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Effect of bubble motion on local heat transfer around a tube across horizontal in-line and staggered tube bundles in bubbly and intermittent flows

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Abstract
Cross-flow boiling in horizontal tube bundles occurs in kettle type evaporators. Convective heat transfer due to the motion of vapor bubbles is an important factor for boiling heat transfer in the evaporator under low heat flux conditions. To clarify the liquid agitation effect on heat transfer, the local heat transfer around a tube in in-line and staggered tube bundles was investigated in two-phase flows under adiabatic and atmospheric pressure conditions. Air and tap water were used as the working fluids. The test section was a vertical duct with inner dimensions of 90 × 90 mm^2. In-line and staggered tube bundles each containing eight rows and five columns, were used as test sections. For both bundles, the tube diameter, d, was 18 mm, and the tube pitch, p, was 22.5 mm (p/d = 1.25). The local heat transfer had the highest values around $\theta = \pm 90°$ where the liquid velocities were high in single-phase, bubbly and intermittent flows for both in-line and staggered arrays. A significant improvement in the heat transfer caused by the bubble motion was present for the in-line array as compared to the staggered array. Owing to the fluctuation of the liquid velocity, the heat transfer coefficient fluctuated significantly under intermittent flow conditions. In the bubbly flow at p/d = 1.25, the average heat transfer coefficient around a tube for the in-line array was higher than that for the staggered array. In contrast, the heat transfer coefficient of the staggered array was high under intermittent flow at a lower gas flow rate. This tendency is different from the results at p/d = 1.5. With a decrease in the relative size between the bubble diameter and the tube gap, there was a high improvement in the heat transfer coefficient due to the liquid agitation in the bubbly flow.

Keywords : Liquid agitation, Tube bundle, In-line, Staggered, Local heat transfer, Pitch-to-diameter ratio

1. Introduction

Kettle type evaporators have been widely used in geothermal binary power plant. Improving the boiling heat transfer in horizontal tube bundles is a critical issue to improve the power generation efficiency and downsize the evaporators. It is known that the effects of liquid agitation caused by the generated vapors are important factors in boiling heat transfer. Therefore, there are many experimental studies regarding the clarification of two-phase flow phenomena in horizontal tube bundles.

Gupta et al. (1995) investigated the boiling heat transfer coefficients in a horizontal tube by changing the horizontal tube rows from 1 to 3 and the pitch-to-diameter ratio, p/d. The heat transfer coefficients were enhanced by the vapor bubbles rising from the lower tubes, and the effect was more significant at lower heat flux conditions. Gupta (2005) conducted measurements of the local and average boiling heat transfer coefficients in 5 rows and 3 columns in-line tube bundles. Rising vapor bubbles enhanced the liquid turbulence around the upper tube, resulting in an increase in the heat

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transfer coefficient. When the heat flux was increased from the lower heat flux, the heat transfer coefficient of the tube reached its maximum value. After the maximum value, it decreased with an increase in the heat flux because of the excessive vapor surrounding the tube. Jensen et al. (1989) compared the boiling heat transfer coefficients between in-line and staggered tube bundles. It was shown that the boiling heat transfer coefficients for staggered bundles were higher than those for in-line bundles at lower heat fluxes. With an increase the heat flux, the difference in the heat transfer coefficients in each bundle decreased. Similar results have been reported by Zhang et al. (2018). From these investigations, it can be seen that the heat transfer coefficient on the downstream tube in the horizontal tube bundle is greatly influenced by the rising bubbles in the nucleate boiling region under low heat flux conditions. This effect changes with the tube arrangement, such as in-line, staggered, and the difference in the p/d.

To clarify the two-phase flow behaviors in horizontal tube bundles, many experimental studies have been conducted under adiabatic two-phase flow conditions regarding flow pattern, void fraction and pressure drop (Kondo and Nakajima, 1980; Ulbrich and Mewes, 1994; Noghrehkar et al., 1999; Mao and Hibiki, 2017). To evaluate the heat transfer coefficients in the tube bundle, it is important to obtain detailed flow structures around the tube. Iwaki et al. (2004, 2005) investigated the velocity fields in single-phase and bubbly flows. They showed that the liquid turbulent intensity in the staggered bundle was slightly larger than that in the in-line bundle in bubbly flow. Therefore, it is expected that the improvement in the heat transfer coefficient due to the liquid agitation is higher in the staggered array than in the in-line array in bubbly flow. Karas et al. (2014) performed experimental studies to measure the heat-transfer coefficient and velocity distributions around a tube in in-line and staggered bundles at p/d = 1.33. They showed that the heat transfer coefficients had the highest values at the frontal part of the tubes or at a place where the liquid velocity was the highest in most cases. Murakawa et al. (2018) measured the void fraction and heat transfer coefficient distributions around tubes in in-line and staggered tube bundles with p/d = 1.5. It was shown that the distribution of the heat transfer coefficient around a tube changes with flow regimes, such as bubbly and intermittent flows. Furthermore, a significant improvement in the local heat transfer by liquid agitation appeared around the bottom and upper positions in the in-line bundle. In the bubbly flow, the average heat transfer coefficients were high for staggered tube bundles; however, the results were the opposite in the intermittent flow owing to the increments in the void fraction in the staggered array.

