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MATHEMATICAL MODEL AND THERMOHYDRAULIC CHARACTERISTICS OF PACKED SCRUBBERS OF CONDENSATION COOLING OF A GAS

A. G. Laptev and E. A. Lapteva UDC 66.048.37+662.986

The authors have solved scientific and technical problems of mathematical modeling and calculation of thermal efficiency, and also of structural characteristics of packed scrubbers of condensation water cooling of a gas in a film countercurrent regime. By simultaneous solution of the heat-balance equation and the expressions for thermal efficiencies of a packed scrubber in the gas and liquid phases, a relation has been established between the thermal efficiencies of condensation cooling of the gas and heating of the water. With an assigned temperature regime of cooling of the gas and its thermodynamic state, a required gas-phase thermal efficiency is computed. To calculate the actual thermal efficiency of a packed bed, use is made of a cellular model of the flow structure in the liquid and gas phases. An expression has been obtained for calculating the height of the packed bed from the assigned thermal efficiency, structural characteristics of the packing, and flow rates of the gas and water. Agreement with the existing experimental data has been shown and a calculation algorithm has been given. Conclusions on the most efficient packing structures have been drawn.

Keywords: *thermal efficiency, random and regular packings, flow structure, film scrubbers.*

Introduction. The cooling of gases is a component part of many technological processes in power engineering and chemical engineering. Here, problems of recuperation of heat are also solved [1–3]. To organize the process of cooling, use is made of various contact apparatuses, including packed countercurrent scrubber economizers in a film regime of interaction of the hot gas with the cooling liquid, most frequently water. Here, the water is heated and is subsequently used in technological processes or for domestic needs. Such apparatuses have a relatively simple structure and a high capacity at a low hydraulic resistance when modern contact devices, i.e., irregular or regular packings, are used. In calculating heat transfer in packed or other apparatuses, use is more frequently made of a model of ideal displacement of flows [1, 4–6], i.e., without the reverse mixing of the gas and the liquid. Thus, the calculated packing height may turn out to be smaller than the required one. Developing engineering methods of calculation of packed scrubbers with novel highly efficient packings with account taken of the flow structure will make it possible to design or modernize them with minimum expenditures of time and funds.

The present work seeks to develop the approach [7] obtained for film-type cooling towers in solving the noted problems of water cooling of the gases in scrubbers with various types of packings with account taken of the flow structure.

Mathematical Model. The basic problem in designing or modernizing heat- and mass-exchange apparatuses is to determine the efficiency of the implemented processes at assigned regime and structural parameters. The inverse problem on selecting structural characteristics of an apparatus at an assigned efficiency and assigned regimes of implementation of the process can also be solved.

The processes of cooling of gases in contact heat transfer are much more complicated than in the surface heat exchanger, since coupled heat and mass transfer occurs. This process has been considered in [1] in greater detail.

In a packed scrubber, we can single out three characteristic zones where contact cooling of the gas occurs. The first zone with height H_0 (Fig. 1) is in the lower part where the entry of the gas is organized and water flows down from the lower part in the form of jets and droplets. In the zone with a packing (basic zone) of height H, we have film flow of the water with contact in countercurrent with the gas. In the upper zone of height H_1, the gas leaves the packed bed, and the liquid phase is fed to the packed bed via a distributor (sprayers or other devices). In the zone with a packing, the contact surface of the

Kazan State Power Engineering University, 51 Krasnosel′skaya Str, Kazan, 420066, Republic of Tatarstan, Russia; email: tvt_kgeu@mail.ru. Translated from Inzhenerno-Fizicheskii Zhurnal, Vol. 95, No. 1, pp. 259–266, January–February, 2022. Original article submitted August 10, 2020.

