VARIATION OF HEAT-TRANSFER INTENSITY AROUND THE PERIMETER OF A HORIZONTAL TUBE IN A FLUIDIZED BED

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An experimental investigation has been made of the variation of local heat-transfer coefficient in the cross section of a horizontal tube in a fluidized bed by the method of measuring local temperature difference and local heat flux power.

The conditions of flow of a fluidized medium over a horizontal tube are known to be different in different zones, and one may therefore expect variation of the local heat-transfer coefficient α around the section of the tube for any velocity ω of the fluidizing agent. The experiments of Planovskii and Nikolaev [1] first demonstrated the difference in α values in the back, side, and front zones of a horizontal tube.* These tests, performed on thin-walled stainless steel tubes, (φ 30 mm) were limited by the fluidization number $W \leq 3.66$ (no maximum value of α_m was attained). The authors observed the largest α values on the sides of the tube, lesser values on the front part, and least on the back part. With increase of air velocity, the differences of α at the various points on the perimeter of the tube section diminished from 65% near the onset of fluidization to 35% when W = 3.66. To determine the local heat-transfer coefficient, the authors of [1] measured the local temperature differences Δt between the fluidized bed and different points around the tube section. Further, the desired α values were calculated from the known quantity Q_m of heat transferred (the power of the electric heater inside the tube) and the tube surface area. It was assumed there that transfer of heat by conduction around the tube section along the wall does not introduce appreciable error because of the small thermal conductivity of the wall. Moreover, the authors ignored the possible asymmetry of heat flux, thus causing some inaccuracy in the results obtained.* Evidently, the various local temperature differences between the bed and points on the tube surface correspond to different heat fluxes. Thus, calculation of the heat-transfer coefficient according to the equation

$$\alpha = Q_m / F \Delta t \tag{1}$$

is not quite legitimate (Q_m here is the total heat flux; F is the tube surface area). It is methodologically correct to calculate α according to the formula

$$\alpha = Q_l / f \Delta t, \tag{2}$$

where Q_l is the local quantity of heat arriving at the surface region f under examination. In the general case here** $Q_l/f \neq Q_m/F$. For the reason described, the numerical values of [1] and the character of the distribution of α presented by the author around the tube section should be corrected.

A similar investigation was made in the Processes and Equipment Department of the Lomonosov Institute of Fine Chemical Technology. With the aim of obtaining truly local values of α, local values of Q_l were measured for fluidization by air of a bed of quartz sand (d_e = 0.35 mm; shape factor 0.84; velocity of onset of fluidization ω_0 = 0.13 kg/m^2 · sec) in an equipment of 380 × 380 mm area. The height of the static bed H_0 was 500 mm; the distance of the tube from the perforated distributor with an effective cross section of 1% was 435 mm. The experiment encompassed the maximum values of α_m.

*This has been studied in the literature from the standpoint of self-screening [2, 3].

*A similar error occurs in [4].

**Similar arguments must be taken into account when measuring instantaneous values of the heat transfer coefficient [5, 6].

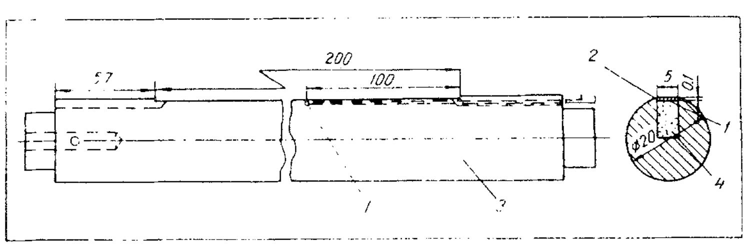

Fig. 1. Schematic of element: 1) hot junction of differential thermocouple; 2) nichrome ribbon; 3) wooden rod; 4) glass fiber.

The element for measuring local heat-transfer coefficient was a wooden rod 20 mm in diameter in which was milled a channel of 200-mm length and 5-mm width (Fig. 1). The channel was packed with glass fiber (with very low thermal conductivity). At the top of the channel there was a nichrome ribbon of thickness 0.1, width 5, and length 200 mm with the hot junction of a differential copper-constantan thermocouple soldered to it (facing the glass fiber). The thermocouple cold junction was located in the bed at a distance of 10 mm from the end of the tube. The ends of the thermocouple were led off to a potentiometer. The method of mounting of the element allowed it to be rotated about its axis and to be fixed every 30° in 12 positions, with an accuracy of not less than 3°.

