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Modeling of Heat and Mass Transfer and Absorption-Condensation Dust and Gas Cleaning in Jet Scrubbers

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Additional information is available at the end of the chapter

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1. Introduction

The process of complex cleaning of gases, injected into the atmosphere, for instance, by thermal power plants, metallurgical, chemical or other industrial enterprises, from dust and harmful gaseous admixtures by means of their irrigation by wash liquids (water or specially selected water solutions) is considered. This process can be implemented in gas pipes or gas-cleaning apparatuses (direct flow or counter flow jet scrubbers) [1]. The process of gas cleaning from dust and gas admixtures is carried out in the following manner. The fluid dispersed by jets is introduced into the dust-vapor-gas flow in the form of droplets, interacts with it, and under nonisothermal conditions the increased moisture content leads to intensive condensation of liquid vapors on particles, their significant enlargement and efficient absorption of liquid droplets due to collisions of the latter with particles [2]. Simultaneously, the liquid droplets and condensate on particles absorb harmful gas components, dissolving them and removing from the vapor-gas flow.

The authors failed to find the mathematical description of this complex process in literature. From the engineering point of view the importance of development of generalized mathematical models, which reflect properly the interaction of heat and mass transfer with the effects of gas components removal and dust capture by the droplets of irrigating liquid in jet scrubbers and reactors, is undisputable, and it is determined by significant opportunities for optimization of operation conditions and constructions of energy-intensive and large-scale equipment in various industries both in terms of reducing of material and energy costs.

2. Problem statement, main equations and assumptions

In the current work we suggest the model for mathematical description of the above process with the following assumptions:

1. droplets and particles are considered monodispersed with equivalent sizes, equal to mass-median by distributions;
2. concentrations of droplets of irrigating liquid, dust particles and harmful gas components are low, what allows us to use the Henry's law for equilibrium of gas components in liquid and gas phases at the interface and assume that the solution in droplet is ideal;
3. the mean-mass temperature of droplets and temperature of their surfaces are equal because of their small sizes [3];
4. the typical time of gas component dissolution in droplet is significantly less than the typical time of mass transfer processes, commonly occurring in the apparatus;
5. the motion velocities of particles with condensate on their surface ("formations") and vapor-gas flow are equal;
6. the moisture content in the flow can be high, what requires consideration of the Stefan correction in mass transfer equations for evaporation-condensation process on droplets and "formations";
7. we do not take into account the evaporation-condensation correction for the resistance and heat transfer coefficients of droplets and "formations", it is insignificant and becomes obvious only at the initial stages of the process at high moisture contents [4];
8. in equation of droplet motion we take into account variability of its mass;
9. the radiant component in the process of heat transfer is neglected because of low temperatures of droplet, "formations" and flow;
10. mutual coalescence of droplets and "formations" is not taken into account, and merging of droplets and "formations" due to collision is the basis of condensation-inertial mechanism of dust capture in jet scrubbers [2].

Under the above conditions equations of model system will take the following form:

Motion equation of a mass-median droplet with variable mass

$$\frac{d\vec{V}_d}{d\tau} = \vec{R}_d + \vec{g} - \frac{\vec{V}_d}{m_d} \frac{dm_d}{d\tau}; \quad (1)$$

equation of heat transfer between droplet and vapor-gas flow

$$c_f m_d \frac{dT_d}{d\tau} = \alpha_d \pi \delta_d^2 (T_d - T) + \sum r_i \frac{dm_{id}}{d\tau} + c_\delta \rho_\delta V_c \frac{\pi \delta_d^2}{4} \eta_{Stk} (T_\delta - T_0); \quad (2)$$

equation of mass transfer between droplet and the i -th component of vapor-gas flow

$$\frac{dm_{i\delta}}{d\tau} = -\beta_{id} \pi \delta_d^2 (\rho_{id} - \rho_i); \quad (3)$$

equation of mass transfer between “formation” and the i -th component of vapor-gas flow

$$\frac{dm_{i\delta}}{d\tau} = -\beta_{i\delta} \pi \delta^2 (\rho_{i\delta} - \rho_i); \quad (4)$$

continuity equation for i -th reacting components, including vapor of liquid

$$\frac{\partial \rho_i}{\partial \tau} + \text{div}(\rho_i \vec{U}) = -\frac{dm_{id}}{d\tau} n_d - \frac{dm_{i\delta}}{d\tau} n_\delta; \quad (5)$$

continuity equation for (mass concentration) of non-reacting component of the vapor-gas mixture

$$\frac{\partial \rho_g}{\partial \tau} + \text{div}(\rho_g \vec{U}) = 0; \quad (6)$$

continuity equation for (mass concentration) of “formations”

$$\frac{\partial \rho_\delta}{\partial \tau} + \text{div}(\rho_\delta \vec{U}) = \sum \frac{dm_{i\delta}}{d\tau} n_\delta - \rho_\delta V_c \frac{\pi \delta_d^2}{4} \eta_{Stk} n_d; \quad (7)$$

continuity equation for (mass concentration) of droplets

$$\frac{\partial \rho_d}{\partial \tau} + \text{div}(\rho_d \vec{V}_d) = \frac{dm_d}{d\tau} n_d; \quad (8)$$

equation of heat transfer between “formation” and vapor-gas flow

$$\tilde{n}_\delta m_\delta \frac{dT_\delta}{d\tau} = -\alpha_\delta \pi \delta^2 (T_\delta - T) + \sum r_i \frac{dm_{i\delta}}{d\tau}; \quad (9)$$

equation of convective heat transfer between vapor-gas flow and droplets and “formations”

$$\rho \frac{d\tilde{n}(T - T_0)}{d\tau} = \alpha_d \pi \delta_d^2 (T_d - T) n_d + \alpha_\delta \pi \delta^2 (T_\delta - T) n_\delta; \quad (10)$$

general rate of droplet mass change due to evaporation-condensation and absorption of removed gas components (droplet collision is assumed unlikely) and “formation” absorption

$$\frac{dm_d}{d\tau} = \sum \frac{dm_{id}}{d\tau} + \rho_\delta V_c \frac{\pi \delta_d^2}{4} \eta_{\text{Stk}} \quad (11)$$

general rate of “formation” mass change (“formation” collision is assumed unlikely)

$$\frac{dm_\delta}{d\tau} = \sum \frac{dm_{i\delta}}{d\tau}; \quad (12)$$

continuity equation for (mass concentration) of dry particles

$$\frac{\partial \rho_p}{\partial \tau} + \text{div}(\rho_p \vec{U}) = -\rho_p V_c \frac{\pi \delta_d^2}{4} \eta_{\text{Stk}} n_d. \quad (13)$$

The following closure relationships shall be added to equations (1-13):

for the force of droplet aerodynamic resistance per a unit of droplet mass,

$$\vec{R}_d = -\xi \frac{(\vec{V}_d - \vec{U})}{\tau_d}; \quad (14)$$

where relative coefficient of droplet resistance is $\xi = \xi / \xi_c$, $\xi_c = 24 / \text{Re}_d$,

$$\tilde{\xi} = 1 + 0,197 \text{Re}_d^{0,63} + 2,6 \cdot 10^{-4} \text{Re}_d^{1,38} (0,1 \leq \text{Re}_d \leq 3 \cdot 10^5) [5], \quad (15)$$

$$\tau_d = \frac{\rho_f \delta_d^2}{18\mu}, \text{Re}_d = \frac{V_c \delta_d \rho}{\mu}, V_c = |\vec{V}_d - \vec{U}|; \quad (16)$$

coefficient of “formation” entrainment according to the empirical formula of Langmuir–Blodgett with Fuchs correction on engagement effect [1]

$$\eta_{\text{Stk}} = \left(\frac{\text{Stk}}{\text{Stk} + 0,5} \right)^2 + 2,5 \frac{\delta}{\delta_d}, \quad (17)$$

$$\text{Stk} = \tau_\delta \frac{V_c}{\delta_d}, \tau_\delta = \frac{\rho_{i\delta} \delta^2}{18\mu} (\rho_{f\delta} \approx \rho_f), \quad (18)$$

where ρ_{fb} is efficient density of “formation”;

mass transfer coefficient of the i -th component with droplets both via evaporation-condensation and absorption-desorption [2]

$$\text{Nu}_{id} = \frac{\beta_{id} \delta_d}{D_i} = 2(1 + 0,276 \text{Re}_d^{0,5} S c_i^{0,33}) K_{ci}, S c_i = \frac{\mu}{\rho D_i}; \quad (19)$$

Stefan correction on increased moisture content

$$K_{\tilde{m}} = 1 + \frac{P_{id} + P_i}{2B}; \quad (20)$$

barometric (total) pressure

$$B = P_g + \sum P_i; \quad (21)$$

density of vapor-gas mixture

$$\rho = \rho_g + \sum \rho_i; \quad (22)$$

state equation for gas components and vapor of liquid

$$\rho_g = \frac{M_g P_g}{RT}, \rho_i = \frac{M_i P_i}{RT}, \rho_{id} = \frac{M_i P_{id}}{RT_d}, \rho_{i\delta} = \frac{M_i P_{i\delta}}{RT_\delta}; \quad (23)$$

diffusion coefficient of the i -th component in non-reacting component of the vapor-gas flow (we assume that its fraction in the flow is predominant)

$$D_i = D_{i0} \frac{B_0}{B} \left(\frac{T}{T_0} \right)^{1,75}, \quad B_0 = 0.1 \text{ MPa}, \quad T_0 = 273 \text{ K}; \quad (24)$$

coefficient of droplet heat transfer according to Drake's formula

$$\text{Nu}_d = \frac{\alpha_d \delta_d}{\lambda} = 2 + 0,459 \text{Re}_d^{0,5} \text{Pr}^{0,3}, \quad \text{Pr} = \frac{\mu c}{\lambda}; \quad (25)$$

countable concentrations of droplets and "formations"

$$n_d = \frac{\rho_d}{m_d}, \quad (26)$$

$$n_\delta = \frac{\rho_\delta}{m_\delta}; \quad (27)$$

heat and mass transfer coefficients of "formations"

$$\alpha_\delta = 2 \frac{\lambda}{\delta} \quad (\text{Nu}_\delta = 2), \quad \beta_{i\delta} = 2 \frac{D_i}{\delta} \quad (\text{Nu}'_{i\delta} = 2), \quad (28)$$

heat capacity of the vapor-gas mixture

$$c = \frac{\sum \rho_i c_i}{\rho}; \quad (29)$$

specific heat of gas absorption with the made assumptions [6]

$$r_i = M_i^{-1} R T^2 \frac{d \ln m_{px,i}}{dT}, \quad (30)$$

it can be assumed for water vapors that $r_d \approx 2500 \text{ kJ/kg}$ [2-4];

according to Henry's law for partial saturation pressure at the interface between i -th gas components, the equilibrium condition is [6]

$$P_{id} = m_{px,i} x_{id}, \quad P_{i\delta} = m_{px,i} x_{i\delta}; \quad (31)$$

$$x_{id,\delta} = \frac{\frac{c_{mid,\delta}}{M_i}}{\frac{c_{mid,\delta}}{M_i} + \frac{1}{M_{dis}}}, \quad (32)$$

where $x_{id,\delta}$ is a molar part, equal to the number of moles of dissolved gas per the total number of moles in solution, M_{dis} is the molar mass of dissolvent.

