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Droplet Transfer in Two-Phase Annular Mist Flow*

(Part 1, Experiment of Droplet Transfer Rate and Distributions of Droplet Concentration and Velocity)

By Sadahiro NAMIE** and Tatsuhiro UEDA***

The experimental results are described on the transfer rate of liquid droplet onto the duct wall, the distributions of the droplet volumetric fraction, the droplet and gas velocities, and the pressure loss for a two-phase (Air -Water) annular mist flow through a horizontal rectangular duct. The mean diameter of droplets was in a range from 27 to 40 microns in this experiment.

The droplet transfer coefficient onto the wall derived from the experimental results indicates that it is affected by the droplet concentration as well as the gas velocity. Therefore, for investigating the mechanism of droplet transfer, the effect of the eddy diffusivity of the gas flow containing the droplets is discussed from the relation between the distribution of velocity and that of shear stress in the gas stream.

1. Introduction

The droplet transfer phenomena onto a duct wall in a two-phase mist flow are of major importance in heat transfer of a high quality two phase flow in connection with dryout of heating surface and the heat transfer coefficient in the liquid deficient flow region. The characteristics of axial droplet transfer in a gas phase stream and especially of the droplet deposition rate onto the duct wall which is an important factor for controlling liquid film flow rate along the duct wall have recently been investigated as an interesting problem.

The first experimental study of these phenomena was performed by Alexander et al.⁽¹⁾ with droplets sprayed axially through a nozzle into a gas flow in a 45 mm diameter tube. They calculated the droplet transfer coefficient onto the tube wall by measuring the variation of total droplet flow rate along the tube. The data of the coefficient obtained in downstream part of the tube where the droplet distribution was estimated to be fairly developed gave relatively high values. This result seems to be caused by the surviving effect of the radial velocity of droplets given initially at the nozzle.

Recently, Cousins et al.⁽²⁾ and Paleev et al.⁽³⁾ measured the droplet transfer rate in a vertical tube

(10 mm in inside diameter) and a horizontal rectangular duct (4 mm by 48 mm) respectively for the mist flow produced from an annular mist flow by sucking off the liquid film through the wall. Though there is some scatter in their data, Paleev et al. suggested an empirical equation for the droplet transfer rate. The equation indicates that the droplet transfer coefficient is approximately proportional to the gas velocity and decreases with an increase of the duct size. Furthermore, it should be noticed that they pointed out the effect of droplet concentration on the coefficient.

On the other hand, Friedlander et al.⁽⁴⁾ performed an experiment on the flow of a mixture of gas and certain small solid-particles in vertical tubes of three diameters (5 mm, 13 mm and 25 mm). In the case of solid particles it is comparatively easy to obtain uniform diameter of particles, but the technique for measuring the particle transfer rate onto the wall is rather difficult because the reflection rate of particle at the wall surface increases as the gas velocity increases and as particle size becomes larger.

Though there are other studies⁽⁵⁾ using solid particles, the sizes of those utilized are less than 8μ and smaller than those of droplets in the mist flow except a part of Friedlander's data. These results with small solid particles show that the transfer coefficient increases with that of the gas velocity and particle size and decreases with an increase of the duct size.

In this report, the characteristics and the mecha-

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** Ministry of Transport, Mitaka, Tokyo.

*** Professor, Faculty of Engineering, University of Tokyo.

nism of droplet transfer onto a duct wall were studied experimentally for an annular mist flow of air-water mixture in a horizontal rectangular duct. The droplet transfer coefficient onto the wall is defined and derived from the experimental results of the droplet transfer rate. In this experiment, the distributions of droplet concentration, droplet velocity and gas velocity, the droplet size and the pressure drop along the duct were measured. The effect of the turbulence of gas phase on the droplet transfer is discussed from the relation between the velocity distribution and the shear stress distribution in the gas stream.

2. Experimental apparatus and techniques

2.1 Experimental apparatus

Figure 1 shows a schematic diagram of the experimental equipment. The test duct is made of acrylic resin and has a rectangular cross section 10 mm high and 60 mm wide. The experiment was carried out on a two-component air-water mixture. The air supplied from compressor is introduced into the test duct after passing through two surge tanks, an air cooler and a calibrated orifice. The water from a pump is also introduced into the test duct through a calibrated orifice and a porous sinter-plate mounted on the bottom surface of the duct. Air-water annular mist flow is formed in a droplet-generating section, the upstream half of the test duct approximately 2.5 m long, and droplets generated from the liquid film surface flow into a test section, the downstream half of the test duct, together with the gas flow. The liquid film along the duct wall is mostly sucked off by a vacuum pump through porous sinter plates mounted on the upper and lower surfaces of

the duct at the end of the droplet-generating section, and then flows into a water tank through a calibrated rotormeter.

The amount of liquid sucked off through the porous sinter plates was controlled so as to leave a small amount of liquid film flow on the wall of the test section. Although the main purpose of this procedure is to prevent the reflection of the droplets deposited on the duct wall, this system was also used to study the effect of film flow on the droplet transfer rate by controlling the liquid film flow rate along the test section. Through the porous sinter-plates, a small amount of air was also blown off together with the liquid film, the fraction of the air flow rate withdrawn to the total air flow rate was less than 2%, so it seems that there was no significant effect on the main flow.

In this experiment, two kinds of ducts, 0.5 m and 1.0 m long, were prepared, and by connecting and exchanging them, the droplet flow rate and other quantities were measured at four positions of the test section spaced by 0.5 m interval, as marked in Fig. 1 as (1)~(4).

