Route Selectivity for Gas-Liquid Flow in Horizontal *T* Junctions

Jan M. H. Fortuin
Jaap Hart
Peter J. Hamersma
Laboratorium voor Chemische Technologie
Universiteit van Amsterdam
1018 WV Amsterdam, The Netherlands

Since the publication of a paper by Oranje (1973) the fascinating phenomenon of unequal-phase splitting during gas-liquid flow in *T* junctions of pipelines has drawn considerable attention. The so-called "route selectivity" of either gas or liquid is important as it affects the design of headers of reboilers and the performance of downstream equipment of pipelines. Recent reviews concerning two-phase flow splits have been published by Azzopardi (1986) and Lahey (1986). These monographs show that the two-phase flow split in junctions is still a poorly understood phenomenon, although some models are available. Most of these models are highly empirical and therefore they are applicable only for limited flow conditions (Saba and Lahey, 1984; Seeger et al., 1986; Ballyk et al., 1988).

Models for the two-phase flow split which recently have been published are the so-called "geometrical models" (Shoham et al., 1987; Hwang et al., 1988). With these models it is possible to calculate the fraction of the cross-sectional area of the gas flow and that of the liquid flow in the inlet, from which the gas and liquid are taken off into the branch. Generally, this concept is complex and application requires matching parameters and different formulations for various flow patterns in the inlet and geometries of the junction. Moreover, these models often describe experimental results poorly.

In the present paper a model will be introduced describing the route selectivity for gas-liquid flow with *small* liquid holdup values ($\epsilon_L < 0.06$), in horizontal regular dividing *T* junctions. The model is based on the assumption that both the gas flow and the liquid flow in the inlet are split up into two streams, an inlet-to-run stream and an inlet-to-branch stream (see Figure 1). Application of the steady-state macroscopic mechanical energy balances to each of these four streams leads to the so-called "Double Stream Model." This model can be used to predict the phase splitting in dividing junctions, such as the ones occurring in manifolds and gas-liquid separators, gas distribution networks and in distribution networks, which is applied for steam injection in oil wells.

Double-Stream Model

In a steady-state *single-phase* fluid flow the Bernoulli equation holds: i.e., along a streamline in the fluid the sum of the reductions of pressure, kinetic energy and potential energy of the fluid is equal to the frictional energy loss.

Considering a steady-state continuous separated *two-phase* flow of gas and liquid of constant densities ρ_G and ρ_L in a dividing *T* junction, the Bernoulli equation may be applied to:

Inlet-to-run gas flow ('1' to '2'; see also Figure 1):

$$(P_1 - P_2)_G + {}^1\!/_2\rho_G(w_{G1}^2 - w_{G2}^2) + \rho_G g(z_{G1} - z_{G2}) = k_{1,2} \cdot {}^1\!/_2\rho_G w_{G1}^2 \quad (1)$$

Inlet-to-branch gas flow ('1' to '3'):

$$(P_1 - P_3)_G + {}^1\!/_2\rho_G(w_{G1}^2 - w_{G3}^2) + \rho_G g(z_{G1} - z_{G3}) = k_{1,3} \cdot {}^1\!/_2\rho_G w_{G1}^2 \quad (2)$$

Inlet-to-run liquid flow ('1' to '2'):

$$(P_1 - P_2)_L + {}^1\!/_2\rho_L(w_{L1}^2 - w_{L2}^2) + \rho_L g(z_{L1} - z_{L2}) = k'_{1,2} \cdot {}^1\!/_2\rho_L w_{L1}^2 \quad (3)$$

Correspondence concerning this paper should be addressed to J. M. H. Fortuin.