The purpose of this study is to evaluate the change in the heat transfer coefficient due to convection caused by the bubble motion in bubbly and intermittent flows. Local heat transfer coefficients around a tube in in-line and staggered tube bundles with p/d = 1.25 were investigated in two-phase flows under adiabatic conditions. The local heat transfers were measured at the tubes in intervals of 45°, and changes in the heat transfer coefficient with respect to the bubble motions were evaluated. Furthermore, the results were compared with our previous investigation conducted at p/d = 1.5.

2. Experimental setup and heat-transfer measurement around a tube

The flow loop used in this study was the same as that used in our previous study (Murakawa et al., 2018). The flow loop consisted of a test channel, overflow tank, pump and gas injector. The test channel was a vertical duct with a cross-section of 90 × 90 mm^2. The total length of the channel was 1,700 mm. The working fluids used were tap water and air. Water was circulated through the test channel using a pump. Air was injected through a gas injector made of porous media that was located at the bottom of the test channel. Figure 1 shows a schematic of the tube bundle test section. Two different tube bundles, in-line and staggered, were used for the measurements. The both bundles consisted of eight rows as shown in the Fig. 1. The test section and the tubes were made of acrylic resin. For both in-line and staggered bundles, the tube length was 90 mm, and the outer diameter, d, was 18 mm. The tube pitch, p, was 22.5 mm and the pitch-to-diameter ratio, p/d, was 1.25. The bubble injectors were placed 660 mm below from bottom tubes, and the distance between the flow inlet and the bottom tubes of the bundles was approximately 800 mm. The local heat transfer around a tube was measured at the test section, which is shown as the measurement position in the figure. It has been reported that the void fraction and the velocity distribution do not change significantly along the flow direction except around inlet and outlet regions (Dowlati et al. 1990; Iwaki et al., 2004). Thus, in this study, local heat transfer was evaluated at the fourth row for the in-line tube bundle and the third row for the staggered tube bundle.

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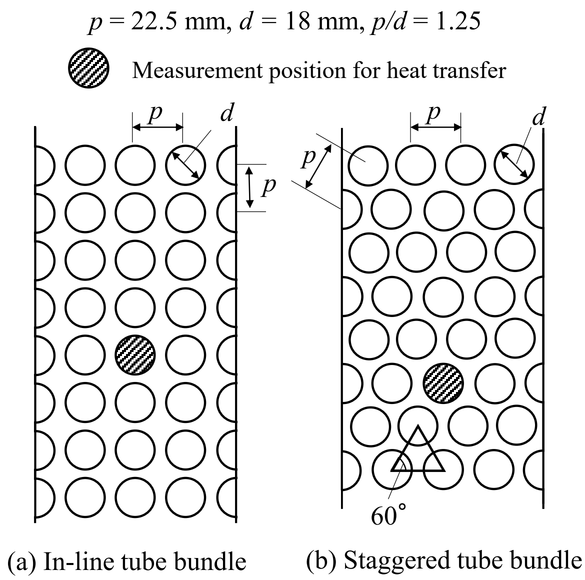

Fig. 1 Schematic of the test section bundles of (a) in-line and (b) staggered with pitch-to-diameter ratio (p/d) of 1.25, containing 8 rows and 4 columns. Local heat transfer was measured at the center tube as shown in the figure.

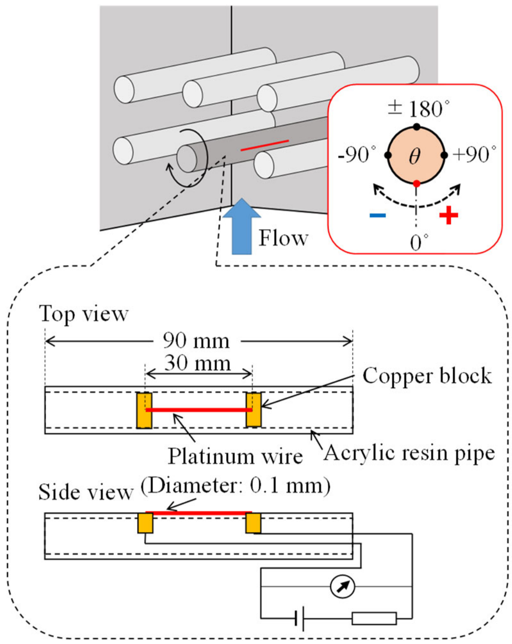

Fig. 2 Schematic of the measurement system for local heat transfer coefficient around a tube. A platinum wire was placed on the tube and was connected with a DC power supply for heating. The local heat transfer coefficients were measured in each angle, θ.

