Heat and mass transfer under the film condensation from vapour-gas-dust mixtures

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Abstract

Results of the theoretical and experimental investigations of the film condensation from vapour-gas-dust mixtures have been presented. Main factors which influence the drift of particles of disperse phase to the surface of the condensate film have been investigated by numerical experiment. For the particle located nearby the interface the time of its capture by the liquid film has been evaluated. It is shown that dust component influences the regime of condensation and changes the heat and mass transfer intensities at the initial site of cooled wall.

1 Introduction

In many technological systems vapour-gas mixture being directed to the condenser contains disperse solid phase. The characteristic example is the condensation of phosphorus from the hot furnace gas. In that case a condensation occurs both on the cooled walls and on the surfaces of dust particles in the vapour-gas flow. Many works are devoted to the second phenomenon [1]. However, the process of catching dust particles by the falling condensate film has not been studied practically. Even though we have now a good deal of experimental data available [1], their satisfactory interpretation is absent. This problem is relevant as it is a question of the quality of condensation products.

In this paper we present results of our theoretical and experimental investigations, analysis of which allows to clarify this problem and to offer methods for calculating the heat and mass transfer under the film condensation from vapour-gas mixtures containing a disperse solid phase.
2 Mathematical model

2.1 Particles motion nearby the surface of condensate film

The model for describing the motion of small particles in the gas flow with complex structure has been carried out in [2]. By means of those methods some important engineering problems have been solved. However, assumptions used in [2] are quite correct for particles with diameter \( d \approx 0.001 \text{ m} \), but they can't be considered as well-founded for dust particles whose diameters are lesser by two (or more) orders.

First, in that case the adequate description for the particle movement is given by Stokes law. Second, nearby the condensate film surface the role of lifting force can essentially increase [2]. Finally, the role of Stefan flux may also be significant provided the equality of orders of the Stefan flux velocity and relative velocities of particles.

Hence, the expression for lifting force \( F_\perp \) induced by great gradients of gas velocity in the boundary layer near the interface is given by [3]:

\[
F_\perp = K \operatorname{sgn} \left( \frac{\partial U_x}{\partial y} \right) \mu d^2 \sqrt{\frac{1}{\nu} \left| \frac{\partial U_x}{\partial y} \right| (U_x - V_x)}. \tag{1}
\]

Let's suppose that particle flies up the boundary layer with settling velocity. Thus

\[
3(C_1) \pi \mu d (V_x - U_x) - \frac{\pi d^3}{6} \rho g = 0,
\]

which gives for the lifting force

\[
F_\perp = -\frac{K \rho g}{18(C_1)} d^4 \sqrt{\frac{1}{\nu} \left| \frac{\partial U_x}{\partial y} \right|} \operatorname{sgn} \left( \frac{\partial U_x}{\partial y} \right). \tag{2}
\]

Assuming the linear profile of gas velocity nearby the interface we obtain [3]

\[
\frac{\partial U_x}{\partial y} = \left( U_{g \infty} - U_t \right) \frac{\sqrt{U_{g \infty} - U_t}}{K_1} \frac{\rho_g}{\mu_g L}.
\]

Hence, acceleration of the transversal motion of the dust particle can be written as

\[
a_\perp = -\frac{K d \rho g}{3\pi(C_1)} \left( \frac{1}{\nu K_1} \sqrt{\frac{\rho_g}{\mu_g L}} \left| U_{g \infty} - U_t \right| \left( U_{g \infty} - U_t \right) \right)^{1/2} \operatorname{sgn} \left( U_{g \infty} - U_t \right). \tag{3}
\]

From (3) it follows that transversal acceleration is proportional to the diameter of the particle. This acceleration is directed towards the film surface if
\( (U_{g,\infty} - U_I) < 0 \) (it is correct, as a rule, in the case of condensation from a rest vapor-gas system), and directed away from the film surface vice versa. For the particle initially located at the distance \( S \) from the interface we can obtain the following estimate for the time of its drift towards the film surface

\[
\tau \approx \sqrt{\frac{2SL}{dgRe_{rel}^3}} \tag{4}
\]

Some results of numerical experiments are shown in Fig. 1.

\[\tau^1\] - drift time of the particle with diameter \( 10^{-3} \) mm under the relative Reynolds number in the boundary layer: \( \text{Re}_{rel} = 5 \)

1- \( \text{Re}_{rel} = 5 \); 2- \( \text{Re}_{rel} = 10 \); 3- \( \text{Re}_{rel} = 20 \); 4- \( \text{Re}_{rel} = 40 \).

Figure 1: Relative time of the dust particle drift towards the film surface

Evaluation for the diameter of dust particles which may be carried away by Stefan flux follows from the known expression [1]. It reads

\[
d \leq \sqrt{\frac{9D\mu}{2g\delta (\rho_d - \rho_s)}} \ln \frac{P - p_S}{P - p_{\infty}}. \tag{5}
\]
2.2 Penetration of hydrophobic particles into the condensate film

Conditions for the dust particle to be caught by the condensate film can be obtained for the following reasons.