The equation for mass concentration of dissolved i -th gas component in the kilogram per 1 kg of dissolvent in the droplet and "formation" is written as

$$\frac{dc_{mid,\delta}}{d\tau} = \frac{dm_{id,\delta}}{d\tau} \frac{6}{\pi \delta_{d,f}^3 \rho_f}. \quad (33)$$

Diameters of specific spherical volume of dissolvent for "formation" δ_f and droplet δ_d are calculated by equations:

$$\frac{d\delta_f^3}{d\tau} = \frac{6}{\pi} \frac{1}{\rho_f} \frac{dm_{v\delta}}{d\tau}, \quad \delta_d = \sqrt[3]{\frac{6m_d}{\pi\rho_f}}; \quad (34)$$

"formation" diameter is

$$\delta = \sqrt[3]{\frac{6m_\delta}{\pi\rho_s} + \delta_0^3}, \quad (\rho_s \approx \rho_f), \quad (35)$$

where ρ_s is solution density, kg/m³.

The reactive force in equation (1) is neglected because of evaporation-condensation and absorption [2]. In equation (2) specific heat capacity c_f is taken constant and equal to specific heat capacity of dissolvent because of low concentrations of absorbed dust and absorbed gases. For small particles and significant amount of condensate on them [2] we will take c_δ equal to specific heat capacity of dissolvent c_f . In this equation the first summand in the right determines convective heat transfer between the droplet and flow, the second summand de-

termines the total heat of phase transitions due to evaporation-condensation and absorption-desorption of gas components, and the third summand determines heat introduced by “formations” into the droplet due to their absorption at collision. In equations (3) and (4) $dm_{id}/d\tau$, $dm_{i\delta}/d\tau$ are the rates of droplet and “formation” mass change due to the processes of evaporation-condensation or absorption-desorption of the i -th gas component. In relationships (16) dynamic viscosity of the vapor-gas flow μ is calculated by generalized Wilkey’s formulas, in our case on the basis of research performed in [2] and [4] with consideration of low concentrations of reacting gas components we will determine μ by Sutherland formulas [2, 4] for a non-reacting component of the vapor-gas mixture. The coefficient of mixture heat conductivity λ will be calculated similarly by Sutherland formula [2, 4] in formulas (25) and (28). The diffusion components of the vapor-gas flow will be determined by their dependences on temperature in the non-reacting component by formula (24). The correction for Stefan flow of gas components is not taken into account because of their low concentrations ($K_{ci}=1$). In the current study we will consider water as the absorbent and $m_{px,i}$ will be taken from tables depending on temperature [6]. If there are no data for some gases in [6], for instance, for SO_2 , we suggest to recalculate volumetric 1 and weight q_s solubility [7] as the limit ones by $m_{px,i'}$, this will be described in detail in this work.

3. Numerical implementation of the model, comparison of calculation results with experimental data

As it is shown in [2], in most technically implemented situations it is possible to use a single-dimensional model for calculation of heat and mass transfer in irrigation chambers, what is determined by the vertical position of apparatuses (hollow jet scrubbers HJC); at their horizontal position it is determined by high velocities of cleaned gases, dust particles and droplets (Venturi scrubber VS), when the gravity force, influencing the flow components and causing its 2D character, is low in comparison with the inertia forces.

The calculation scheme of the problem for the vertical construction of apparatus is shown in Fig. 1a). The scheme of interaction between a droplet of washing liquid dispersed by the jets with vapor-gas flow and dust particles is shown in Fig. 1b).

The hollow jet scrubber HJS can have direct-flow and counter-flow construction. In the direct-flow scheme the initial parameters of the vapor-gas flow, irrigating liquid and dust are set on one side (inlet) of apparatus, and the resulting parameters are achieved at the apparatus outlet. In the counter-flow scheme the parameters of vapor-gas flow and dust are set on one side of apparatus, the parameters of irrigating liquid are set on the opposite side (at apparatus outlet). Scheme 1a) is attributed to the counter-flow. From the point of numerical implementation the direct-flow scheme is the Cauchy problem, and the counter-flow scheme is the boundary problem. Let’s perform calculations for the direct-flow scheme according to the known experimental data for generalized volumetric mass transfer coefficients, shown in [6, p. 562], for different gases absorbed on dispersed water. The calculation scheme is shown in Fig. 2 (it is conditional, the construction can differ).

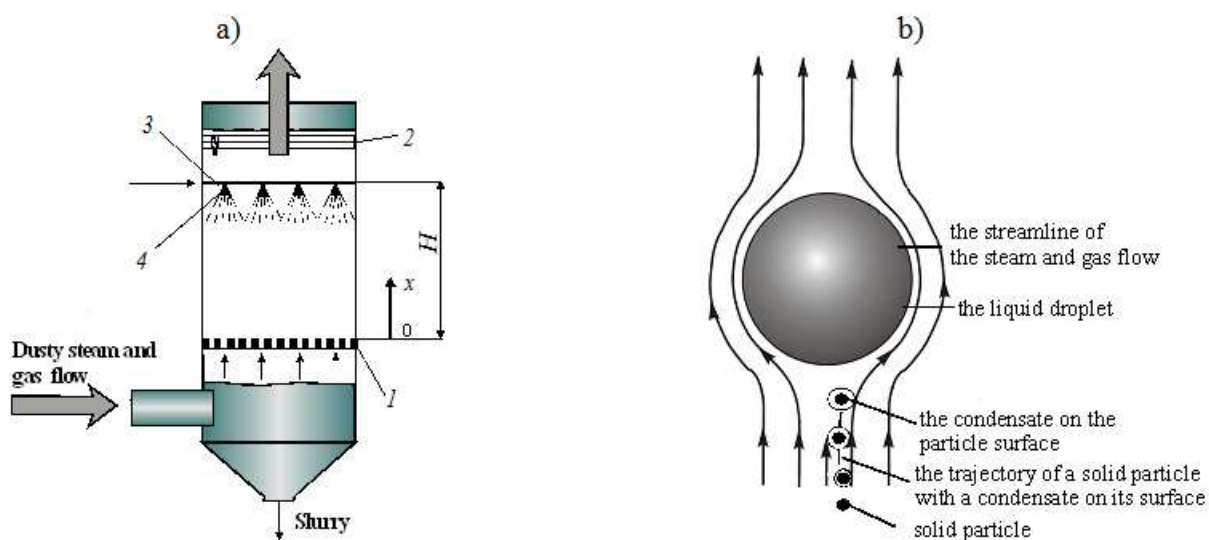


Figure 1. The HJS scheme: 1 – gas-distributing grate, 2 – droplet catcher, 3 – water collector, 4 – jets, H – scrubber operation height; b) the scheme of droplet interaction with the flow and dust particle

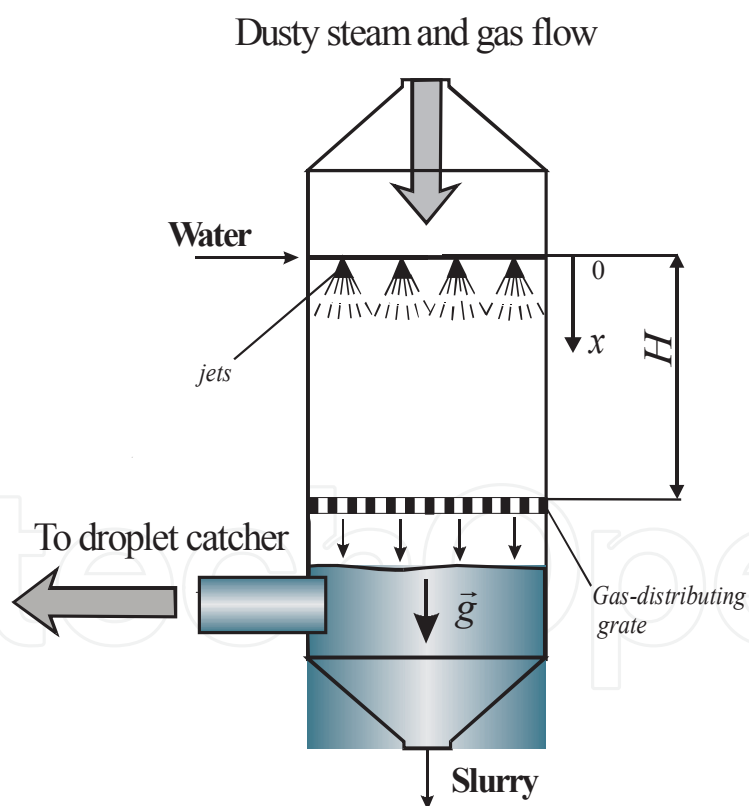


Figure 2. The scheme of direct-flow HJS

The problem will be solved in the stationary statement. The boundary conditions are set at $x=0$ ($\tau=0$) in the following manner:

$$\left. \begin{array}{l} \text{for the vapor-gas flow } U=U_0, \quad d=d_0, \quad d_i=d_{i0}, \quad T=T_{00}; \\ \text{for dispersed liquid } V_d=V_{d0}, \quad \delta_d=\delta_{d0}, \quad q=q_0, \quad T_d=T_{d0}; \\ \text{for dust } \rho_p=\rho_{p0}, \quad \delta=\delta_0. \end{array} \right\} \quad (36)$$

Continuity equations (6) and (8) in stationary single-dimensional case can be reduced to the following, as in [2], analytical dependences:

$$U=U_0 \frac{T}{T_{00}} \frac{B-\sum P_{j0}}{B-\sum P_j} \approx U_0 \frac{T}{T_{00}} \frac{K_{ave} + d_\Sigma}{K_{ave} + d_{\Sigma 0}}, \quad (37)$$

$$\rho_d = \rho_{d0} \frac{V_{d0}}{V_{dx}} \frac{m_d}{m_{d0}}, \quad \rho_{d0} = q \rho_f \frac{U_0}{V_{d0}}, \quad m_d = \rho_f \frac{\pi \delta_d^3}{6}, \quad m_{d0} = \rho_f \frac{\pi \delta_{d0}^3}{6}, \quad (38)$$

where

$$K_{ave} = \frac{\sum_{i=1}^k K_i}{k}, \quad K_i = \frac{M_i}{M_g}, \quad (39)$$

k is the number of reacting components, including liquid vapors, M_g is molecular mass of a non-reacting component of gas;

$$d_\Sigma = \frac{\sum_{i=1}^k K_i P_i}{B - \sum_{i=1}^k P_i} \approx K_{ave} \frac{\sum_{i=1}^k P_i}{B - \sum_{i=1}^k P_i}, \quad (40)$$

efficiency of dust capture and gas component removal is determined by relationships:

$$\eta_p = 1 - \frac{(\rho_p U)_H}{\rho_{p0} U_0}, \quad (41)$$

$$\eta_i = 1 - \frac{(\rho_{d,i} d_i U)_H}{\rho_{d,i0} d_{i0} U_0}. \quad (42)$$

3.1. Calculation with the use of m_{px} [6] or Henry's coefficient E [7]

Calculation results on absorption of CO₂ by water droplets in the direct-low hollow jet scrubber are shown in Fig. 3. There are no any restrictions for calculations by solubility limit (concentration of gas dissolving in the droplet). However, the solubility limits exists as the experimental fact for gases, presented in tables of Hand-books as absorption coefficient α , in volumetric fractions reduced to 0 °C and pressure of 0.1 MPa or in the form of solubility coefficient q_p in mass fractions to solution or dissolvent [8, 9]. Thus, in [9] for $\alpha(T)$, m³ of gas/ m³ of water, for CO₂ and CH₄ the following data are shown (Table 1).

