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ON THE DEVELOPMENT OF THE ANNULAR-DISPERSED FLOW

There is presented a hypothesis about what the mechanism of the annular-dispersed flow development is. A mass balance of the liquid occurring in the form of film and droplets is also determined. Yet investigations of the mass transfer of droplets of liquid from the gas phase core to film have been thoroughly analysed. Thereafter a model of division of a stream of liquid into droplets and film during the annular-dispersed flow development has been put forward. Last, but not least, methodology and results of the research verifying that model have been presented.

NOMENCLATURE

A_F, A_R	– film flow area, gas core flow area (m^2),
C, $\bar{C}$	– mass concentration, average mass concentration of liquid phase (kg/m^3),
d_K	– diameter of droplet (m),
D_A	– diameter of appararus (m),
g	– acceleration due to gravity (m/s^2),
k_K	– mass transfer coefficient for droplets (m/s),
$\dot{M}$	– mass flow rate (kg/s),
n	– number of apparatus section examined,
$Nz = \dot{V}_c/\dot{V}_g$	– liquid to gas flow rate ratio (m^3/m^3),
r	– radius coordinate (m),
t	– time (s),
$u_g^* = \sqrt{\frac{\tau_w}{\rho_g}}$	– friction gas velocity (m/s)
$\dot{V}$	– volume flow rate (m^3/s),
w, $\bar{w}$	– velocity, average velocity (m/s),
W	– gas humidity (kg of H_2O/kg of dry gas),
x	– distance from liquid injection section (m).

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GREEK SYMBOLS

α – quantity defined by equation (17),
δ – diffusion coefficient (kg/(m·s)),
δ_F – film thickness (m),
Δ – difference, increase,
η – viscosity coefficient (kg/(m·s)),
ξ – quantity defined by equation (32),
ρ – density (kg/m^3),
σ – surface tension (N/m),
τ_w, τ_i – near the wall, interfacial shear stress (N/m^2),
τ^+ – nondimensional time of droplet relaxation,
τ_p – quantity defined by equation (33).

SUBSCRIPTS

A – apparatus,
B – influence of inertia,
c – liquid phase or position in apparatus axis,
F – film,
g – gas phase,
gs – dry gas,
i – interfacial surface,
K – droplets,
KD – deposited droplets,
KG – generated droplets,
pw – water vapour,
∞ – fully developed flow.

SUPERSCRIPTS

R – radial movement of droplets,
$*$ – equilibrium quantities.

1. INTRODUCTION

Heat and mass transfer processes occurring during contact between liquid and gases basically depends on the two-phase flow. One of the less known structures is the annular-dispersed flow, especially the process of its development. Further knowledge on the subject would enable simulation and, consequently, designing of many processes and apparatuses. It could also contribute to development of methods of heat and mass exchangers designing, high-velocity wet dedusters, reactors for liquid wastes and postsorptive solution oxygenation, but it could also be useful in the combustion theory, or in simulating emergency states of cooling systems in nuclear reactors.

2. A SHORT PHYSICAL ANALYSIS

In the course of a vertical (taking place from the bottom upwards) cocurrent annular-dispersed flow, drops of liquid and its film flowing up the apparatus walls move in the same direction. The liquid injected to a stream of gas flowing at a considerable velocity (20–50 m/s) disintegrates into droplets. The droplets which subject to a drag of the gas dominating over gravity and inertia forces (an aerostatic uplift may be neglected) are being accelerated in the axial direction (from the bottom upwards). Simultaneously, a considerable concentration of droplets of liquid in the gas core along with high turbulence of gas cause also their shift towards the apparatus wall. Deposition of droplets on the walls leads to the formation of liquid film.

Both acceleration of droplets in the axial direction and their radial movement towards the walls become a cause of changes in the droplet concentration in gas along the apparatus axis, and these in turn directly have a bearing on local values of the liquid hold-up and unitary interfacial area.

Strong aerodynamic interactions between a stream of gas and the liquid film surface can be a reason of its destabilization consisting in rippling of the surface and generating new droplets which pass in the gas phase core. Thus, the radial movement of liquid droplets can take place in two directions, adaxial and abaxial. The droplet deposition process slightly influences an average axial velocity of their whole set moving in the gas core, whereas generation of droplets from the surface of the film flowing at relatively low velocities (usually ranging from 1 to 4 m/s) causes a decrease in the average axial velocity of the set of droplets.

Thus, droplet movement in the gas phase core along the apparatus axis is characterized by variable axial and radial components of velocity and variable concentration of the droplets.

The liquid film flowing up the apparatus wall is subject to oppositely directed friction forces of gas against the liquid surface and liquid against the wall. The film is practically flowing at a constant axial velocity, and its slight changes can result from its variable thickness.

3. THE LIQUID MASS BALANCE

Let us consider two neighbouring cross-sections, n and $n+1$, of the apparatus. The mass flow rate of liquid $\dot{M}_c$ in the section $n+1$ equals the mass flow rate of liquid in the section n reduced or increased by a change of water vapour mass flux $\dot{M}_{pw}$ in a segment between the sections n and $n+1$, depending on whether evaporation or condensation takes place in the segment n; $n+1$. It can be written down as follows:

$$\dot{M}_c(n+1) = \dot{M}_c(n) \pm \Delta \dot{M}_{pw}(n;\ n+1). \tag{1}$$

As liquid can occur both in the form of droplets in the gas core and in the form of film on the apparatus walls, the form of equation is slightly complex

$$\dot{M}_K(n+1) + \dot{M}_F(n+1) = \dot{M}_K(n) + \dot{M}_F(n) \pm \Delta\dot{M}_{\mathrm{pw}}(n;\ n+1). \tag{2}$$