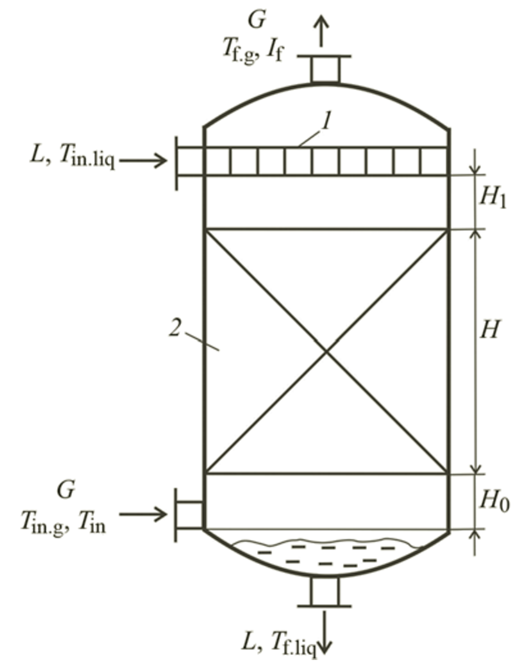


Fig. 1. Diagram of the packed scrubber: 1) distributor of the liquid phase; 2) packed bed.

phases is much larger than in the upper and lower zones with droplet-jet contact. Here, the estimates show that more than 90% of the heat is transferred from the gas to the liquid in the packed zone. However, here the condition that $H >> H_0$ and $H >> H_1$ must be observed, otherwise, we should take into account the heat transfer in the upper and lower zones where there is no packing.

Thus, in forming a mathematical model, we have made the following assumptions:

(1) basic heat and mass transfer (more than 90%) is in the packed zone in a film regime;
(2) basic resistance to heat and mass transfer is concentrated in the gas phase (more than 90%);
(3) phase motion is stationary and stabilized;
(4) the regime of gas motion in the packed bed is turbulent;
(5) the separation and removal of water droplets from the surface of the flowing-down water film is minor.

The transfer of heat in condensation gas cooling ($x > x^*$) occurs due to the contact of the phases (convective and molecular mechanisms of transfer), and also due to the condensation of moisture from the gases onto the surface of the flowing-down water film; this may ensure cooling below the dewpoint. The heat balance at an evaporation coefficient of unity is written in the form

$$Q = Gc_{pg}(T_{in.g} - T_{f.g}) + G(I_{in}x_{in} - I_f x_f) - G(x_{in} - x_f)c_{pliq}T_{f.liq} . \tag{1}$$

Also, the flux of heat transferred from the gas into the liquid phase may be represented by using the equations of heat and mass transfer in countercurrent.

The heat flux during the liquid cooling of the gas in countercurrent phase motion in the column is equal to

$$Q = KF(T_g - T_{liq}) + I_{st}r_g F \beta_g (x - x^*) . \tag{2}$$

In liquid cooling of the gas, the entire heat-transfer resistance, in practice, is concentrated in the gas phase, then we have $K = \alpha_g$.

Applying the Lewis analogy in the gas phase and the known expressions for the enthalpy, that are associated with the gas temperature and the moisture content, we can write expression (2) in the form [8]

$$Q = \beta_x F \Delta I_{av} . \tag{3}$$

Representing the heat-balance equation in the form

$$Q = G(I_{in} - I_f) = \beta_x F \Delta I_{av} , \tag{4}$$

we can compute the required area of contact of the phases F in an assigned temperature regime of cooling of the gas.

Expression (4) has been written with the condition of ideal displacement of the gas, which does not necessarily correspond to the actual hydrodynamic situation in packed columns with various structures of contact elements. It is common knowledge that reverse mixing of the flow reduces the motive force of heat- and mass-transfer processes and the required value of F will be greater than that yielded by Eq. (4). To take account of the reverse mixing, use is made of diffusion and cellular models of the flow structure. Next, we show an example of using the cellular model to calculate the efficiency of cooling of the gas in a trickled packed bed in a film regime.

The thermal efficiency of the process in the phase gas will be written in the form

$$E_g = \frac{T_{in.g} - T_{f.g}}{T_{in.g} - T_{in.liq}} . \tag{5}$$

The thermal efficiency is also defined as the ratio of the difference of enthalpies

$$E_g = \frac{I_{in} - I_f}{I_{in} - I_f^*} . \tag{6}$$

The E_g values have been written from formulas (5) and (6) on the basis of the greatest possible indices of cooling of the gas that were obtained in actual practice.