First, the hydrophilic particles hitting the film surface will be, as a rule, caught by the film. And the hydrophobic particle, which moves with sufficient transverse velocity \( V_\perp \), also can penetrate into the condensate film.

The appropriate scheme of the particle penetrating into the liquid film is shown in Fig. 2.

Figure 2: Scheme of the dust particle penetration into the condensate film

Thus for the penetration could take place the initial kinetic energy \( E_0 \) of the particle had to be sufficient to compensate the work of surface forces \( F_\sigma \).

The above work reads

\[
A_\sigma = \int \frac{d}{2} F_\sigma \sin(\theta + \alpha) dx.
\]

where \( \theta \) is the equilibrium contact angle. Here we suggest that for the small dust particle the distinction between equilibrium and dynamic contact angles can be ignored.

Then it reads

\[
A_\sigma = 2\pi \sigma \int \frac{d}{2} \sqrt{\frac{d^2}{4} - x^2} \sin(\theta + \alpha) dx =
\]
After rearranging we obtained the formula for the work of complete penetration of a hydrophobic particle into the condensate film:

\[ W = \frac{\pi \sigma d^2}{3} (2 \sin \theta + \cos \theta) . \]  

(7)

The initial kinetic energy of particle is given by

\[ E_0 = \frac{m_d V_x^2}{2} = \frac{\pi d^3}{12} \rho (a \tau)^2 . \]

From this it follows:

\[ (a \tau) \geq \sqrt{\frac{4 \sigma}{\rho d}} (2 \sin \theta + \cos \theta) . \]  

(8)

Thus the estimate for the vapour-gas mixture velocity gradient nearby the surface of the condensate film looks as follows

\[ \nabla U_s \sim \frac{\sqrt{4 \sigma(2 \sin \theta + \cos \theta) / \rho}}{d \text{Re}_{rel}^{3/8}} . \]

The capture time can be calculated from the momentum equation for the particle which penetrates into the liquid film.

The equation reads

\[ m \dot{x} = \pm F(x) , \]  

(9)

where

\[ F(x) = 2 \pi \sigma \left[ \frac{d^2 - 4 x^2}{2d} \sin \theta + \frac{xyd^2 - 4 x^2}{d} \cos \theta \right] . \]  

(10)

Initial conditions:
\[ x(0) = 0, \]
\[ \dot{x}(0) = a_1 \tau \]

From (9) it follows
\[
\int_{\frac{d}{2}}^{\frac{d}{2}} \left( 2 \int F(\xi) d\xi + C \right)^{-\frac{1}{2}} dx = \tau_p, \quad (11)
\]

where
\[ C = a_1 \tau - \frac{2\sigma \cos \theta}{\rho d}. \]

Let's designate as \( \tau^* \) the total time for drifting and penetrating the dust particle into the condensate film, i.e.
\[ \tau^* = \tau + \tau_p. \quad (12) \]

As the time \( \tau^* \) is a function of the particle diameter, the density of the distribution of dust particles by sizes inside the liquid film will differ from the density of particles distribution nearby the film surface in the vapour-gas phase \( P_N(d) \).

Let \( F(d) \) be the function of distribution of the dust particles by sizes in the vapour-gas phase such as
\[ F(d) = \int_{0}^{d} P_N(s) ds. \quad (13) \]

Thus if \( \Delta \tau \) is the average characteristic time of vapour-gas phase flow through the apparatus then re-normalized density of distribution of the particles inside the liquid film can be written as
\[ p_N^*(d^*) = \frac{F(d)}{\int_{0}^{d^*} F(s) ds}, \quad (14) \]

where \( d^* \) is defined from (11) provided that
\[ \tau^*(d^*) = \Delta \tau. \quad (15) \]
So, we have

\[
F_N^*(d) = \frac{\int_0^d F(s)ds}{\int_0^d F(s)ds}.
\]  

(16)

Figure 3 depicts the results of comparing the experimentally obtained functions of distribution of the dust particles by sizes both in the vapour-gas phase (curve 1) and in the condensate film (curve 2) with distribution obtained by computer simulation (curve 3). It’s obvious that presented model correctly describes the shift of \( F^*(d) \) curve.

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**3 Heat and mass transfer: experimental results and interpretation**

The principal aspect of the problem discussed is linked to the experimentally found sharp decrease of the heat transfer coefficient under the condensation from vapour-gas-dust mixtures in comparison with that coefficient under the condensation from pure vapour-gas systems. It’s found also that small particles in vapour-gas mixture highly influence the temperature drop at the liquid-vapour interface. In our experiments the maximum temperature drop measured nearby the condensate film surface increased from 4 K for pure vapour-gas mixture (air-glycerin vapour) to 10 K for vapour-gas-dust mixture. Besides, under the con-
condensation from vapour-gas-dust mixture the temperature of cooled wall was higher than under the condensation from pure two-phase system.