2.2 Method for measuring droplet transfer rate

At each position of the test section described above, the droplet flow rate was measured by a droplet flow-rate-measuring device shown in Fig. 2(a). In this device fixed at the exit of the test section, a droplet-sampling duct which has a little smaller rectangular cross section than that of the test duct is supported. Through this sampling duct, the droplets in the main flow were introduced with the gas flow into an exit cyclone separator and then the amount of droplets separated there was measured with a messycylinder.

The droplet transfer rate onto the duct wall was determined from the decrease of the measured total droplet flow rate along the test section. Although the droplet transfer rate is also obtained from the increase of the liquid film flow rate on the wall along the test section, the error caused by this method is larger than that by the method described above. Since the mean liquid film thickness was approximately 0.2 mm in this experiment, as shown in Fig. 2(a), both the upper and lower gaps 0.8 mm wide between the sampling duct and the test duct were provided to avoid introducing the crest of liquid film into the separator. By measuring the distribution of the droplet flow rate across the test section, it was confirmed that the amount of droplets drained off through those gaps together with the liquid film flow was very small in comparison with the total droplet flow rate.

Fig. 1 Experimental apparatus

2.3 Measurement of velocity and droplet distributions

For measuring the distributions of gas velocity, droplet velocity, and droplet concentration, an equipment shown in Fig. 2(b) was used. This equipment consists of a movable probe essentially the same as the "isokinetic probe" used by Silvestri⁽⁶⁾. Through the inner tube of this probe, air and droplets mixture was withdrawn at the same velocity as that of the main flow in the test duct, and the flow rates of air and droplets were measured with a rotometer and a messycylinder respectively, after being separated. The probe consists of the inner tube for withdrawing fluid and the outer one for measuring a static pressure in the inner one. In order to achieve the isokinetic withdrawing condition, the pressure just behind the entrance of the inner tube was kept equal to that in the main flow during the measurement by adjusting the suction valve ① of a vacuum pump. By closing the valve ② and a three way cock and opening the valve ③, this probe was also used for measuring the total pressure of mist flow. The method for calibrating this equipment will be described in the latter part of this paper.

Besides this method, the droplet velocity was also determined by a photographic method, in which distances were measured between two images of each droplet on a photofilm exposed to the light of two stroboscopes placed above and under the exit of the

test section and adjusted to be charged at an interval of $50\mu\text{sec}$ by an electrical delay circuit. The light from stroboscopes is introduced through 2 mm slits onto the center plane of the width of the mist flow discharged from the test section, and is caught on the photofilm after being reflected by the surface of the droplets.

2.4 Measurement of droplet diameters

Liquid droplets were caught in an oil film on a glass plate placed behind a shutter with a small slit at the exit of the duct, and diameters of droplets were determined through microscopic measurement. As a droplet-catching oil, a lubricating and a silicon oil, suggested by Tanasawa et al.⁽⁷⁾ in their report, were used.

Besides these measurements mentioned above, a pressure drop was also obtained through four pressure taps located axially on the duct wall. In the preliminary test conducted with a single phase flow, good agreement was obtained between the value of pressure drop measured and that calculated from Blasius' equation.

The range of variables covered in this experiment is listed in Table 1. The symbols in Table 1, $\bar{u}_g$, Q_{d1} and Q_{f1} , denote respectively the mean gas velocity, and the flow rates of liquid droplets and liquid film at the position (1), i.e., at the entrance of the test section. With regard to the data in No. 41, 51 and 61, only the liquid droplet transfer rates were measured.

3. Droplet transfer rate onto the wall and the coefficient of droplet transfer

The liquid droplet flow rate Q_{d1} at the entrance of the test section is controlled by adjusting the liquid flow rate Q_l introduced into the duct inlet, and the liquid film flow rate Q_{f1} on the wall of the test section may be changed by adjusting the withdrawn liquid flow rate through the porous sinter-plates at the end

Table 1 Experimental region

No.	$\bar{u}_g$ m/sec	Q_{d1} l/min	Q_{f1} l/min
1~4	43	0.50	0.16~1.15
10~16	34	0 ~1.15	0.35~0.50
20~25	43	0 ~1.55	0.53~0.60
30~34	63	0 ~1.75	0.32~0.51
41	36	0.45	0.56
51	50	0.45	0.52
61	58	0.45	0.63

(a) Device for measuring droplet flow rate

(b) Equipment for measuring distributions of velocities and droplet concentration

Fig. 2

of the droplet-generating section.

First, the effect of the liquid film flow rate Q_{f1} on the droplet transfer rate was investigated experimentally by fixing the values of the gas velocity $\bar{u}_g$ and the droplet flow rate Q_{d1} . The result is shown in Fig. 3. Each curve plotted in Fig. 3 shows the variation of the droplet flow rate Q_d against the distance x from the entrance of the test section for the respective values of Q_{f1} . It is recognized from this result that the decreasing rate of the droplet flow rate Q_d , which is equal to the droplet transfer rate onto the wall, is not so much affected by the film flow rate Q_{f1} for the range where the value Q_{f1} is less than 1 l/min. Meanwhile the decreasing rate near the entrance of the test section in the case of Q_{f1} being extremely small value ($Q_{f1}=0.16$ l/min.) shows a lower value. This seems to be caused by the fact that a relatively large amount of air and droplets near the surface of the liquid film was withdrawn together with the liquid film in order to reduce the liquid film flow rate of test section Q_{f1} .

On the other hand, the result that the droplet transfer rate appears to decrease in the case of Q_{f1} being larger than 1 l/min. is due to the generation of new droplets from the liquid film surface in the test section. The critical film flow rate, above which droplet-generation from the film surface occurred, was determined against the air flow rates in the preliminary experiment carried out by flowing a small amount of liquid in the test duct. All measurements described later were performed in the conditions of the film flow rate Q_{f1} being less than the critical value.