We write the equations of heat balance

$$Q = Lc_{pliq}(T_{f.liq} - T_{in.liq}) + Q_{ev} = G(I_{in} - I_f) , \tag{7}$$

of the thermal efficiency of water heating

$$E_{liq} = \frac{T_{f.liq} - T_{in.liq}}{T_{in.g} - T_{in.liq}} \tag{8}$$

and of the efficiency of water evaporation

$$E_x = \frac{x_{in} - x_f}{x_{in} - x_f^*} . \tag{9}$$

From expressions (6)–(8), we obtain a relationship between the thermal efficiencies for the liquid and gas phases in the form

$$\frac{E_{liq}}{E_g} = \frac{\rho_g w_g (I_{in} - I_f^*)}{\rho_{liq} c_{pliq} q_{liq} (T_{in.g} - T_{in.liq}) + Q_{ev}/S} . \tag{10}$$

With account of the thermal efficiency (9), we find the heat flux with evaporated water from the expression

$$Q_{ev} = Gc_{pliq} T_{f.liq} E_x (x_{in} - x_f^*) . \tag{11}$$

The value of Q_{ev} is much less than Q and may be disregarded.

The outlet temperature of the gas is calculated from the existing expression for the enthalpy

$$T_{f.g} = \frac{I_f - r_0 x_f}{c_{pwat} + c_{pst} x_f} . \tag{12}$$

By analogy with the theory of mass-transfer processes, from expression (4) we write the ratio

$$\frac{\beta_x F}{G} = \frac{I_{in} - I_f}{\Delta I_{av}} = N_g . \tag{13}$$

The contact surface of the phases in the packed bed is defined as $F = SH_{\text{in}}a_\text{v}\psi_\text{w}$, and the flow rate of the gas, as $G = w_0 S_\text{f}\rho_\text{g}$. Then the number of transfer units of (13) will be written in the form

$$N_\text{g} = \frac{\beta_x H a_\text{v}\psi_\text{w}}{\rho_\text{g} w_0} . \qquad (14)$$

According to the cellular flow-structure model, if the number of cells of full mixing is known for both the gas phase (n) and the liquid one (m), for the countercurrent at $n > m$ it is written [9]

$$E_\text{g} = 1 - \left(1 + \frac{N_\text{g} m}{n}\right)^{-n/m} . \qquad (15)$$

Hence the height of the packed bed at $E_\text{g} < 1$ is equal to

$$H = \frac{w_0 n \rho_\text{g}}{m \beta_x a_\text{v}\psi_\text{w}} \left[\left(\frac{1}{1 - E_\text{g}}\right)^{m/n} - 1\right]. \qquad (16)$$

At $m > n$, we have

$$E_\text{g} = 1 - \left(1 + \frac{N_\text{g} n}{m}\right)^{-m/n} . \qquad (17)$$

Then the height of the packed bed is

$$H = \frac{w_0 m \rho_\text{g}}{n \beta_x a_\text{v}\psi_\text{w}} \left[\left(\frac{1}{1 - E_\text{g}}\right)^{n/m} - 1\right]. \qquad (18)$$

With a nonuniform inlet profile of the gas, the obtained height H from formulas (16) and (17) should be increased by $\sim 7d_\text{e}$ (i.e., $H' = H + 7d_\text{e}$), which enables us to indirectly take account of the nonuniformity of the phase distribution at entry into the packed bed [10].

Parameters of the Model. The number of full-mixing cells in expressions (14)–(16) is related to a modified Péclet number of the flow structure by the known approximate relation $n \approx \text{Pe}/2$ ($\text{Pe} > 10$), where Pe_g in the gas phase may be computed from the modified Taylor formula for random packings [10]

$$\text{Pe}_\text{g} = 0.52 \frac{H}{d_\text{e}} (\text{Re}_{\text{e.g}}/\xi)^{0.25} . \qquad (19)$$

For regular packings without flow swirlers, the Péclet number for the gas phase is equal to [10]

$$\text{Pe}_\text{g} = 0.43 \frac{H}{d_\text{e}\sqrt{\xi}} ; \qquad (20)$$

and in the liquid phase for annular packings [9], to

$$\text{Pe}_{\text{liq}} = \frac{\bar{u}_{\text{av}} H}{D_{\text{m.liq}}} = A\, \text{Re}_\text{d}^p\, \text{Ga}_\text{d}^k \frac{H}{d} , \qquad (21)$$

where the Re_d and Ga_d numbers are calculated from the nominal size of the packing d and the average real velocity of the liquid $\bar{u}_{\text{av}} = q_{\text{liq}}/\varepsilon_{\text{liq.d}}$:

$$\text{Re}_\text{d} = \frac{\bar{u}_{\text{av}} d}{\nu_{\text{liq}}} , \quad \text{Ga}_\text{d} = \frac{g d^3}{\nu_{\text{liq}}^2} .$$

For rings (annuli) 6–25 mm in size in formula (21) we have $A = 1.9$, $p = 0.5$, and $k = -0.33$.