Because

$$\dot{M}_K = w_K C_K A_K, \tag{3}$$

$$\dot{M}_F = \bar{w}_F \rho_c A_F, \tag{4}$$

$$A_K = \frac{\pi D_A^2}{4} - A_F = \frac{\pi}{4}(D_A - 2\delta_F)^2 \tag{5}$$

and

$$\dot{M}_{\mathrm{pw}} = \dot{V}_{\mathrm{gs}}\rho_{\mathrm{gs}}W = \dot{M}_{\mathrm{gs}}W, \tag{6}$$

finally we get

$$\begin{aligned} & w_K(n+1)C_K(n+1)\frac{\pi}{4}[D_A - 2\delta_F(n+1)]^2 \\ & + \bar{w}_F(n+1)\rho_c\pi\delta_F(n+1)[D_A - \delta_F(n+1)] \\ = {} & w_K(n)C_K(n)\frac{\pi}{4}[D_A - 2\delta_F(n)]^2 + \bar{w}_F(n)\rho_c\pi\delta_F(n)[D_A - \delta_F(n)] \\ & + \dot{M}_{\mathrm{gs}}[W(n+1) - W(n)]. \end{aligned} \tag{7}$$

The balance equation can be also written in a slightly different form. It expresses a change in the mass flow rate of the droplets in the axial direction by means of mass flux of the droplets in the radial direction $\dot{M}_K^R$

$$\dot{M}_K(n+1) = \dot{M}_K(n) \pm \dot{M}_K^R(n;\ n+1) \pm \Delta\dot{M}_{\mathrm{pw}K}(n;\ n+1), \tag{8}$$

and similarly, a change in the mass flow rate of the film in the axial direction as

$$\dot{M}_F(n+1) = \dot{M}_F(n) \pm \dot{M}_K^R(n;\ n+1) \pm \Delta\dot{M}_{\mathrm{pw}F}(n;\ n+1). \tag{9}$$

A gain in the mass flow rate of droplets in the axial direction takes place as a result of droplets generation from the film surface, whereas a decrease results from their deposition on that surface.

The opposite phenomenon takes place in the case of changes in the mass flow rate of the film.

In order to use practically the balance equations presented above, one should describe a series of physical quantities by functions of the time t of the movement of liquid elements and their present distance x from the injection section. This refers among others to the mass flow rate of liquid in the form of the droplets $\dot{M}_K$ and the film

$\dot{M}_F$ or their fraction in the overal liquid mass flow rate, axial velocity of the droplets w_K and the average film velocity $\bar{w}_F$.

4. DIVISION OF THE LIQUID MASS FLUX INTO DROPLETS AND FILM

4.1. A HYPOTHETICAL COURSE OF THE PROCESS

It can be supposed that radial movement of droplets from places of higher concentration to these of lower is caused generally by turbulence of a two-phase droplets–gas mixture. Equalization of the droplet concentrations in the gas transporting them is similar to the turbulent diffusion of gaseous substances. Thus, for example, a mass flux of diffusing droplets depends on difference in their concentrations (i.e., the difference between the concentrations of droplets at the film surface and in the gas core) and the interfacial area (in this case, the film surface). If we deal with the dynamic equilibrium, characterized by equality of the mass flux of droplets from the gas core and those generated from the film and occurring during the fully developed flow, we can then assume that exchange of droplets takes place only in a thin layer of the gas near the film, as it is not connected with a change of droplet concentration in the gas core. Then we can write that

$$\begin{aligned}\dot{M}_K^R(n;\ n+1) &= |\dot{M}_{KG}^R(n;\ n+1)| - |\dot{M}_{KD}^R(n;\ n+1)| \\ &= k_K\pi(D_A - 2\bar{\delta}_F)[x(n+1) - x(n)](\bar{C}_K^* - \bar{C}_K) = 0. \end{aligned} \tag{10}$$

When the flow is not fully developed, there is no dynamic equilibrium previously mentioned because

$$\dot{M}_K^R(n;\ n+1) \neq 0. \tag{11}$$

As there is no relevant report in the literature on the subject, it is difficult to formulate explicitly the right-hand side of the dependence (11). It is sure, however, that $\bar{C}_K^* \neq \bar{C}_K$, which is related to the inevitable radial shift of the liquid droplets inside the gas core. As a result, the droplets can reach considerable velocities in the radial direction, and because their mass is substantial (for example, in comparison to the diffusing gas molecules), the value of the radial mass flux of the droplets has to be influenced by an inertial factor. Thus, the hypothesis that in the course of the not fully developed flow the mass flux of droplets in the radial direction can be expressed as

$$\dot{M}_K^R(n;\ n+1) = k_K\pi(D_A - 2\bar{\delta}_F)[x(n+1) - x(n)](\bar{C}_K^* - \bar{C}_K) \pm \dot{M}_B^R(n;\ n+1) \tag{12}$$

seems probable, while the inertial component is probably a function of the radial velocity and the droplet concentration

$$\dot{M}_B^R = f(w_K^R, C_K). \tag{13}$$

A significant obstacle in defining the mass flux of droplets moving in the radial direction is diversity of meaning and fragmentary state of the data that have hitherto been obtained. The data concerns the droplet mass transfer, intensity of which is to be characterized by the droplet mass transfer coefficient k_K. The reports that have been done so far ([1]–[3], [5]–[12]) refer exclusively to the fully developed flow. HEWITT [1] analysing the vertical, fully developed annular-dispersed flow claims that according to the results of the experiments carried out jointly with COUSINS [9], the mass transfer coefficient of droplets is independent of the gas velocity, the tube diameter and the gas density, and in the range of parameters examined it amounts to about 0.15 m/s. The problem discussed here was considered in a more detailed way by GANIČ and MASTANAJAH [13], [14]. The ultimate result of their considerations is a quite complicated dependence of the following form