Figure 4 shows data on the variation of droplet flow rate Q_d against the distance x for various droplet flow rates Q_{d1} in the conditions of Q_{f1} described above. The logarithmic plot of the droplet flow rate Q_d against the distance x turns out approximately as a straight line. So it is confirmed that the decreasing rate of droplet flow, i.e., the droplet transfer rate onto the duct wall, is approximately in proportion to the

Fig. 3 Variation of the droplet flow rate along the axis (Effect of liquid film flow)

droplet flow rate or to the mean droplet concentration at each position of the duct. But it also reveals that since the inclinations of these lines are not same, the decreasing rate is not constant for a wide range of droplet concentrations and becomes larger as the concentration becomes lower.

The definition of the droplet transfer coefficient onto the wall:

As described above, the droplet transfer rate onto the duct wall is not affected by the liquid film flow rate, i.e., the state of the gas liquid interface unless droplet-generation from the liquid film occurs. From this result, it seems that the mechanism of the transportation of large particles like droplets in this experiment is considerably different from that of the usual mass transfer which is controlled by a molecular motion. However since the mechanism of droplet transfer is not clarified yet, the following definition, the similar to the molecular mass transfer, is used here taking into account the fact that the droplet transfer rate is in proportion to the average droplet concentration over the duct for a narrow range of the concentration

$$\dot{m}_0 = K\bar{C} \quad \dots\dots\dots (1)$$

where $\dot{m}_0$ is the droplet transfer rate onto the duct wall per unit surface area in unit time ($\text{kg}/\text{m}^2\text{sec}$), and K and $\bar{C}$ are the coefficient of droplet transfer (m/sec) and the mean droplet concentration over the duct (kg/m^3) respectively.

Putting W_d , W_g (kg/sec) as the flow rates, $\bar{u}_d$, $\bar{u}_g$ (m/sec) as the mean velocities and γ_d , γ_g (kg/m^3)

Fig. 4 Variation of the droplet flow rate along the axis (Effect of droplet concentration)

as the specific weights of droplet and gas phase respectively, the mean droplet concentration $\bar{C}$ is given by the following equation.

$$\bar{C} = \frac{W_d}{\frac{W_g}{\gamma_g} \bar{u}_g + \frac{W_d}{\gamma_d}} = \gamma_g \frac{W_d}{W_g} \frac{\bar{u}_g}{\bar{u}_d} \quad (2)$$

On the other hand, neglecting the droplet transfer to the side walls of the duct, the relation between the droplet transfer rate and the droplet flow rate in the gas core is expressed as follows;

$$\dot{m}_0 = -\frac{1}{2B} \frac{dW_d}{dx} \quad (3)$$

Substituting Eqs. (2) and (3) into Eq. (1), the following expression is obtained for the droplet coefficient K .

$$K = -\frac{\bar{W}_g}{2B\gamma_g\bar{S}} \frac{d(\ln W_d)}{dx} \quad (4)$$

Since the volumetric fraction of droplets and the liquid film in this experiment are very small, W_g may be almost equalized to $BH\gamma_g\bar{u}_g$. Therefore, K is rewritten as follows;

$$K = -\frac{H\bar{u}_g}{2\bar{S}} \frac{d(\ln Q_d)}{dx} \quad (4)'$$

where H is the height of duct (m), B is the width of duct (m), $\bar{S} = \bar{u}_g/\bar{u}_d$ is the ratio of mean velocities and Q_d is the volumetric flow rate of droplets (m³/sec).

The droplet transfer coefficient K defined above is obtained from a slope of Q_d -line in Fig. 4, the gas velocity $\bar{u}_g$ and the measured value of $\bar{S}$ revealed in the later Fig. 12. Figure 5 shows the result of K obtained against $\bar{C}$. Though the transfer coefficient K seems to be approximately proportional to the mean gas velocity $\bar{u}_g$ in a range of low droplet concentrations $\bar{C}$, K decreases as the value of $\bar{C}$ increases and is not necessarily proportional to $\bar{u}_g$ for high values of $\bar{C}$. This phenomenon is considered to be caused by the effect of the droplet concentration on the turbulence of the gas flow as described later.

The value of K measured by Alexander et al.⁽¹⁾ is about two times larger than that obtained in this experiment, which seems to be due to the effect of

Fig. 5 Droplet transfer coefficient

the radial velocity component of droplets given at the spraying nozzle as described previously.

Although the value of K obtained by Cousins et al.⁽²⁾ for a 10 mm inside diameter tube has some scatter over a range of 0.10–0.45(m/sec), the result shows a trend like Fig. 5 if the data were rearranged against the droplet concentration.

The effect of droplet diameter:

The mean diameter of droplets (an arithmetic mean diameter) in this experiment changed over a range of 27–40 μ according to the condition of the flow rates of gas and liquid as shown in Fig. 6. In order to investigate the effect of droplet diameter on the droplet transfer, an experiment was performed by adding a surface active agent to the water phase in the case of $\bar{u}_g = 43$ m/sec. While the mean diameter decreased from about 32 μ to 23 μ in this procedure (as shown in Fig. 6 by a dashed number), there was no remarkable change though the transfer coefficient shows a trend of increasing as the mean diameter decreases.

