

A Discussion on Wet Gas Flow Parameter Definitions

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1. Introduction

For more than a decade, the technical papers presented at flow meter conferences have included papers that have discussed the issues of wet gas flow metering. This is directly related to the continued increase in the development and use of wet gas flow in the natural gas production industry. Wet gas meter technologies had previously been developed primarily for steam measurement within the power generation industry. However, the steam industry's research into this topic had wound down by the early 1980s. When the natural gas production industry restarted this research on the strength of earlier publications, some terminology was taken from these records and other terminology was created by the new research. However, there has never been any attempt to unify the terminology. Therefore, the result is that researchers, meter manufacturers and meter users are free to produce their own definitions for commonly used terms, which often do not match those of others, and as a result of this there has been confusion and misunderstanding.

There are now a few documents that can be referenced but there is no guarantee that the definition given by any one document will match that of any other document. Furthermore, the authors of wet gas flow metering papers presented at previous North Sea Flow Measurement Workshops sometimes have different definitions for the same term and this has gone unquestioned or un-remarked upon by the audiences.

It is not beneficial to any research topic for no common language to be agreed upon. In some instances at NSFMW question periods, arguments have initiated because the two parties did not realise they were not using the same definitions. In this paper there is an attempt to explain the different terminologies being used and the authors offer suggested definitions that they believe to be the most appropriate for stated technical reasons as a starting point for a debate.

2. Definitions of Two phase Flows, Multiphase Flows and Wet Gas Flows

There can be confusion between the terms *two phase flow*, *multiphase flow* and *wet gas flow*. With no agreed standard to define these terms it is not surprising that different organizations and individuals have created their own definitions and that these do not always agree.

In his presentation at the NSFMW in 2006, Jamieson [1] suggested that industry should adopt the term *unseparated flow* to encompass all hydrocarbon production fluid flows upstream of any process equipment. Unseparated flow is a term that can describe any flow whether it is a single phase homogenous flow or not. This is seen by the authors as a reasonable suggestion. However, even if this term is adopted there is still considerable scope for confusion when trying to define sub-sets of unseparated flow, i.e. two phase flow, multiphase flow and wet gas flow.

In classical physics there are three phases: solids, liquids and gases. It could therefore be reasonably assumed then that the term *multiphase flow* would be used to describe a flow of more than any two possible phases, i.e. solids, liquids and gases. Likewise, it could be reasonably assumed then that the term *two phase flow* would be used to describe a flow that consisted of any two of the three possible phases, i.e. a gas / liquid flow, a gas / solid flow, or a liquid / solid flow. However, in the terminology that has developed in the hydrocarbon production industry these assumptions can in fact be wrong. Flow meters labelled as multiphase flow meters do not typically measure combined flows of solids, liquids and gases but rather the flows of gas and liquid where the liquid phase has more than one component, with different liquid components labelled as different phases. In reality this generally means natural gas, hydrocarbon liquids and water. The multiphase meters on the market do not typically have the ability to indicate or measure the flow of any solids entrained in the gas and liquid flow but they do differentiate the liquid component flowrates, although it is recognised that devices are available for the detection of solid components within multiphase flow streams.

The hydrocarbon production industry commonly defines *two phase flow* as exclusively gas / liquid flows. Furthermore, it is common, but by no means universal, to assume that two phase flow is the flow of a gas with one liquid component. However, with the strict use of the scientific word “phase” a two phase flow of gas and liquid could consist of several different components of each phase (i.e. the gas and the liquid). Almost universally gas is denoted as one phase regardless of the number of components in the gas, whereas it is common to describe a liquid flow of hydrocarbon liquid and water as a flow of two “phases”.

In the flow metering industry terminology an unseparated flow could be used as an all-encompassing term to describe any flow of any combination of gases, liquids and solids (as proposed by Jamieson). It is proposed here that *two phase flow*, which is currently ill-defined by the hydrocarbon production industry, should be any gas and liquid flow. That is, two phase flow is the combined flow of gases and liquids where the phases can consist of one or more components. This then is a subset of unseparated flow. Two phase flow is all unseparated flow that is not a single phase flow and has no solids flowing. As industry is certain to keep the almost universally accepted terminology of *multiphase flow*, it could be said that multiphase flow is a flow of gas and liquid with more than one liquid component. That is, it is a subset of two phase flow as it is a gas and liquid flow where there is more than one liquid component. Figure 1 shows this graphically.

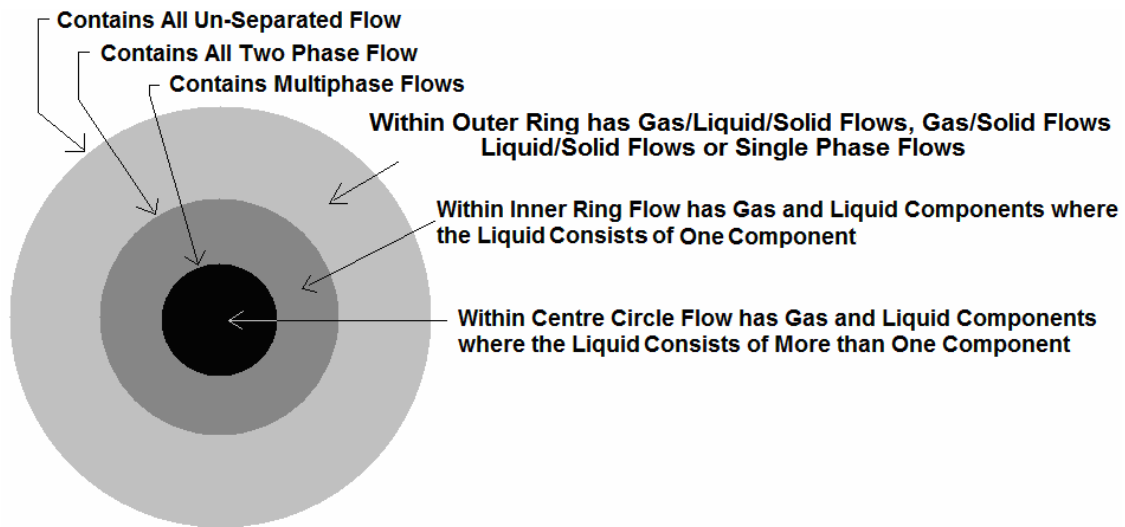


Figure 1. Pictorial view of suggested definitions

In many production flows that are called multiphase flow there could be solid particulates entrained in the liquid and gas flow (e.g. sand, hydrate crystals, etc.). In such cases, the flow would be unseparated flow outside the subsets of two phase or multiphase flow described above. Industry still tends to call this situation multiphase flow which is one of the inconsistencies the adoption of the term unseparated flow could help resolve.

No mention has yet been made of the relative quantities of the phases and the phase components. *Wet gas* is a term that has possibly caused more confusion than the two phase and multiphase flow terms. The details of different definitions given to this term over the years will be dealt with in turn but for the current discussion we will say the term loosely means a relatively small amount of liquid (of any composition) by volume in a flow that is predominantly of gas by volume. All definitions discussed later are on how to quantify relative magnitudes of gas and liquid flowrates and not on what liquid components are flowing. A wet gas flow is always a two phase flow and depending on the liquid phase composition it may be a multiphase flow.

There is sometimes confusion on just what is meant by the flow metering industry with regards to the liquid phase. Gas flows that include components that are evaporated liquids (such as humid air containing water vapour) but have no condensed liquids flowing with them are not usually considered to be two phase, multiphase or wet gas flows from the measurement point of view. One way to explain this is to consider an aircraft flying in humid air (i.e. a relative humidity greater than 0% and less than 100%). The aircraft would not be considered to be flying through a wet gas. If it was flying in the rain that would be a different story.

The Norwegian Society for Oil and Gas Measurement [2] explain this issue in the following way:

“Hydrocarbon gases that contain heavy components that will condense during further processing (but at a particular pressure and temperature behaves as a pure gas) are not considered to be a wet gas from the measurement point of view.”