$$k_K = \frac{u_g^* \dfrac{C_c}{C_K}}{\left\{2.5 \ln \left[\dfrac{1+2\left(1-\dfrac{30}{r_0^+}\right)^2}{1-\left(1-\dfrac{30}{r_0^+}\right)^2}\right] + \dfrac{1}{0.75}\sqrt{\dfrac{1}{1+3\dfrac{\tau^+}{r_0^+}\alpha}}\right\}} \, \frac{1}{1+3\dfrac{\tau^+}{r_0^+}\alpha}, \tag{14}$$

where

$$\tau^+ = \frac{d_K^2 \rho_g \rho_c u_g^{*2}}{18\eta_g^2}, \tag{15}$$

$$r_0^+ = \frac{r_0 u_g^* \rho_g}{\eta_g}, \tag{16}$$

$$\alpha = 6.3 \cdot 10^{-4} Re_g^{0.509}. \tag{17}$$

The dependence (14), however, has been formulated basing on the assumption that the flow is fully developed, and there are no inlet effects. Also the latest papers [15]–[17] devoted to the radial movement of droplets (or particles) do not explain a manner of defining the coefficient k_K when one deals with a not fully developed flow.

Thus, there are no premises so far that could allow us to estimate at least the values of the mass transfer coefficient of droplets in the course of the annular-dispersed flow development. So it is not possible, at the present level of knowledge, to model mathematically the mass transfer of droplets on the basis of equation (12).

Using an overall physical analysis of the development process of the annular-dispersed flow, one can foresee its qualitative course.

Let us assume that liquid is injected into the apparatus and directed to the gas phase where it almost immediately disintegrates into the droplets. Such a situation arises when a high concentration of droplets in the gas core C_K is accompanied by the lack of the liquid film on the internal wall of the apparatus, and the concentration of droplets in the layer near the wall attains the value of $C_K^* = 0$. The difference between the droplet concentrations is a reason of their movement towards the wall. The radial movement of droplets towards the wall takes place continually, which causes not only deposition of the droplets that are near the wall, but also acceleration of the droplets that are deep in the gas core. As a result of droplet deposition on the apparatus wall, there appears liquid film efficiently increasing its thickness and flowing cocurrently with the gas. The increasing thickness of the film facilitates its deformations. These are caused by shear stresses existing on the film surface as a result of the gas friction. Consequently, the deformation increase leads to intensification of the droplet generation from the liquid film. These droplets pass into a thin boundary layer of gas, causing in this way increase in the value of the droplet equilibrium concentration C_K^*. Thus, when the droplet deposition overbalances their generation, we are to deal with a decrease in the droplet concentration in the gas core, $(dC_K/dt) < 0$ and an increase in the droplet equilibrium concentration $(dC_K^*/dt) > 0$.

At the moment of levelling of the two concentrations C_K^* and C_K, the velocities reached by the droplets in the radial direction cause inertial movement of the droplets towards the wall, and their deposition is continued for some time. It brings about a further increase both in the film thickness and the droplet equilibrium concentration $(C_K^* > C_K)$. The state lasts until

$$|(|\dot{M}_{\mathrm{KG}}^R| - |\dot{M}_{\mathrm{KD}}^R|)| < |\dot{M}_B^R|. \tag{18}$$

At the moment when $|(|\dot{M}_{\mathrm{KG}}^R|-|\dot{M}_{\mathrm{KD}}^R|)| = |\dot{M}_B^R|$ and $\dot{M}_K^R = 0$, a relation between the two droplet concentrations will be as follows: $C_K^* > C_K$. The next moment a shift of droplets from the boundary layer to the gas phase core starts. If we assume that the droplet movement in the abaxial direction is a possitive direction and the adaxial is a negative one, then this phase of the annular-dispersed flow development begins when $\dot{M}_K^R < 0$. In this period, the equilibrium concentration of droplets diminishes, $(dC_K^*/dt) < 0$, the film thickness diminishes, and acceleration of the droplets in the adaxial direction takes place. When the equilibrium concentrations becomes equal to the droplet concentration in the gas core, $C_K^* = C_K$, the inertial component of the droplet mass flux $\dot{M}_B^R < 0$ will cause further adaxial movement of droplets in the gas core and further generation of droplets in the film surface.

When $|(|\dot{M}_{\mathrm{KD}}^R|-|\dot{M}_{\mathrm{KG}}^R|)|$ will reach the value of $|\dot{M}_B^R|$, the droplet mass flux in the radial direction will temporarily disappear ($\dot{M}_K^R = 0$), and $C_K > C_k^*$. Thus, the process of the abaxial droplet movement will start anew and the phenomena described above will repeat again. Variation in the droplet radial mass flux would have a sinusoidal course if there were no damping factors, like gas friction and energy dissipation resulting from generation of droplets in the film surface. During damping of oscillation, the fraction of the inertial component $\dot{M}_B^R$ in the balance equation (12) diminishes, and the values of the droplet concentration C_K^* and C_K become closer and closer to each other. At a certain distance from the section of liquid injection, the flow becomes fully developed, $\dot{M}_B^R$ approaches zero and a definite level of concentration $C_{K\infty}^* = C_{K\infty}$ is established. It does not mean, however, that the inverse processes of generation and deposition of droplets have stopped. They still continue, but they remain in the state of dynamic equilibrium, which can be expressed as follows:

$$|\dot{M}_{\mathrm{KG}}^R| = |\dot{M}_{\mathrm{KD}}^R|, \tag{19}$$

so

$$|k_K C_K^*| = |k_K C_K|. \tag{20}$$

Probably, mass transfer of droplets still takes place only in the boundary layer, and principally the droplets from the gas core do not take part in the process, which is responsible for the lack of the inertial component in the balance (19).