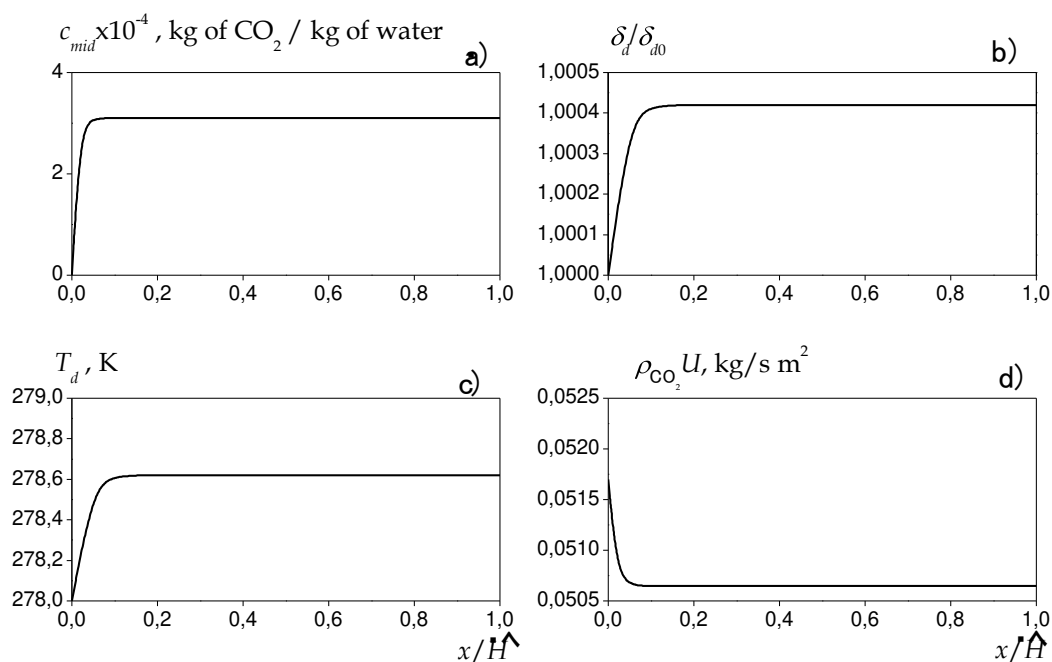


Figure 3. Calculation of CO₂ absorption in direct-flow jet scrubber: $H=12.75$ m; $q=0.015$ m³/m³; $\delta_{d0}=700$ μ m; $V_{d0}=24.5$ m/s; $U_0=0.25$ m/s; $T_{d0}=278$ K; $T_{00}=293$ K; $d_0=0.02$ kg/kg of dry air; $d_{CO_2}=0.2$ kg of CO₂/kg of dry air; $\eta_{CO_2}=0.022262$

T, K	273	283	293	303	313	323	333	353	373
α_{CO_2}	1.713	1.194	0.878	0.665	0.530	0.436	0.359
$\alpha_{CH_4} \cdot 10^3$	55.6	41.8	33.6	27.6	23.7	21.3	19.5	17.7	17.0

Table 1. Absorption coefficients for CO₂ and CH₄ α , m³/m³ of water at $B=0.1$ MPa and $T_0=273$ K

It follows from this table that the limit value of CO₂ concentration in a water droplet is

$$\begin{aligned}
 c_{mco_2, \lim} &= \alpha_{co_2} \rho_{co_2}(T_0, B_0) = \alpha_{co_2}(T) \frac{M_{co_2} B_0}{R T_0} = \alpha_{co_2} \frac{44 \cdot 101325}{8,314 \cdot 10^3 \cdot 273} = \\
 &= 1,964 \cdot 10^{-3} \cdot \alpha_{co_2}(T), \frac{\kappa g CO_2}{\kappa g water},
 \end{aligned}
 \quad (43)$$

where $\alpha_{co_2}(T)$ is the table value of absorption coefficient for CO_2 (Table 1).

This value shall limit concentration of CO_2 dissolved in the droplet. It can be seen in Fig. 3a) that according to Table 1 calculated value of c_{mid} does not reach the solubility limit and it is one order lower. Thus, $c_{mid, \lim}(T_d = 278,63 \text{ K, see Fig. 3c}) = 2.81 \cdot 10^{-3} \text{ kg of } CO_2/\text{kg of water}$. It can be seen in Fig. 3b) that as a result of water vapor condensation and CO_2 absorption the size of droplet increases insignificantly, less than by 0.05 %, i.e., a small amount of water vapors condenses on the droplet and a small amount of CO_2 is absorbed by the droplet, Fig. 3a). The calculated amount of the mass of gas component absorbed by liquid droplets in the scrubber is determined by formula, $\text{kg/h} \cdot \text{m}^2$,

$$W_{i,th} = \rho_{d.a.,0} d_{i0} U_0 \eta_i 3600. \quad (44)$$

For calculated situation with consideration of partial density of dry air at the inlet $\rho_{d.a.,0} = 1.0363 \text{ kg/m}^3$, gas content $d_{CO_2,0} = 0.2 \text{ kg/kg of dry air}$, $U_0 = 0.25 \text{ m/s}$ and $\eta_{CO_2} = 0.022262$ according to formula (44) we will obtain

$$W_{i,th} = 1,0363 \cdot 0,2 \cdot 0,25 \cdot 0,022262 \cdot 3600 = 4,153 \text{ kg/h} \cdot \text{m}^2. \quad (45)$$

Let's compare the obtained result with the value achieved via the empirical volumetric mass transfer coefficient, shown in [6, p. 562] (in our nomenclature):

$$\beta_{iv} = 720 U_0^{0,9} Q^{0,45} H^{-0,65} \left(\frac{D_{i,0}}{D_{NH_3,0}} \right)^{0,67}, 1/\text{h}. \quad (46)$$

Here Q is irrigation density, m/h , $D_{NH_3,0} = 0,198 \cdot 10^{-4} \text{ m}^2/\text{s}$ is coefficient of methane diffusion in air at $B_0 = 0.1 \text{ MPa}$ and $T_0 = 273 \text{ K}$, H is calculated scrubber height, m .

Let's write down the value of obtained coefficient per an area unit of apparatus cross-section via coefficient β_{iv} in the following form, $\text{kg/h} \cdot \text{m}^2$:

$$W_{i,e} = \beta_{iv} \Delta \rho_i l, \quad (47)$$

where $\Delta\rho_i$ is calculated concentration pressure on the way of gas component obtaining $x=0$, $x=l$, where l corresponds to the coordinate, where thermodynamic equilibrium is achieved for the i -th component in the flow.

Substituting (46) into (47) and assuming $l=H$, as it was made at treatment of experimental data in [6], we get

$$W_{i,e} = 720U_0^{0,9} (3600U_0q)^{0,45} H^{0,35} \Delta\rho_i \left(\frac{D_{i,0}}{D_{NH_3,0}} \right)^{0,67}, \text{ kg/h} \cdot \text{m}^2, \quad (48)$$

where q is irrigation coefficient, m^3 of water/ m^3 of vapor-gas flow at apparatus inlet.

Let's transform formula (48), and finally for calculation we obtain dependence

$$W_{i,e} = 28685,9U_0^{1,35} q^{0,45} H^{0,35} \Delta\rho_i \left(\frac{D_{i,0}}{D_{NH_3,0}} \right)^{0,67}. \quad (49)$$

For the considered situation $D_{i,0}=D_{CO_2,0}=0,138 \cdot 10^{-4} \text{m}^2/\text{s}$ [6].

Let's take the average experimental data of [6] as the calculation working height of absorber $H_{ave}=(H_{min} \quad H_{max})^{1/2}=(4,3 \quad 12)^{1/2} \approx 7 \text{ m}$, then for calculated value $\Delta\rho_{CO_2}=\rho_{d.a.,0}d_{CO_2,0}\eta_i=1,0363 \cdot 0,2 \cdot 0,022262=0,004614 \text{ kg/m}^3$

$$W_{i,e} = 28685,9 \cdot (0,25)^{1,35} (0,015)^{0,45} (7)^{0,35} 0,004614 \left(\frac{0,138}{0,198} \right)^{0,67} \approx 4,772 \text{ kg/h} \cdot \text{m}^2 \quad (50)$$

since $D_i=D_{i,0} \frac{B_0}{B} \left(\frac{T}{T_0} \right)^{1,75}$ and total multiplier $\frac{B_0}{B} \left(\frac{T}{T_0} \right)^{1,75}$ in (50) is reduced.

Calculated (45) and experimental (50) results differ by $\Delta \approx 13 \%$. If we take $H=4,3 \text{ m}$, then $W_{i,e}=4,024 \text{ kg/h m}^2$ and $\Delta \approx 3 \%$. At $H=12 \text{ m}$, $W_{i,e}=5,76 \text{ kg/h m}^2$ and $\Delta=28 \%$.

It follows from formula (42) for calculated concentration difference at apparatus inlet and outlet that

$$\Delta\rho_{i,cal} = \left(\frac{U_{out}}{U_0} - 1 \right) \rho_{i,in} + \Delta\rho_i,$$

where calculated value is $\Delta\rho_{i,cal}=\eta_i\rho_{i,in}$. Thus, for $U_{out}=U_0$ $\Delta\rho_{i,cal}=\eta_i\rho_{i,in}=\Delta\rho_i$, i.e., the calculated value of concentration difference coincides with real value $\Delta\rho_i$. Therefore, we can make a conclusion that efficiency of dust capture and mass transfer shall be calculated not

by the measured difference of dust concentrations and extracted gas components at the inlet and outlet, but by difference of their mass fluxes in accordance with the law of mass conservation, and if velocities at the inlet and outlet are equal or close efficiencies can be calculated by real concentration difference. Distribution of mass flux of CO_2 along the scrubber height is shown in Fig. 3d). It is obvious from Figs. 3c) and 3d) that the process of absorption completes long before the flow escape from the scrubber, for the given version of calculation at $x/H \approx 0.1$ (≈ 1.3 m). Hence, the residual height of the scrubber is excessive, and it can not be determined experimentally.