3. Defining Wet Gas Flow

It is now appropriate to discuss a quantitative definition of wet gas flow. Different researchers and organizations have used various parameters for describing the relative quantity of liquid in a flow predominately of gas; even those that used the same parameters did not necessarily use the same definition for what was the maximum quantity of liquid that would be the cut-off between wet gas and general two phase flow. One of the most commonly used parameters to describe the relative quantity of liquid in a two phase flow is called the *Lockhart-Martinelli parameter* (denoted here as X_{LM}). However, it is crucial that this term is precisely defined as it has been given different definitions over the years, and without a precise definition of the Lockhart-Martinelli parameter, any definition of wet gas flow using this term will be ambiguous. Five different Lockhart-Martinelli parameter definitions have been used, and this has caused considerable confusion. The history of these different definitions and the confusion this has caused is now explained.

3.1. The history of the Lockhart-Martinelli parameter

The term *Lockhart-Martinelli parameter* so often used by the oil and gas industry has a protracted origin. There are now four different equations for this term (and arguably five definitions as one of these equations can be derived in two different ways). To understand these definitions, where they come from, how they relate to each other, why there has been so much confusion and why industry is slowly beginning to favour one of these terms, the discussion starts 58 years ago.

Lockhart and Martinelli [3] discussed the pressure losses occurring in two phase flow in unit lengths of straight, constant area, smooth pipes. This work produced a parameter which Lockhart and Martinelli denoted as X . It was defined, for a unit length of smooth pipe, as the square root of the ratio of the head losses (i.e. friction induced pressure losses) if the liquid and gas phases of a two phase flow flowed alone in the conduit, equation (1),

$$X = \sqrt{\frac{\Delta P_{l,hl}}{\Delta P_{g,hl}}} . \quad (1)$$

Where $\Delta P_{g,hl}$ and $\Delta P_{l,hl}$ are the head losses of the gas and liquid phases of a two phase flow if they were to flow alone in the unit length of pipe. It can be seen that the original term *Lockhart-Martinelli parameter* was not developed as a wet gas flow metering tool but rather as a pressure loss predictor for unit lengths of smooth pipe with two phase flow. Lockhart and Martinelli derived these pressure losses using a generalised Blasius friction factor equation which is applicable only to smooth pipes. For cases where both the liquid and gas phases flowing alone in the pipe had Reynolds numbers greater than

2000 Lockhart and Martinelli stated that the parameter X was found from equation (2). No upper Reynolds number limit was given although it is known that the Blasius friction factor equation is not generally applicable at high Reynolds numbers. However, the maximum Reynolds number in the Lockhart and Martinelli data set was 124,000, so the use of a generalized Blasius friction factor was a reasonable method for Lockhart and Martinelli to derive the pressure loss in their data.

$$X^2 = \frac{\Delta P_{l,hl}}{\Delta P_{g,hl}} = \left(\frac{m_l}{m_g} \right)^{1.8} \frac{\rho_g}{\rho_l} \left(\frac{\mu_l}{\mu_g} \right)^{0.2}, \quad (2)$$

where m_g and m_l are the gas and liquid mass flowrates, ρ_g and ρ_l are the gas and liquid densities, and μ_g and μ_l are the gas and liquid dynamic viscosities.

However, unlike the flow conditions Lockhart and Martinelli were discussing, in most hydrocarbon production flows the pipes can not be presumed to be smooth and usually conditions are such that if the gas and liquid flowed alone in the pipe, for the gas flow at least, the Reynolds numbers would be too high for the application of a Blasius friction factor equation to be appropriate. Therefore, if the original Lockhart-Martinelli parameter definition is to be used in these cases, a more appropriate method is needed for predicting the head losses for the phases flowing alone.

For higher Reynolds numbers, the head loss for a single phase flow per unit length of pipe is predicted in general hydraulic theory by the product of the fluid density and a factor called the “major losses” (usually denoted by the symbol h_l). For horizontal flow this major loss is defined by equation (3),

$$\frac{P_1 - P_2}{\rho} = \frac{\Delta P_{hl}}{\rho} = h_l, \quad (3)$$

where P_1 and P_2 are the upstream and downstream pressures across a unit length of pipe, ΔP_{hl} is the head loss for a single phase flow along a unit length of pipe. For turbulent single phase pipe flows h_l is calculated by equation (4),

$$h_l = f \frac{L \bar{U}^2}{D 2}, \quad (4)$$

where L and D are the unit length of pipe and pipe diameter, \bar{U} is the average flow velocity of the single phase flow and f is the friction factor which is traditionally predicted by the Moody¹ diagram which shows curves of friction factor against Reynolds number for smooth pipes and for pipes of increasing relative roughness, denoted by the term e/D .

¹ Moody published the Moody Diagram (building on the work of Rouse) in 1944, so this work was known and available to Lockhart and Martinelli. This, and the fact that the highest Reynolds number used in the Lockhart and Martinelli data set was 124,000, shows that Lockhart and Martinelli were not aiming their research at the Reynolds numbers range typical to modern day natural gas production flows. Lockhart and Martinelli would therefore have been unlikely to have agreed that their parameter was suitable for such industrial applications.

For two phase, multiphase and wet gas flows when one phase is imagined to flow in the pipe alone the associated average velocity is called the superficial velocity. The liquid superficial velocity (\bar{U}_{sl}) and the gas superficial velocity (\bar{U}_{sg}) are calculated by equations (5a) and (5b),

$$\bar{U}_l = \frac{m_l}{\rho_l A}; \quad (5a)$$

$$\bar{U}_g = \frac{m_g}{\rho_g A}, \quad (5b)$$

where A is the cross sectional area of the pipe.

For wet gas flows, the average liquid velocity if the liquid flowed alone is significantly smaller than the gas velocity if the gas flowed alone. It is also true that usually the liquid viscosity μ_l is order(s) of magnitude greater than the gas viscosity μ_g . Therefore, from the Reynolds number definition, equation (6),

$$\text{Re} = \frac{\text{Inertia Forces}}{\text{Viscous Forces}} = \frac{\rho \bar{U} D}{\mu} = \frac{4m}{\pi \mu D}, \quad (6)$$

it can be seen for a wet gas flow that the liquid Reynolds number if the liquid flowed alone is much lower than the gas Reynolds number if the gas flowed alone.

The friction factor is, for a given pipe roughness, solely a function of the Reynolds number. In this case it is always the same pipe in question with two different flows (i.e. that of the liquid and the gas phases of the wet gas flow as if they were flowing alone in that particular pipe) and therefore the relative pipe roughness (e/D) is constant. Therefore, from the Moody diagram for any given pipe roughness, it can be seen that the difference in Reynolds numbers between the gas and liquid flows means different friction factor values, say f_l and f_g , for the liquid and gas friction factors and because $\text{Re}_l \ll \text{Re}_g$ the condition $f_l > f_g$ exists. From equations (3) and (4) the single phase major pressure loss in a pipe is given by

$$\Delta P = \rho h_l = \rho f \frac{L \bar{U}^2}{D} . \quad (7)$$

Substituting equation (7) into equation (1) for liquid and gas gives the original Lockhart and Martinelli parameter definition (i.e. the square root of the single phase pressure drop ratio) calculated using an appropriate friction factor calculation for rough pipe and the Reynolds numbers in use with typical wet natural gas production flows, X_M ,

$$X_M = \sqrt{\frac{\Delta P_{l,hl}}{\Delta P_{g,hl}}} = \frac{\sqrt{\rho_l f_l \frac{L \bar{U}_l^2}{D \cdot 2}}}{\sqrt{\rho_g f_g \frac{L \bar{U}_g^2}{D \cdot 2}}} = \sqrt{\frac{f_l}{f_g}} \frac{\bar{U}_l}{\bar{U}_g} \sqrt{\frac{\rho_l}{\rho_g}}. \quad (8)$$

Substituting equations (5a) and (5b) into equation 8 gives

$$X_M = \sqrt{\frac{\Delta P_{l,hl}}{\Delta P_{g,hl}}} = \sqrt{\frac{f_l}{f_g}} \frac{m_l}{m_g} \sqrt{\frac{\rho_g}{\rho_l}}. \quad (9)$$

Equation (9) is the mathematical expression for the original meaning of the Lockhart-Martinelli parameter as published by the engineers R.W. Lockhart and R.C. Martinelli in 1949 using the appropriate calculation method for typical flow range encountered in natural gas production. Equation (9) and equation (2) are essentially the same, with the only difference being the method of calculating the friction factors.