4. Experimental results of distributions of gas velocity, droplet velocity and droplet concentration

Using the equipment described above, local values of the gas and droplet velocities and droplet concentration were calculated from the following equations similar to those used by Silvestri⁽⁶⁾.

$$G_d = A_s u_d \gamma_d f_d, \quad A_s = \pi d_s^2/4 \quad (5)$$

$$G_g = A_s u_g \gamma_g (1 - f_d) \div A_s u_g \gamma_g \quad (6)$$

where A_s (m²), d_s (m) are the cross area and diameter of the inner tube of the probe, G_d , G_g (kg/sec) are the local flow rates, u_d , u_g (m/sec) are the local velocities of droplet and gas at the probe position respectively and f_d represents the volumetric fraction of droplets in the gas flow. ($f_d \ll 1$, in this experiment). Furthermore, a total pressure p_t (kg/m²) measured with the probe, under the condition of the exit valve ② being closed, is expressed as follows;

Fig. 6 Mean diameter of droplets

$$p_t = \frac{1}{2} \rho_g u_g^2 (1 - f_d) + c \frac{G_d}{g A_s} u_d$$

$$\doteq \frac{1}{2} \rho_g u_g^2 + c \frac{G_d}{g A_s} u_d \quad \dots\dots\dots (7)$$

where ρ_g (kgsec²/m⁴) is the gas density, $\dot{g}$ (m/sec²) is an acceleration of the gravity and c is a coefficient which is related to a momentum exchange of droplets in colliding with liquid surface in the probe. Therefore, if G_d , G_g and p_t are measured, we can obtain u_g from Eq. (6), u_d from Eq. (7) assuming c equals 1.0, and f_d from Eq. (5) substituting the values of u_g and u_d . But in actual measurements, the following precautions are necessary.

First, before measuring G_g , the pressure at the inlet of the probe should be equalized to a static pressure in the main flow near the probe. However, in this probe the pressure is measured at the pressure tap located on the wall of the inner tube, 2 mm behind the entrance of the probe, instead of the pressure in front of it. This measured pressure is smaller than that of the front because of the effects of the wall friction and the flow contraction in the probe. So, if the measurements were performed to equalize this measured pressure to that in the main flow, the flow rate G_g would be underestimated. Then, a preliminary experiment was conducted for a single phase flow in a circular tube to calibrate the flow rate measured by the probe with that obtained by a Pitot tube. In Fig. 7 the calibrated curve is shown. Furthermore, in order to check whether this calibration was applicable or not for the condition of gas flow entraining droplets, a comparison of the integral value of G_g measured by the probe with the inlet gas flow rate was also performed in the case of an annular mist flow through this circular tube. Agreement was good and the difference between these values was less than $\pm 2 \sim 3\%$.

On the other hand, for measuring G_d , the pressure in the inner tube has not so significant an effect as in the case of measuring G_g . The measured value of G_d remained almost constant within about 5% for a difference up to ± 400 mmAq between the static

Fig. 7 Calibrated curve for measuring gas flow rate

pressure in the probe and that in the main flow.

Second, we may substitute 1.0 for a constant c in Eq. (7), only in the case where the droplets colliding with the surface of water filling the probe transfer their momentum completely to the water. The state of the water surface at the entrance of the probe was, therefore, observed using a probe made of glass. Although the surface of water filling the probe was disturbed by a collision of droplets over a range of 2~3 mm long from the tip of the probe, remarkable reflection of droplets at the entrance of the probe was not observed. Hence, it seems reasonable to assume $c \doteq 1.0$.

For measuring the droplet velocity u_d , a photographic method was also used as described previously. A comparison of the results obtained by these two methods is shown in Fig. 8. In general, good agreement was obtained between the "isokinetic" and the photo methods.

Figures 9, 10, and 11 show the results of the distributions of the gas velocity u_g , the droplet velocity u_d and the droplet volumetric fraction f_d respectively. The values of u_g , f_d in Figs. 9 and 11 were obtained by the isokinetic method and u_d in Fig. 10 by the photo method.

The gas velocity distributions are steeper than that of a single phase flow, because the gas flow in an annular flow is influenced by a rough surface effect due to the liquid film. The distributions have a trend of becoming a little flatter as the droplet concentration $\bar{C}$ increases.

The droplet velocity is higher than 90% of the gas velocity and has a fairly flat distribution in the

Fig. 8 Comparison of results obtained by isokinetic probe and photo method

direction of the duct height. The droplets are slightly accelerated along the duct. This may be interpreted as follows: the droplets generated from the liquid film near the end of the droplet-generating section are not yet accelerated sufficiently, and hence, the mean velocity of droplets is relatively low at the cross section No. (1).

The droplet volumetric fraction varies considerably in the direction of duct height and the distributions are a little biased to the lower side of the duct. This distortion of the distribution seems to be depending on the remaining effect of the distribution in the droplet-generating section, where the liquid film on the lower surface is thicker than that on the upper surface. The gravity effect on the droplets is considered to be insignificant because the distortion becomes smaller along the distance x as indicated by data No. 23 shown in Fig. 11.

Figure 12 shows the data of the ratio of mean velocities $\bar{S} = \bar{u}_g / \bar{u}_d$. Though the ratio $\bar{S}$ varies with the conditions of the gas velocity and the measurement position, most of the experimental data fall in a range of 1.0 to 1.1.