At this stage, an additional question should be posed: on what level should the mentioned state of dynamic equilibrium become fixed, that is what values do the droplet concentrations $C_{K\infty}^* = C_{K\infty}$ and the volume fraction of the film in the total amount of liquid flowing through the apparatus $(\dot{V}_F/\dot{V}_c)_\infty$ reach? It seems that it is easier to answer these questions as the quantities in question correspond to the state of the full development of the flow which has been thoroughly investigated. The most precious in this respect is the research of HUTCHINSON and WHALLEY [8] concerning the values of droplet equilibrium concentration $C_{K\infty}^*$. They presented their results in the form of a graphic dependence based on a dimensionless criterion $S = \tau_i \delta_F / \sigma$. In order to facilitate the use of Hutchinson's and Whalley's results, especially in the field of mathematical modelling of annular-dispersed flows, MELOCH described them [18] by means of the following dependence

$$C_{K\infty}^* = \left(\frac{\tau_i \delta_F}{\sigma}\right)^{0.885} \left[103 \left(\frac{\tau_i \delta_F}{\sigma}\right)^{0.885} - 1\right] + 0.885 \cdot 10^{-2}, \tag{21}$$

while the shear stresses taking place on the interface can be expressed as

$$\tau_i = \frac{\frac{D_A}{2} - \delta_F}{2}\left(-\frac{dp}{dx} - \rho_g g\right) \cong -\frac{D_A}{4}\left(\frac{dp}{dx} + \rho_g g\right). \tag{22}$$

Assuming the annular character of the film flow, we can write down that

$$\dot{V}_F \cong \left[\frac{\pi D_A^2}{4} - \frac{\pi(D_A - 2\delta_F)^2}{4}\right]\bar{w}_F, \tag{23}$$

so

$$\frac{\dot{V}_F}{\dot{V}_c} \cong \frac{\pi \bar{w}_F}{\dot{V}_c}\delta_F(D_A - \delta_F), \tag{24}$$

whereas

$$\left(\frac{\dot{V}_F}{\dot{V}_c}\right)_\infty \cong \frac{\pi \bar{w}_{F\infty}}{\dot{V}_c}\delta_{F\infty}(D_A - \delta_{F\infty}). \tag{25}$$

Moreover, taking into account that

$$\dot{V}_K \rho_c = (\dot{V}_c - \dot{V}_F)\rho_c \cong \frac{\pi(D_A - 2\delta_F)^2}{4} C_K w_K, \tag{26}$$

we can arrive at a dependence

$$\frac{\dot{V}_F}{\dot{V}_c} = 1 - \frac{\pi}{4\dot{V}_c\rho_c}(D_A - 2\delta_F)^2 C_K w_K \tag{27}$$

and so

$$\left(\frac{\dot{V}_F}{\dot{V}_c}\right)_\infty = 1 - \frac{\pi}{4\dot{V}_c\rho_c}(D_A - 2\delta_{F\infty})^2 C^*_{K\infty} w_{K\infty}. \tag{28}$$

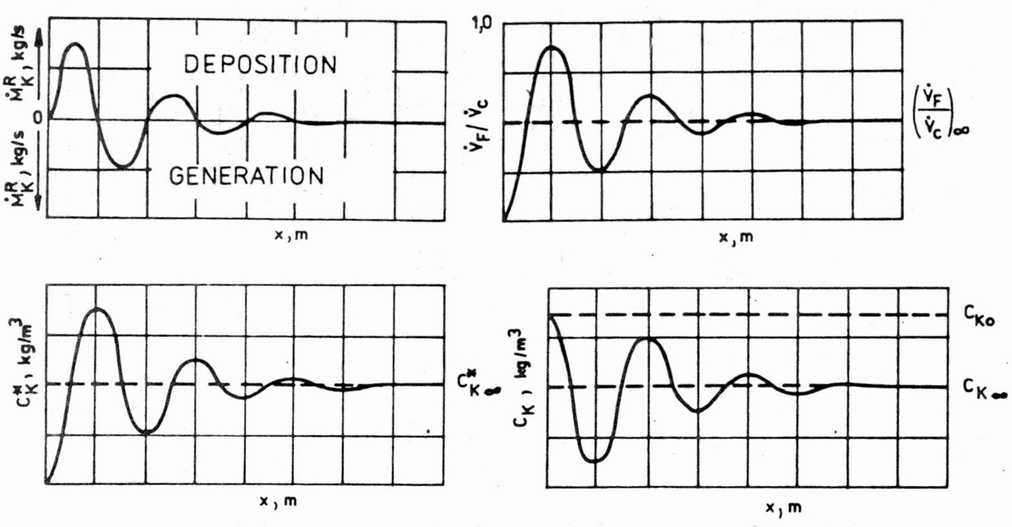

Fig. 1. A hypothetical development of the vertical annular-dispersed flow when a liquid is introduced into the gas core

In this way, a system of three equations (21), (25) and (28) has been established, and its solution allows us to estimate the value in question $(\dot{V}_F/\dot{V}_c)_\infty$ and other, equally significant, parameters of the fully developed annular-dispersed flow, like $C^*_{K\infty} = C_{K\infty}$ and $\delta_{F\infty}$.