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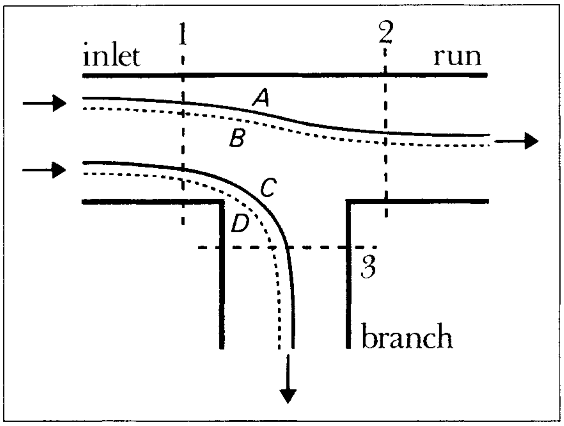

Figure 1. Fluid flow through a regular dividing *T* junction according to the double-stream model.

Streamlines:
A = inlet-to-run gas flow
B = inlet-to-run liquid flow
C = inlet-to-branch gas flow
D = inlet-to-branch liquid flow

Inlet-to-branch liquid flow ('1' to '3'):

$$(P_1 - P_3)_L + \tfrac{1}{2}\rho_L(w_{L1}^2 - w_{L3}^2) + \rho_L g(z_{L1} - z_{L3}) = k'_{1,3} \cdot \tfrac{1}{2}\rho_L w_{L1}^2 \quad (4)$$

where $k_{1,2}$, $k'_{1,2}$, $k_{1,3}$ and $k'_{1,3}$ represent the friction loss coefficients for dividing flow in a pipe tee and where w_{Gi} and w_{Li} (i = 1,2,3 for inlet, run and branch, respectively) represent the root mean squares of the axial velocities of gas in the gas phase and of liquid in the liquid phase. Subtracting Eq. 1 from Eq. 2, Eq. 3 from Eq. 4, and assuming:

$$(P_2 - P_3)_G = (P_2 - P_3)_L \quad (5)$$

result, after division by $\tfrac{1}{2}\rho_L w_{L1}^2$, in:

$$\frac{\rho_G w_{G1}^2}{\rho_L w_{L1}^2}\left(\frac{w_{G2}^2}{w_{G1}^2} - \frac{w_{G3}^2}{w_{G1}^2}\right) - \left(\frac{w_{L2}^2}{w_{L1}^2} - \frac{w_{L3}^2}{w_{L1}^2}\right) + \frac{2g}{w_{L1}^2}\left\{\frac{\rho_G}{\rho_L}(z_{G2} - z_{G3}) - (z_{L2} - z_{L3})\right\} = \frac{\rho_G w_{G1}^2}{\rho_L w_{L1}^2}(k_{1,3} - k_{1,2}) - (k'_{1,3} - k'_{1,2}) \quad (6)$$

If for each of the separated phases the velocity profiles in the inlet, run and branch are similar, we may assume that
In the gas flows:

$$w_{G2}^2/w_{G1}^2 = \langle v_{G2}\rangle^2/\langle v_{G1}\rangle^2$$

and
$$w_{G3}^2/w_{G1}^2 = \langle v_{G3}\rangle^2/\langle v_{G1}\rangle^2 \quad (7)$$

In the liquid flows:

$$w_{L2}^2/w_{L1}^2 = \langle v_{L2}\rangle^2/\langle v_{L1}\rangle^2$$

and
$$w_{L3}^2/w_{L1}^2 = \langle v_{L3}\rangle^2/\langle v_{L1}\rangle^2 \quad (8)$$

Further, we define the ratio κ of the kinetic energies of gas and liquid in the inlet:

$$\kappa = \frac{\rho_G w_{G1}^2}{\rho_L w_{L1}^2} = \alpha\frac{\rho_G\langle v_{G1}\rangle^2}{\rho_L\langle v_{L1}\rangle^2} = \alpha\frac{\rho_G u_{G1}^2}{\rho_L u_{L1}^2} \cdot \frac{\epsilon_{L1}^2}{(1 - \epsilon_{L1})^2} \quad (9)$$

where α is dependent on the velocity profiles of the gas flow and liquid flow in the inlet; $\alpha = 1$ if these profiles are similar.

Equation 6 describes the route selectivity of gas-liquid flow through a pipe tee. This equation can considerably be reduced for gas-liquid flows with *small* liquid holdup values ($\epsilon_L < 0.06$) through a horizontal *regular* dividing *T* junction ($D_1 = D_2 = D_3$).