Previous comparison can be made in the relative form, what will prove the validity of calculation of mass transfer coefficient as a measure determining process intensity, on the basis of model in comparison with its experimental expression [6]:

$$\frac{W_{i,th}}{W_{i,e}} = \frac{\rho_{d.a.,0} d_{i0} U_0 \eta_i 3600}{28685,9 U_0^{1,35} q^{0,45} H^{0,35} \Delta \rho_i \left(\frac{D_{i,0}}{D_{\text{NH}_3,0}} \right)^{0,67}} = \frac{0,1255}{q^{0,45} (H U_0)^{0,35} \left(\frac{D_{i,0}}{D_{\text{NH}_3,0}} \right)^{0,67}}, \quad (51)$$

where it is assumed that $\Delta \rho_i = \eta_i \rho_{d.a.,0} d_{i0}$. Thus, for our case

$$\frac{W_{i,th}}{W_{i,e}} = \frac{0,1255}{0,616 \cdot 0,151 \cdot H^{0,35} \cdot 0,785} = \frac{1,72}{H^{0,35}} = 0,87 \text{ at } H=7 \text{ m and } \frac{W_{i,th}}{W_{i,e}} = 1,03 \text{ at } H=4.3 \text{ m.}$$

Here $\Delta \rho_i$ has the meaning of efficient drop of gas concentration, not real, but corresponding to extraction of the gas component due to absorption on a liquid droplet. Real drop of CO_2 concentrations at the inlet and outlet at the example of Fig. 4 is even negative: $\Delta \rho_{\text{CO}_2} = \rho_{\text{CO}_2,in} - \rho_{\text{CO}_2,out} = 0,1245 - 0,2099 = -0,0854 \text{ kg/m}^3$, here $U_{in} = U_0 = 0.25 \text{ m/s}$, $U_{out} = 0.1439 \text{ m/s}$.

Extractions per a total volume of apparatus can be presented as, kg/h,

$$\Delta G_{i,th} = \beta_{iv,th} \Delta \rho_i \pi D^2 H; \quad (52)$$

$$\Delta G_{i,e} = \beta_{iv,e} \Delta \rho_i \pi D^2 H. \quad (53)$$

On the other hand

$$\beta_{iv,th} = \frac{\Delta G_{i,th}}{\Delta \rho_i \pi D^2 H} = \frac{\rho_{d.a.,i0} d_{i0} U_0 \eta_i 3600 \pi D^2}{\Delta \rho_i \pi D^2 H}.$$

Hence, with consideration of formulas (46) and (53) we will obtain the relationship for volumetric mass transfer coefficients (theoretical and experimental ones)

$$\frac{\beta_{iv,th}}{\beta_{iv,e}} = \frac{3600U_0\Delta\rho_i\pi D^2}{28685,9U_0^{1,35}q^{0,45}H^{-0,65}\Delta\rho_i\left(\frac{D_{i,0}}{D_{NH_3,0}}\right)^{0,67}\pi D^2H},$$

after elementary reductions in numerator and denominator this corresponds to formula (51). Here D is apparatus diameter.

Calculations results for the same situation as in Fig. 3 are shown in Fig. 4, but for the increased moisture content $d_0=0.5$ kg/kg of dry air. The theoretical value of absorbed CO_2 is:

$$W_{i,th}=0,6226 \cdot 0,2 \cdot 0,029136 \cdot 3600=3,265 \text{ kg/h}\cdot\text{m}^2.$$

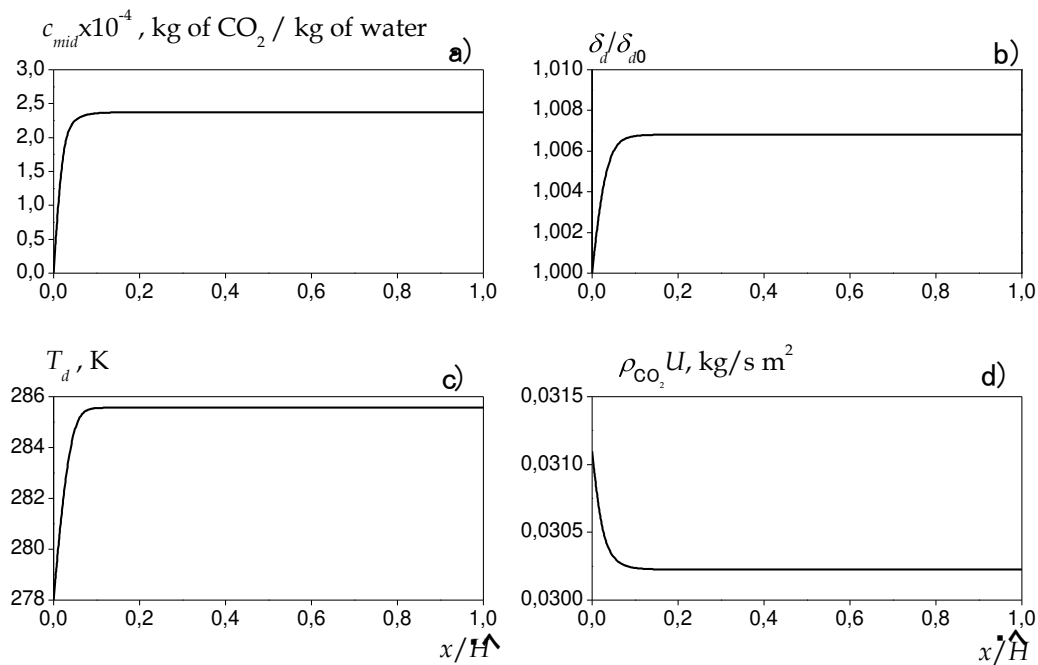


Figure 4. Fig. 4. Calculation of CO_2 absorption in the direct-flow jet scrubber: $H=12.75$ m; $q=0.015$ m^3/m^3 ; $\delta_{d0}=700$ μm ; $V_{d0}=24.5$ m/s; $U_0=0.25$ m/s; $T_{d0}=278$ K; $T_{00}=293$ K; $d_0=0.5$ kg/kg of dry air; $d_{CO_2}=0.2$ kg of CO_2 /kg of dry air; $\eta_{CO_2}=0.029136$

Calculation by formula (48) for height $H=4.3$ m gives the following

$$W_{i,e}=28685,9 \cdot (0,25)^{1,35}(0,15)^{0,45}(4,3)^{0,35}0,00363 \cdot 0,785=3,165 \text{ kg/h}\cdot\text{m}^2,$$

what differs from the theoretical value by 3 %. Here $\Delta\rho_{CO_2}=0,00363$ kg/ m^3 by calculation ($\Delta\rho_{CO_2}=0,62226 \cdot 0,2 \cdot 0,029136$). We should note that even for the increased moisture contents the size of droplets increases slightly due to condensation and absorption (less than by 1 %) (Fig. 4b). For calculated scrubber height $H_{ave}=7$ m $W_{i,e}=3,75$ kg/h \cdot m 2 ($\Delta=15.6$ %).

According to comparison, the model agrees well with the experimental data.

In calculations tabular data m_{px} for water solution of CO_2 [6] were approximated by temperature dependence T ,

$$m_{px} = (2,389T^2 - 994,6T + 100765) 10^4, \text{Pa.} \quad (54)$$

Partial pressures of saturated water vapors on droplet and "formation" surfaces were calculated by formula [2] (the partial pressure of saturated vapors of gas components were not taken into account)

$$P_{sd,\delta} = P_{cr} \exp \left(A_1 \ln \frac{T_{d,\delta}}{T_{cr}} + A_2 f_2 \right), \quad (55)$$

where

$$f_2 = \frac{4 \left(\frac{T_{d,\delta}}{T_{cr}} - 1 \right)}{\frac{T}{T_{cr}}} + f_1 - 5, 3 \ln \frac{T_{d,\delta}}{T_{cr}},$$

$$f_1 = \left(\frac{T_{d,\delta}}{T_{cr}} - 1 \right) \left[\frac{\left(\frac{T_{d,d}}{T_{cr}} + 1 \right)^2}{5} + 0, 5 \right],$$

$$P_{cr} = 221.29 \cdot 10^5 \text{ Pa}; t_{cr} = 374.1 \text{ }^\circ\text{C}; A_1 = 7.5480; A_2 = 2.7870.$$

For hydrogen sulfide m_{px} for water solution [6] was approximated by dependence:

$$m_{px} = (-0,0251T^2 + 148,73T - 36374) 10^4, \text{Pa.} \quad (56)$$

Calculation results on absorption of hydrogen sulfide on a water droplet from the vapor-gas flow are shown in Fig. 5.

Theoretical value of $W_{i,th}$ for H_2S ($\rho_{d.a.,0} = 1,0029 \text{ kg/m}^3$) is

$$W_{i,th} = 1,0029 \cdot 0,2 \cdot 0,25 \cdot 0,062478 \cdot 3600 = 11,28 \text{ kg/h m}^2. \quad (57)$$

Calculation by formula (48) with experimental mass transfer coefficient gives for $H=4.3 \text{ m}$

$$W_{i,e} = 28685,9 \cdot 0,1539 \cdot 0,151 \cdot 1,66615 \cdot 0,01253 \cdot 0,7426 = 10,335 \text{ kg/h m}^2, \quad (58)$$

where $\left(\frac{D_{H_2S}}{D_{NH_3}}\right)^{0,67} = \left(\frac{0,127}{0,198}\right)^{0,67} = 0,7426$, $\Delta\rho_{H_2S} \approx \eta\rho_{SO_2,0} \approx 0,22031 \cdot 1,0029 \cdot 0,2 = 0,01253$ kg/m³. Difference between results of (57) and (58) is $\Delta \approx 8\%$. For calculated height $H=7$ m $W_{i,e}=12,257$ kg/h m² and $\Delta \approx 8\%$ on the other hand. In calculations for H₂S the limit of concentration (solubility) in water is not exceeded (solubility for 20 °C is about $3.85 \cdot 10^{-3}$ kg of H₂S/kg of water) (see Fig. 5a)). According to the diagrams, here absorption is completed at 1.3 – 1.5 m from the scrubber inlet.