5. Shear stress distribution

Let us consider the shear stress at the surface, at height h from the center of the duct, as shown in Fig. 13. In a case where the surface is within the

Fig. 9 Gas velocity distribution

gas flow ($h < h_i$), the momentum equation for x -direction of the hatched element in the figure is,

$$\frac{1}{g} \{ d(W_g \bar{u}_g)_h + d(W_d \bar{u}_d)_h - \bar{u}_d d(W_d)_h \} = -2Bdx_g - 2hBdp$$

Fig. 10 Droplet velocity distribution

Fig. 11 Droplet volumetric fraction in gas flow

where $(W)_h$ and $(\bar{u})_h$ denote the flow rate and the mean velocity through the flow area $(2h \times B)$ respectively, and the subscript "g" represents the gas phase and "d" the liquid phase. Further, u_d^x is the x -directional velocity component of the droplet across the surface at height h , and p is the static pressure. Hence, from the above equation, the shear stress τ_g (kg/m^2) is rewritten as follows;

$$\tau_g = -h \frac{dp}{dx} - \frac{1}{2g} \times \left\{ \frac{d(W_g \bar{u}_g)_h + d(W_d \bar{u}_d)_h - u_d^x d(W_d)_h}{B dx} \right\} \dots (8)$$

On the other hand, in a case where the surface in consideration is within the liquid film flow ($h > h_i$), the shear stress τ_f is expressed as,

$$\tau_f = -h \frac{dp}{dx} - \frac{1}{2g} \times \left\{ \frac{d(W_g \bar{u}_g)_h + d(W_d \bar{u}_d)_h + d(W_f \bar{u}_f)_h}{B dx} \right\} \dots (9)$$

where the subscript "f" represents a liquid film, and here $d(W_g \bar{u}_g)_h$ and $d(W_d \bar{u}_d)_h$ are equal to $d(W_g \bar{u}_g)$ and $d(W_d \bar{u}_d)$ respectively.

The shear stress on the gas-liquid interface, which is denoted by a subscript "i", is obtained from Eqs. (8) and (9) as,

$$\tau_{gi} = -h_i \frac{dp}{dx} - \frac{1}{2g} \times \left\{ \frac{d(W_g \bar{u}_g) + d(W_d \bar{u}_d) - u_d^x d(W_d)}{B dx} \right\} \dots (10)$$

$$\tau_{fi} = -h_i \frac{dp}{dx} - \frac{1}{2g} \times \left\{ \frac{d(W_g \bar{u}_g) + d(W_d \bar{u}_d)}{B dx} \right\} \dots (11)$$

The difference between the above two equations, $u_d^x d(W_d)$, reveals the momentum exchanged by the droplets injected into the liquid film. The shear stress at the duct wall τ_0 is also expressed as,

$$\tau_0 = -\frac{H}{2} \frac{dp}{dx} - \frac{1}{2g} \times \left\{ \frac{d(W_g \bar{u}_g) + d(W_d \bar{u}_d) + d(W_f \bar{u}_f)}{B dx} \right\} \dots (12)$$

Since a two component flow system is used in

Fig. 12 Ratio of mean velocities

this experiment, the momentum of the gas flow is approximately constant along the axial direction. The momentum change of the liquid film flow is also negligibly small comparing with the other terms of the above equations, i.e.,

$$d(W_g \bar{u}_g)_h = 0, \quad d(W_f \bar{u}_f)_h = 0 \dots (13)$$

Further, the measured results have indicated that the droplet velocity distribution is fairly uniform. Hence, $(u_d)_h$ and u_d^x may be assumed to be equal to the mean droplet velocity, i.e.,

$$u_d^x = (\bar{u}_d)_h = \bar{u}_d \dots (14)$$

Figure 14 shows these relations described above. On the basis of these assumptions, the gas shear stress τ_g , Eq. (8), becomes,

$$\tau_g = -h \frac{dp}{dx} - \frac{1}{2gB} \frac{(W_d)_h d\bar{u}_d}{dx} \dots (8)'$$

Therefore, substituting the measured values of the static pressure variation along x direction and of $(W_d)_h d\bar{u}_d$ into Eq. (8)', the shear stress τ_g in the gas flow containing droplets can be calculated. It was shown from this calculation that the gas shear stress decreased as the droplet concentration increased, as shown later in Fig. 15.

6. Consideration of the liquid droplet transfer

The liquid droplet transfer is significantly affected by the droplet concentration as described in section 3, which suggests that the mechanism is considerably different from that of usual mass transfer phenomena.

Therefore, taking into account the fact that the shear stress in the gas flow and also the gas velocity distribution are affected by the droplet concentration, the state of the gas turbulent flow suspending liquid

Fig. 13 Force balance in the duct flow

Fig. 14 Shear stress distribution in the duct flow

droplets was studied and the results was compared with that of the droplet transfer coefficient.

For the single phase turbulent flow in a duct except in the region adjacent to duct wall, as is well known, there is the following relation between the shear stress, the velocity fluctuations u' , v' and Prandtl's mixing length l .

$$\tau_g = \rho_g \overline{u'v'} = \rho_g l^2 \left| \frac{du_g}{dy} \right| \left| \frac{du_g}{dy} \right| \dots\dots\dots (15)$$

where y is the distance from duct wall. This is also expressed in terms of eddy diffusivity ε (m^2/sec) as,

$$\tau_g = \rho_g \varepsilon \frac{du_g}{dy} \dots\dots\dots (16)$$

Hence, introducing a local friction velocity $u_g^* = \sqrt{\tau_g/\rho_g}$, l and ε are expressed as follows;

$$\frac{1}{l} = \frac{1}{u_g^*} \frac{du_g}{dy}, \quad \varepsilon = l^2 \frac{du_g}{dy} \dots\dots\dots (17)$$

Then, the values l and ε can be calculated from the shear stress τ_g and the velocity gradient (du_g/dy) . For example, if the velocity distribution is assumed to follow a logarithmic law, i.e.,

$$\left. \begin{aligned} u_g/u_{g0}^* &= 5.5 + 2.5 \ln(y u_{g0}^*/\nu_g) : \\ &\quad \text{for smooth surface} \\ u_g/u_{g0}^* &= 8.5 + 2.5 \ln(y/k_s) : \\ &\quad \text{for rough surface} \end{aligned} \right\} \dots\dots\dots (18)$$

where $u_{g0}^* = \sqrt{\tau_{g0}/\rho_g}$, ν_g and k_s are the wall friction velocity, the kinetic viscosity and an equivalent roughness respectively, then, the velocity gradient for both (smooth and rough) cases becomes,

$$\frac{du_g}{dy} = \frac{2.5}{y} u_{g0}^* = \frac{2.5}{y} \sqrt{\frac{\tau_{g0}}{\rho_g}} \dots\dots\dots (19)$$