Murdock [4] used a parameter which has subsequently been denoted by X by some engineers applying Murdock's work (although Murdock himself never used this symbol). This parameter was used by Murdock during the discussion of the behaviour of orifice plate meters in general two phase flow. For clarity, in this paper $X_{Murdock}$ will denote this Murdock parameter. Murdock described the relative amount of liquid and gas in a given pipe two phase / wet gas flow by using the square root of the ratio of the differential pressures ΔP_l and ΔP_g that would be read by a given orifice plate meter if the liquid and gas phases flowed alone, equation (10),

$$X_{Murdock} = \sqrt{\frac{\Delta P_{l,m}}{\Delta P_{g,m}}}, \quad (10)$$

where subscript m means meter-induced pressure drop.

The standard differential pressure (DP) meter equation is

$$\text{for gas:} \quad m_g = EA_t \varepsilon C_{d,g} \sqrt{2\rho_g \Delta P_{g,m}}; \quad (11a)$$

$$\text{and for liquid:} \quad m_l = EA_t C_{d,l} \sqrt{2\rho_l \Delta P_{l,m}}, \quad (11b)$$

where E is the velocity of approach, A_t is the orifice area, ε is the expansibility factor (denoted as Y by Murdock), $C_{d,g}$ and $C_{d,l}$ are the discharge charge coefficients of gas and liquid, and $\Delta P_{g,m}$ and $\Delta P_{l,m}$ are the differential pressures which would be observed if the gas and liquid phases flowed alone through the meter. Equation (10) can also be written in the form

$$X_{Murdock} = \sqrt{\frac{\Delta P_{l,m}}{\Delta P_{g,m}}} = \frac{C_{d,g} \varepsilon}{C_{d,l}} \frac{m_l}{m_g} \sqrt{\frac{\rho_g}{\rho_l}}. \quad (12)$$

Murdock never called this parameter the Lockhart-Martinelli parameter although some in industry now confuse the terms.

Chisholm [5, 6] published a general two phase flow correlation for orifice plate meters and used a similar, but not identical, parameter denoted by upper case X. For clarity, in this paper $X_{Chisholm}$ will denote this Chisholm parameter. This is shown as equation (13) where x is the flow quality as described by equation (14). The equation for Chisholm's correlating parameter (13) can be rewritten as equation (15),

$$X_{Chisholm} = \frac{1-x}{x} \sqrt{\frac{\rho_g}{\rho_l}}; \quad (13)$$

$$x = \frac{m_g}{m_g + m_l}; \quad (14)$$

$$X_{Chisholm} = \frac{m_l}{m_g} \sqrt{\frac{\rho_g}{\rho_l}}. \quad (15)$$

Chisholm derived equation (15) similarly to the way that Murdock derived equation (12). However, unlike Murdock, Chisholm called the parameter he derived "...the Lockhart-Martinelli correlating group". The reference he cites as his reference 13 is the Lockhart-Martinelli paper, reference [3] in this paper.

Chisholm therefore appears to have erroneously called the parameter defined in equation (15) the Lockhart-Martinelli parameter. Returning to equation (15) it can be seen that it is the same as equation (12) only if there is the added assumption that

$$C_{d,g} \varepsilon = C_{d,l}. \quad (16)$$

This is an approximation that could be treated as valid under some, but not all, two phase flow conditions. However, although equation (15) can be found from equation (12) by making this assumption, Chisholm did not state as much, but it can also be formed by an alternative derivation. For any given pipe cross sectional area and any given two phase flow imagine that the phases were to flow alone. Then, if the square root of the ratio of the inertia of these gas and liquid phase flows across that given cross section of the pipe is taken, this would give a parameter which can be denoted as X_{LM} ,

$$X_{LM} = \sqrt{\frac{\text{Liquid Flowing Alone Inertia}}{\text{Gas Flowing Alone Inertia}}} = \sqrt{\frac{\rho_l U_{sl}^2 D^2}{\rho_g U_{sg}^2 D^2}}, \quad (17)$$

with *Liquid Flowing Alone Inertia* = $\rho_l U_{sl}^2 D^2$ and *Gas Flowing Alone Inertia* = $\rho_g U_{sg}^2 D^2$. Therefore, substituting in equations (5a), (5b) and equation (17) and comparing the result to equation (15) gives,

$$X_{LM} = \sqrt{\frac{\rho_l U_{sl}^2 D^2}{\rho_g U_{sg}^2 D^2}} = \frac{U_{sl}}{U_{sg}} \sqrt{\frac{\rho_l}{\rho_g}} = \frac{m_l}{m_g} \sqrt{\frac{\rho_g}{\rho_l}} = X_{Chisholm}. \quad (18)$$

The importance of this simple result in terms of wet gas flow metering is considerable.

The original Lockhart-Martinelli parameter, X , is only applicable to smooth pipes and low Reynolds numbers. The associated equation developed here that is valid for typical wet gas natural gas production flow conditions, equation (9), is dependent on the friction factors and therefore the condition of the particular pipe in question. Two identical wet gas flow conditions flowing in pipes that only differ by having different relative roughness values, will have different Lockhart-Martinelli parameters. Hence, as a tool to define a wet gas flow so that a simple and direct comparison between flows can be made this is clearly an impractical parameter to use.

The Murdock parameter, $X_{Murdock}$, that some have mistakenly called the Lockhart-Martinelli parameter, is not only dependent on the relationship of meter discharge coefficient with Reynolds number and the gas expansibility factor, but is solely useable for the case of wet gas flow metering with DP meters. It is also dependent on the geometry of the DP meter in question. It is not directly applicable to non-DP meter wet gas flow metering systems. Hence, this is not a practical tool to help characterise a wet gas flow so that a simple and direct comparison between flows can be made.

However, the Chisholm parameter $X_{Chisholm}$, when derived by this different method, unlike the original Lockhart-Martinelli parameter, X , is shown to be independent of the relative pipe roughness. Further, the Chisholm parameter $X_{Chisholm}$, unlike Murdock's parameter, $X_{Murdock}$, is independent of any DP meter technology. It is simply the square root of the ratio of the liquid inertia to the gas inertia at any given cross section of the pipe if the phases of a two phase flow flowed alone. It can therefore be used to describe the liquid to gas content of any wet gas flow, in any roughness of pipe, whether any type of meter is present or not. This parameter is therefore a very useful non-dimensional way of describing the relative amount of liquid in a gas flow.

By the 1990's some in industry were calling Chisholm's version of the parameter $X_{Chisholm}$ the Lockhart-Martinelli parameter, as Chisholm had erroneously done. However, the developed version of the original Lockhart-Martinelli parameter, the Murdock parameter and the Chisholm parameter are only the same if equation (19a) holds. This in fact is never the case. All real two phase flows follow equation (19b),

$$\sqrt{\frac{f_l}{f_g}} = \frac{C_{d,g}\mathcal{E}}{C_{d,l}} = 1; \quad (19a)$$

$$\sqrt{\frac{f_l}{f_g}} \neq \frac{C_{d,g}\mathcal{E}}{C_{d,l}} \neq 1. \quad (19b)$$

For a wet gas flow condition, the square root of the ratio of the friction factors of the liquid and gas flows if these phases were to flow alone in a straight unit length of pipe is not equal to unity; the ratio of the product of gas discharge coefficient and expansibility of the gas flow flowing alone to the liquid discharge coefficient of the liquid flowing alone is not equal to unity. Therefore, the Lockhart-Martinelli, Murdock and Chisholm parameters that have all been commonly denoted by X are all different. The term Lockhart-Martinelli parameter is now extensively used by the metering fraternity to describe either the Murdock or Chisholm parameters with little in the literature to state

that these are two different definitions and neither of these are the actual Lockhart-Martinelli parameter as defined by Lockhart and Martinelli.