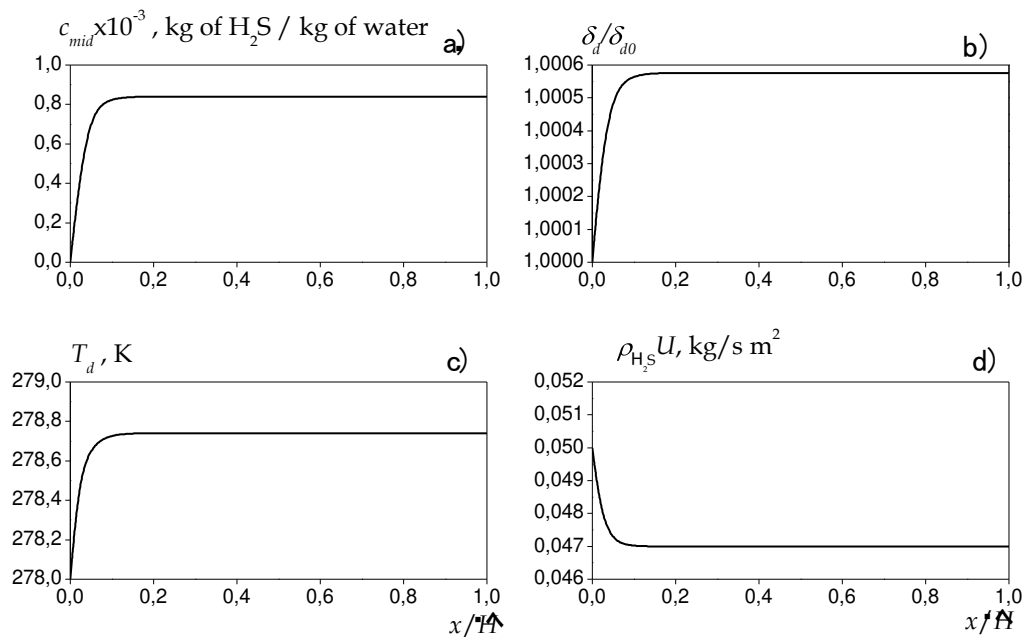


Figure 5. Calculation of hydrogen sulfide absorption: $H=12.75$ m; $q=0.015$ m³/m³; $\delta_{d0}=700$ μm; $V_{d0}=24.5$ m/s; $U_0=0.25$ m/s; $T_{d0}=278$ K; $T_{00}=293$ K; $d_0=0.02$ kg/kg of dry air; $d_{H_2S,0}=0.2$ kg of H₂S/kg of dry air; $\eta_{H_2S}=0.062478$

3.2. Calculation of absorption by solubility of l and q_s

If there are no tabular data for m_{px} (or E) of any gas, and solubility information is available in the hand-book, for instance, for l , m³ of gas/m³ of water and for q_s , g of gas/100 g of water, we can relate l and q_s to the limit density of saturated gas on the droplet surface $\rho_{id,lim}$, kg/m³, taking into account that the process of its dissolution occurs in droplet volume fast, i.e., the typical time of gas dissolution is significantly less than the typical time of droplet stay in the working volume of scrubber. Then,

$$\rho_{id,\lim} = 10 \frac{q_s}{l}. \quad (59)$$

Thus, in [8, p. 260-261] there are tabular data for SO₂ for l and q_s , where we have shown re-calculation of $\rho_{SO_2d,\lim}$ by formula (59) in the last line of Table 2:

$t, ^\circ\text{C}$	0	10	20	30	40
l	79.8	56.7	39.4	27.2	18.8
q_s	22.8	16.2	11.3	7.8	5.41
$\rho_{SO_2d,\lim} = 10 \frac{q_s}{l}$	2.8571	2.8571	2.8680	2.8676	2.8777

Table 2. Volumetric l and weight q_s solubility coefficients for SO

According to this Table, $\rho_{SO_2d,\lim} = 2.8655 \approx 2.9 \text{ kg/m}^3$ and it is almost constant value.

First, for this case we calculate m_{px} :

$$P_{id,\lim} = P_{SO_2d,\lim} = \rho_{SO_2d,\lim} \frac{RT_d}{M_{SO_2}} = 2,8655 \frac{8,314 \cdot 10^3}{64} T_d; \quad (60)$$

$$(m_{px})_{SO_2} = \frac{P_{SO_2d,\lim}}{x_{SO_2d,\lim}}; \quad (61)$$

$$x_{SO_2d,\lim} = \frac{\frac{q_s 10^{-2}}{M_{SO_2}}}{\frac{q_s 10^{-2}}{M_{SO_2}} + \frac{1}{M_{water}}}. \quad (62)$$

As a result, the following approximation was obtained by formula (61) for SO₂

$$(m_{px})_{SO_2} = 2976,58 T^2 - 1594158 T + 215090898, \text{ Pa.}, \quad (63)$$

Knowing $(m_{px})_{SO_2}$ we determine specific heat of SO₂ absorption by water:

$$RTM_{\text{SO}_2}^{-1} \frac{d \ln(m_{px})_{\text{SO}_2}}{dt} = r_{\text{SO}_2}(T_d), \text{ J/kg.} \quad (64)$$

Following calculation is performed by the general scheme (formulas (31)–(35)).

Results of calculation are shown in Fig. 6 at ventilation of air humidity $d_0=0.02$ kg/kg of dry air, $d_{\text{SO}_2,0}=0.2$ kg/kg of dry air, $d_{\text{CH}_4,0}=0.2$ kg/kg of dry air. Other parameters are shown in captions to the figure. It was obtained for this calculation version that $\eta_{\text{SO}_2}=51.1\%$, $\eta_{\text{CH}_4}=0.08\%$. The limit value of SO_2 concentration in a droplet is not achieved even for CH_4 . According to tabular data on absorption coefficient α , m^3/m^3 of water (see Table 1):

$$c_{\text{mid},\text{lim}}=0, 717 \cdot 10^{-6} \alpha, \text{ kg of CH}_4/\text{kg of water.}$$

According to calculation of extracted SO_2 for the given case ($\rho_{d,a,0}=0, 8121 \text{ kg/m}^3$):

$$W_{i,\text{th}}=0, 8121 \cdot 0, 2 \cdot 900 \cdot 0, 51072=74, 656 \text{ kg/h} \cdot \text{m}^2,$$

$$W_{i,e}=28685, 9 \cdot 0, 1539 \cdot 0, 151 \cdot (12)^{0,35} \cdot 0, 8121 \cdot 0, 2 \cdot 0, 51072 \cdot 0, 715=94, 34 \text{ kg/h} \cdot \text{m}^2$$

for $H=12$ m,

$$W_{i,e}=28685, 9 \cdot 0, 1539 \cdot 0, 151 \cdot (7)^{0,35} \cdot 0, 8121 \cdot 0, 2 \cdot 0, 51072 \cdot 0, 715=78, 13 \text{ kg/h} \cdot \text{m}^2$$

$$\text{at } H_{\text{ave}}=7 \text{ m } (H_{\text{ave}}=\sqrt{4, 3 \cdot 12} \approx 7 \text{ m}). \text{ Here } \left(\frac{D_{\text{SO}_2,0}}{D_{\text{NH}_3,0}} \right)^{0,67} = 0, 715.$$

Comparison of $W_{i,\text{th}}$ and $W_{i,e}$ for SO_2 proves good agreement between theory and experiment.

According to calculation, Fig. 6c), methane is not absorbed by water. However, even for methane comparison of calculation with experiment yields satisfactory agreement:

$$W_{i,\text{th}}=0, 1624 \cdot 900 \cdot 0, 00079541=0, 11626 \text{ kg/h} \cdot \text{m}^2;$$

$$W_{i,e}=28685, 9 \cdot 0, 1539 \cdot 0, 151 \cdot 0, 1624 \cdot 0, 00079541 \cdot 0, 715=0, 1026 \text{ kg/h} \cdot \text{m}^2$$

at $H=4.3$ m, $\Delta=11.75\%$. At $H=7$ m, $W_{i,e}=0, 1217 \text{ kg/h} \cdot \text{m}^2$ and $\Delta=4.5\%$.

We should note that absorber height H in experimental dependence for β_{iv} is taken improperly. The optimal and calculated height of setup should equal path l , where the process of component extraction is completed. In most cases of calculations, it completed earlier at the height less than the accepted height of absorber $H=12.75$ m. Therefore, at comparison of calculation and experimental data in experimental dependence for mass transfer coefficient we have varied the calculated height in the range of the heights of tested setups from 4.3 to 12 m [6].

For CH_4 m_{px} is approximated by dependence

$$m_{px} = (-47, 154T^2 + 35490T - 5962310) 10^4, \text{ Pa.} \quad (65)$$

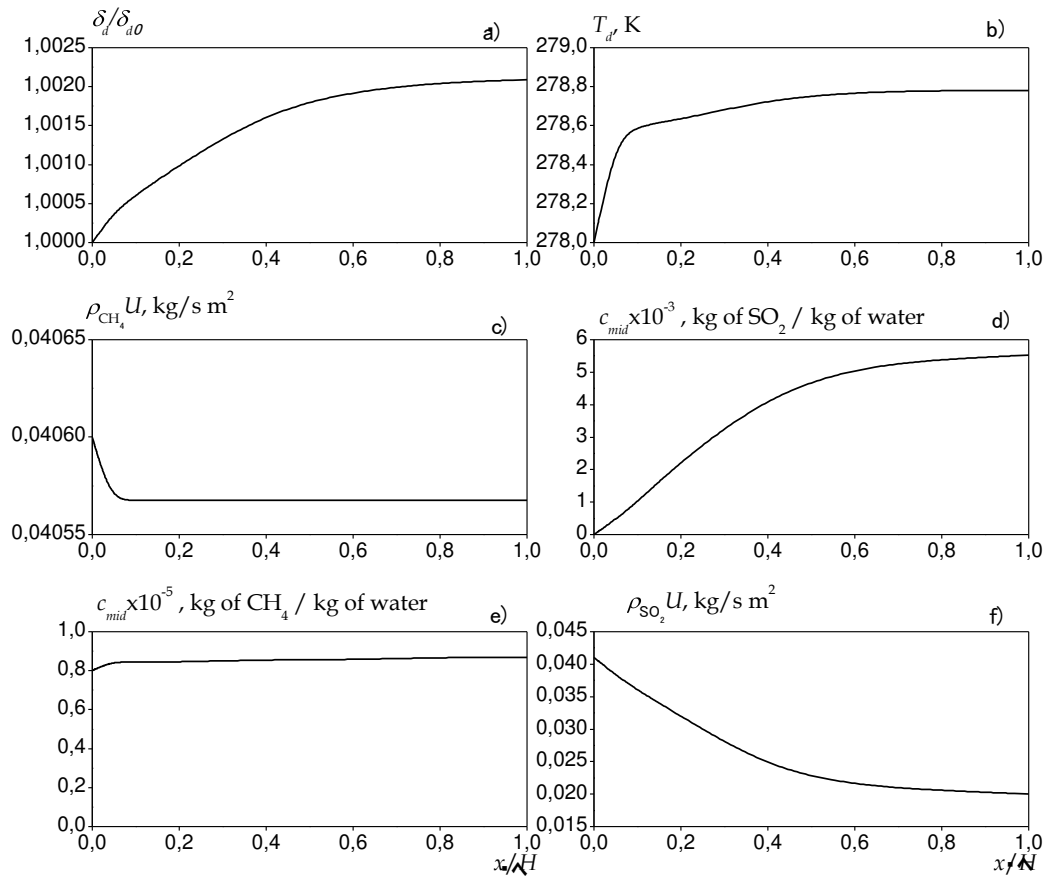


Figure 6. Calculation of SO_2 and CH_4 absorption in direct-flow jet scrubber: $H=12.75$ m; $q=0.015$ m³/m³; $\delta_{d0}=700$ μm; $V_{d0}=24.5$ m/s; $U_0=0.25$ m/s; $T_{d0}=278$ K; $T_{00}=293$ K; $d_0=0.02$ kg/kg of dry air; $d_{\text{SO}_2,0}=0.2$ kg/kg of dry air; $d_{\text{CH}_4,0}=0.2$ kg/kg of dry air; $\eta_{\text{SO}_2}=0.51064$; $\eta_{\text{CH}_4}=0.00079541$

3.3. Calculation of combined absorption-condensation dust-gas cleaning

Calculations of combined condensation dust capture and absorption extraction of hydrogen sulfide from the vapor-air flow in direct-flow hollow scrubber are shown in Fig. 7. Calculated parameters are shown below the figure. According to Fig. 7a), even at increased moisture content the size of droplets increases weak due to condensation. Therefore, for similar processes the equation of droplet motion can be calculated with a constant mass.