The dynamic holdup of the liquid is equal, according to the expression, to

$$\varepsilon_{\text{liq.d}} = A\,\text{Re}_{\text{liq}}^{0.64}\ \text{Ga}^{p}\ ,$$

where for random annuli, $A = 0.747$ and $p = 0.42$, and for packed annuli, $A = 0.226$ and $p = 0.35$;

$$\text{Re}_{\text{liq}} = \frac{4q_{\text{liq}}}{a_{\text{v}}\nu_{\text{liq}}}\ ,\quad \text{Ga} = (a_{\text{v}}\theta)^{-3}\ ,\quad \theta = (\nu_{\text{liq}}^{2}/g)^{1/3}\ ;$$

for the metallic random packing Inzhekhim-2012 [10], we have

$$\text{Pe}_{\text{liq}} = 2.348H\,\text{Re}_{\text{liq}}^{0.428}/d_{\text{e}}\ . \tag{22}$$

For other packing structures, the calculated expressions of Pe have been presented in [9–12]. At the value of the Péclet number $\text{Pe} \le 10$, the numbers of cells n and m are found from the formulas

$$n = (\text{Pe}_{\text{g}} + 1.25)/2.5;\ m = (\text{Pe}_{\text{liq}} + 1.25)/2.5\ . \tag{23}$$

The coefficient of mass transfer in the gas phase β_{g} is found experimentally or from the expression obtained with the model of a turbulent boundary layer [10]

$$\text{Sh}_{\text{g}} = 0.175\ \text{Re}_{\text{e.g}}^{0.75}\text{Sc}_{\text{g}}^{0.33}(\xi/2)^{0.25}\ . \tag{24}$$

Expression (24) obtained for random packings at $\text{Re}_{\text{e.g}} > 40$ can also be used for regular packings with corrugations, notchings, swirlers, etc.

Examples of Calculation of Hydrodynamic Characteristics. ***Example 1.*** The initial data are as follows: $T_{\text{in.g}} = 45^{\text{o}}\text{C}$ and $T_{\text{in.liq}} = 20^{\text{o}}\text{C}$; relative air humidity $\varphi = 40\%$, $x_{\text{in}} = 0.022$ kg/kg; air flow rate $G = 1.65$ kg/s; at $S = 1.0\ \text{m}^2$, $w_0 = 1.5$ m/s, and $T_{\text{f.liq}} = 25^{\text{o}}\text{C}$, the assigned efficiency of cooling is $E_{\text{g}} = E_x = 0.8$ (80%).

The thermodynamic parameters are as follows: $I_{\text{in}} = 102$ kJ/kg; at $T_{\text{in.liq}} = 20^{\text{o}}\text{C}$ and $\varphi = 100\%$, we have $x_{\text{f}}^{*} = 0.022$ kg/kg and $I_{\text{f}}^{*} = 56{,}050$ J/kg.

The calculated values are as follows: $I_{\text{f}} = I_{\text{in}} - E_{\text{g}}(I_{\text{in}} - I_{\text{f}}^{*}) = 65{,}640$ J/kg; $x_{\text{f}} = x_{\text{in}} = E_x(x_{\text{in}} - x_{\text{f}}^{*}) = 0.0158$ kg/kg; $Q = G(I_{\text{in}} - I_{\text{f}}) = 52{,}358$ W. Water flow rate $L = Q/[c_{\text{pliq}}(T_{\text{f.liq}} - T_{\text{in.liq}})] = 2.5$ kg/s or water concentration $q_{\text{liq}} = L(\rho_{\text{liq}}S)\cdot 3600 = 9.0\ \text{m}^3/(\text{m}^2\cdot\text{h})$. Outlet temperature of the gas (12) $T_{\text{f.g}} = 25^{\text{o}}\text{C}$; thermal efficiency (5) $E_{\text{g}} = 0.8$; i.e., the E_{g} values obtained from formulas (5) and (6) are coincident.