Figure 2 depicts the measurement system of the local heat transfer. A platinum wire with a diameter of 0.1 mm and a length of 30 mm was placed on the tube surface in the center region of the tube. A DC power supply was connected to the platinum wire for heating. The temperature of the wire was determined based on the electrical resistance. Based on the temperature difference between the water and platinum and the heat input, the local heat transfer in each rotated position was evaluated. The signals were sampled at a sampling rate of 100 Hz, and the time-averaged heat transfer coefficient was calculated from the data for 2 min. The measurement uncertainty of heat transfer can be evaluated as about 6–15%, calculated based on the uncertainty of a standard resistance (2972A04, Yokogawa Test & Measurement Corp.) with an accuracy of ±30ppm and the uncertainty of a data logger (8423, Hioki Corp.) with an accuracy of ±0.1% at full scale. However, uncertainty was improved by inducing a small current to the platinum wire to measure the fluid temperature and calibrate the heat transfer coefficient in each measurement.

The experiments were performed at 22 ± 0.5 °C. The superficial gas velocity, J_G, varied up to 1.0 m/s. The superficial liquid velocities, J_L, were set to 0.1, 0.2, and 0.3 m/s for the flow observations and to 0.2 m/s for measuring the local heat transfer coefficients. Superficial velocities were defined as the minimum cross-sectional area.

3. Results and discussions

3.1 Flow regime maps

Flow observations were carried out at the test section using a high-speed camera (FASTCAM SA 1.1, Photron Limited, Japan) with a frame rate of 1,000 fps. For a lower J_G, bubbles with a diameter smaller than the tube gap flowed in the tube gaps, and the flow was defined as bubbly flow. With an increase in the gas flow rate, larger bubbles were observed intermittently. These large bubbles caused a flow oscillation, resulting in a local reversal flow. When the reversal flow was observed, the flow was defined as an intermittent flow. Based on these criteria, flow pattern maps were obtained, as shown in Fig. 3. The dashed lines in the maps represent the criteria between the bubbly, transition of bubbly to intermittent, and intermittent flows obtained at p/d = 1.5 in our previous investigation (Murakawa et al., 2018). The flow regime transitions from the transition region to intermittent flows were confirmed at almost the same gas flow rate for both bundles. J_L had little effect on the flow pattern transition except for the transition from the bubbly

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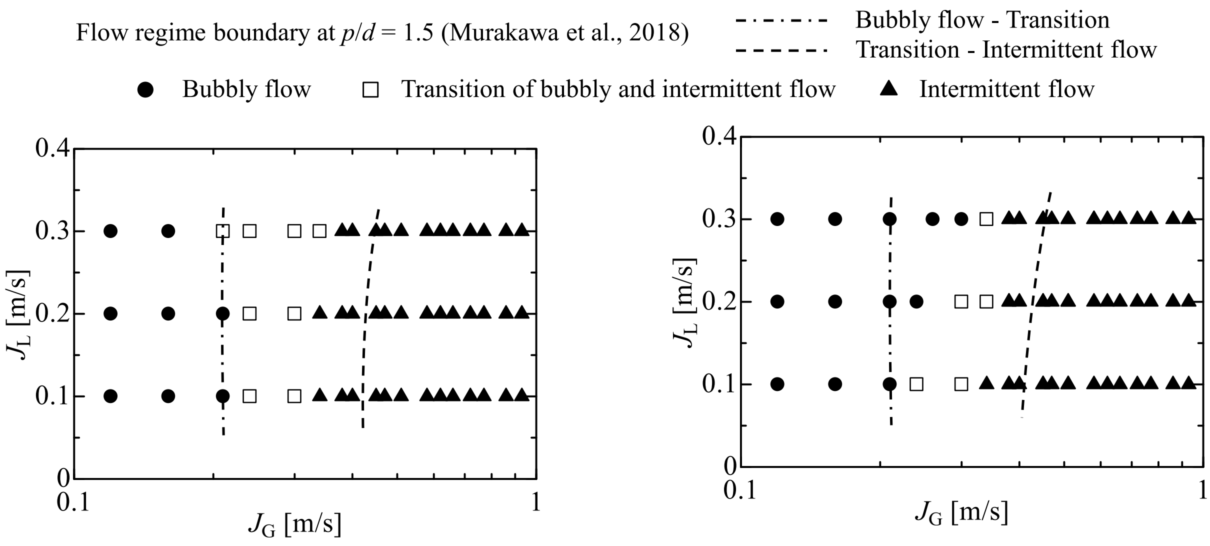

(a) In-line tube bundle (b) Staggered tube bundle

Fig. 3 Flow pattern map in the horizontal tube bundles at p/d = 1.25 obtained from the visual observations. The dashed lines are the criteria between the bubbly, transition, and the intermittent flows obtained at p/d = 1.5 (Murakawa et al., 2018).

to transition region at J_L = 0.3 m/s in staggered bundle. For J_G > 0.34 m/s, intermittent flows were observed for both bundles. To compare with the results at p/d = 1.5, the transition to intermittent flows started at a slightly lower J_G when p/d = 1.25. As the tube gap decreased, the ratio of the bubble diameter to the tube gap increased. As a result, it is considered that the generation of large bubbles due to the collision of small bubbles occurred at a lower J_G, resulting in the transition from bubbly to intermittent flows, which occurred at a slightly lower J_G in this study.