The mechanism of the annular-dispersed flow development presented above has been illustrated in figure 1.

If the above considerations prove right, the formulation of quantitative dependences between basic parameters of the developed flow becomes indispensable to evaluation of real interfacial area and to correct designing of heat and mass transfer processes accompanying this flow.

4.2. THE FILM FRACTION IN THE TOTAL FLOW RATE OF LIQUID

In order to estimate the fraction, the flow rate of the liquid existing in the form of droplets or film should be measured. Bearing in mind the radial gradients of velocity and of droplet concentration in gas, the flow rate of droplets should be examined in many points of each cross-section of the apparatus. This would make the experiment labour-consuming and expensive. Therefore, we decided to take careful measurement applying the method of suction through porous fragments of the wall [4].

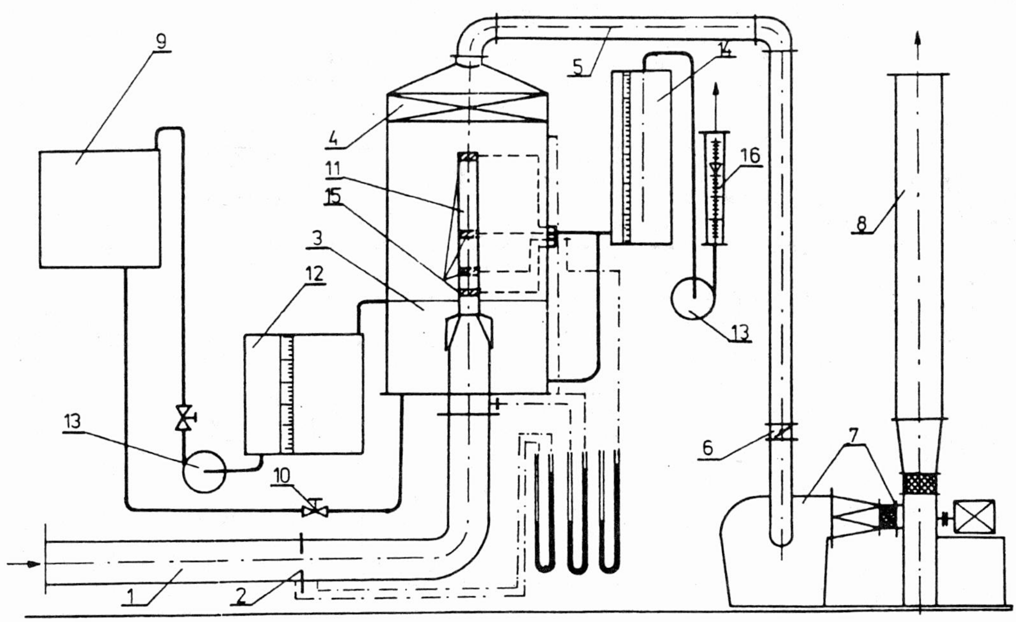

Fig. 2. Diagram of the stand for measuring the liquid film flow rate
1 – surveyor pipeline, *2* – flowmeter, *3* – injection contactor, *4* – droplet separator, *5*, *8* – pipelines, *6* – throttle, *7* – fans, *9* – tank, *10* – valve, *11* – contacting pipe, *12*, *14* – surveyor tanks, *13* – pumps, *15* – porous fragments of the pipe, *16* – flowmeter

A diagram of a testing stand is presented in figure 2. The structure of porous elements of the wall, serving for the liquid film suction, is presented in figure 3. The amount of the sucked liquid (delivered by means of a vacuum pump) had been being increased to a moment when right behind the porous surface we noticed a complete decay of the film. Then the proper measurement took place. We also checked if a further increase in the vacuum pump delivery did not cause any increase in the flow rate of the sucked liquid.

The experiment was carried out in three apparatuses with diameters of 26, 50 and 70 mm.

In each of the apparatuses, measurements were being taken in four measuring sections, about $2D_A$, $4D_A$, $8D_A$ and $16D_A$ away from the liquid injection section.

The measurement results obtained were presented in the form of variation of the film flow rate fraction in the total flow rate of the liquid introduced into the apparatus $\dot{V}_F/\dot{V}_c$. Exemplary variations $\dot{V}_F/\dot{V}_c$ as the function of the distance from the section of liquid injection for some parameters of the two-phase flow were presented in figure 4.