Example 2. In the second example, we take $\varphi = 80\%$, and the other values (of temperature and flow rate of the gas and of efficiency) will remain as in the first example.

The thermodynamic parameters are as follows: $x_{\text{in}} = 0.047$ kg/kg, $I_{\text{in}} = 167$ kJ/kg, $I_{\text{f}} = 78.3$ kJ/kg, and $x_{\text{f}} = 0.218$ kg/kg.

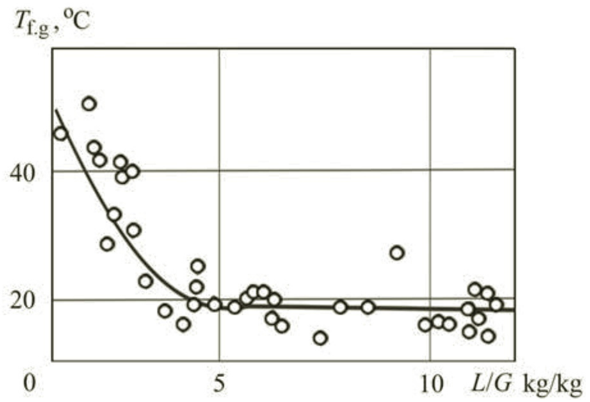
The calculated values are as follows: $Q = 145.82$ kW, $L = 17.4$ kg/s, $q_{\text{liq}} = 62.6\ \text{m}^3/(\text{m}^2\cdot\text{h})$, and $T_{\text{f.g}} = 25.5^{\text{o}}\text{C}$. The values of E_{g} from formulas (5) and (6) are coincident, in practice.

Example 3. We take the gas temperature $T_{\text{in.g}} = 90^{\text{o}}\text{C}$, and the other values will remain as in the first example.

We obtain $x_{\text{in}} = 0.176$ kg/kg, $I_{\text{in}} = 58.0$ kJ/kg, $I_{\text{f}} = 156$ kJ/kg, $x_{\text{f}} = 0.047$ kg/kg, $Q = 601.5$ kW, $L = 28.8$ kg/s, $q_{\text{liq}} = 103.6\ \text{m}^3/(\text{m}^2\cdot\text{h})$, and $T_{\text{f.g}} = 36.3^{\text{o}}\text{C}$. The efficiency obtained from formula (5) is $E_{\text{g}} = 0.768$, i.e., 4% lower than that assigned from formula (6) $E_{\text{g}} = 0.8$, which is, probably, due to a certain error in determining the thermodynamic parameters of moist air.

Thus, on the basis of the use of the heat-balance equations, the expressions for thermal efficiencies, and the cellular model, we can compute the thermodynamic parameters of cooling of the gas and heating of the cooling water in a film-type packed column.

Calculation Results. A comparison with experiment [1] has been made of the temperature of escaping gases and the moisture content at exit from the scrubber as functions of the L/G ratio for a packing of ceramic Raschig rings with dimensions 35 × 35 × 4 mm ($a_{\text{v}} = 140\ \text{m}^2/\text{m}^3$) with bed height $H = 1.0$ m. The gas velocity varied within the limits of 0.4–1.9 m/s, and the water concentration, in the range 3–55 $\text{m}^3/(\text{m}^2\cdot\text{h})$. The initial water temperature was ~12$^{\text{o}}$C. The



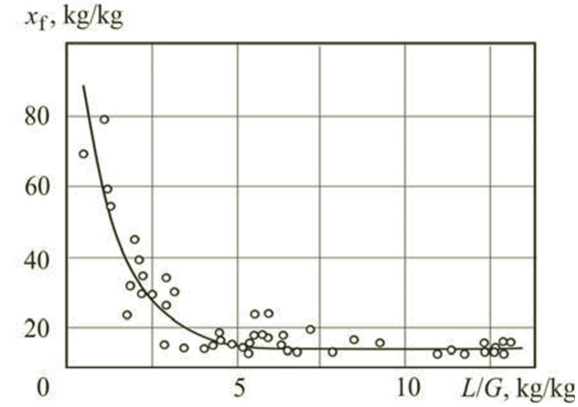


Fig. 2. Temperature of escaping gases vs. reflux-to-product ratio for the packing of ceramic rings with dimensions 35 × 35 × 4 mm at an initial gas temperature of 250–280°C and a packing height of 1000 mm; solid line, calculation from the mathematical model, points, experiment [1].