3.2 Distribution of local heat transfer coefficients around a tube

Figures 4 and 5 represent the time-averaged heat transfer coefficients around the tube at p/d = 1.25 and 1.5 at J_L = 0.2 m/s, respectively. The vertical axis represents the local heat transfer coefficients, h, normalized by the average heat transfer coefficient over the tube in a single-phase flow for the in-line tube bundle at p/d = 1.5, $\overline{h_L}$. The horizontal axis represents the rotation angle θ. It can be confirmed that the distributions of heat transfer are not axisymmetric, especially for single-phase and bubbly flows. This is due to the non-axisymmetric of flow, and similar results have been confirmed in the literature (Iwaki et al., 2004, 2005). For the in-line bundle in a single-phase flow, h takes the smallest values around θ = ±135° and the largest values around θ = ±90°. Approximately, the same results were obtained for p/d = 1.25 and 1.5. Conversely, distributions of h in the single-phase flow are slightly different between p/d = 1.25 and 1.5 for the staggered bundle. The smallest values are confirmed at approximately ±180° for both p/d, whereas the angle at which h takes the maximum values is different. By changing the p/d from 1.5 to 1.25, θ which takes the maximum values of h, varied from ±45° to ±90°. Therefore, it can be confirmed that p/d affects the flow distribution around the tube, resulting in a change in the local heat transfer.

In the bubbly flow, there is a remarkable improvement of the heat transfer in in-line tubes due to the liquid agitation caused by the bubble motion, particularly at approximately θ = 0° and ±45°. As shown in Fig. 6, a small number of bubbles flowed between the tube rows, resulting in an improvement in the heat transfer. Although the bubbles flowed near the bottom of the tubes, a significant improvement in h was also confirmed at θ = ±135° and ±180°. However, this effect is greater for p/d = 1.25 than for p/d = 1.5. This implies a large liquid agitation effect for a smaller vertical gap if approximately the same diameter bubbles flow in the horizontal gap. As a result, the improvement of the heat transfer coefficient by the bubble agitation effect was greater for lower p/d at the same superficial velocities. In contrast, the improvement in the heat transfers due to the motion of the bubbles was less for the staggered array than for the in-line array. Furthermore, h increased at θ = 0° but not at θ = ±45° at p/d = 1.25, whereas it slightly increased for every angle of θ at p/d = 1.5. It is thought that the effect of collisions between bubbles around the top of the tube suppressed the liquid turbulence, leading to a decrease in h around θ = ±45°. The bubble agitation effect was less for the staggered array than for the in-line array. This is because the bubbles move smoothly in