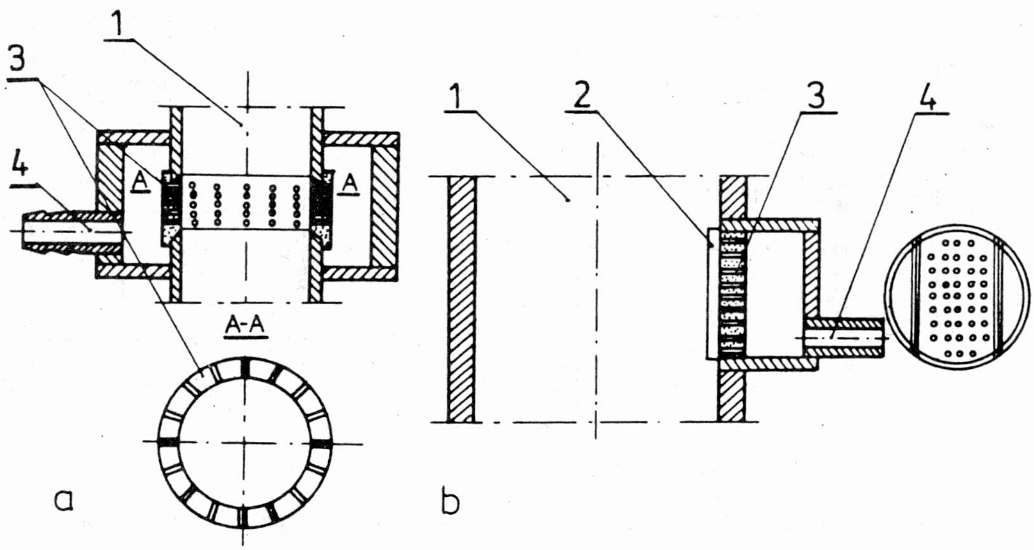

Fig. 3. Structures of the porous elements of contacting pipe
a – pipe of 26 mm in diameter, *b* – pipes of 50 and 70 mm in diameters, *1* – contacting pipe, *2* – stop wall, *3* – porous fragment of the pipe (methylene polymetacrylate sinter with 0.5 mm openings), *4* – film suction pipe stub

The result of the experiment seem to confirm the hypothesis that the annular-dispersed flow development proceeds according to the course presented above, and the variation of the film fraction in the total liquid flow rate can be characterized as oscillation with damping.

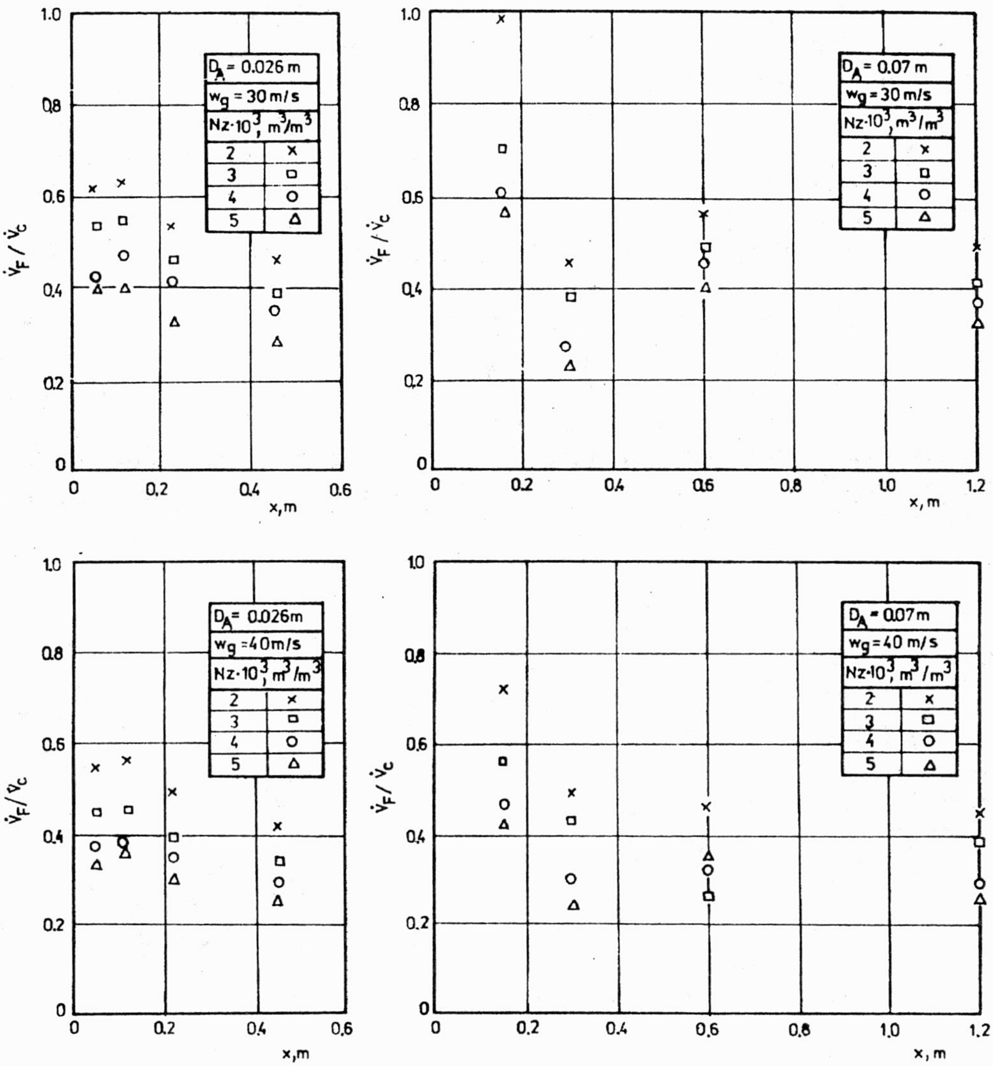

Fig. 4. Exemplary $\dot{V}_F/\dot{V}_c$ variabilities versus distance from the liquid injection section

If the annular-dispersed flow development is analysed in the aspect of dynamics of a certain system in the domain of a distance x, then the variation of the fraction $\dot{V}_F/\dot{V}_c$ constitutes a response of a system described by a linear differential equation of the second order to a step disturbance (figure 5). The interference of the system is caused by injection of a definite stream of liquid into a stream of gas. The response is, of course, different depending on the rate of the disturbance, i.e. on the value of flow rate of the liquid injected into the gas and on a damping degree of the system. The response can be modelled by solution of the mentioned differential equation in the form

$$\frac{\dot{V}_F}{\dot{V}_c} = \left\{\frac{\dot{V}_F}{\dot{V}_c}\right\}_\infty \left\{1 - \exp\left(-\xi\frac{x}{\tau_p}\right) \cdot \left[\cos\left(\frac{x}{\tau_p}\sqrt{1-\xi^2}\right) + \frac{\xi}{\sqrt{1-\xi^2}}\sin\left(\frac{x}{\tau_p}\sqrt{1-\xi^2}\right)\right]\right\}. \tag{29}$$

