Use of Dimensional Analysis for Evaluation of Corrosion Inhibitor Effectiveness in Wet Gas Pipelines

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Abstract

Effective corrosion inhibitor treatment in an upstream environment in oil and gas production requires a quite detailed picture of the detailed physical processes present in the pipeline. This is particularly challenging in wet-gases, where most multiphase pipeline models have proven inadequate, and even those models that can be successfully applied to wet-gas pipelines do not produce all the output required for design and implementation of an effective corrosion inhibitor protocol. The Froude number, a dimensionless number generally under-utilized in multiphase pipeline design, has proven to be a very powerful qualitative and even quantitative tool for the evaluation of liquid-liquid mixing, droplet production, and drop-drop and drop-wall interactions. As such, it is the ideal correlating parameter to predict and quantify corrosion and corrosion inhibitor effectiveness.

Introduction

Wet gas systems, or gas-condensates, arise when gas produced from a gas reservoir drops through the retrograde condensation region of the phase envelope as the pressure drops through the well tubing. This is the opposite of what occurs in an oil system, where *gas* comes out of solution as the pressure drops through the well and production tubing. The reason for the differing behavior is the fact that gas systems always lie to the right of the system critical point (i.e., in the retrograde condensation region), whereas oil systems always fall to the left.

Multiphase flow can be categorized as either low-liquid-loading (wet gas) systems, or high-liquid-loading (oil) systems. The demarcation point between the two could be considered as a GOR of 1000 scf/bbl, or alternatively 1000 bbl/mmscf. Gas condensates generally have a small amount of condensing liquid - anywhere from 1-100 bbl/mmscf, giving a volumetric gas to liquid *in situ* volumetric flow rate ratio of typically around 100-200 to 1.

Wet gas systems exhibit many qualities not seen in oil systems:

- While the liquid hold-up over most of the operating range is quite high, as the flow rate decreases below a critical value, the liquid volume fraction in wet gas systems can go from 1-2% to 50-60% full of liquid over a very small rate range.
- Liquid volume fraction is very strongly dependent on pipeline angle – angles changes as small as 0.1 degree can change the liquid hold-up by an order of magnitude.
- At high rates, wet gas systems can carry nearly all of the liquid in the pipeline as droplets (measurements indicate as much as 98% of the liquid is carried as droplets at high rates).

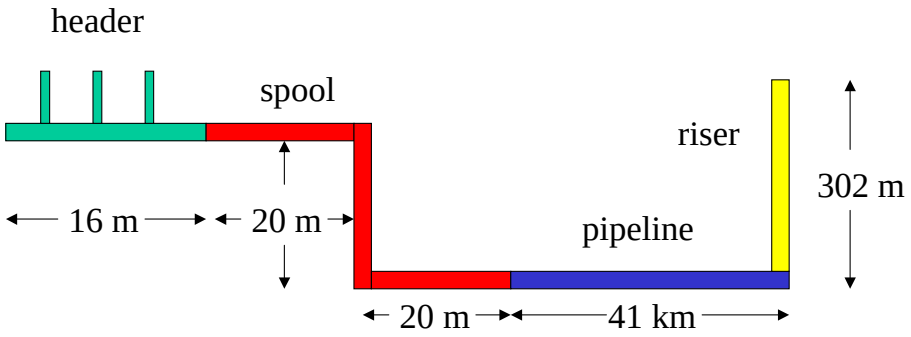
Thus, prediction of the critical rate where the liquid pipeline inventory starts to increase markedly at low production rates, and prediction of the droplet fraction at high production rates are absolutely critical to effective modeling of wet gas pipelines.

Wet gas systems often operate in the stratified or stratified-mist flow regime, with liquid moving primarily at the bottom of the line, and gas moving primarily at the top of the line. In order to achieve effective corrosion inhibitor coverage, there must be adequate mixing of the inhibitor with the aqueous phase, and sufficient droplet production to get corrosion inhibitor chemical to the top of the line. Further, there must be good organic-aqueous liquid mixing at the bottom of the pipe over the range of flow rates encountered in production to ensure continued replenishment of the inhibitor film. Both droplet creation and aqueous-organic liquid mixing are a strong function of the Froude number.

In this paper, a wet-gas multiphase production system is analyzed using the OLGA2000 pipeline simulator, which gives reasonable results across all liquid loadings, from oil to wet gas. Additional analysis performed on the output of the OLGA program in order to answer questions specific to corrosion prediction and inhibition. The focus of this paper is on organic/aqueous mixing, droplet production/interaction, and mixing of the corrosion inhibitor with the bulk liquid.

Wet Gas System

The schematic below gives the configuration of the header, spool, pipeline, and riser for the wet gas system. The header is 16 m long, and the heating medium spool is 60 m long. The pipeline is 41 km, and the riser to the platform is 302 m. The production header is constructed of a corrosion resistant alloy, and the corrosion inhibitor is injected into the header.