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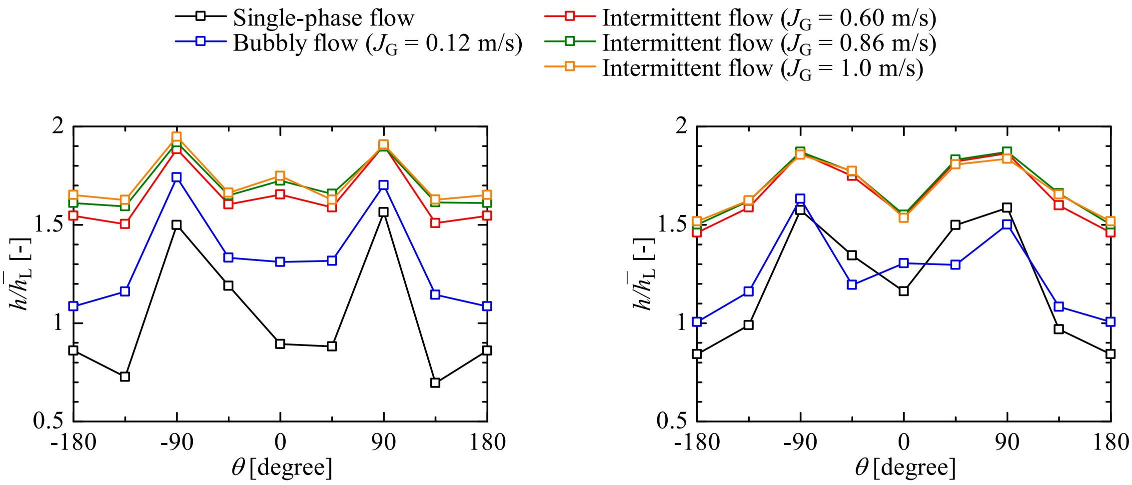
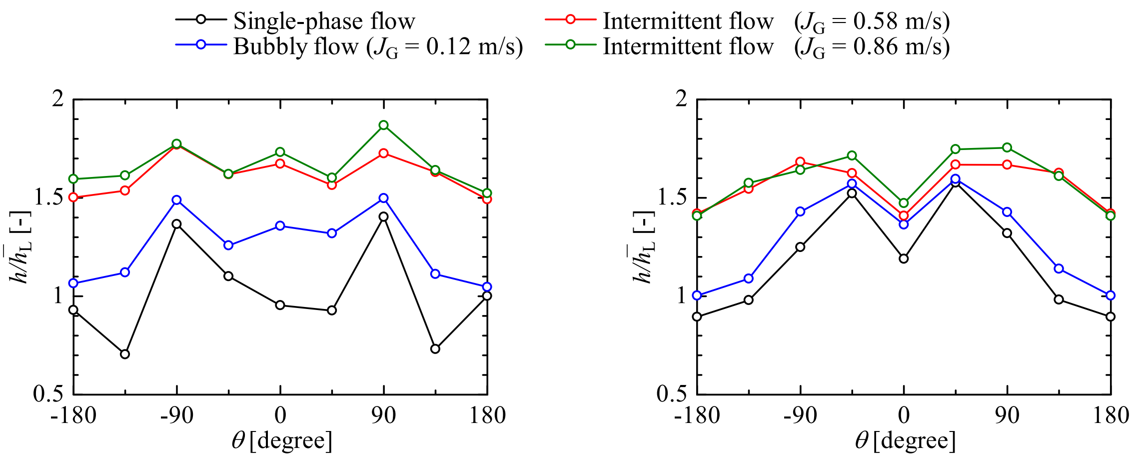
the tube gap for the staggered array. Therefore, the liquid agitation caused by bubble motion is lower for staggered than for in-line bundles in bubbly flow.



(a) In-line tube bundle (b) Staggered tube bundle

Fig. 4 Time-averaged heat transfer coefficients around a tube under different flow conditions at J_L = 0.2 m/s (p/d = 1.25).



(a) In-line tube bundle (b) Staggered tube bundle

Fig. 5 Time-averaged heat transfer coefficients around a tube under different flow conditions at J_L = 0.2 m/s (p/d = 1.5, Murakawa et al., 2018).

(a) In-line tube bundle (b) Staggered tube bundle

Fig. 6 Flow visualization around a tube in bubbly flows at J_G = 0.12 m/s for in-line and staggered tube bundles at p/d = 1.25.

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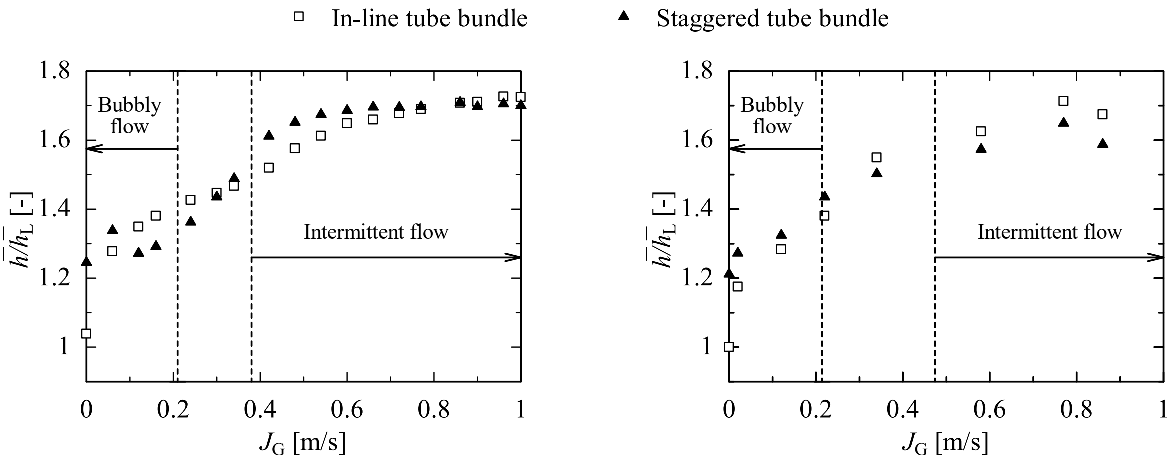
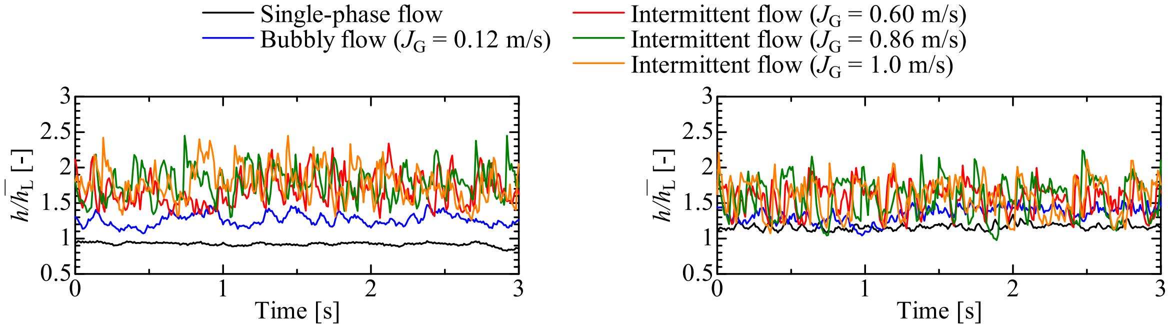
Under intermittent flow conditions, further improvement of the heat transfer due to liquid agitation caused by bubble motion was confirmed. In particular, the improvement of h at $\theta = 0°$ is better for the in-line tube bundle than for the staggered bundle. Large bubbles intermittently pass through the tube bundle, resulting in a flow oscillation. It is considered that the oscillation increased the liquid agitation effect throughout the tube and improved the heat transfer coefficient. As a result, the distribution of h around the tube was flattened in comparison to the single and bubbly flow conditions. However, the heat transfer coefficients are further improved by increasing the gas flow rate of the in-line bundle under intermittent flow conditions, whereas the difference of h between the conditions of J_G = 0.60 m/s and 0.86 m/s was not significant for the staggered array.