Defining the mass intake fractions of 'incompressible' gas and liquid in branch and run:

$$\lambda_G = \frac{u_{G3}}{u_{G1}} = \frac{\langle v_{G3}\rangle}{\langle v_{G1}\rangle} \cdot \frac{\epsilon_{G3}}{\epsilon_{G1}} \qquad \lambda_L = \frac{\langle v_{L3}\rangle}{\langle v_{L1}\rangle} \cdot \frac{\epsilon_{L3}}{\epsilon_{L1}}$$
$$1 - \lambda_G = \frac{\langle v_{G2}\rangle}{\langle v_{G1}\rangle} \cdot \frac{\epsilon_{G2}}{\epsilon_{G1}} \qquad 1 - \lambda_L = \frac{\langle v_{L2}\rangle}{\langle v_{L1}\rangle} \cdot \frac{\epsilon_{L2}}{\epsilon_{L1}} \quad (10)$$

and assuming that *approximately:*

$$\epsilon_{L1} = \epsilon_{L2} = \epsilon_{L3} \quad z_{G2} = z_{G3}$$
$$z_{L2} = z_{L3} \quad k'_{1,3} - k'_{1,2} = k_{1,3} - k_{1,2} \quad (11)$$

Equation 6 can be rearranged, using Eqs. 7–11, to:

$$\alpha \cdot \frac{\rho_G\langle v_{G1}\rangle^2}{\rho_L\langle v_{L1}\rangle^2} \cdot (1 - 2\lambda_G) - (1 - 2\lambda_L) = \alpha \cdot \frac{\rho_G\langle v_{G1}\rangle^2}{\rho_L\langle v_{L1}\rangle^2} \cdot (k_{1,3} - k_{1,2}) - (k_{1,3} - k_{1,2}) \quad (12)$$

Defining:

$$\kappa_0 = \frac{\rho_G\langle v_{G1}\rangle^2}{\rho_L\langle v_{L1}\rangle^2} \quad (13)$$

$$\lambda_0 = \tfrac{1}{2}(1 + k_{1,2} - k_{1,3}) \quad (14)$$

and substitution of Eqs. 13 and 14 into Eq. 12 results in:

$$\lambda_L = \lambda_0 + \alpha\kappa_0(\lambda_G - \lambda_0) \quad (15)$$

Equation 15 predicts the branch liquid mass intake fraction λ_L for gas-liquid flow through a horizontal regular dividing *T* junction. According to Eq. 15, λ_L is a function of:

- The branch gas mass intake fraction λ_G
- The geometry of the junction, affecting the friction loss coefficients $k_{1,2}$ and $k_{1,3}$ and λ_0
- The ratio of kinetic energies of gas and liquid in the inlet (κ_0)
- The velocity profiles of gas flow and liquid flow in the inlet, affecting α

With the Double-Stream Model, a so-called "route-selectivity diagram" can be constructed, where λ_L is plotted vs. λ_G for given values of κ. Such a route-selectivity diagram is represented in

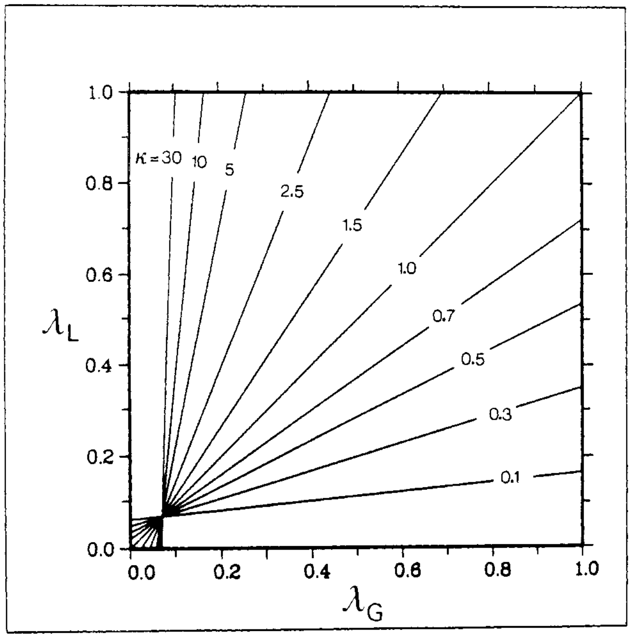

Figure 2. Route selectivity diagram for gas-liquid flow through a horizontal regular sharp-edged dividing *T* junction.