An increase in the size of “formations” is more significant due to condensation of water vapors on them: for $\delta_0=0.01$ μm it is 2.1, for $\delta_0=0.1$ μm it is 2.3, and for $\delta_0=1$ μm it is 32 and more for the same total concentration of dust at the inlet of 1.72 g/m³. In the first case, particles are not caught, in the second case, about 5.76 % of particles are caught, and in the third case, 100 % of particles are caught at the inlet to the apparatus. For this version of calculation the stable state by concentrations of H_2S dissolved in droplets and in condensate on “formations”

occurs far from the flow escape from the scrubber. For particles water vapor condensation at flow escape from the scrubber has been also competed already (see Fig. 7g). Therefore, in this case the height of absorber above 1.5 m is excessive, and in construction it can be limited by 2 m. According to Figs. 7d) and 7e), concentration of H_2S dissolved in condensate on the particle and in droplets increases, but it does not exceed the solubility limit (in this case it is about $3.85 \cdot 10^{-3}$ kg of H_2S /kg of water).

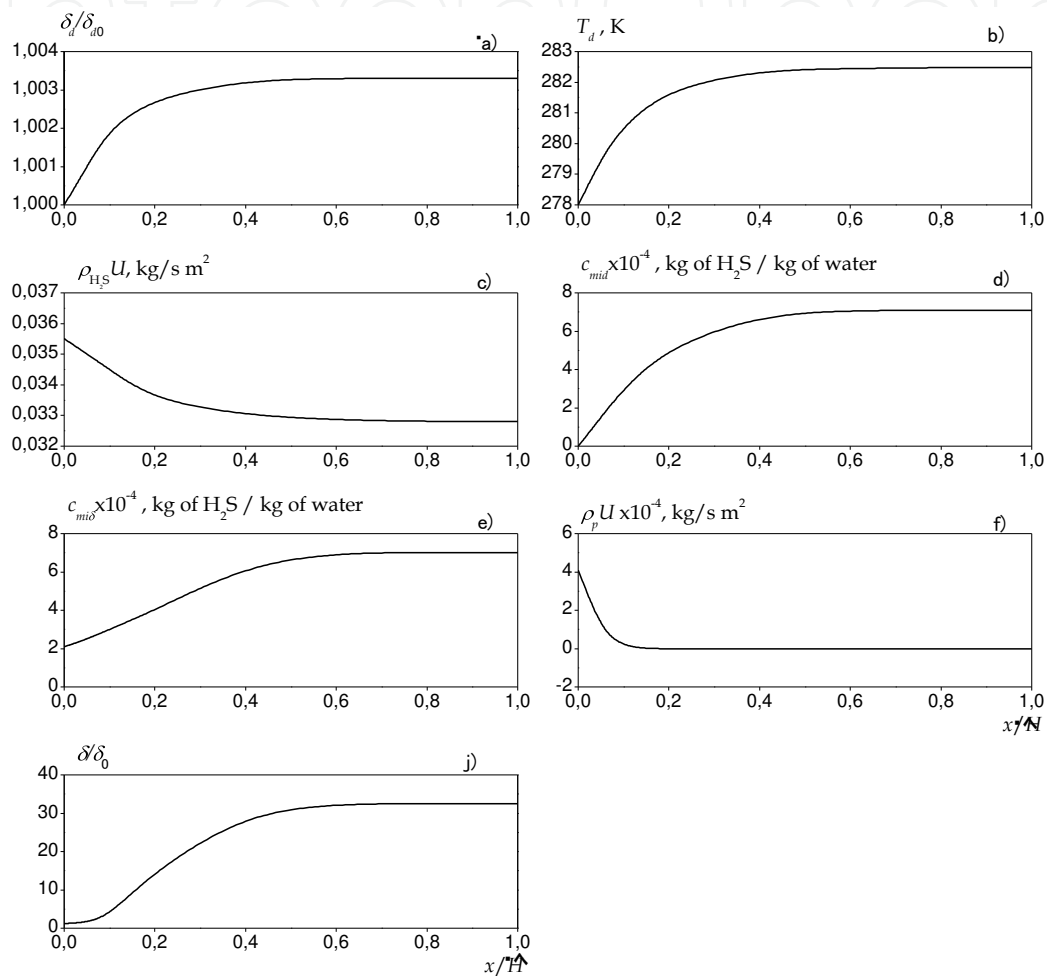


Figure 7. Calculation of combined air cleaning from submicron dust and hydrogen sulfide in direct-flow scrubber: $H = 2$ m; $q = 0.015$ m³/m³; $\delta_{d0} = 700$ μm; $V_{d0} = 24.5$ m/s; $U_0 = 0.25$ m/s; $T_{d0} = 278$ K; $T_{00} = 333$ K; $d_0 = 0.2$ kg/kg of dry air; $d_{H_2S,0} = 0.2$ kg/kg of dry air; $\rho_{p0} = 1.72$ g/m³; $\delta_0 = 1$ μm; $\eta_{H_2S} = 0.075411$; $\eta_p = 1.0$

3.4. Calculation of absorption and condensation dust capture in Venturi scrubber

Calculation results on H_2S absorption and condensation capture of dust with different sizes in Venturi scrubber are shown in Fig. 8. As an example the Venturi scrubber with following parameters was chosen for calculations: diameter of Venturi tube mouth $d_m = 0.02$ m, diffuser

length $l=0.2$ m, diffuser opening angle $\alpha=6^\circ$ ($\alpha=6-7^\circ$, $l/d_m=10-15$ are recommended for normalized Venturi tube [6, 10]), vapor-gas flow velocity in the tube mouth $U_0=80$ m/s, initial velocity of droplets in the tube mouth $V_{d0}=4$ m/s, irrigation coefficient $q=0.015$ m³/m³, temperature of the vapor-gas flow and droplets in the tube mouth $T_{00}=333$ K and $T_{d0}=278$ K, respectively, concentration of dust particles at the inlet $\rho_{p0}=1.72$ g/m³, size of dust particles $\delta_0=0.1$ μ m, moisture content in water vapor at the inlet was set $d_0=0.2$ kg/kg of dry air, gas content $d_{H_2S,0}=0.1$ kg/kg of dry air. Efficient of H_2S extraction and dust capture were determined $\eta_{H_2S}=0.072959$ and $\eta_p=0.52904$, respectively.

The mean-mass size of droplets in the tube mouth was calculated by Nukiyama-Tanasava formula [1]:

$$\delta_{d0} = \frac{0,585}{U_0 - V_{d0}} \sqrt{\frac{\sigma_f}{\rho_f}} + 53,4 \left(\frac{\mu_f}{\sqrt{\rho_f \sigma_f}} \right)^{0,45} q^{1,5}, \text{ m}, \quad (66)$$

where ρ_f (kg/m³), μ_f (Pa•s), σ_f (N/m) and q (m³/m³) are density, dynamic viscosity, surface tension coefficient of pneumatically atomized liquid, and irrigation coefficient.

Velocity U was calculated with consideration of diffuser expansion angle [2, 11].

Dependences of droplet size along the diffuser length are presented in Fig. 8a). It can be seen that firstly condensation of water vapors occurs intensively, then this process stops at the length of $x/l \approx 0.2$, and the size of droplets stays constant up to the scrubber outlet. At this, the quantitative droplet size changes slightly along the diffuser length (it stays almost constant: the maximal increase is a little bit higher than 0.3 %).

A change in droplet temperature due to convective heat transfer between droplets and vapor-gas flow, thermal effects of water vapor condensation on droplets, and gas dissolution is shown in Fig. 8d). A change in mass concentration of H_2S dissolved in a droplet is shown in Fig. 8c). It is obvious that absorption is almost completed at the length of tube diffuser 1 for this version of calculation. The same circumstance is illustrated by mass concentration of H_2S in "formation" condensate along the diffuser in Fig. 8c). According to the figure, the solubility limit on "formations" and droplets is not achieved as in the hollow jet scrubbers. A change in "formation" size due to water vapor condensate on their surfaces is illustrated in Fig. 8f). It can be seen that firstly water vapors condense very intensively, then at the distance of about $x/l \approx 0.1$ this process completes, the size increases more than twice and stays constant until the leaving from the scrubber. Efficiency of dust capture in this version is up to 53 %. Calculation at the same parameters of the vapor-gas flow and dust at the scrubber inlet with mouth $d_i=0.1$ m and constrictor length $l=1$ m gives $\eta_p=0.77729$, $\eta_{H_2S}=0.074965$. It follows from the diagrams in this figure that for the calculated version it is practically reasonable to be limited by diffuser length $x/l \approx 0.4$ ($x=0.08$ m), where the processes of dust capture (Fig. 8g)) and absorption are completed (Fig. 8b)). Therefore, the residual length of 0.12 m is

excessive. Figs. 8b) and 8g) illustrate distributions of dust and H₂S mass fluxes along the diffuser of Venturi tube.