Fig. 3. Moisture content of escaping gases vs. reflux-to-product ratio. Notation is the same, as in Fig. 2.

calculated value of the outlet gas temperature was found with formulas (15) and (16) for E_g (depending on the number of cells), and also with formulas (6), (9), and (12). First, we found the value of E_g, from expression (6), the value of I_f and, from (9), x_f at $E_x = E_g$, and thereafter the temperature of escaping gases $T_{f.g}$ from (12). From Figs. 2 and 3, it can be seen that the disagreement with the experiment is within the investigation error. Thus, the adequacy of the presented mathematical model has been confirmed.

Calculation of the Packing Height. For comparative characteristics of the heat-transfer efficiency and hydraulic characteristics, consideration was given to a number of random and regular packings. A detailed description of the structure, and also the hydraulic characteristics of the considered packings have been given in [9–13].

Initially, we investigated the most popular of packings: Raschig rings. For example 1, at w_0 = 1.5 m/s and the water concentration q_{liq} = 9 $m^3/(m^2 \cdot h)$, the packing of Raschig rings of diameter 15 mm or smaller is not suitable as far as the hydrodynamic regime is concerned, i.e., it is in the regime of suspension and flooding. Therefore, calculations were done for metallic rings with dimensions 25 × 25 × 0.5 mm (a_v = 220 m^2/m^3 and d_e = 0.017 m) and 50 × 50 × 0.8 mm (a_v = 110 m^2/m^3 and d_e = 0.035 m).

For the packing of 25 × 25 × 0.5 mm rings, we have obtained $Re_{e.g}$ = 1893, ξ = 3.53, β_x = 0.074 kg/($m^2 \cdot s$), ψ = 0.523, bed height H = 0.32 m, and pressure difference ΔP_{op} = 345 Pa,and for the packing of 50 mm rings, $Re_{e.g}$ = 3733, ξ = 3.1, β_x = 0.058 kg/($m^2 \cdot s$), ψ = 0.621, H = 0.675 m, and ΔP_{op} = 288 Pa. If there are no pressure-difference constraints, it is the 25 mm packing that is the most suitable.

Next, we consider modern random regular packings.

Metallic Moebius rings of 40 × 40 mm (a_v = 191 m^2/m^3 and d_e = 0.0185 m). We have obtained $Re_{e.g}$ = 1850, ξ = 4.5, β_x = 0.05, β_x = 0.071 kg/($m^2 \cdot s$), $\psi \approx 0.6$, H = 0.36 m, and ΔP_{op} = 150 Pa.

Metallic random packing Inzhekhim-2012, nominal dimensions of the rings 24 and 45 mm. For 24 mm (a_v = 166 m^2/m^3 and d_e = 0.023 m), we have obtained $Re_{e.g}$ = 2455, ξ = 3.5, β_x = 0.067 kg/($m^2 \cdot s$), $\psi \approx 0.65$, H = 0.36 m, and ΔP_{op} = 110 Pa. For 45 mm (a_v = 101 m^2/m^3 and d_e = 0.0386 m), we have obtained $Re_{e.g}$ = 4120, ξ = 0.22, β_x = 0.058 kg/($m^2 \cdot s$), $\psi \approx 0.7$, H = 0.64 m, and ΔP_{op} = 100 Pa.

Metallic segmented regular roll packing Inzhekhim (a_v = 480 m^2/m^3 and d_e = 0.0075 m). We have obtained $Re_{e.g}$ = 960, ξ = 0.22, β_x = 0.05 kg/($m^2 \cdot s$), $\psi \approx 0.35$, H = 0.35 m, and ΔP_{op} = 40 Pa.

Metallic regular roll corrugated packing Inzhekhim (a_v = 250 m^2/m^3 and d_e = 0.0125 m). We have obtained $Re_{e.g}$ = 1333, ξ = 0.47, β_x = 0.047 kg/($m^2 \cdot s$), $\psi \approx 0.45$, H = 0.55 m, and ΔP_{op} = 37 Pa.