These authors of this paper promote the use of the Chisholm parameter due to its independence from pipe friction factors and DP meter characteristics. However, it is understood that the term Lockhart-Martinelli parameter is entrenched in industry and therefore it is suggested that the term should remain and the preferred definition be denoted by X_{LM} , that is,

$$X_{LM} = \sqrt{\frac{\text{Superficial Liquid Inertia Force}}{\text{Superficial Gas Inertia Force}}} = \frac{m_l}{m_g} \sqrt{\frac{\rho_g}{\rho_l}} = \frac{Q_l}{Q_g} \sqrt{\frac{\rho_l}{\rho_g}}. \quad (20)$$

This definition is also written in terms of the gas and liquid volume flowrates Q_g and Q_l as this is also a commonly used equation.

3.2. Is there really an incompatibility in definitions within the industry?

Unfortunately, it is not simply the case that most engineers use one definition and only those not familiar with the industry mistakenly revert to another, as shown by some examples. In these examples it must be remembered that with no formal standard agreed upon engineers are free to use what ever definition they wish and therefore nobody is incorrect. They are simply using the same term to define different phenomena. Examples:

1. If an engineer did not know the definition of the Lockhart-Martinelli parameter it would be reasonable practice to look up the original paper and assume that the given term is the universally used definition as this is common practice in the scientific community. The original Lockhart-Martinelli parameter definition was briefly discussed by Falcone [7] at the NSF MW in 2006. The paper indicated that there are other definitions but did not explain further.
2. Murdock's original plot of the gas flowrate error of a meter vs. $X_{Murdock}$ [4], is now widely referred to as a Murdock plot and is the almost universal method of plotting DP meter wet gas data. However, nearly all recent graphs use gas flowrate error of a meter vs. $X_{Chisholm}$ instead, with no mention of this inconsistency being made. The Murdock parameter definition is indirectly suggested by Geach and Jamieson [8] who made the comment, "the correlations used to correct for liquid entrainment in wet gas require both the gas and liquid discharge coefficients which are not the same." However, this would only be true if the correlation were designed for use with the Murdock parameter $X_{Murdock}$.
3. In many papers and publications, the Chisholm parameter is used in plots and correlations and called the Lockhart-Martinelli parameter without any comment being made regarding the difference, e.g. de Leeuw [9], Hall and Steven [10].

3.3 Impact of the different definitions in practical applications

This is not academic trivia. There is a direct relevance to the current practices involving wet gas flow metering as there can be significant differences between the four equations. A worked example will indicate this point.

Consider a wet natural gas flow in a 4 inch schedule 80 pipe (inside bore diameter of 0.09718 m) with a relative roughness level of 0.001. The pressure is 50 bara at 300K and the natural gas molecular weight is 17.4. The gas density is therefore approximately 38 kg/m³ and the gas viscosity is approximately 0.01124 cP. The gas flowrate is 400 m³/hr (approximately 17.5 MMSCFD or 4.23 kg/s). In this example, the liquid flowrate is 1.937 kg/s (1317 US barrels per day or 8.72 m³/hr), the density is 800 kg/m³ and the liquid viscosity is 1.92 cP. The flow is through an ISO 5167 Part 2 standard orifice plate meter of beta 0.7 with D and D/2 tappings fitted.

What are the values of the original Lockhart-Martinelli parameter, X , the roughness-modified Lockhart-Martinelli parameter, X_M , the Murdock parameter $X_{Murdock}$, and the Chisholm parameter $X_{Chisholm}$, and how do they compare to each other?

Applying the original Lockhart-Martinelli parameter, equation (2), unaware that it is not applicable to these rough pipes with high Reynolds numbers would give

$$X = \sqrt{\frac{\Delta P_{l,hl}}{\Delta P_{g,hl}}} = \sqrt{\left(\frac{m_l}{m_g}\right)^{1.8} \frac{\rho_g}{\rho_l} \left(\frac{\mu_l}{\mu_g}\right)^{0.2}} = 0.180. \quad (21)$$

The Reynolds number of the gas phase if it flowed alone in the pipe is approximately 4.93×10^6 . The Reynolds number of the liquid phase if it flowed alone in the pipe is approximately 1.32×10^4 . From the Moody diagram for a relative roughness of 0.001, the gas friction factor f_g is 0.0196 and the liquid friction factor f_l is 0.032. Therefore the original Lockhart-Martinelli parameter when modified for the case of rough pipes and high Reynolds numbers is

$$X_M = \sqrt{\frac{\Delta P_{l,f}}{\Delta P_{g,f}}} = \sqrt{\frac{f_l}{f_g} \frac{m_l}{m_g} \sqrt{\frac{\rho_g}{\rho_l}}} = 0.128. \quad (22)$$

From Table A.2 of ISO 5167 Part 2 the orifice plate meter discharge coefficient at a Reynolds number of 4.93×10^6 is 0.608 and at a Reynolds number of 1.32×10^4 is 0.639. Using ISO 5167 part 2 the expansibility factor in this example is 0.9887. Therefore Murdock's parameter from equation (12) is

$$X_{Murdock} = \sqrt{\frac{\Delta P_{l,m}}{\Delta P_{g,m}}} = \frac{C_{d,g} \varepsilon}{C_{d,l}} \frac{m_l}{m_g} \sqrt{\frac{\rho_g}{\rho_l}} = 0.094. \quad (23)$$

Chisholm's parameter is calculated from equation (15) and (17),

$$X_{Chisholm} = X_{LM} = \frac{m_l}{m_g} \sqrt{\frac{\rho_g}{\rho_l}} = 0.100. \quad (24)$$

Therefore, for the same wet gas flow conditions,

Calculation	Symbol	Value
Original Lockhart-Martinelli	X	0.180
Roughness modified Lockhart-Martinelli	X _M	0.128
Murdock	X _{Murdock}	0.094
Chisholm	X _{Chisholm}	0.100

Taking X_{LM} as the most appropriate definition it can be seen that the original Lockhart-Martinelli parameter is 80% higher, the modified original parameter for this particular pipe roughness is 28% higher and the Murdock parameter for this particular DP meter is 6% lower. Hence, the difference between these parameters is not trivial.

3.4. Correlation sensitivity to Lockhart-Martinelli parameter errors

The error in the gas flowrate prediction when using a gas DP meter wet gas correlation when there is an error in the Lockhart-Martinelli parameter definition is, while still significant, not as serious as the errors in the Lockhart-Martinelli parameter itself due to the sensitivity of the correlations to the Lockhart-Martinelli parameter. To show this point the above worked example is continued by applying Chisholm's[6] orifice plate wet gas correlation,

$$Over - reading = \frac{m_{g,apparent}}{m_g} = \sqrt{1 + \left\{ \left[\left(\frac{\rho_g}{\rho_l} \right)^{1/4} + \left(\frac{\rho_l}{\rho_g} \right)^{1/4} \right] X_{LM} \right\}^2 + (X_{LM})^2}, \quad (25)$$

where $m_{g,apparent}$ is the uncorrected gas flowrate prediction. With a gas to liquid density ratio of 0.0475 the associated over-readings (or positive error in the gas flowrate prediction) are:

Calculation	Symbol	Value	Gas flowrate over-reading
Original Lockhart-Martinelli	X	0.180	22.6%
Roughness modified Lockhart-Martinelli	X _M	0.128	16.1%
Murdock	X _{Murdock}	0.094	12.0%
Chisholm	X _{Chisholm}	0.100	12.7%

Chisholm used $X_{LM} = X_{Chisholm} = 0.100$, so the Chisholm correlation is saying the over-reading is 12.7%. However, an incorrect choice of definition can cause an error in the final gas mass flowrate prediction. In this case a 0.63% over-estimation for

$X_{Murdock} = 0.094$, a 3% under-estimation for $X_M = 0.128$ and an 8.8% under-estimation for $X = 0.180$. This is a bias, not an uncertainty.