Even small deposition of dust at the initial site seems to lead to transition from drop condensation regime with high heat transfer coefficients to the film condensation with much lesser heat transfer intensity. It's known that regime of drop condensation will be stable under the conventional Reynolds number till $1.8 \cdot 10^{-2}$. In that case there is no continuous liquid film along the cooled surface. Therefore, it is accepted to calculate above conventional Reynolds number from formula $\text{Re} = \frac{w R_c}{v_l}$. In the absence of dust in the vapour-gas mixture the average number $n_i$ of condensate drops per unit of cooled surface depends only on surface wetting at the initial site but $n_i$ doesn't depend on the heat flux. It follows from theoretical consideration, and experimental data corroborate the theoretical conclusions [1]. However, if the system contains disperse solid phase, the situation will essentially change. The higher temperature of saturated vapour, the smaller critical diameter of dust particle by which the drop regime of condensation at the initial site becomes unstable. In the presence of small dust particles stuck to the cooled surface the emergence of condensate germs also becomes easier. Thus even if there is not complete surface wetting, the condensate drops start to spill over the cooled surface (Fig. 4).

![Figure 4: Drop regime of condensation from pure vapour-gas systems -a) and film regime of condensation from vapour-gas-dust mixtures - b)](image)

Transition to the film condensation regime leads to the sharp decrease of heat transfer coefficients. Indeed, heat transfer coefficients calculated by the known methods have the order of $10^3$, whereas experimentally measured heat transfer coefficients under the condensation from vapour-gas-dust mixture have the order of not more than $10^2$. Moreover, experiments show that influence of non-condensable components on the heat transfer intensity is higher under the drop condensation than under the film regime. It's explained by the fact that while decreasing the partial pressure of condensing component the Laplace pressure dif-
ference becomes more considerable. Experiments show that decrease of heat transfer coefficients at the top point of cooled wall corresponds to the calculations made with allowance for the transition from drop to the film condensation regime (Fig. 5). Owing to Stefan flux the capture of smallest particles by condensing phase is most intensive just at the initial site.

Vapour-gas mixture consumption $G_c = 12.5 \, \text{m}^3/\text{h}$; water consumption $G_b = 0.1 \, \text{m}^3/\text{h}$; $y_V = 0.4$. Dust concentration: $\bullet - g_d = 2.5 \, \text{gr/m}^3$; $\nabla - g_d = 2 \, \text{gr/m}^3$; $\circ - g_n = 0.5 \, \text{gr/m}^3$; $\ast - g_n = 0$, $T = 350 \text{K}$; $\diamond - g_n = 0$, $T = 345 \text{K}$.

Figure 5: Average (a) and local (b) heat transfer coefficient under the condensation from vapour-gas-dust mixtures.

Influence of the concentration $y_V$ of the condensable component in the vapour-gas-dust mixture at the inlet of the apparatus is illustrated in Fig. 6. Increase of the initial concentration $y_V$ from 0.1 to 0.9 led to increasing the average heat transfer coefficient $\overline{\alpha}$ from 50 to 140 $\text{W/(m}^2\text{K)}$ and to increasing the average mass transfer coefficient $\overline{\beta}$ from $0.8 \cdot 10^{-9}$ to $12 \cdot 10^{-9} \, \text{kg/(m}^2\text{s} \cdot \text{Pa})$. Influence of the dust particles finds expression in the decrease of the heat transfer coefficient under the condensation and, quite the reverse, in the increase of the mass transfer coefficient.

The observed increase of mass transfer coefficients under the condensation from vapour-gas-dust system can be explained by the contribution of the phase transition taking place on the dust particles surfaces. Indeed, while measuring the condensate consumption at the outlet of condenser it is difficult to separate the condensate forming only on the cooled wall.
1- $\bar{\alpha}$ calculated in accordance with the inlet temperature; 2- $\bar{\alpha}$ - in accordance with the average temperature; 3- mass transfer coefficient $\bar{\beta}$.

Figure 6: Average heat and mass transfer coefficients as functions of the concentration of condensable component in the vapour-gas-dust mixture

4 Conclusions

The transversal acceleration of the dust particles in vapour-gas-dust mixtures nearby the condensate film surface is proportional to the diameter of the particle. This acceleration is directed towards the film surface in the case of condensation from the rest (or low Reynolds flowing) vapour-gas-dust system. As a result, dust particles penetrate into the condensate film and stimulate the transition of condensation regime from drops to the film. This phenomenon leads to decreasing the heat transfer coefficient under the condensation.

References