In Fig. 4, the results of calculating the height of the bed of packings and the pressure difference are presented in a histogram. The packings numbered 3, 4, and 6 have the greatest advantage. However, it should be taken into account that random packing Nos. 3 and 4 are flooded under high loads, and packing No. 6 has been investigated at the maximum

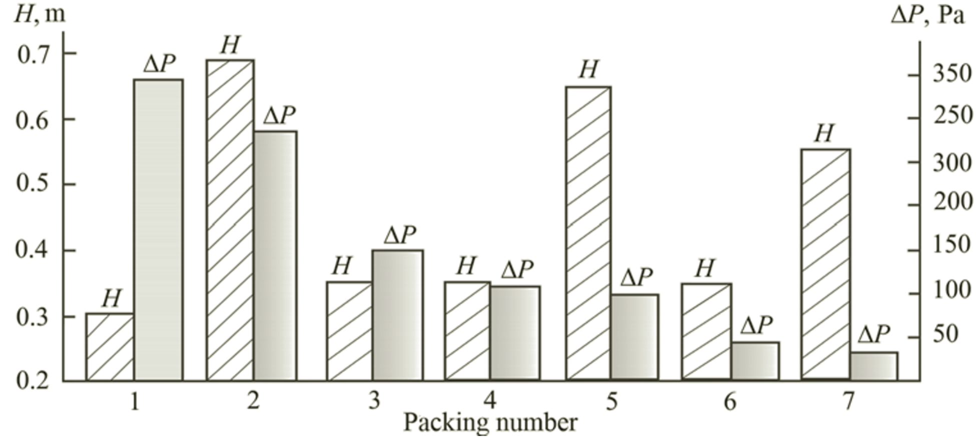


Fig. 4. Histograms of calculations of packing heights and pressure differences. Random packings: 1) Raschig rings of diameter 25 mm; 2) Raschig rings (50 mm); 3) Moebius rings (40 mm); 4) Inzhekhim-2012 (24 mm); 5) Inzhekhim-2012 (45 mm). Regular packings: 6) segmented regular; 7) roll corrugated. The assigned thermal efficiency is $E_g = 0.8$.

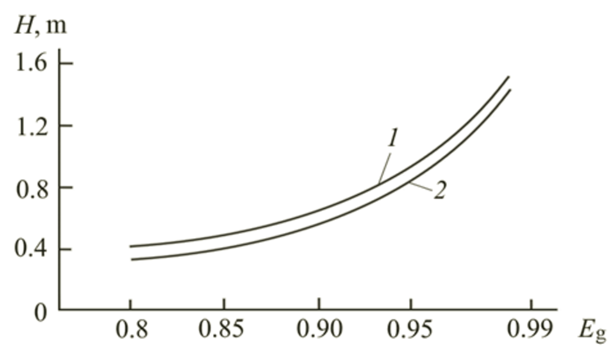


Fig. 5. Required heights of the regular packing vs. assigned efficiency of cooling of flue gases at the gas velocity $w_0 = 2$ m/s, the initial gas temperature $T_{in.g} = 150^oC$, and the initial water temperature $T_{in.liq} = 12^oC$: 1) roll corrugated packing ($d_e = 0.0125$ m); 2) regular perforated packing ($d_e = 0.0075$ m).

water concentration $q_{liq} = 30$ m^3/(m^2·h) and the air velocity to $w_0 = 4.5$ m/s. Under such loading, suspension and flooding begin at $w_0 > 2.5$ m/s. Packing No. 7 has been investigated [12] in the interval of gas velocities $0.6 \le w_0 \le 6.0$ m/s and water concentrations $10 \le q_{liq} \le 20$ m^3/(m^2·h). At the gas velocity $w_0 = 2.5$ m/s, suspension and flooding occur at $q_{liq} \ge 90$ m^3/(m^2·h). We can draw the conclusion on the fields of application of packings depending on hydraulic loads.

Figure 5 gives the required values of the height of regular packings versus the assigned thermal efficiency. It follows from the calculations that for deep cooling of flue gases, it is required that H be more than 1.5 m.

Thus, from the obtained results presented in the examples with expressions (17) or (18) to calculate the packed-bed height H, we can determine the structural and regime characteristics of a column of water cooling of the gas with an assigned type of packing and select the most rational variant.