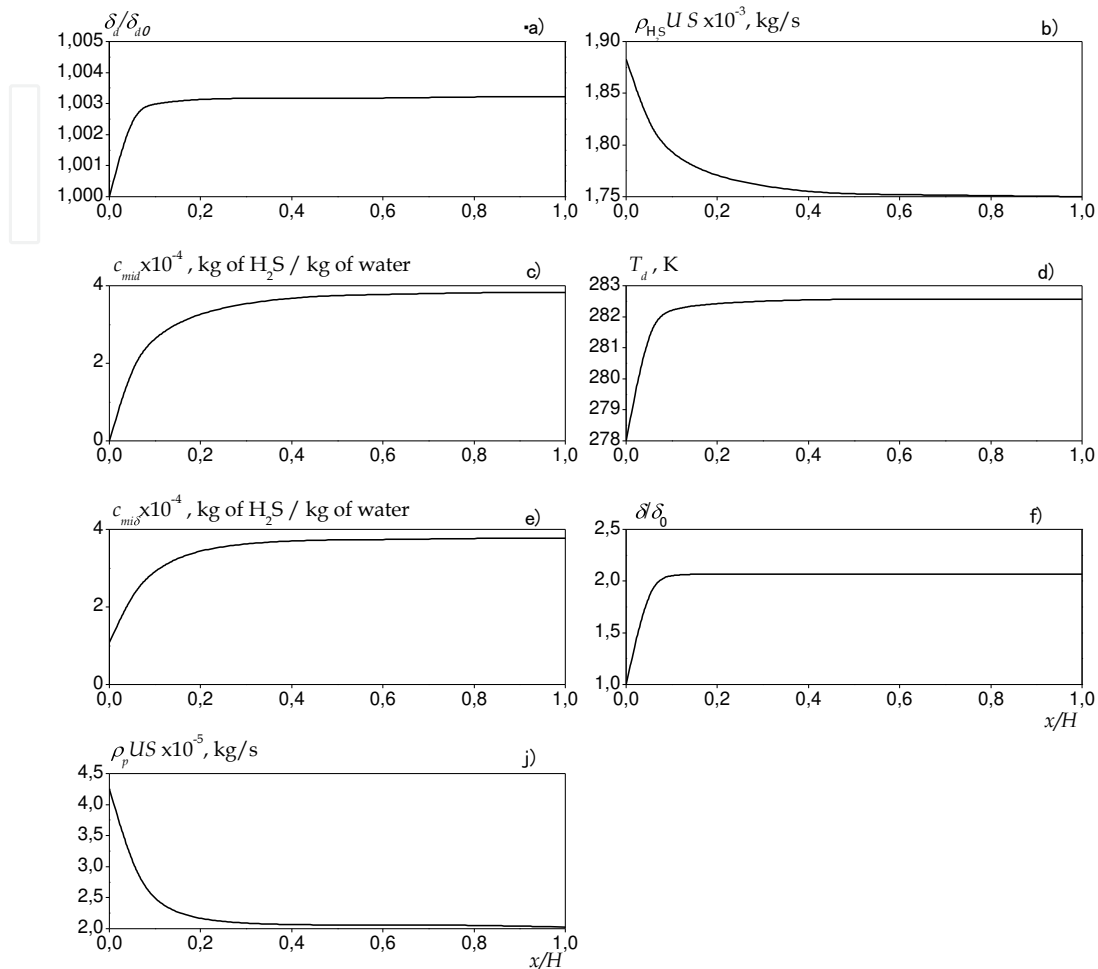


Figure 8. Calculation of H₂S absorption and dust capture in Venturi scrubber (calculated parameters are presented in the text)

It is necessary to note that these calculation versions do not meet the conditions of optimal scrubber operation; they only illustrate the character of complex gas cleaning. To determine the optimal regimes, a series of calculation on the basis of suggested model should be carried out and analyzed for the specific industrial conditions.

Let's turn to comparison of calculation results with the known experimental data. The experimental volumetric mass transfer coefficient is shown in [6] for NH₃ absorption in the Venturi tube with mouth diameter $d_m=0.02$ m. There no geometrical and other parameters. This coefficient is presented as

$$\beta_{iv,e}^{\text{NH}_3} = 260 U_0^{1,56} q_l^{0,57}, \quad (67)$$

where q_l is irrigation coefficient in l/m^3 , U_0 is in m/s , and β_{iv} is in $1/\text{h}$.

The experimental value of H_2S absorbed in Venturi tube is expressed by formula, kg/h ,

$$\Delta G_{i,e} = \beta_{iv,e}^{\text{H}_2\text{S}} \Delta \rho_{i,e} V_{dif}, \quad (68)$$

where V_{dif} is diffuser volume, $\Delta \rho_i$ is calculated drop of gas concentration along the diffuser length, kg/m^3 , corresponding to experimental data.

The theoretical value of mass of absorbed gas, kg/h , is

$$\Delta G_{i,th} = \beta_{iv,th}^{\text{H}_2\text{S}} \Delta \rho_{i,th} V_{dif} \approx \Delta \rho_{i,th} U_0 \frac{\pi d_m^2}{4} 3600. \quad (69)$$

Then

$$\frac{\Delta G_{i,th}}{\Delta G_{i,e}} = \frac{U_0 \frac{\pi d_m^2}{4} 3600 \Delta \rho_{i,th}}{\beta_{iv,e}^{\text{NH}_3} V_{dif} \left(\frac{D_{i,0}}{D_{\text{NH}_3,0}} \right)^{0,67} \Delta \rho_{i,e}}, \quad (70)$$

where (see Fig. 9) the volume of truncated cone is

$$V_{dif} = \frac{\pi}{24} \frac{d_m^3}{\text{tg} \frac{\alpha}{2}} \left[\left(\frac{2l}{d_m} \text{tg} \frac{\alpha}{2} + 1 \right)^3 - 1 \right]. \quad (71)$$

Substituting (67), (71) into (70), at $\alpha=6^\circ$, $l=0.2 \text{ m}$, $d_m=0.02 \text{ m}$, $U_0=80 \text{ m/s}$, $q_l=15 \text{ l/m}^3$ we get the following, assuming that $\Delta \rho_{i,th} \approx \Delta \rho_{i,e}$,

$$\frac{\Delta G_{i,th}}{\Delta G_{i,e}} = 0,9457, \quad (72)$$

where for H_2S $\left(\frac{D_{i,0}}{D_{\text{NH}_3,0}} \right)^{0,67} = 0.7426$, i.e., the difference between the theory and experiment is less than 6 %, what is a good agreement, considering the assumed parameters for normalized Venturi tube.

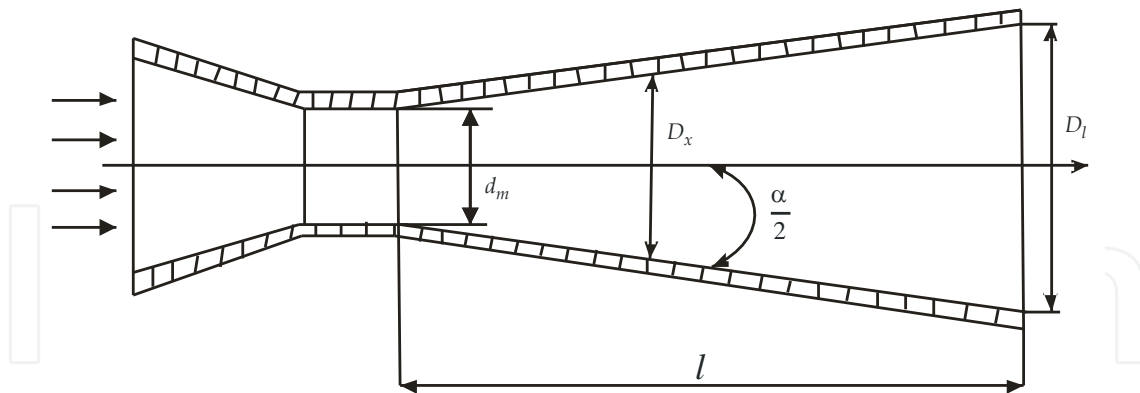


Figure 9. The scheme of Venturi tube

The amount of absorbed H_2S for the version of calculation in Fig. 8 ($\eta_{H_2S}=0.072959$, $\rho_{d.a.,0}=0.7541 \text{ kg/m}^3$) is

$$W_{i,th} = 0,7541 \cdot 0,1 \cdot 0,072959 \cdot 80 \frac{3,14(0,02)^2}{4} 3600 = 0,5 \text{ kg/h.}$$

For the scrubber with $d_m=0.1 \text{ m}$, $l=1 \text{ m}$ at the same dust and gas parameters at the inlet $W_{i,e}=12,78 \text{ kg/h}$.

The experimental values of efficiency of condensation capture of submicron dust are compared with results of model calculation in [2, 11, 12] at the example of deposition of ash particles from cracking gases under the industrial conditions in hollow jet scrubbers [13]; good agreement is achieved.

4. The choice of the value of calculated concentration difference for the absorbed gas component

Let's consider this important question in detail as an addition to iss. 2 at the example of water absorption of SO_2 , comparing calculation and experimental data [6] on volumetric mass transfer coefficient.

It follows from equation (3) that

$$W_{i,th} = - \int_0^H \beta_{id} \pi \delta_{d0}^2 (\rho_{id} - \rho_i) n_d dx, \text{ kg/s} \cdot m^2, \quad (73)$$

where, according to formulas (26) and (38)

$$n_d = 6q \frac{U_0}{V_{dx}} \frac{1}{\pi \delta_{d0}^3}. \quad (74)$$

In (73) and (74), according to calculation results, it is assumed that $\delta_d = \delta_{d0}$.

Let's put β_{id} from (19) to (73) at $K_{ci}=1$ and n_d from (74), we obtain, proved by estimates,

$$0, 276 \text{Re}_d^{0,55} \text{Sc}_i^{0,33} \gg 1, V_{dx} \gg U,$$

$$W_{i,th} \approx -3,312 \frac{qU_0}{\delta_{d0}} \int_0^H \frac{(\rho_{id} - \rho_i) dx}{\text{Re}_d^{0,45} \text{Sc}_i^{0,67}}, \quad (75)$$

where

$$\text{Re}_d \approx \frac{V_{dx} \delta_{d0} \rho}{\mu}. \quad (76)$$

Lets' turn dependence (74) to the following form using the theorem about an average for integral:

$$W_{i,th} \approx 3,312qU_0 \frac{(\bar{D}_{\text{NH}_3})^{0,67}}{\delta_{d0}^{1,45} \bar{V}_d^{0,45}} (\bar{\nu})^{-0,22} \left(\int_0^H \Delta \rho_{id} dx \right) \left(\frac{D_{i,0}}{D_{\text{NH}_3,0}} \right)^{0,67}, \text{ kg/s} \cdot \text{m}^2, \quad (77)$$

where $\bar{\nu} = (\bar{\mu} / \bar{\rho})$ is the average value of kinematic viscosity of the vapor-gas flow in the scrubber volume, m^2/s , \bar{D}_{NH_3} is the average value of diffusion coefficient of methane NH_3 , \bar{V}_{dx} is average velocity of droplets on the 0–H way at motion from scrubber inlet to the outlet, m/s , $\Delta \rho_{id} = |\rho_{id} - \rho_i|$.

Expressing velocity U_0 and V_{dx} in m/h , and assuming $\delta_{d0} = 7 \cdot 10^{-4} \text{m}$ ($700 \text{ } \mu\text{m}$),

$$\bar{D}_{\text{NH}_3}(T_{d,ave} \approx 278, 5 \text{ K}) = 0,198 \cdot 10^{-4} \left(\frac{285}{273} \right)^{1,75} = 0,205 \cdot 10^{-4} \text{m}^2/\text{s} \quad (\text{see Fig. 10 b)),$$

$\bar{\nu} = (285 \text{ K}) \approx 1,5 \cdot 10^{-6} \text{ m}^2/\text{s}$, $\bar{V}_{dx} \approx 5,25 \text{ m/s}$ (see Fig. 10 e)), $(D_{\text{SO}_2,0} / D_{\text{NH}_3,0})^{0,67} = 0,715$, we ob-

tain for calculation parameters of Fig. 10 $W_{i,th} \approx 83 \text{ kg/h} \cdot \text{m}^2$, where $\int_0^H \Delta \rho_{id} dx = 0,6437 \text{ kg/m}^2$

is obtained via model calculation (see Fig. 11).