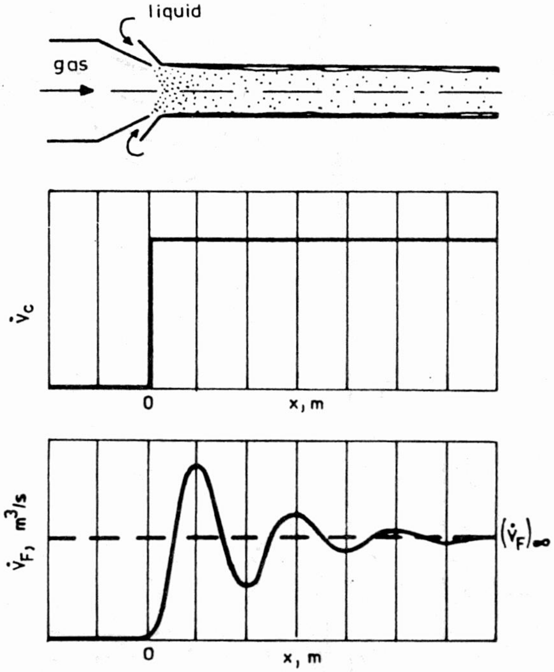

Fig. 5. The system response (in the form of film flow rate $\dot{V}_F$ variability) to step disturbance resulting from the application of a given liquid flow rate $\dot{V}_c$ in the apparatus

Careful observation of the function $\dot{V}_F/\dot{V}_c = f(x)$ in figure 4 allows the statement that not only amplitude is damped, but frequency of changes of the fraction $\dot{V}_F/\dot{V}_c$ as well. Thus, a mathematical description of the courses analysed can be made more precise as follows

$$\frac{\dot{V}_F}{\dot{V}_c} = \left\{\frac{\dot{V}_F}{\dot{V}_c}\right\}_\infty \left\{1 - \exp\left(-\xi\frac{x}{\tau_p}\right) \cdot \left[\cos\left(\frac{x^n}{\tau_p}\sqrt{1-\xi^2}\right) + \frac{\xi}{\sqrt{1-\xi^2}}\sin\left(\frac{x^n}{\tau_p}\sqrt{1-\xi^2}\right)\right]\right\}. \tag{30}$$

The detailed forms of the dependences $n = f_1(D_A, Nz, w_g)$, $\xi = f_2(D_A, Nz, w_g)$ and $\tau_p = f_3(D_A, Nz, w_g)$ have been presented below