The calculation algorithm is as follows:

(1) we assign the values of E_g (5) and E_x (13), $E_g = E_x$. For example, $E_g = E_x = 0.9$;

(2) we assign the initial values of temperatures $T_{in.g}$ and $T_{in.liq}$; the gas flow rate is G;

(3) from the tabulated data or the known expressions, we find I_{in}, x_{in}, x_f^*, and I_f^* (at $\varphi = 100\%$ and the value $T_{in.liq}$);

(4) we compute $x_f = x_{in} - E_x(x_{in} - x_f^*)$;

(5) we compute $I_f = I_{in} - E_g(I_{in} - I_f^*)$;

(6) we find the heat flux $Q = G(I_{in} - I_f)$;

(7) from the expression $Q = Lc_{pliq}(T_{f.liq} - T_{in.liq})$, we find the mass flow rate L at the assigned outlet temperature $T_{f.liq}$ or, at the assigned flow rate, the temperature $T_{f.liq}$ (depending on the posed problem);

(8) we compute the model′s parameters $Pe_{in.g}$, $Pe_{in.liq}$, and Sh_g for the selected type of packing in a film regime;

(9) we find the packed-bed height H from expression (16) or (18) (depending on the number of cells);

(10) we compute the pressure difference of the packed bed.

If the obtained results meet the requirements of the terms of reference for design of a scrubber, the calculation is completed, if not, the alternative type of packing is selected.

Conclusions. In designing or modernizing heat- and mass-exchange apparatuses in various industries, use is made of analytical, numerical, or approximate methods to model the processes. Each of these methods has its advantages and drawbacks. An advantage of the approximate method of mathematical modeling of the cooling of gases in scrubbers, that has been considered in this article, lies in carrying out calculations rapidly and determining quite reliably the regime and structural characteristics of packed apparatuses in a film regime. The presented mathematical model allows computations based on the use of results of the hydraulic characteristics of packed scrubbers. This reduces considerably the terms and costs in designing or modernizing the apparatuses in various industries.

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NOTATION

a_v, specific surface of the packing, m^2/m^3; c_{pg} and c_{pliq}, specific heats of the gas and the liquid, J/(kg·K); D_g, diffusion coefficient of the steam in the gas, m^2/s; D_{st}, coefficient of reverse mixing, m^2/s; d_e, equivalent diameter of the packing, m; F, area of contact of the phases, m^2; G, mass flow rate of the gas, kg/s; H, height of the packed bed, m; I_{in} and I_f, initial and final values of the enthalpy of the gas, J/kg; I_f^*, value of the enthalpy of the gas at the water temperature $T_{in.liq}$ (at the water inlet), J/kg; I_{st}, enthalpy of the steam (in water cooling), J/kg; K, heat-transfer coefficient, $W/(m^2{\cdot}K)$; L, mass flow rate of water, kg/s; N_g, thermal number of units of transfer; $Pe_g = Hw_g/D_{st.g}$, Péclet (Bodenstein) number for the gas phase; $Pe_{liq} = \overline{u}_{av}H/D_{st.liq}$, Péclet number for the liquid phase; Q, heat flux, W; Q_{ev}, heat flux of the evaporated water, W; q_{liq}, water concentration, $m^3/(m^2{\cdot}s)$ or $m^2/(m^2{\cdot}h)$; Re, Reynolds number; S, cross-sectional area of the column, m^2; $Sh_g = \beta_g/D_g$, Sherwood number; w_0, full-cross-section gas velocity S, m/s; w_g, gas velocity in the bed, m/s; x, moisture content of the gas, kg/kg; x^*, moisture content of the saturated gas, kg/kg; α_g, coefficient of heat transfer in the gas phase, $W/(m^2{\cdot}K)$; β_g, coefficient of mass transfer in the gas phase, m/s; β_x, mass-transfer coefficient referred to the enthalpy difference, $kg/(m^2{\cdot}s)$; ΔI_{av}, average enthalpy difference in the gas phase, J/kg; $\varepsilon_{liq.d}$, dynamic component of the liquid holdup in the bed, m^3/m^3; ν_g, coefficient of kinematic viscosity of the gas, m^2/s; ξ, coefficient of hydraulic resistance of the packing; ρ_g and ρ_{liq}, densities of the gas and the liquid, kg/m^3; ψ_w, coefficient of wettability of the packing surface ($\psi_w \le 1$). Subscripts: g, gas; liq, liquid; f, final value; in, initial value; e, equivalent.

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