This figure has been calculated with the Double Stream Model (Eq. 15), assuming an average value of $\lambda_0 = 0.07$.

Figure 2 for gas-liquid flow through a horizontal regular *sharp-edged* dividing *T* junction. This figure has been obtained with Eq. 15, taking $\lambda_0 = 0.07$.

According to Eqs. 11 and 14 the values of λ_0, $k'_{1,3} - k'_{1,2}$ and $k_{1,3} - k_{1,2}$ can be calculated if the *single-phase* friction loss coefficients $k_{1,2}$ and $k_{1,3}$ are known. The values of $k_{1,2}$ and $k_{1,3}$ can be calculated with correlations published by Gardel (1957), which are applicable to sharp-edged and rounded, regular and reduced *T* and *Y* junctions. Generally, the accuracy of calculated k-values is limited (Denn, 1980). If for a given *T* junction no experimental single-phase k-values are available, Gardel's correlations can be applied for obtaining reasonable approximations.

Effect of Velocity Profile on α

Under industrial *low liquid holdup* conditions the gas flow is turbulent, but the liquid flow can be either laminar or turbulent. If in the inlet both the gas flow and the liquid flow are turbulent, the velocity profiles of these phases are similar, resulting in $\alpha = 1$. However, if in the inlet the gas flow is turbulent and the liquid film is laminar, $w_{L1}^2 = \langle v_{L1}^3 \rangle / \langle v_{L1} \rangle \neq \langle v_{L1} \rangle^2$, and $\alpha \neq 1$. If in this case the laminar liquid film has a *parabolic* velocity profile, it can be derived that $\alpha = 0.65$. Laminar liquid film flow in the inlet occurs if (Bird et al., 1960):

$$Re_{L1} = 4\Gamma_1/\eta_L < 2{,}000 \quad \text{where} \quad Re_{L1} = Re_{SL1}/\Theta_1 \qquad (16)$$

In Eq. 16, Γ_1 is the mass flow rate of the liquid film in the inlet per unit width of wetted wall, and Θ_1 is the fraction of the tube wall wetted by the liquid film. Correlations for obtaining Θ_1 values have been published elsewhere (Hart et al., 1989).

Liquid Holdup ϵ_L

From the above it is clear that knowledge of the *liquid holdup* ϵ_{L1} in the inlet is crucial for determining the liquid route selectivity. For the determination of *small* liquid holdup values ($\epsilon_{L1} < 0.06$) in the inlet of a horizontal *T* junction, the following relation can be applied (Hart, 1988; Hart et al., 1989):

$$\frac{\epsilon_{L1}}{1 - \epsilon_{L1}} = \frac{u_{L1}}{u_{G1}} \{1 + [10.4\, Re_{SL1}^{-0.363} (\rho_L/\rho_G)^{1/2}]\} \qquad (17)$$

Experimental Verification

In Figure 3 a comparison has been made between solid lines obtained with the present model (Eqs. 15, 16 and 17, and $\lambda_0 = 0.07$) and markers referring to experimental results obtained in our laboratory with a 51-mm-diameter pipeline system containing a sharp-edged horizontal regular dividing *T* junction. For the lines A, B and C the liquid film flow in the inlet is laminar (Re_{L1}, < 2,000 and $\alpha = 0.65$), and for the line D the liquid film flow in the inlet is turbulent ($Re_{L1} > 2{,}000$ and $\alpha = 1$). It can be seen that a good agreement is found between experimentally determined and calculated values concerning the liquid route selectivity. Note that, if $\kappa > 15$, an increase of λ_G from 0.05 to 0.15 results in a sudden increase of λ_L from 0 to 1 ('flip-flop' effect).