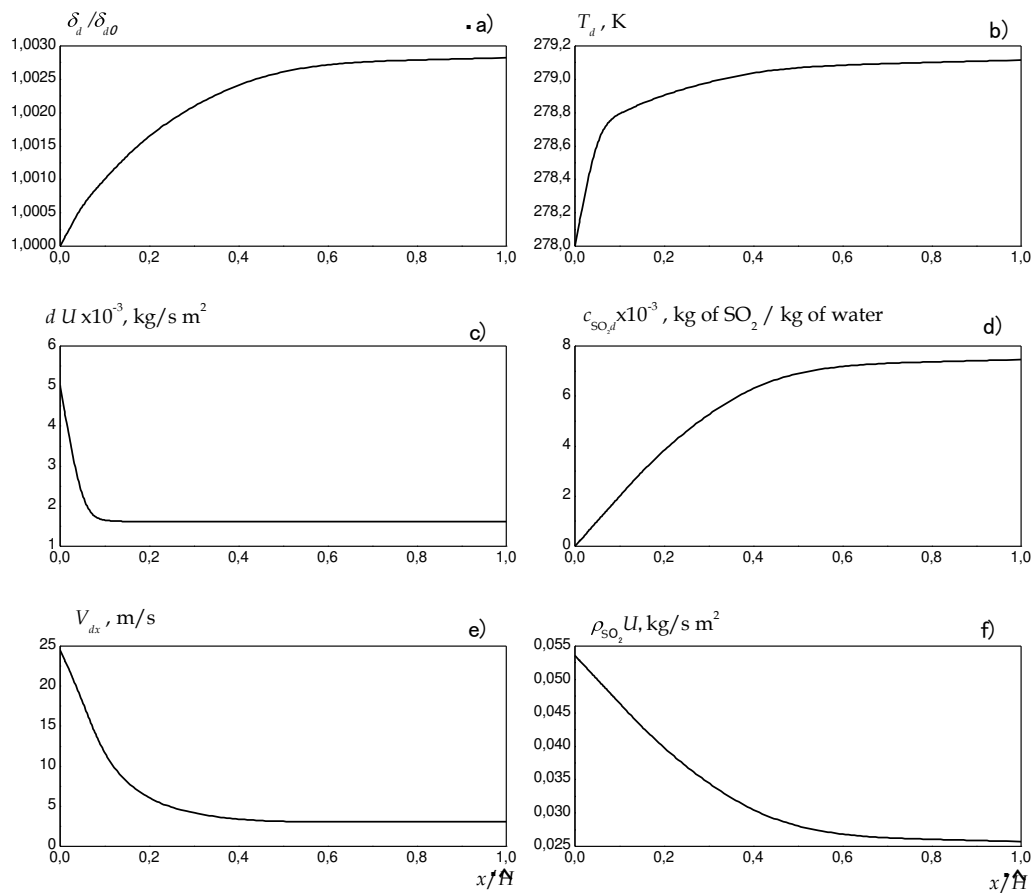


Figure 10. Calculation of SO₂ absorption in direct-flow jet scrubber: $H=12.75$ m; $q=0.015$ m³/m³; $\delta_{d0}=700$ μm; $V_{d0}=24.5$ m/s; $U_0=0.25$ m/s; $T_{d0}=278$ K; $T_{00}=293$ K; $d_0=0.02$ kg/kg of dry air; $d_{SO_2,0}=0.2$ kg/kg of dry air; $\eta_{SO_2}=0.51722$

We should note that multiplier $(D_{i,0}/D_{NH_3,0})^{0.67}$ was included into formula (76) as a correction like to was dome for empirical dependence (46).

Numerical calculation by the model give the value of SO₂ extraction

$$W_{i,th} = \rho_{d,a,0} d_{SO_2,0} U_0 \eta_{SO_2} 3600 = 1,0743 \cdot 0,2 \cdot 0,25 \cdot 0,51722 \cdot 3600 = 100 \text{ kg/h} \cdot \text{m}^2. \quad (78)$$

The difference is 17 %, what is a sequence of simplifications and averaging in dependence (76).

If we assume average concentration difference in accordance to average experimental height $H=7$ m $\Delta\rho_i=0,6437/7=0,092$ kg/m³, then

$$W_{i,e} = 28685,9 U_0^{1,35} q^{0,45} H^{0,35} \Delta\rho_i \left(\frac{D_{i,0}}{D_{NH_3,0}} \right)^{0,67} = 28685,9 \cdot 0,1539 \cdot 0,151 \cdot 1,976 \cdot 0,092 \cdot 0,715 = 86,6 \text{ kg/h} \cdot \text{m}^2.$$

The difference with $W_{i,e}=83$ kg/h•m² is 4.2 %.

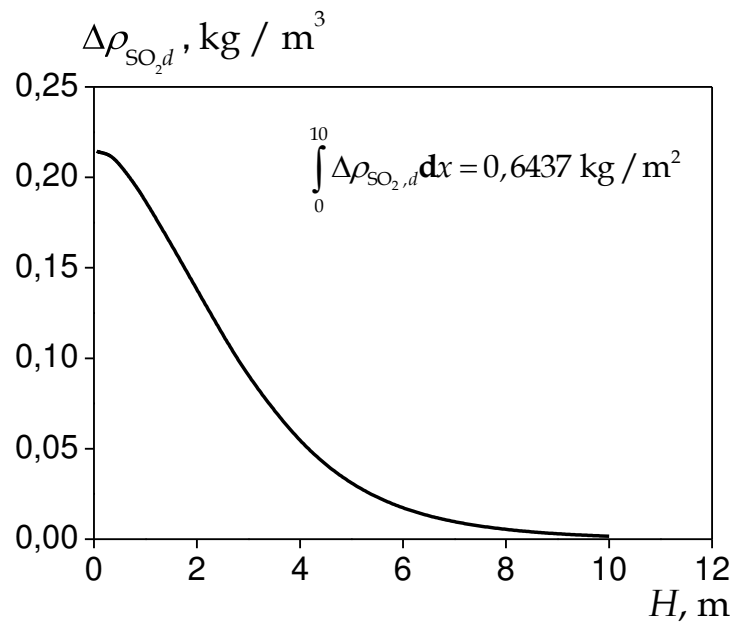


Figure 11. Distribution of $\Delta\rho_{SO_2,d}$ along the scrubber height for calculation parameters of Fig. 10.

If we take $\Delta\rho_i = \rho_{d,a,0} d_{SO_2,0} \eta_{SO_2} = 0,111 \text{ kg/m}^3$, then $W_{i,e} = 104.5 \text{ kg/h} \cdot \text{m}^2$, what differs from result of (77) by similar 4.3 % with accuracy of estimation error. This proves the fact that calculated volumetric mass transfer coefficient agrees empirical dependence (46) of [6].

On the basis of analysis performed the calculated concentration difference should be recommended for practical application as the most appropriate

$$\Delta\rho_i = \rho_{i,0} - \rho_{i,H} \frac{U_H}{U_0} \quad (79)$$

at determination of the value of extracted gas component by formula (49), thus, it is necessary to measure $\rho_{i,0}$, $\rho_{i,H}$ and U_0 , U_H at apparatus inlet and outlet. Calculation of moisture content is shown in Fig. 10 c), at this, it was obtained that $U_H = 0,2234 \text{ m/s}$, $\rho_{d,a,0} = 1,0743 \text{ kg/m}^3$.

5. Conclusions

The suggested physical-mathematical model of complex heat and mass transfer and condensation-absorption gas cleaning from dust and harmful gaseous components is confirmed by the known experimental data and can be used for engineering calculations and optimization of construction and operation parameters of hollow jet scrubbers of direct and counter flow types. This was proved by its numerical implementation for the specific conditions. Calcula-

tions on absorption of some gases (CO_2 , H_2S , SO_2 , CH_4) on water droplets, dispersed by coarse centrifugal nozzles in hollow direct-flow jet scrubber and pneumatic Venturi scrubber from wet air is shown in the current paper together with calculation of combined absorption-condensation air cleaning from H_2S and various-sized fine dust in these apparatuses at an increased moisture content. The system of model equations is written at some certain conditions for the multicomponent vapor-gas mixture with particle. This makes it possible to use this system for calculation of complex gas cleaning from several harmful gas components and several fractions of dust particles and investigate regularities of this process.

Nomenclature

\vec{V}_d vector of droplet velocity

\vec{U} vector of vapor-gas low velocity

\vec{g} vector of gravity acceleration

m_d droplet mass (variable value due to evaporation-condensation and absorption), kg

c_f specific heat capacity of liquid, J/kg•K

T_d mean mass temperature of droplets, K

α_d heat transfer coefficient of droplet, W/m²•K

δ_d mass-median size of droplet, m

T temperature of vapor-gas flow, K

r_i specific heat of absorption, evaporation-condensation, J/kg

c_δ specific heat capacity of “formation”, J/kg•K

ρ_δ mass concentration of “formations” in the vapor-gas flow, kg/m³

$V_c = |\vec{V}_d - \vec{U}|$ module of relative droplet velocity, m/s

η_{Stk} coefficient of “formation” capture by droplets

T_δ mean mass temperature of “formations”, K

β_{id} coefficient of droplet mass transfer with the i -th component of vapor-gas flow by concentration difference, m/s

ρ_{id} and ρ_i partial densities (mass concentrations) of saturated vapors of dissolvent and gas components near droplet surface and far from it (in the flow), kg/m³

n_d and n_δ calculated concentrations of droplets and “formations” in the flow, $1/\text{m}^3$

ρ_g partial density of non-reacting gas component, kg/m^3

ρ_d mass concentration of droplets, kg/m^3

α_δ heat transfer coefficient of “formation”, $\text{W}/\text{m}^2\cdot\text{K}$

δ size of “formation”, m

ρ density of vapor-gas flow, kg/m^3

c specific heat capacity of vapor-gas flow, $\text{J}/\text{kg}\cdot\text{K}$

ρ_p mass concentration of dry dust particles in the flow, kg/m^3

ρ_f density of liquid (droplets), kg/m^3

μ dynamic viscosity of vapor-gas flow, $\text{Pa}\cdot\text{s}$

M_i molar masses of components of the vapor-gas mixture, kg/kmole

$R=8,314 \text{ kJ}/\text{kmole}\cdot\text{K}$ universal gas constant

D_i diffusion coefficient of mixture component, m^2/s

P_i partial pressure of the i -th component of the vapor-gas mixture, Pa

$P_{id,\delta}$ partial saturation pressures of mixture components, calculated by droplet and “formation” temperature, Pa

$m_{px,i}$ constants of phase equilibrium of solutions of i -th components of extracted gases, Pa

$x_{id,\delta}$ mole fractions of gas components dissolved in a droplet and “formation” condensate

$c_{mid,\delta}$ mass fractions of gas components in droplet and “formation” dissolvent, kg/kg of dissolvent

M_{dis} molar mass of dissolvent, kg/kmole

$dm_{v\delta}/d\tau$ rate of “formation” mass change due to evaporation-condensation of liquid, kg/s

δ_0 initial size of dust particles, m

ρ_s density of solution on “formation” due to condensation of liquid vapors and absorption of gas components, kg/m^3

d moisture content, kg of vapors/ kg of dry non-reacting component of vapor-gas mixture

d_i gas content, kg of reacting gas component/ kg of dry non-reacting component of vapor-gas mixture

$q=Q_f/Q_{sg0}$ irrigation coefficient

Q_f volumetric flow rate of liquid, m³/s

Q_{sg0} volumetric flow rate of vapor-gas mixture at apparatus inlet, m³/s

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