The prediction of a DP meter wet gas correlation is therefore affected by which parameter is being used. Different DP meters and correlations will have different sensitivities to varying Lockhart-Martinelli parameter values. For example, a Venturi meter is more sensitive to the Lockhart-Martinelli parameter than an orifice meter. For the case of a Venturi the above results would most likely have shown greater biases, although the calculation in equation (23) requires the discharge coefficient which for a Venturi requires a flow calibration. The risk of a relatively large bias if the Murdock parameter is applied where a Lockhart-Martinelli parameter should be applied is somewhat mitigated by the Venturi meter's tendency to have a relatively small shift in discharge coefficient across a large Reynolds number range compared to the orifice plate.

These are errors and not uncertainties. Typical DP meter wet gas correlations correct the over-reading to an uncertainty of 2% so small differences may not be noticeable but the added error in this case is avoidable. As this example is randomly chosen it is possible that in many production wet gas flow metering cases the error could be larger.

3.5. Another definition for the Lockhart-Martinelli parameter

The Lockhart-Martinelli parameter can be defined as the square root of the ratio of the gas densimetric Froude number and the liquid densimetric Froude number. This is in fact the same parameter as equation (18). The definition of the gas densimetric Froude number is the square root of the gas flow inertia if the gas flowed alone to the liquid phase gravitational force. The definition of the liquid densimetric Froude number is the square root of the liquid flow inertia if the liquid flowed alone to the liquid phase gravitational force. This is shown by equations (26) and (27),

$$Fr_g = \sqrt{\frac{\text{Superficial Gas Inertia}}{\text{Liquid Gravity Force}}} = \frac{U_{sg}}{\sqrt{gD}} \sqrt{\frac{\rho_g}{\rho_l - \rho_g}} = \frac{m_g}{A\sqrt{gD}} \sqrt{\frac{1}{\rho_g(\rho_l - \rho_g)}}; \quad (26)$$

and

$$Fr_l = \sqrt{\frac{\text{Superficial Liquid Inertia}}{\text{Liquid Gravity Force}}} = \frac{U_{sl}}{\sqrt{gD}} \sqrt{\frac{\rho_l}{\rho_l - \rho_g}} = \frac{m_l}{A\sqrt{gD}} \sqrt{\frac{1}{\rho_l(\rho_l - \rho_g)}}, \quad (27)$$

where g is the gravitational constant. Taking the ratio of the liquid densimetric Froude number and gas densimetric Froude number gives the Chisholm parameter,

$$X_{LM} = \frac{Fr_l}{Fr_g} = \frac{\sqrt{\frac{\text{Gas Superficial Inertia}}{\text{Liquid Gravity Force}}}}{\sqrt{\frac{\text{Liquid Superficial Inertia}}{\text{Liquid Gravity Force}}}} = \sqrt{\frac{\text{Gas Superficial Inertia}}{\text{Liquid Superficial Inertia}}} = \frac{m_l}{m_g} \sqrt{\frac{\rho_g}{\rho_l}}. \quad (28)$$

3.6. Definitions of wet gas flow conditions

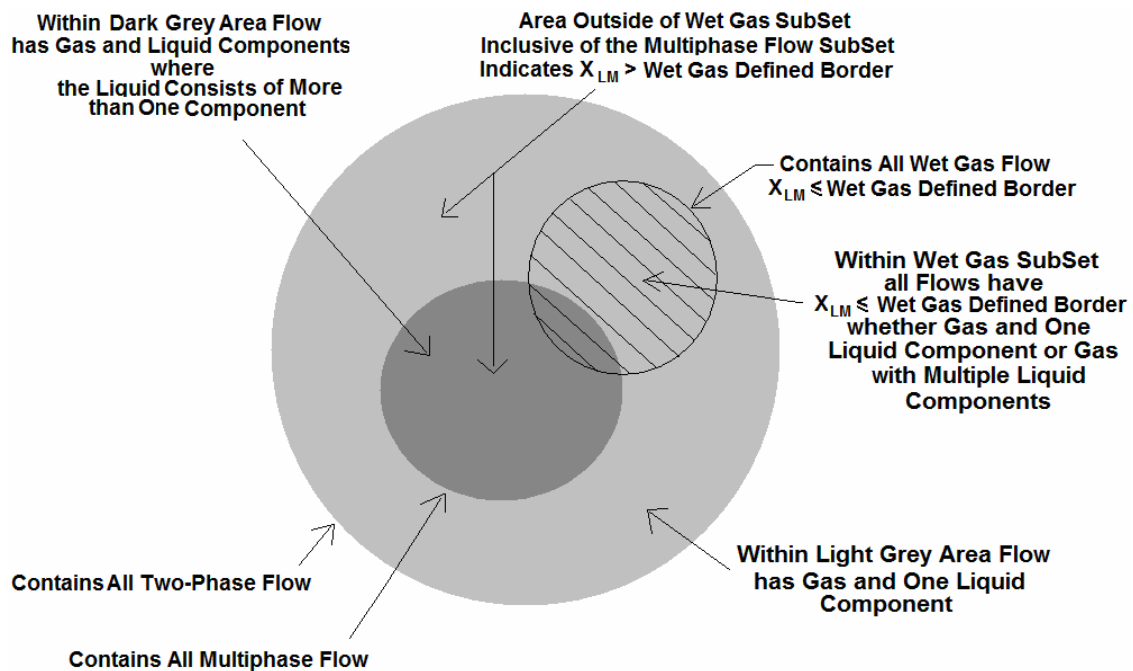


Figure 2. Wet gas as a subset of two phase and multiphase flow

There are several different wet gas flow definitions. For example, many engineers in the steam industries typically call any saturated steam with a quality, x , defined in equation (14), less than unity, wet steam (i.e. a wet gas), but others say wet steam is any steam quality greater than 0.5. Some in the oil and gas industry call wet gas any flow with a gas volume fraction (GVF) greater than 90%. Others have said wet gas flow should be defined as having a Lockhart-Martinelli parameter of 0.35 or 0.3, although they don't all confirm what their definition of Lockhart-Martinelli parameter has been used. (The $X_{LM} \leq 0.35$ definition is an early definition for wet gas used by Shell that came about due to experimental evidence suggesting that slugging in the pipe can become significant at $X_{LM} > 0.35$.) There are other definitions over the years, but recently, with the exception of those that continue to talk of $GVF > 90\%$, the trend is to define wet gas flow as any two phase flow (i.e. any number of liquid components) by using a limiting value of the Lockhart-Martinelli parameter. Figure 2 shows a pictorial view of wet gas.

Just at what value of Lockhart-Martinelli parameter the border between wet gas and general two phase flow should be set is a matter of some debate. Three organisations promote a border in the region of a Lockhart-Martinelli parameter of 0.3. Those are the Norwegian Society for Oil and Gas Measurement [2], the American Petroleum Institute (API) [11²] and the American Society of Mechanical Engineering (ASME³) [12].

² It is understood by the authors that at the time of writing API has not yet ratified this document.

³ At the time of writing this ASME document is under ballot and is due for general release in early 2008.

It is difficult to find a technical review that promotes the use of another wet gas defining parameter with technical reasoning. All three publications listed above give a description of the Lockhart-Martinelli parameter that is fundamentally the same, i.e. equation (18).

The ASME document states that the wet gas definition is $X_{LM} \leq 0.30$. The Norwegian Society for Oil and Gas Measurement defines wet gas as a two phase flow with a maximum Lockhart-Martinelli parameter value of approximately 0.30. That is, they state $X_{LM} \leq 0.30$ as an approximate border between a wet gas flow and a general two phase flow but they stop short of declaring this an absolute border and allow room for individual interpretation.

The API document [11] has a more detailed approach. The definition of wet gas flow is separated into three “types”:

API Type I Wet Gas:

“Wet gas Type I is defined as the region with Lockhart-Martinelli number X equals or less than 0.02...” That is $X_{LM} \leq 0.02$.

API Type II Wet Gas:

“Type II wet gas is defined as the region above Type I ... [and] ... constrained to the liquid content limit by the following Lockhart-Martinelli relationship equal to or less than 0.3.” That is $0.02 < X_{LM} \leq 0.3$.