Figure 7 compares the area-averaged heat transfer coefficient around a tube, $\bar{h}$, between in-line and staggered bundles. Bubbly and intermittent flow regions are approximately shown in the figures. In the single-phase flow condition, the staggered tube bundle had a higher $\bar{h}$ than that of the in-line bundle. For almost every condition, $\bar{h}$ increased with an increase in the gas flow rate. For the in-line bundle, the difference in $\bar{h}$ between single-phase and bubbly flows was large, and it increased almost linearly with the gas flow rate from bubbly to intermittent flow conditions. For the staggered bundle, $\bar{h}$ increased with an increasing gas flow rate in the bubbly flow. Note that at p/d = 1.25, $\bar{h}$ at J_G = 0.06 m/s was higher than that in single-phase and at J_G = 0.12 m/s in the bubbly flow. This is because h was high at $\theta = \pm 90°$. However, further investigation is needed to clarify this. The magnitude relationship of $\bar{h}$ for the in-line and staggered bundles differs depending on p/d. At p/d = 1.25, $\bar{h}$ for the in-line bundle was larger in the bubbly flow, whereas $\bar{h}$ for the staggered array became larger than that of the in-line bundle under the intermittent flow condition. It is confirmed that $\bar{h}$ tends to converge to a certain value, even if J_G increases in the intermittent flow. Therefore, $\bar{h}$ for in-line bundle became larger than that for the staggered bundle at J_G > 0.90 m/s. This tendency was not observed at p/d = 1.5.



(a) p/d = 1.25 (b) p/d = 1.5

Fig. 7 Comparison of area-averaged heat transfer coefficient around a tube between in-line and staggered (J_L = 0.2 m/s). The flow regimes are approximately shown for covering the regions for both in-line and the staggered arrays.



(a) In-line tube bundle (b) Staggered tube bundle

Fig. 8 Examples of time-series of local heat transfer at $\theta = 0°$ for each bundle at p/d = 1.25.

3.3 Instantaneous local heat transfer coefficient

Figure 8 shows examples of the time-series of local heat transfer coefficients during a period of 3 s at θ = 0° to compare the single-phase, bubbly, and intermittent flows at p/d = 1.25. It can be seen that h is approximately constant in single-phase and bubbly flows for both in-line and staggered bundles. The fluctuation becomes more intense as J_G increases. As large bubbles pass through the tube bundle, the liquid velocity is repeatedly accelerated and decelerated, giving rise to considerable fluctuations in h. Therefore, the transition to the intermittent flow causes a large fluctuation in h. In order to understand the change in the fluctuation of h, the probability density function (PDF) of h in each angle and condition is calculated from the time-series data over a period of 2 min as shown in Fig. 9. The vertical dashed lines represent the average values in each condition.