Introducing a shear stress distribution in a single phase duct flow $\tau_{g0} = \tau_g \{H/(H-2y)\}$, the above equation becomes as follows;

$$\frac{du_g}{dy} = \frac{2.5}{y} \sqrt{\frac{H}{H-2y}} \sqrt{\frac{\tau_g}{\rho_g}} = \frac{2.5}{y} \sqrt{\frac{H}{H-2y}} u_g^* \dots\dots\dots (20)$$

Hence, from Eq. (17) the mixing length l is expressed

only by the distance from the wall y , i.e., $l = (y/2.5) \times \sqrt{(H-2y)/H}$. Then, Eq. (20) indicates that the velocity gradient du_g/dy is related as a linear line against τ_g/ρ_g on the logarithmic diagram for the fixed value of y .

Figure 15 shows the experimental data of du_g/dy against τ_g/ρ_g obtained at the position of $2y/H=0.4$, together with the relations for a single phase flow. The solid line (1) represents the result calculated from Eq. (20). The line (2) also shown in Fig. 15 is the result obtained from the $(1/7)$ power law velocity distribution, $u_g/u_{g0}^* = 8.57(y u_{g0}^*/\nu_g)^{1/7}$, i.e.,

$$\left. \begin{aligned} \frac{du_g}{dy} &= \frac{1.225}{y} \sqrt{\frac{H}{H-2y}} (y^+)^{1/7} u_g^* \\ l &= \frac{y}{1.225} \sqrt{\frac{H-2y}{H}} \left(\frac{1}{y^+} \right)^{1/7} \end{aligned} \right\} \dots\dots\dots (21)$$

The data plots for the single phase flow in both cases with and without a liquid film (data for $Q_{d1}=0$) fall in near the line (2).

On the other hand, the data plots for the gas flow containing liquid droplets deviate to the upper side of the line for a single phase flow, i.e., to such direction that the eddy diffusivity or the mixing length in the gas flow decreases, as the droplet concentration increases. This trend was the same with the data corresponding to the other distances from the duct wall.

It has been shown in Fig. 5 that the liquid droplet transfer coefficient K decreases with an increase of the droplet concentration. This trend of the value K for the liquid droplet concentration coincides with that of the eddy diffusivity ε as mentioned above. Then, the droplet transfer coefficient K was plotted against the values of ε for the gas flow suspending liquid droplets calculated from Eq. (17) and measured values of u_g^* and du_g/dy . The result is shown in Fig. 16, in which a good correlation between K and

Fig. 15 Relation between velocity gradient and shear stress

Fig. 16 Correlation between droplet transfer coefficient and eddy diffusivity in gas flow

ε is obtained. Therefore, it may be concluded from this result that with the decrease of the gas mean velocity $\bar{u}_g$ and with the increase of the droplet concentration the eddy diffusivity in a gas phase flow ε decreases and that this decrease of the value ε causes reduction of the droplet transfer coefficient K .

This result seems also to suggest that the transport phenomena of large particles like droplets in this experiment are affected more strongly by the state of the turbulence in the gas core region than by the state of the flow near the duct wall.

For comparison, is also shown in Fig. 16 the calculated result from the data by Friedlander et al.⁽⁴⁾, which were obtained for solid particles of 32 μ diameters in the solid particle gas flow system.

7. Conclusions

Experimental results are presented on the transfer rate of liquid droplet onto the duct wall and the distributions of droplet concentration, droplet velocity and gas velocity for the annular mist flow through a rectangular duct, and the following conclusions are obtained. In this experiment, the mean diameter of liquid droplets was in a range from 27 to 40 microns, and the measurements were also performed in the case where only the mean size of droplets was changed from 32 to 23 microns by adding a surface active agent to the liquid phase.

(1) The liquid droplet transfer coefficient K is affected not only by the gas velocity but also by the mean droplet concentration.

(2) The ratio of the mean velocity of the gas phase to that of the droplets is between 1.0 and 1.1,

and the velocity distribution of the liquid droplet is fairly flat over the cross section of the duct, while the distribution of the droplet volumetric fraction is not uniform.

(3) The eddy diffusivity ε of the gas flow containing liquid droplets decreases as the droplet concentration $\bar{C}$ increases. This fact is considered as the reason why the droplet transfer coefficient decreases with an increase of the droplet concentration.

On the basis of these results, it seems that the droplet transport phenomena are strongly affected by the state of the turbulence in the gas core flow.

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References

- (1) Alexander, L.G. and Coldren, C.L., *Indust. & Engng. Chem.*, Vol. 43 (1951), p. 1325.
- (2) Cousins, L.B. and Hewitt, G.F., *Symp. on Two-Phase Flow*, Vol. 2 (1965), C 401.
- (3) Paleev, I.I. and Filippovich, B.S., *Int. J. Heat & Mass Transf.*, Vol. 9 (1966), p. 1089.
- (4) Friedlander, S.K. and Johnstone, H.F., *Indust. & Engng. Chem.*, Vol. 49 (1957), p. 1151.
- (5) Beal, S.K., *Nucl. Sci. & Engng.*, Vol. 40, (1970), p. 1.
- (6) Silvestri, M., *Advanc. in Heat Transf.*, Vol. 1, (1964), p. 355.
- (7) Tanasawa, Y. and Takeno, S., *Trans. Japan Soc. Mech. Engrs.* (in Japanese), Vol. 35, No. 276 (1969), p. 1741.