API Type III Wet Gas:

“All the regions above the boundaries defined by the Lockhart-Martinelli relationship of 0.3 can be designated as a ‘Multiphase Measurement System’ ... ”. That is $X_{LM} > 0.3$.

The API Type I definition appears to be based solely on single phase differential pressure (DP) meter design wet gas performance and is not general to all wet gas flows. It is accepted by some in industry that at $X_{LM} \leq 0.02$ most DP meters give small gas flowrate errors and a DP meter can therefore be used instead of a more sophisticated wet gas or multiphase meter. At $X_{LM} > 0.02$ the use of a wet gas meter becomes thought of by many as a more valid option. However, the Lockhart-Martinelli parameter value of 0.02 is not universally important to all gas flow meter types and hence it could be argued it is inappropriate to use this value to give a general non-meter specific definition of wet gas flow. Furthermore, it is debatable whether this $X_{LM} \leq 0.02$ limit for DP meters is a particularly precise value for when a wet gas meter shows advantage over a standard gas meter. Figure 3 shows a typical wet gas test result on a 4 inch, 0.4 beta Venturi meter, reproduced from Steven [13]. The solid points are the uncorrected wet gas flow rate predictions from the meter and the hollow points the de Leeuw [9] Venturi correlation corrected gas flowrate predictions for a known liquid flowrate. It can be seen that at a Lockhart-Martinelli parameter of 0.02 the over-reading can be in the order of 5%. It is arguable that a 5% gas flowrate error could be improved upon by some wet gas flow meter devices.

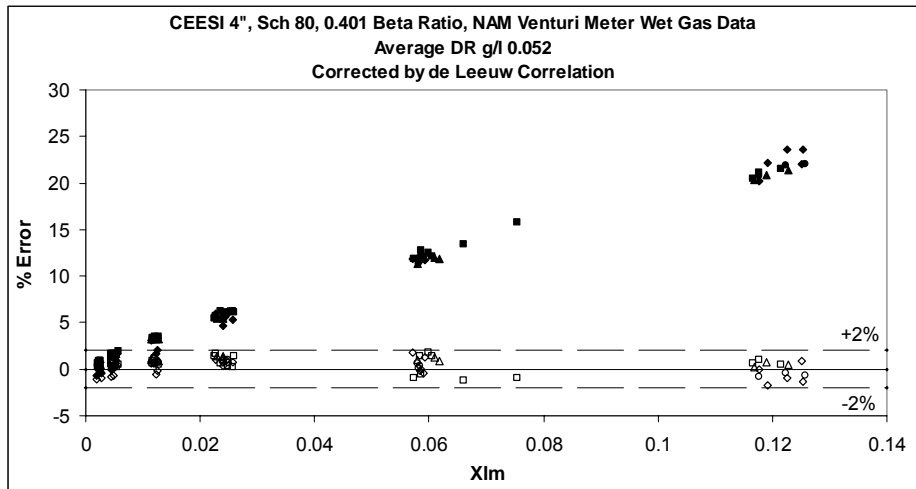


Figure 3. CEESI wet gas test data on a 4 inch 0.4 beta Venturi meter

The API Type II definition is bound at the lower end by the Type I definition which has the above issues.

The API Type III is ambiguous as there is no upper liquid content limit. According to this definition, an infinitely small quantity of gas in an infinitely large quantity of liquid is still technically a Type III wet gas flow. A diagram in the API report [11] tends to suggest the upper limit of Type III is perhaps a GVF of 80% but this is not precisely stated.

Therefore, the API gave no upper liquid content limit to wet gas flow. The Norwegian Society for Oil and Gas Measurement left the upper liquid quantity limit of wet gas flow deliberately vague by calling the border a Lockhart-Martinelli parameter of approximately 0.3. The border of the Lockhart-Martinelli parameter of 0.3 is rather arbitrary. In terms of fluid flow it has no special significance, with no fundamental physical condition change between a flow just above and just below this 0.3 value. Some wet gas and some multiphase flow metering technologies may have operating envelopes that straddle this arbitrary border. For example, one particular wet gas meter design could be claimed to operate in the range $0 \leq X_{LM} \leq 0.35$. Some multiphase meter technologies claim to operate at $GVF \leq 0.95$, which, depending on the flow conditions, could mean a minimum Lockhart-Martinelli parameter of less than 0.3.

To the authors of this paper, it seems far more important that operators using wet gas meters agree on the definition of terms to be used when describing flow ranges and mathematical equations than it is to give a precise scientific definition of the border between wet gas flow and general two phase / multiphase flow.

3.7. Equivalent wet gas parameters and conversions

A common requirement from the increasingly accepted wet gas definition of approximately $X_{LM} \leq 0.30$ is an ability to express this boundary in the units normally encountered in production situations.

The Lockhart-Martinelli parameter is typically asked to be expressed in terms of one of five common alternative parameters. These are:

- Gas volume fraction (GVF), the gas volume flowrate to the total volume flowrate as shown by equation (29a);
- Liquid volume fraction (LVF), the liquid volume flowrate to the total volume flowrate as shown by equation (29b);
- flow quality (sometimes termed the dryness fraction), and defined by equation (14);
- liquid to gas mass flowrate ratio, the ratio of the gas mass flowrate to liquid mass flowrate;
- liquid to gas volume ratio, the ratio of the gas volume flowrate to liquid volume flowrate.

GVF and LVF are defined by

$$GVF = \frac{Q_g}{Q_g + Q_l}; \quad (29a)$$

and

$$LVF = \frac{Q_l}{Q_g + Q_l}. \quad (29b)$$

Unfortunately, conversion isn't as straightforward as a direct equivalent numerical answer as these parameters are all related via the gas to liquid density ratio. The various parameters are related as shown in the following equation series:

$$X_{LM} = \frac{m_l}{m_g} \sqrt{\frac{\rho_g}{\rho_l}} = \frac{Q_l}{Q_g} \sqrt{\frac{\rho_l}{\rho_g}} = \frac{1 - (GVF)}{(GVF)} \sqrt{\frac{\rho_l}{\rho_g}} = \frac{1 - x}{x} \sqrt{\frac{\rho_g}{\rho_l}} = \frac{(LVF)}{1 - (LVF)} \sqrt{\frac{\rho_l}{\rho_g}}; \quad (30a)$$

$$GVF = \frac{1}{1 + \left(\frac{m_l}{m_g} * \frac{\rho_g}{\rho_l} \right)} = \frac{1}{\left(\frac{Q_l}{Q_g} \right) + 1} = 1 - (LVF) = \frac{\sqrt{\frac{\rho_l}{\rho_g}}}{X_{LM} + \sqrt{\frac{\rho_l}{\rho_g}}} = \frac{\left(\frac{\rho_l}{\rho_g} \right)}{\left(\frac{\rho_l}{\rho_g} \right) + \left(\frac{1 - x}{x} \right)}; \quad (30b)$$

$$x = \frac{1}{1 + \left(\frac{m_l}{m_g} \right)} = \frac{1}{1 + \left(\frac{\rho_l}{\rho_g} \frac{Q_l}{Q_g} \right)} = \frac{1}{1 + X_{LM} \sqrt{\frac{\rho_l}{\rho_g}}} = \frac{1}{1 + \left(\left(\frac{1 - (GVF)}{(GVF)} \right) * \frac{\rho_l}{\rho_g} \right)} = \frac{1}{1 + \left(\left(\frac{(LVF)}{1 - (LVF)} \right) * \frac{\rho_l}{\rho_g} \right)}; \quad (30c)$$

$$\frac{m_l}{m_g} = X_{LM} \sqrt{\frac{\rho_l}{\rho_g}} = \frac{\rho_l}{\rho_g} \frac{Q_l}{Q_g} = \frac{1-(GVF)}{GVF} \frac{\rho_l}{\rho_g} = \frac{LVF}{1-LVF} \frac{\rho_l}{\rho_g} = \frac{1-x}{x}; \quad (30d)$$

$$\frac{Q_l}{Q_g} = X_{LM} \sqrt{\frac{\rho_g}{\rho_l}} = \frac{\rho_g}{\rho_l} \frac{m_l}{m_g} = \frac{1-(GVF)}{GVF} = \frac{LVF}{1-(LVF)} = \left(\frac{1-x}{x}\right) \frac{\rho_g}{\rho_l}; \quad (30e)$$

$$LVF = 1-(GVF) = \frac{X_{LM} \sqrt{\frac{\rho_g}{\rho_l}}}{1 + X_{LM} \sqrt{\frac{\rho_g}{\rho_l}}} = \frac{\left(\frac{1-x}{x}\right) \frac{\rho_g}{\rho_l}}{\left(\frac{1-x}{x}\right) + \frac{\rho_l}{\rho_g}} = \frac{\frac{\rho_g}{\rho_l}}{\frac{\rho_g}{\rho_l} + \frac{m_g}{m_l}} = \frac{\frac{\dot{Q}_l}{\dot{Q}_g}}{1 + \frac{\dot{Q}_l}{\dot{Q}_g}}. \quad (30f)$$