In the bubbly flow for the in-line array, h increases significantly at each angle owing to the motion of the bubbles in comparison to that in the single-phase flow. In this condition, it can be seen that the range of h in the PDF is wider within θ = 0°–90° than that at θ = 135° and 180°. In the region of θ = 0°–90°, the fluctuation of h became large as the bubbles passed near the tube surface for the in-line array, as confirmed by the visual observation shown in Fig. 6. Meanwhile, in the range of θ = 135°–180°, bubbles flow in a region slightly away from the tube surface. Therefore, it is considered that the velocity fluctuation of the liquid phase caused by bubble motion is relatively small, and the distribution of the PDF became narrow. For a staggered array, the range of h is approximately the same as that for the in-line array at θ = 0° in bubbly flow. However, because h in the single-phase flow is higher than that in the in-line array, the increase in h due to the liquid agitation is relatively smaller than that for the in-line array. At θ = 45° and 90°, it can be confirmed that the PDF distribution shifted to lower values than in the single-phase flow. The bubbles flow near the surface of the tube, resulting in an increase in the fluctuation of the liquid velocity and widening of the distribution of the PDF. However, it is clear that these fluctuations reduce the local heat transfer, that is, the negative effect, in comparison to that in the liquid single-phase flow. This means that the bubble motion suppresses liquid agitation, resulting in a decrease in h. At θ = 180°, h is larger than that in the single-phase flow. However, the average value in the staggered array was slightly lower than that in the in-line array. In this region, it is shown that the magnitude of the liquid velocity is small in both single-phase and bubbly flows (Iwaki et al., 2005). Therefore, it is considered that the liquid agitation effect caused by the bubble motion did not have a large effect, and the increase in h in the bubbly flow with respect to the single-phase flow did not become very large.

In intermittent flow, the tendency of the PDF distribution differs for each angle and array. For the in-line array at θ = 0°, the distribution shifts to a larger h from that in the bubbly flow, although the distribution becomes slightly wider. The same results were obtained for other values of θ, except θ = 90°. At θ = 90° for the in-line array, the PDF was distributed over a wide range, which starts from a lower value than that in a single-phase flow. As large bubbles pass through the bundle, the velocities of the gas and liquid phases oscillate. Although there is a moment when h becomes high owing to the acceleration of the liquid phase, there is additionally a moment when the liquid phase velocity becomes zero and a reverse flow occurs locally. At such moments, the heat transfer coefficient decreased. As a result, it is considered that the distribution of the PDF becomes wider. Even if the gas flow rate was increased in the intermittent flow, the average heat transfer coefficient did not improve at θ = 90° in the in-line array. For the staggered array, the moment when the heat transfer coefficient decreases in the intermittent flow is confirmed at θ = 0°–90°. These angles are in the region where the void fraction is high. For such angles, no significant improvement in the time-averaged h was confirmed, even if the gas flow rate is increased in the intermittent flow.

The increase in $\bar{h}$ was confirmed by the transition from the bubbly to intermittent flow. However, near the region where the local void fraction is high, h did not increase significantly with the gas flow rate in the intermittent flows. The maximum value of h in the time-series data tended to increase as the gas flow rate increased in the intermittent flow. However, at angles close to the region where the void fraction is high, the increase in the time-averaged h with J_G was suppressed because of the longer time of the lower heat transfer due to the flow oscillation. It has been reported that the area-averaged void fraction in a staggered bundle is higher than that in an in-line bundle (Murakawa et al., 2018). Owing to the higher void fraction for the staggered bundle, the effect on liquid-phase agitation is limited even if the gas flow rate increases. Therefore, it is considered that $\bar{h}$ did not change significantly even if the gas flow rate increased in the intermittent flow for the staggered array. The difference in h between p/d = 1.25 and 1.5 appears at approximately θ = 90° in the bubbly flow for the in-line array. Because the tube gap is smaller at p/d = 1.25 than that at p/d = 1.5, the bubble agitation effect was greater at p/d = 1.25, which led to a higher $\bar{h}$ in the bubbly flow. The transition to the intermittent flow at p/d = 1.25 occurred at a slightly lower gas flow rate than that at p/d = 1.5.