Moreover, experimental results obtained from literature (Shoham et al., 1987) also agree well with the Double Stream Model, as has been shown in Figure 4. In this figure line A represents a

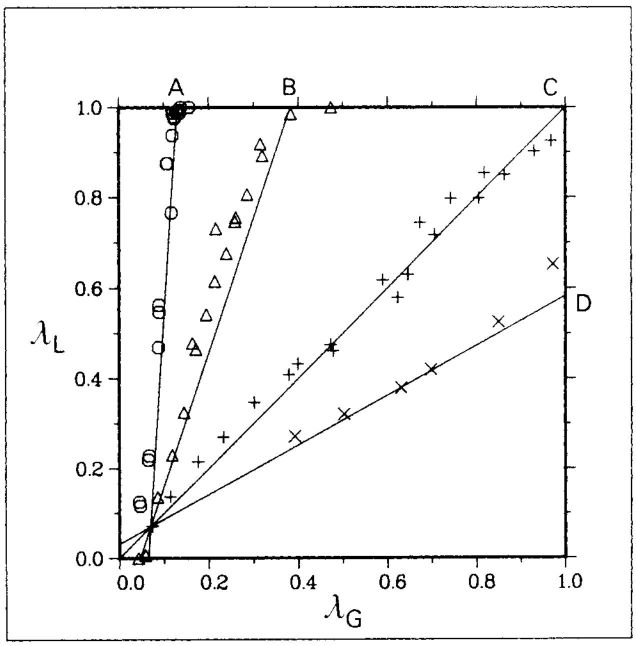

Figure 3. Liquid branch mass intake fraction λ_L as a function of gas branch mass intake fraction λ_G for air-water flow through a horizontal regular sharp-edged *T* junction.

At room temperature and atmospheric pressure; $u_{G1} = 12.4\ \mathrm{m \cdot s^{-1}}$ $D_1 = D_2 = D_3 = 0.051$ m.

O, A: $u_{L1} = 0.000157\ \mathrm{m \cdot s^{-1}}$ ($\epsilon_{L1} = 0.0019$, $Re_{L1} = 113$, $\kappa = 15.2$)
Δ, B: $u_{L1} = 0.00156\ \mathrm{m \cdot s^{-1}}$ ($\epsilon_{L1} = 0.0077$, $Re_{L1} = 552$, $\kappa = 2.97$)
+, C: $u_{L1} = 0.00724\ \mathrm{m \cdot s^{-1}}$ ($\epsilon_{L1} = 0.020$, $Re_{L1} = 1{,}563$, $\kappa = 1.00$)
X, D: $u_{L1} = 0.0313\ \mathrm{m \cdot s^{-1}}$ ($\epsilon_{L1} = 0.051$, $Re_{L1} = 4{,}029$, $\kappa = 0.055$)

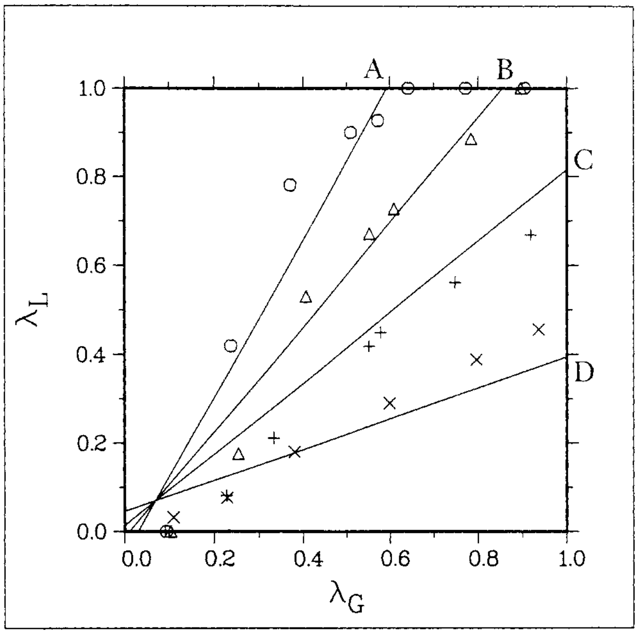

Figure 4. Liquid branch mass intake fraction λ_L as a function of gas branch mass intake fraction λ_G for air-water flow through a horizontal regular sharp-edged dividing *T* junction.