It can be seen from equation (30a) that setting the upper limit of wet gas flow at a Lockhart-Martinelli parameter of 0.3 does not set a value for any of the alternative five parameters since they are related through the gas to liquid density ratio. It is a fair and understandable request from meter users and operating engineers to have the seemingly abstract concept of the Lockhart-Martinelli parameter expressed in units they can more readily understand, but an example shows how this is not necessarily possible. Taking the GVF as an example, a conventional wet gas definition is to say that wet gas flow has a GVF limit of 90%. Engineers get asked whether this $GVF > 90\%$ flow is approximately the same as a $X_{LM} \leq 0.30$ flow. From equation (30a), if $X_{LM} = 0.30$ and $GVF = 0.90$ then this would define the density ratio, ρ_g/ρ_l as 0.137174. This is clearly a meaningless statement. In fact, setting any two of the six parameters also then sets the gas to liquid density ratio value. (More details were supplied by one of the authors to ASME [12].) There is therefore no equivalent expression to $X_{LM} \leq 0.30$ without including a function of the gas to liquid density ratio.

Figure 4 shows a graphical form of equations (30). The abscissa is the gas to liquid density ratio and the ordinate is the Lockhart-Martinelli parameter. There are four coloured series of lines. These are black for the quality, blue for the GVF, orange for the liquid to gas mass flowrate and green for the liquid to gas volume flowrate. (Since $GVF = 1 - LVF$, the LVF lines overlap the GVF lines and so are not shown.)

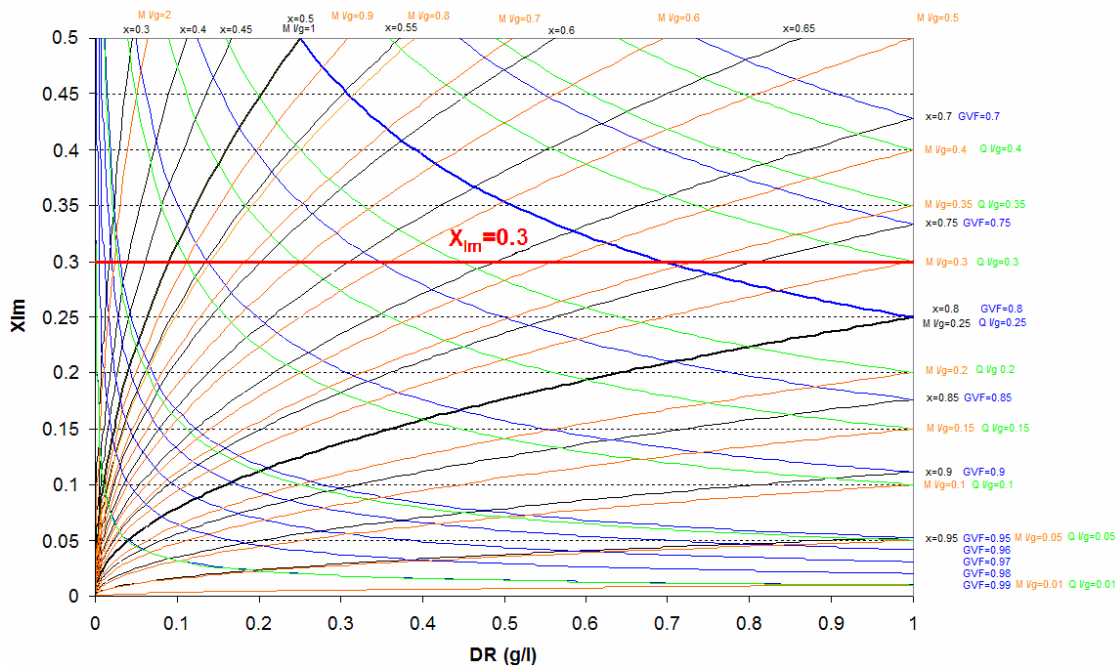


Figure 4. Sample wet gas / two phase flow parameter comparison graph

As an example take a gas to liquid density ratio of 0.2 and a quality of 0.8. Finding this point in Figure 4 immediately allows the other parameters to be read off for quick comparisons, that is:

$$X_{LM} = 0.11; GVF = 0.955 = 1 - LVF; LVF = 0.045;$$

$$m_l/m_g = 0.25 \text{ and } Q_l/Q_g = 0.05.$$

Such graphs are useful only when the flow conditions fit the graph. In general two phase flows the Lockhart-Martinelli parameter can be much greater than 0.5 and a larger scale is needed. Also in most real world flows the gas to liquid density ratio is less than 0.2 and a span of up to 0.2 gives more resolution. Therefore ASME [12] gives a series of such graphs and this figure is given as an example only.

4. The Difference Between the Gas Volume Fraction and the Gas Void Fraction

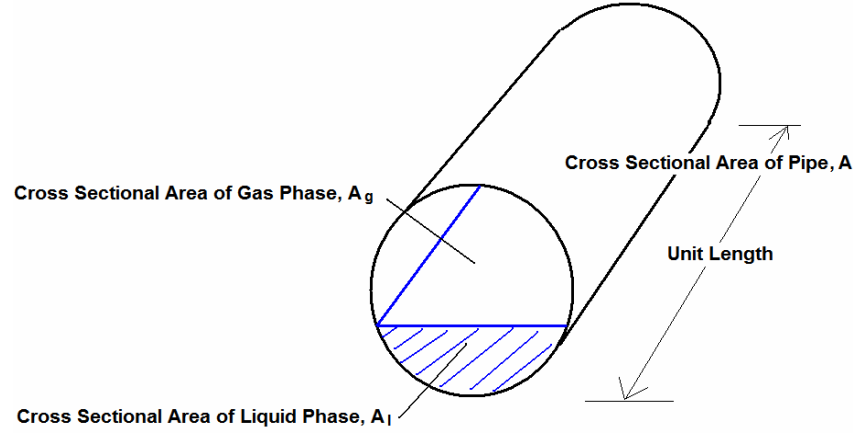


Figure 5. A sketch of stratified flow in a pipe

The *gas volume fraction* is defined as the ratio of the gas volume flowrate to the total volume flowrate, given by equation (31). The *liquid volume fraction* is defined as the ratio of the liquid volume flowrate to the total volume flowrate, given by equation (32). The name gas volume fraction has led to some confusion in the industry as this is sometimes mistakenly thought to be the actual ratio of the volume of the gas to the pipe volume along a unit section of pipe in two phase flow at any given instant in time. The *gas void fraction* (α_g) is defined as the ratio of the cross-sectional area occupied by the gas to the total cross-sectional area, at any given instant in time, and this is shown in equation (33),

$$GVF = \frac{Q_g}{Q_g + Q_l} ; \quad (31)$$

$$LVF = \frac{Q_l}{Q_g + Q_l} ; \quad (32)$$

$$\alpha_g = \frac{V_g}{V_g + V_l} = \frac{A_g L}{A_g L + A_l L} = \frac{A_g}{A_g + A_l} . \quad (33)$$

V_g is the gas volume in a unit volume of constant area pipe, V_l is the liquid volume in a unit volume of constant area pipe, A_g is the gas cross-sectional area, A_l is the liquid cross-sectional area and L is a unit length of constant area pipe, as shown in Figure 5.