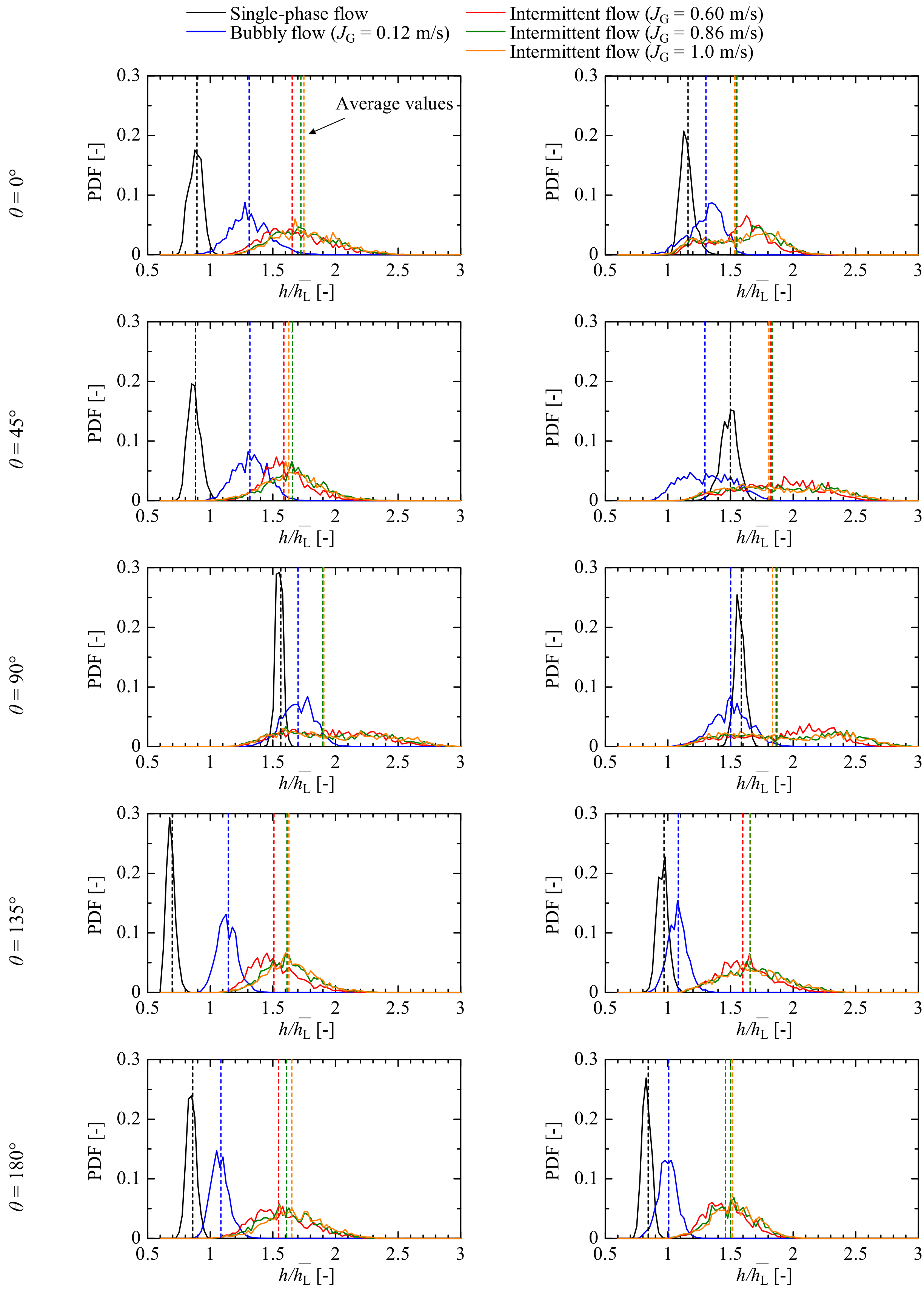


Fig. 9 Probability density function, PDF, of the heat transfer in each angle calculated from the time-series signal for 2 min at p/d = 1.25. The dashed lines represent the average values of h.

Therefore, $\bar{h}$ for the staggered array in the intermittent flow was larger than that for the in-line array at p/d = 1.25 with J_G < 0.8 m/s, but the opposite results were confirmed at p/d = 1.5. It is considered that the in-line array is better in the intermittent flow for $\bar{h}$ at higher gas flow rates, *i.e.*, J_G > 0.8 m/s, owing to the liquid agitation around the bottom and upper parts of the tube.

4. Conclusions

The local heat transfer around a tube in in-line and staggered tube bundles with p/d = 1.25 were investigated under bubbly and intermittent flow conditions. The local heat transfer takes the highest values around $\theta = \pm 90°$, where the liquid velocity is the highest in single-phase, bubbly, and intermittent flows for both in-line and staggered arrays. A significant improvement of the heat transfer caused by the liquid motion appeared for the in-line array rather than for the staggered array. Owing to the fluctuation of the liquid velocity, the heat transfer coefficient fluctuated significantly around the positions where many bubbles flowed near the tube under intermittent flow conditions. As the liquid velocity decreased to zero, the instantaneous local heat transfer coefficient significantly decreased. Under such conditions, the local time-averaged heat transfer coefficient hardly changed, even if the gas flow rate increased in the intermittent flow. As a result, the heat transfer coefficient converged to an approximately constant value with an increase in the gas flow rate in the intermittent flow for the staggered array. However, further improvement of the heat transfer was confirmed for the in-line array under such conditions. In the bubbly flow at p/d = 1.25, the average heat transfer coefficient around a tube for the in-line array was found to be higher than that for the staggered array. In contrast, the heat transfer coefficient of the staggered array was higher in intermittent flow with a lower gas flow rate. This tendency is different from the results at p/d = 1.5. A decrease in the relative size between the bubble diameter and the tube gap led to a significant improvement in the heat transfer coefficient due to the liquid agitation in the bubbly.

Based on this investigation, it is better to change the tube arrangement of the heat exchanger depending on the flow conditions for optimization. At p/d = 1.25, it is better to install an in-line bundle for upstream where the flow is bubbly flow, and for downstream where the flow is intermittent flow with high quality, and a staggered bundle is better for the other region.

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