Discussion

K. AKAGAWA (Kobe University):

(1) While the shear stress in the single phase flow varies linearly with the distance from the duct wall, the τ_g -distribution in this experiment calculated from Eq. (8)' substituting the measured values of $(W_d)_h$ and $\bar{u}_d$ seems to be a different type depending on the h -directional distribution of $(W_d)_h$ (or f_d in Fig. II). What is the distribution type of the shear stress τ_g in the gas flow suspending droplets?

(2) The authors have mentioned that the eddy diffusivity ε is calculated from Eq. (17), i.e.,

$$\varepsilon = l^2 \left(\frac{du_g}{dy} \right) = u_g^{*2} \left/ \left(\frac{du_g}{dy} \right) = \left(\frac{\tau_g}{\rho_g} \right) \right/ \left(\frac{du_g}{dy} \right)$$

and "from the measured values of u_g^* and du_g/dy " (as described on 2nd line from the bottom in the right-hand column of page 1576). The calculated value of ε seems to vary with the distance y . If so, what is

the type of the distribution of ε ? And to what position does the value ε in Fig. 16 correspond?

(3) If the value of ε varies with the distance y , and if that value in Fig. 16 is only the local one, the results of Fig. 16 might have no generality.

Furthermore, what is the physical meaning of the dependence of the relation between the values of K and ε on the gas velocity as shown in Fig. 16?

Y. MORI and K. HIJIKATA (Tokyo Institute of Technology):

(4) The authors have mentioned that a large part of the liquid film in the droplet-generating section was sucked off through the porous sinter-plates provided on the upper and lower walls of the duct in order to allow a small amount of liquid film flow in the test section. Are there any problems in leaving the liquid film flow on the side walls of the duct?

(5) The writers ask the authors to show the results on the relation between the critical flow rate of the liquid film and the air flow rate.

(6) Transport phenomena of liquid droplets to the liquid film on the duct wall are considered to consist of a diffusion phenomenon of liquid droplets in the gas flow and a capture phenomenon of droplets at the liquid film surface. And it seems that generally the droplet transfer rate by the former phenomenon is related to the difference of droplet concentration and the latter to the droplet concentration itself. In the definition like Eq. (1), therefore, the mean gradient of droplet concentration in the duct or the droplet concentration at the liquid film surface should be used instead of the mean concentration $\bar{C}$. Furthermore, since there is no physical consideration on the transport process of droplets, the meaning of K in Eq. (1) seems to be only an experimental constant, and the name of the droplet transfer coefficient might not always be appropriate.

(7) It seems in introducing Eq. (3) that an error caused by the droplet transfer to the side walls of the duct should be discussed.

(8) Though the authors have concluded about the dependence of the value K on the mean gas velocity $\bar{u}_g$ for high values of $\bar{C}$ that "the phenomenon is caused by the effect of the droplet concentration on the turbulence of the gas flow", do they consider that the experimental data are sufficient to introduce this conclusion?

(9) It seems appropriate to use a nondimensional Weber number as an ordinate of Fig. 6 instead of the mean droplet diameter.

(10) The authors have mentioned that the droplets were caught on a glass plate for measuring the diameters. What is the type of the distribution curve of the droplet diameter?

(11) Since the distribution of droplet velocity is as shown in Fig. 10, the discussion on the mean value over the cross section of the duct, $\bar{S}$ (Fig. 12), seems not to be so meaningful. Furthermore, what is the reason for the value $\bar{S}$ approaching unity as $\bar{u}_g$ decreases as shown in Fig. 10?

(12) When the flow rates of liquid droplet and the gas phase per the y -directional unit length are represented as w_d , w_g respectively, $d \int_{-h}^h (w_g u_g) dh$ should be substituted for the term $d(W_g \bar{u}_g)_h$ in Eq. (8). There are no reasons to equalize $d \int_{-h}^h (w_g u_g) dh$ to $d(W_g \bar{u}_g)_h$ which means $d \left(\int_{-h}^h w_g dh \times \frac{1}{2h} \int_{-h}^h u_g dh \right)$. If the term $\int_{-h}^h (w_g u_g) dh$ does not vary along the x direction, that term, the value of which is equal to

zero, should be omitted from Eq. (8). Furthermore, please show the order of magnitude for that term in the right hand side of Eq. (8) in order to confirm the adequacy of the assumption of Eq. (13).

(13) The results of Fig. 15 seem to be fairly affected by the method for obtaining the value of τ_g . What method did the authors use?

(14) It seems unreasonable that the value of K , which is used to remove the influence of $\bar{C}$ from the droplet transfer rate, depends on the value $\bar{C}$ itself as described in the conclusion.

Author's closure

(1) The shear stress distribution in the gas flow containing droplets is illustrated as a solid line in Fig. 14. The value of the term $(1/2 gB) \{ (W_d)_h d\bar{u}_d / dx \}$ increases rapidly with the distance from the duct axis near the center of the duct, and the rate of increasing becomes low as the distance increases. But, since the absolute value of the term is small in this experiment and about 10% of the value of the first term in the right hand side of Eq. (8)', the solid line does not deviate so far from the dotted line which reveals the linear distribution of the first term.