The gas volume fraction is not the same parameter as the gas void fraction unless the average gas velocity and average liquid velocity are the same. The slip ratio denoted here by K , which is defined as the ratio of the average gas and liquid velocities, equation (34), must be unity, for the GVF and the gas void fraction parameters to be equivalent,

$$K = \bar{U}_g / \bar{U}_l . \quad (34)$$

This statement may be further explained by deriving an expression to relate the GVF to the cross-sectional area occupied by the gas and the liquid. Starting with equation (35),

$$Q = A\bar{U}, \quad (35)$$

where Q is the volume flowrate, A the area that the flow in question flows through and \bar{U} is the average velocity of the flow. Equation (36) can now be derived,

$$GVF = \frac{Q_g}{Q_g + Q_l} = \frac{A_g \bar{U}_g}{A_g \bar{U}_g + A_l \bar{U}_l} = \frac{A_g}{A_g + \left(\frac{\bar{U}_l}{\bar{U}_g}\right) A_l} = \frac{A_g}{A_g + \left(\frac{1}{K}\right) A_l}; \quad (36)$$

Whereas the void fraction is defined by equation (33) which is only equivalent to equation (36) with $K = 1$.

The expression *hold-up* can be defined as the ratio of the liquid to total cross sectional area and this is shown as equation (37). Hold up is therefore the same as the liquid to pipe volume ratio for a steady flow in a constant area pipe,

$$Liquid\ holdup = \alpha_l = \frac{A_l}{A_g + A_l}. \quad (37)$$

Clearly from equations (33) and (37),

$$\alpha_g + \alpha_l = 1. \quad (38)$$

In most two phase flows, the slip, while difficult to measure in practice, is known to often be considerable, so an approximation of unity is not valid in the vast majority of cases.

A numerical example helps show this point, continuing with the worked example started in section 3.3. Chisholm created a slip model [6] which stated that for stratified flow the slip of a wet gas was

$$K = \frac{\bar{U}_g}{\bar{U}_l} = \left(\frac{\rho_l}{\rho_g}\right)^{1/4}. \quad (39)$$

There are many slip models in the literature and they should only be applied according to the limits of the models assumptions and the suitability of the model to the application in question. These authors make no claim on the validity of any particular slip model, and Chisholm's slip model used here only as an example. In the example, the gas density is 38 kg/m³ and the liquid density is 800 kg/m³, so the slip according to Chisholm's expression is

$$K = \frac{\bar{U}_g}{\bar{U}_l} = \left(\frac{800}{38}\right)^{1/4} = 2.14. \quad (40)$$

The meaning of this value is that the average gas velocity is 2.14 times that of the average liquid velocity. From equations (33), (36) and (39),

$$\alpha_g = \frac{1}{1 + \left\{ K \left(\frac{1}{GVF} - 1 \right) \right\}}. \quad (41)$$

Using the example values of 400 m³/hr for the gas flowrate and 8.72 m³/hr for the liquid flowrate, the GVF and LVF are

$$GVF = \frac{Q_g}{Q_g + Q_l} = 0.979; \quad (42a)$$

$$LVF = 1 - GVF = \frac{Q_l}{Q_g + Q_l} = 0.021; \quad (42b)$$

and the void fraction is

$$\alpha_g = \frac{1}{1 + \left\{ K \left(\frac{1}{GVF} - 1 \right) \right\}} = \frac{1}{1 + \left\{ 2.14 * \left(\frac{1}{0.979} - 1 \right) \right\}} = 0.956. \quad (43)$$

Hence the hold up is

$$\alpha_l = 1 - 0.956 = 0.044. \quad (44)$$

The Armand coefficient, C_A is a very rarely used parameter that indicates the difference between GVF and gas void fraction. It was quoted by Chisholm in his derivation of equation (25). This is shown in equation (45) along with the value in this example,

$$C_A = \frac{\alpha_g}{GVF} = \frac{0.956}{0.979} = 0.977. \quad (45)$$

The Armand coefficient would take a value of 1 where there is no slip, i.e. $K=1$.

So it should be understood that the GVF and the gas void fraction are not the same parameter. In this particular example a small difference in GVF and gas void fraction is noticeable but it is important to note that other flow conditions and other slip models can have greater differences between the parameter values. Furthermore, sometimes engineers are more interested in the liquid flowrate of a wet gas flow. In these cases the LVF is often found by equation (42b). If the GVF in this example has been erroneously described as the gas void fraction the difference between a GVF of 0.979 and gas void fraction of 0.956 is a small difference of -2.4%. However, if the GVF was taken as 0.955 when calculating the LVF, this would give a value of 0.044 instead of the actual 0.021, i.e. a difference of 110%!

The relevance of this discussion regarding the difference between GVF and gas void fraction is that it is generally the gas void fraction, or the liquid holdup that is the parameter measured. This is because the measuring device will see the fluids as they

flow in the pipe, unless somehow the gas and liquid have been homogenised to remove all slip. The GVF on the other hand is typically derived from other measured parameters. Either the slip, in addition to the holdup, must also be directly measured, or it must be modelled. If the GVF could be independently measured in some way the value would be available for input into equation (30a) and hence for known gas and liquid densities the Lockhart-Martinelli parameter would be known. This in turn would allow the use of any single phase meter wet gas correlation based on the Lockhart-Martinelli parameter to predict the gas and liquid flowrates.

5. Summary and recommendations

Wet gas can be considered a subset of two phase flow, and is a condition described as a relatively small amount of liquid (of any composition) in a flow that is predominantly gas by volume. The upper boundary of this condition can be numerically described in dimensionless terms as having a *Lockhart-Martinelli parameter* of approximately 0.30.

The Lockhart-Martinelli parameter has been defined differently by several frequently-cited authors in the field over the last 6 decades, and by subsequent interpretation of those authors' works. The present authors recommend the definition for the Lockhart-Martinelli parameter of the correlating parameter defined by Chisholm, which is independent of pipe roughness and friction factor, and independent of any type of differential pressure meter. This quantity should be denoted by X_{LM} and is defined by

$$X_{LM} = \frac{m_l}{m_g} \sqrt{\frac{\rho_g}{\rho_l}}. \quad (20)$$

Finally, attention has been drawn to the difference between the quantities *gas void fraction*, which represents the actual cross-sectional distribution of the gas and liquid phases within the pipe, and *gas volume fraction* which is a volumetric ratio of the gas and liquid rates.

Nomenclature

α	Holdup (-)
ΔP	Pressure drop (Pa)
ε	Expansibility factor (-)
μ	Dynamic viscosity (Pa.s)
ρ	Density (kg/m ³)
A	Cross-sectional area (m ²)
A_t	Differential pressure meter restriction area (m ²)
C_A	Armand coefficient (-)
C_d	Discharge coefficient (-)

D	Pipe diameter (m)
e	Pipe roughness (m)
E	Velocity of approach (-)
f	Friction factor (-)
Fr	Froude number (-)
GVF	Gas volume fraction (-), commonly expressed as a percentage
K	Slip ratio (-)
L	Unit length of pipeline (m)
LVF	Liquid volume fraction (-), commonly expressed as a percentage
m	Mass flowrate (kg/s)
P ₁	Upstream pressure (Pa)
P ₂	Downstream pressure (Pa)
Q	Volumetric flowrate (m ³ /s)
Re	Reynolds number (-)
U	Velocity (m/s)
x	Quality (-)
X	Dimensionless parameter defined by Lockhart & Martinelli (-)
X _{Chisholm}	Dimensionless parameter defined by Chisholm (-)
X _{LM}	Alternative definition of Chisholm dimensionless parameter (-)
X _M	Roughness-modified version of dimensionless parameter defined by Lockhart & Martinelli (-)
X _{Murdock}	Dimensionless parameter defined by Murdock (-)

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