_m = \overline{\mathrm{Nu}}_{\delta m}/\mathrm{Re}_{\delta m}^{0.8}\,\mathrm{Pr}_m^{0.4} = f\,(x/d_e). \tag{14}$$

Then, the experimental data for bundles of coiled tubes with different $\mathrm{Fr_m}$ numbers (Fig. 4) can be generalized by a power law function

$$c_m = 0.0426\,(x/d_e)^{-0.287}\,. \tag{15}$$

In this case, the criterional dependences on heat transfer for $x/d_e = 3.75$–14 will have the form

$$\overline{\mathrm{Nu}}_{\delta m} = 0.0426\,(x/d_e)^{-0.287}\,\mathrm{Re}_{\delta m}^{0.8}\,\mathrm{Pr}_m^{0.4}\,, \tag{16}$$

$$\mathrm{Nu}_{\delta m} = 0.0426\,(x/d_e)^{-0.287}\left(1 + 0.15\cos\frac{2\pi l}{S}\right)\mathrm{Re}_{\delta m}^{0.8}\,\mathrm{Pr}_m^{0.4}. \tag{17}$$

In analyzing the experimental data on heat transfer in the form

$$\mathrm{Nu} = \mathrm{Nu}\,(\mathrm{Re},\ \mathrm{Fr_m},\ x/d_e,\ T_w/T_f) \tag{18}$$

in Eqs. (1) and (8) the constant factor must be replaced by the expression

$$c = 0.0490\,(x/d_e)^{-0.287} \tag{19}$$

in calculating the local heat transfer along the initial segment ($x/d_e = 3.75$–14).

In the transitional region of the stabilized flow with a $\mathrm{Re}_{\delta m}$ number less than 500 or with the number

$$\mathrm{Re} < 2\mathrm{Re}_{\delta m}\,(1 + 3.6\mathrm{Fr_m}^{-0.357})^{4}\,(T_w + T_{\mathrm{mm.f}})/2T_{\mathrm{mm.f}} \tag{20}$$

the heat-transfer coefficient in the bundle of coiled tubes is determined by the equations [3]:

$$\overline{\mathrm{Nu}} = 83.5\mathrm{Fr_m}^{-1.2}\,\mathrm{Re}^n\,\mathrm{Pr}^{0.4}\,(T_w/T_f)^{-0.55}\,(1 + 3.6\mathrm{Fr_m}^{-0.357}), \tag{21}$$

$$\overline{\mathrm{Nu}}_{\delta m} = 6.47\mathrm{Fr_m}^{-0.845}\,\mathrm{Re}_{\delta m}^{n}\,\mathrm{Pr}_m^{0.4}\,, \tag{22}$$

where

$$n = 0.212\mathrm{Fr_m}^{0.194}\,, \tag{23}$$

which, taking into account the characteristics of heat transfer along the initial segment, will have the form

$$\mathrm{Nu} = 178\,(x/d_e)^{-0.287}\left(1 + 0.15\cos\frac{2\pi l}{S}\right)\mathrm{Fr_m}^{1.2}\,\mathrm{Re}^n\,\mathrm{Pr}^{0.4}\,(T_w/T_f)^{-0.55}\,(1 + 3.6\,\mathrm{Fr_m}^{-0.357}), \tag{24}$$

$$\mathrm{Nu}_{\delta m} = 13.8\,(x/d_e)^{-0.287}\left(1 + 0.15\cos\frac{2\pi l}{S}\right)\mathrm{Fr_m}^{-0.845}\,\mathrm{Re}_{\delta m}^{n}\,\mathrm{Pr}_m^{0.4}\,, \tag{25}$$

while along the segment with stabilized flow for $x/d_e > 14$, the values will differ by the factor

$$\left(1 + 0.15\cos\frac{2\pi l}{S}\right).$$

The results obtained indicate the fact that on the initial segment of the bundle both in the turbulent and transitional regions of the flow, the nature of the effect of the temperature factor T_w/T_f on heat transfer is the same, similar to the effect of the temperature factor on the stabilized flow, and does not depend on the $\mathrm{Fr_m}$ number. It was also discovered that the length of the initial segment in the bundles of coiled tubes with axisymmetrical flow input is $x_i/d_e = 14$ and is practically independent of the $\mathrm{Fr_m}$ number.

Based on the above presentation, we can conclude that the results of the study carried out can be used for calculating local heat transfer along the initial segment and along the segment with stabilized flow in heat exchangers with longitudinal flow past bundles of coiled tubes with axisymmetrical flow input of the heat-transfer agent.

NOTATION

Fr_m, modified Froude number, characterizing the action of centrifugal forces on the flow; Nu, local Nusselt number; $\overline{Nu}$, locally averaged Nusselt number; Re, Reynolds number; Pr, Prandtl number; T_w, temperature of the wall; T_f, flow temperature; u_{mm}, mean mass velocity; d_e, equivalent diameter; S, pitch of the tube coil; d, maximum size of the oval; ρ, density; μ, viscosity; δ, effective thickness of the near-wall layer; x, longitudinal coordinate, measured from the flow inlet into the bundle of tubes; l is the longitudinal coordinate, measured from the outlet of the bundle upwards along the flow; q_0, specific heat flux; λ, thermal conductivity; $T_{mm.f}$, mean mass temperature of the flow; F, area of the throughput section of the bundle.

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HEAT TRANSFER TO AN EMULSION WITH HIGH SUPERHEATING OF ITS DISPERSE PHASE

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A method is described, and results presented, for measurement of the heat-transfer coefficient to an emulsion consisting of ether dispersed in glycerin.

In heat-treating metals, it is necessary to control the cooling rate. The latter depends on the dimensions and thermophysical properties of the specimen, as well as on the heat-transfer coefficient α. The value of α is determined by the velocity and properties of the coolant. If the specimen dimensions and material are given, then the coefficient α depends only on the properties of the coolant. Pure liquids and their mixtures, such as emulsions, are most frequently used at low temperatures. In this case, the cooling rate can be additionally controlled by changing the concentration of the components.

The present article attempts to measure the coefficient of heat transfer to an emulsion when droplets of the disperse phase are possibly superheated on the heat-emitting surface [1].

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