At room temperature and pressure of 3 bar; $u_{G1} = 6.3\ m \cdot s^{-1}$
$D_1 = D_2 = D_3 = 0.051\ m$
O, A: $u_{L1} = 0.00285\ m \cdot s^{-1}$ ($\epsilon_{L1} = 0.012$; $Re_{L1} = 1{,}001$; $\kappa = 1.78$)
Δ, B: $u_{L1} = 0.00914\ m \cdot s^{-1}$ ($\epsilon_{L1} = 0.026$; $Re_{L1} = 2{,}221$; $\kappa = 1.23$)
+, C: $u_{L1} = 0.03048\ m \cdot s^{-1}$ ($\epsilon_{L1} = 0.055$; $Re_{L1} = 4{,}980$; $\kappa = 0.54$)
X, D: $u_{L1} = 0.0585\ m \cdot s^{-1}$ ($\epsilon_{L1} = 0.084$; $Re_{L1} = 7{,}704$; $\kappa = 0.35$)
From Shoham et al. (1987).

laminar flowing liquid film in the inlet; calculation of the conditions represented by the lines B, C and D shows that the liquid film flow in the inlet is turbulent.

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Notation

- D = average internal pipe diameter, m
- g = acceleration due to gravity, $g \approx 9.81\ m \cdot s^{-2}$
- $k_{1,2}, k'_{1,2}$ = frictional loss coefficient for gas flow and liquid flow from site 1 to 2, Figure 1, 1
- $k_{1,3}, k'_{1,3}$ = frictional loss coefficient for gas flow and liquid flow from site 1 to 3, Figure 1, 1
- P = pressure, Pa
- Re_L = Reynolds number of liquid film, 1
- Re_{SL} = superficial Reynolds number of the liquid phase; $Re_{SL} = \rho_L u_L D/\eta_L$, 1
- u = superficial velocity of a fluid, $m \cdot s^{-1}$
- v = time-averaged local axial velocity of a fluid, $m \cdot s^{-1}$
- $\langle v \rangle$ = average value of v in the cross section of one fluid, $m \cdot s^{-1}$
- $\langle v_G \rangle$ = average axial velocity of the gas phase, $\langle v_G \rangle = u_G/\epsilon_G$, $m \cdot s^{-1}$
- $\langle v_L \rangle$ = average axial velocity of the liquid phase, $\langle v_L = u_L/\epsilon_L$, $m \cdot s^{-1}$
- w = root mean square velocity, $w^2 = \langle v^3 \rangle / \langle v \rangle$, $m \cdot s^{-1}$
- z = distance between center of gravity of a fluid and bottom of tube, m

Greek letters

- α = constant dependent on velocity profile in the liquid film in the inlet, 1
- Γ = mass flow rate of liquid film per unit width of wetted wall, wall, $kg \cdot s^{-1} \cdot m^{-1}$
- ϵ_L = fraction of cross-sectional area occupied by the liquid phase (liquid holdup), 1
- ϵ_G = gas holdup, $\epsilon_G = 1 - \epsilon_L$, 1
- η_L = dynamic viscosity of the liquid phase, Pa s
- Θ = fraction of the tube wall wetted by the liquid film, 1
- κ = ratio of kinetic energies based on root mean square velocities of gas and liquid in inlet, Eq. 9, $\kappa = \alpha\kappa_0$, 1
- κ_o = ratio of kinetic energies, based on average axial velocities of gas and liquid in inlet, Eq. 13, 1
- λ_G = mass intake fraction of gas phase into branch, Eq. 10, 1
- λ_L = mass intake fraction of liquid phase into branch, Eq. 10, 1
- λ_0 = parameter defined by Eq. 14, 1
- ρ = density of a fluid, $kg \cdot m^{-3}$

Subscripts

- G = gas
- L = liquid
- ^{1} = at site 1, Figure 1
- ^{2} = at site 2, Figure 1
- ^{3} = at site 3, Figure 1

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