(2) Some of the experimental results of the eddy diffusivity in the gas flow suspending droplets are presented in the Append.-Fig. 1. The ordinate of the figure is expressed as nondimensional value. The shape of the y -directional distribution is almost the same as that of no droplet flow, and the value of the nondimensional eddy diffusivity may be considered, as the first approximation, to be nearly constant along the distance y in a range of $2y/H > 0.2$. Therefore, as a representative value, one corresponding to the position $2y/H = 0.4$ is used in Fig. 16. Data No. 20 in the figure correspond to the case of no droplet (i.e., single phase flow), and solid lines represent the value for a single phase flow by Hinze and one calculated from Eq. (21) based on the $1/7$ power distribution. It is shown, further, that the absolute value of ϵ decreases with an increase of

Append.-Fig. 1 Eddy diffusivity in gas flow (Containing droplets)

the droplet concentration.

(3) Since, as described above, the value of ε in the core region is considered to be approximately constant, consideration on the relation between the values of ε and K seems to have a generality to some extent. Though the value of ε decreases rapidly in the region near the duct wall as shown in the Append.-Fig. 1, the authors consider that the droplet transportation phenomena for the relatively large droplets in this experiment are mainly controlled by the flow in the gas core region.

As for the reason for the dependence of the relation between ε and K on the average gas velocity, they express their opinion in answer (8) to the remarks of Professor Mori and Dr. Hijikata.

(4) Though the explanation was not presented on the point indicated in question (4), the liquid film flowing on the side walls of the duct was also sucked off through the slits provided on the side walls at the sinter-section of the duct.

(5) The critical liquid film flow rate, above which the droplet-generation occurs, is shown in an Append.-Fig. 2 against each gas velocity. The symbol Δ represents the data of the critical value obtained by the duct in this experiment, which correspond to the condition that the total amount of the generated droplets is less than 2~3 cc/min. For reference, the results obtained by Hanratty et al.^(*) are also presented as a dotted line in the figure.

Since the authors consider the mechanism of the droplet transportation is mainly controlled by the diffusion process in the turbulent gas core stream, the local gradient of the droplet concentration seems to have necessarily a dominant meaning. But for practical use it seems appropriate to define K for the difference between the mean droplet concentration and the concentration at the liquid film surface C_0 . Since a small amount of the liquid film was left on the inner wall of the test section in this experiment, the authors supposed that such

relatively large droplets as in the mist flow were absorbed into this liquid film without reflection after being transferred from the gas core to the film surface. Taking into account this condition that the liquid film surface may be a perfect sink for droplets, the authors regarded the value of C_0 as zero. The model of the droplet transport process proposed by the authors will be discussed in the second part of this report.

(7) In this report the authors assumed, as the first approximation, the flow in the duct as one between two parallel flat plates. On the other hand, since the droplet transport phenomena are different from the molecular mass transfer and are more affected by the state of the core flow than by that near the duct wall, the droplets may be also considered to be equally transferred both to the side walls and to the upper and lower walls of the duct. Therefore, it may be rather reasonable to substitute $(B+H)$ for B in Eq. (3). The actual case will be between these two cases. From these considerations, though it is difficult to discuss strictly, the error caused by using Eq. (3) is estimated to be about 17% at the most.

(8) The authors consider that the influence of the droplet concentration on the turbulence of the gas flow generally varies with the gas velocity $\bar{u}_g$. But they also believe that the results in this experiment are insufficient to discuss the degree of the effect as indicated in the discussion and that further investigation on this problem is necessary.

(9) In the case of a study on the mechanism of droplet formation it seems certainly significant to arrange the data for the nondimensional Weber number. But, since there are problems in selecting the representative velocity in this number, only the trend of the droplet diameter has been shown in Fig. 6 instead of the Weber number.

(10) The droplet distribution, one example of which will be presented in the second part of this report, is approximately of logarithmic type.

(11) Though the ratio of the mean velocities is meaningless in studying the behavior of each droplet, it is considered to be reasonable to use a mean value in defining a macroscopic quantity such as the droplet transfer coefficient in Eq. (4).

Further, since in the case of the low value of $\bar{u}_g$ the period until droplets arrive in each measuring position is long and even the droplets, which have not been accelerated sufficiently at the entrance part of the test section, are well accelerated during this period, the value of the ratio $\bar{S}$ seems to approach unity rapidly.

(12) As for the expression for the momentum terms in Eq. (8), it seems more appropriate to have

Append.-Fig. 2 Critical liquid film flow rate

(*) Hanratty, T.J., et al., *Chem. Engng. Sci.*, Vol. 24 (1969), p. 299.

used those like $d \int_{-h}^{+h} w_g u_g dh$. But the value of the momentum term of the gas phase [the second term in the right hand side of Eq. (8)], even if it is calculated by using the term $d \int_{-h}^{+h} w_g u_g dh$ described above, can be neglected and is only 2~3%, at the most, of the first term in the right hand side of the equation, because the measured value of the gas velocity distribution scarcely varies along the direction of the duct axis.

The value of the droplet momentum terms [the third and the fourth terms in the right hand side of Eq. (8)], which is nearly the same for both expressions because of flatness of the droplet velocity distribution as shown in Fig. 10, is about 10% of the first term. Because the value of the liquid film velocity

itself is small, the value of the liquid film term [the fourth term in the right hand side of Eq. (9)] is about 1% of the first term in Eq. (9) and can be neglected.

(13) The authors calculated the value of τ_g from Eq. (8)' by substituting measured values of the pressure drop, the difference of droplet velocity between two positions and the droplet flow rate $(W_d)_h$ through the passage of $2h$ height at the middle of the positions.

(14) The value K in Eq. (1) is defined against the difference of a droplet concentration $\bar{C}$ which is a direct motive of the droplet transportation. The authors consider it is not unreasonable that the value K is affected further by $\bar{C}$ in the case of high values of $\bar{C}$ just as in cases of the heat transfer in condensation and in boiling.