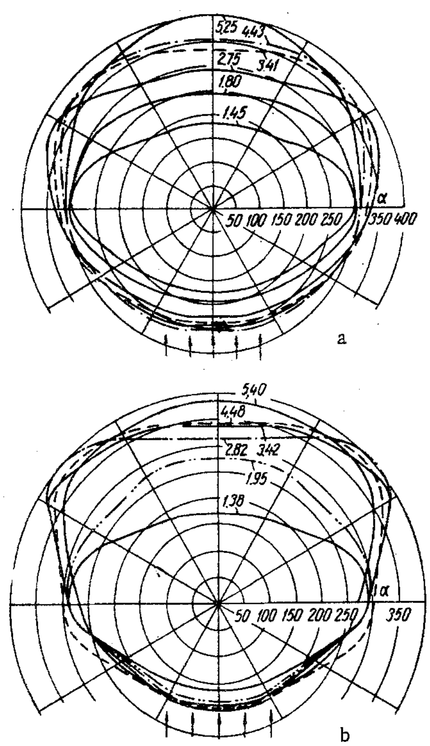

Fig. 2. Distribution of heat transfer coefficient α (kcal/m^2 · hr · °C) around the cross section of a tube 20 mm in diameter: a) single tube; b) central tube in a horizontal row (pitch 40 mm).

When electric current was passed through the nichrome ribbon, practically all the heat was given off through the exposed surface of the ribbon to the fluidized bed (heat losses by conduction to the wood were minimum and did not exceed 2–5%). The temperature difference between the ribbon and the bed was kept constant during the tests, equal to 12.0° C. Thus, to each value of the heat-transfer coefficient at any of the 12 positions there corresponds a known value of the heat (electrical) load on the nichrome ribbon, which permits calculation of local α values, using (2), where f is the area of the nichrome ribbon.

It may be seen from Fig. 2a that, for small fluidization numbers, the values of the heat-transfer coefficient in the front and back* are approximately equal. On the sides of the tube, as in [1], the greatest values of α were observed (the difference reached ~70%), since it is precisely here that the air, passing around the horizontal rod, brings the particles into quite intense motion. In contrast to the above-mentioned paper, however, with increase of fluidization number, the heat-transfer coefficient at the front part first increases sharply due to increased intensity of motion of the particles, and then flattens off and even decreases somewhat. The latter is due to increase in the fraction of time of contact of the surface with the gas bubbles and to the formation of a zone with a low concentration of solid phase in the front part under the tube. Similarly, but in a considerably smaller range, there is variation of the heat transfer coefficient on the sides of the tube; with increase of gas velocity the intensity of particle motion first increases (although to a less degree), but then the effectiveness of contact of the walls with the gas bubbles begins to play a part.

In the back of the tube section, owing to increased mobility of particles forming a "cap," the heat-transfer coefficient increases with increase of velocity of the fluidizing agent. At considerable air velocities it even exceeds α on the sides of the tube. It may be assumed that, for a considerable increase in ω, a decrease in local values of α will be observed even in this region.

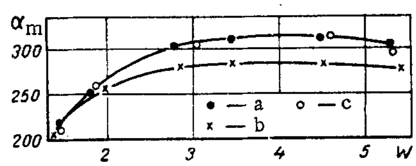

Fig. 3. Dependence of mean integral heat transfer coefficient α_m (kcal/m^2 · hr · °C) on fluidization number W: a and b) for a single tube and the central tube of a horizontal row (pitch 40 mm) from measurement of local heat transfer coefficient and subsequent integration over the circle; c) for a single copper tube of ∅ 20 mm with an internal electric heater and direct measurement of total heat-transfer coefficients.

Figure 2b shows the distribution of α around the tube in the case in which the element is located in the center of a horizontal row of tubes with pitch, i.e., the distance between tube axes, of 40 mm (the location of the tube in the bed was not changed). From comparison of Fig. 2a and Fig. 2b it may be seen that the presence of adjacent tubes in the horizontal row (and

*Values of α obtained at points symmetrical relative to the vertical axis were averaged.

evidently a decrease in pitch) leads to reduction of the heat-transfer coefficient, although the general nature of the variation of α around the tube section is preserved. It should be noted, however, that the presence of the adjacent tubes leads to some displacement of the greatest local values of α toward the back of the tube section. This is evidently due to increase of local air velocity (with unchanged velocity computed in the empty equipment) owing to the considerable constraint of the gas stream by the row of horizontal tubes.

A comparison is made in Fig. 3 of the mean integral values α_m for a single tube and for a tube in a horizontal row.

Thus, the experiment performed permits determination of the true profile of the coefficients of heat transfer between a horizontal tube and a fluidized bed. The question of the influence of the diameter of the horizontal tube on the nature of the distribution of α over the perimeter of the cross section must be left for later elucidation.

Notation

α—coefficient of heat transfer between surface and fluidized bed; ω—velocity of fluidizing agent based on empty equipment; w—fluidization number; α_m—mean value of heat-transfer coefficient; Δt—temperature difference between surface and bed; F—area of heat-transfer surface; f—a zone, part of the heat-transfer surface; d_e—equivalent diameter of fluidized particles; ω_0—velocity at onset of fluidization; H_0—height of bed at velocity ω_0.

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