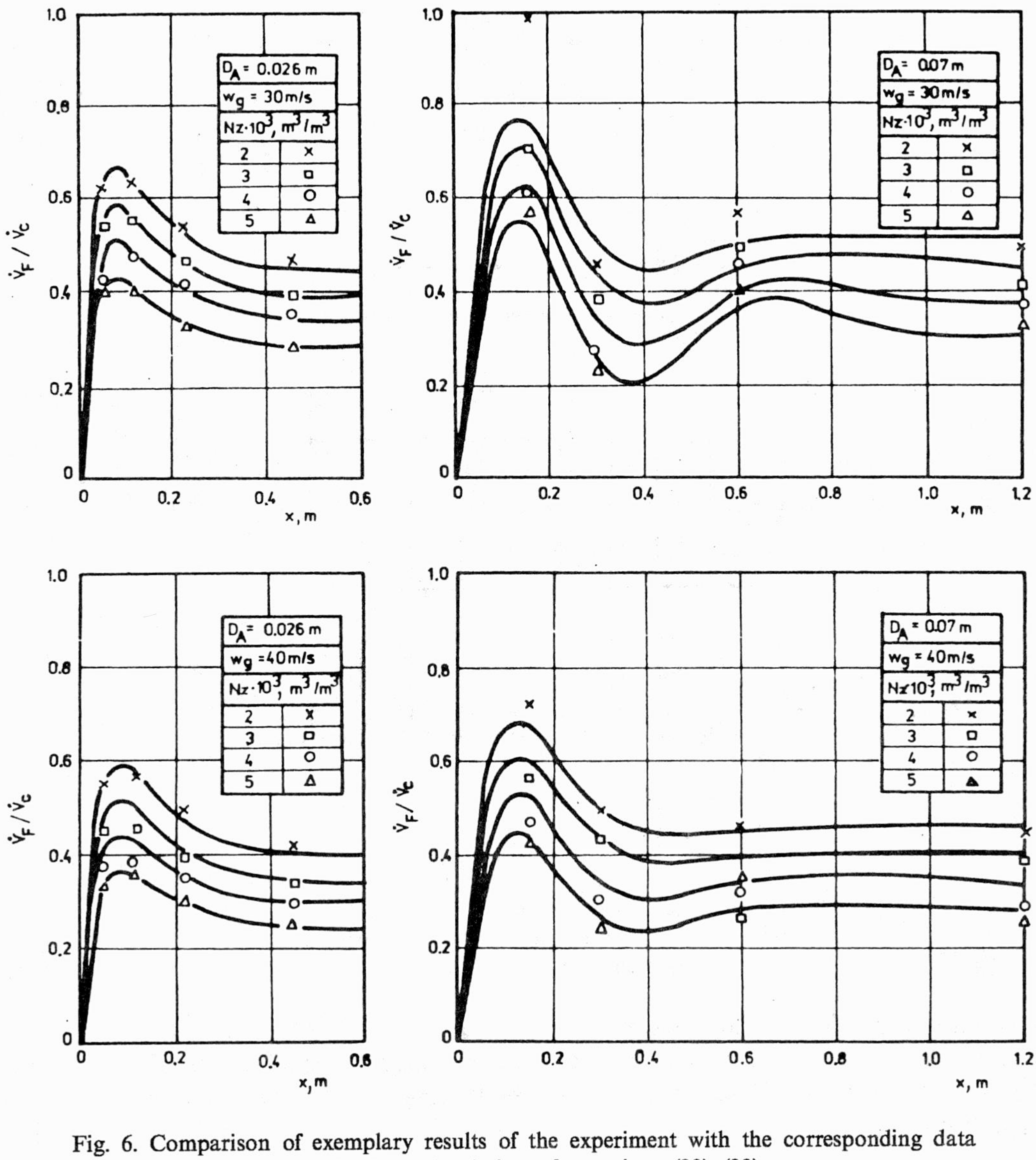

Fig. 6. Comparison of exemplary results of the experiment with the corresponding data obtained by simulation of equations (30)–(33)

$$n = -5.27 \cdot 10^2 D_A^2 + 0.673 \cdot 10^2 D_A - 1.29 + (0.233 \cdot 10^5 D_A^2 - 0.141 \cdot 10^4 D_A - 0.21 \cdot 10^2) Nz + (5.22 D_A^2 - 0.755 D_A + 0.161 \cdot 10^{-1}) w_g, \tag{31}$$

$$\xi = 0.189 \cdot 10^3 D_A^2 - 0.302 \cdot 10^2 + 1.19 + (0.189 \cdot 10^5 D_A^2 - 0.352 \cdot 10^4 D_A + 0.788 \cdot 10^2) Nz - (0.132 \cdot 10^2 D_A^2 - 1.70 D_A + 0.352 \cdot 10^{-1}) w_g, \tag{32}$$

$$\tau_p = 0.136D_A + 0.0664 - (0.575 \cdot 10^2 D_A - 1.49)Nz + (0.0136D_A - 0.354 \cdot 10^{-3})w_g. \quad (33)$$

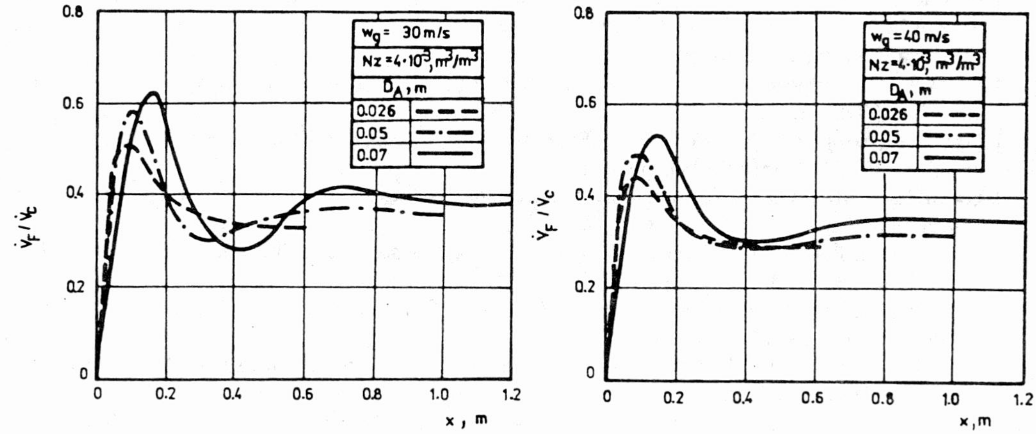

Fig. 7. Exemplary simulations of the variability $\dot{V}_F/\dot{V}_c$ for different apparatus diameters

The equations enable one to describe the experiment results with the inaccuracy, whose value does not exceed $\pm 12\%$, and with the probability of 91%. A comparison of the exemplary experiment results to their respective curves obtained by means of simulation according to the equations (30)–(33) has been presented in figure 6. In figure 7, there are presented the results of simulation of $\dot{V}_F/\dot{V}_c$ variation for different apparatus diameters.

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O ROZWIJANIU SIĘ PRZEPŁYWU PIERŚCIENIOWO-DYSPERSYJNEGO

Przedstawiono hipotezę mechanizmu rozwijania się przepływu pierścieniowo-dyspersyjnego. Określono bilans masy cieczy występującej w formie filmu i kropel. Przeanalizowano dotychczas poznane dane na temat wymiany masy kropel cieczy między rdzeniem fazy gazowej a filmem, a następnie zaproponowano model podziału strumienia cieczy na krople i film podczas rozwijania się przepływu pierścieniowo-dyspersyjnego. Przedstawiono metodykę oraz wyniki badań weryfikujących ten model.

О РАЗВИТИИ ЦИКЛИЧНО-ДИСПЕРСИОННОГО ТЕЧЕНИЯ

Представлена гипотеза механизма развития циклично-дисперсионного течения. Определен баланс массы жидкости, выступающей в виде фильма и капель. Проведен анализ изученных до сих пор данных на тему массообмена капель жидкости между стержнем газовой фазы и фильмом, а затем предложена модель деления потока жидкости на капли и фильм во время развития циклично-дисперсионного течения. Представили методики, а также результаты исследований, проверяющих эту модель.