Corrosion inhibitor is often added to a subsea tie-back at the subsea production header. Should corrosion inhibition be required, the first question to consider is whether there is sufficient mixing of inhibitor by the end of the header than the spool can be constructed

of carbon steel, rather than a corrosion resistant alloy (CRA). If it is found that the mixing in the header is insufficient, then the spool will also be constructed of CRA.

Table 1 gives the pipeline information for the wet gas system evaluated.

Pipeline Service	Produced Gas
Pipeline OD	12.75 inches (323.9 mm)
Pipeline Wall Thickness	0.69 inches (17.5 mm)
Pipeline length	40.5 km
Design Pressure	3765 psig (259 barg)
Pipe Grade	AP 5L X65
Design Gas Production Rate	200 mmscfd
Design Liquid Production Rate	40 bbl/mmscf
Design Produced Water + MEG Rate	5 bbl/mmscf

Table 1 – Wet gas pipeline information

The design condensate loading is 35 bbl/mmscf There is 2.5 bbl/mmscf water production (condensing water), with an additional 2.5 bbl/mmscf of glycol added for hydrate control, for a total aqueous loading of roughly 5 bbl/mmscf.

The pipeline profile used in this study is given in Figure 1. The subsea center is at the left side of the plot (at 0 meters on the x-axis), and the platform is at the right end of the plot (at 40.5 km).

A range of flow rates for this wet gas system, from 20 to 200 mmscfd (full rate), were investigated using the OLGA2000 multiphase pipeline simulator. OLGA runs were done, and the results of those runs were evaluated to answer corrosion-specific questions regarding: bulk mixing of the aqueous and organic phases; droplet-droplet mixing, and; water condensation and top of the line corrosion.

Figure 2 gives the hold-up through the pipeline at 20, 100, and 200 mmscfd. Note that at lower flow rates, the liquid in the pipeline begins to accumulate in the uphill sections of the pipeline. This can clearly be seen by comparing Figure 1 to Figure 2. Liquid accumulation becomes most severe when the Froude number for the flow is less than 1. At this point, gravity forces begin to overwhelm frictional forces, with the end result being higher accumulations of liquid, lower liquid velocities, and potential for solids deposits and under-deposit corrosion.

Definition of the Froude Number for Pipe Flow

The Froude number is a dimensionless group that it thought to be important whenever there is a fluid flow with a free interface. As this is certainly the case for multiphase pipe flow, we could reasonably expect the Froude number to be a good correlating parameter for a number of multiphase phenomena, including corrosion rates and corrosion inhibitor effectiveness.

First, we must define several common terms used in multiphase pipeline flow. The liquid hold-up is defined as:

$$H_L = A_L/A$$

where

A_L = cross-sectional area of pipeline occupied by liquid (m^2);
A = total cross-sectional area of pipeline (m^2).

The superficial gas and liquid velocities are defined as

$$U_{SG} = Q_G/A$$
$$U_{SL} = Q_L/A$$

where

Q_G = volumetric flow rate of gas (m^3/s);
Q_L = volumetric flow rate of liquid (m^3/s).

Thus, U_{SG} and U_{SL} are volumetric fluxes of gas and liquid respectively. The actual gas and liquid velocities can be derived from their respective superficial velocities and the liquid hold-up,

$$U_L = U_{SL}/H_L$$
$$U_G = U_{SG}/(1-H_L)$$

The Froude number is a measure of the ratio of the kinetic to the potential energy of the liquid layer in the pipe, or

$$Fr = ((\rho U^2/2) / (\rho g L))^{1/2} = U/(gL)^{1/2}$$

where,

U = characteristic velocity, (m/s);
g = acceleration due to gravity, (9.8 m/s^2);
L = characteristic length, (m);
ρ = fluid density, (kg/m^3).

For our purposes,

$$Fr = U_L/\mathrm{sqrt}(gh)$$

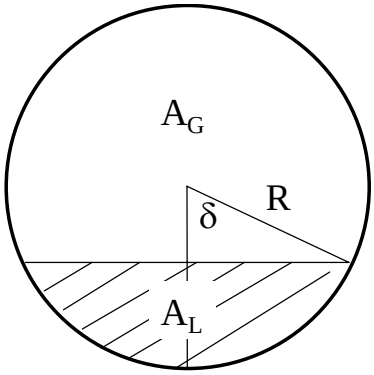
where,

U_L = liquid velocity, (m/s);
h = average liquid film height, (m).

The average height of the liquid film h can be estimated from the following:

$$\delta = \pi/2(1 - (1 - H_L)^{1/3} + H_L^{1/3})$$
$$h = \pi D H_L/(4 \sin(\delta))$$

where the schematic below explains how the angle δ is defined.



$H_L = A_L/A$ $\qquad$ $h = \pi D H_L/(4\sin(\delta))$

Froude number Evaluation of a Wet Gas Pipeline

Uninhibited water can be corrosive in amounts as small 1-2% water cut; without inhibition, corrosion could occur over the entire length of the pipeline. Since the water is present in amounts about 10-20% water cut in this particular wet gas production case, corrosion inhibition will be required.

Historically-used rules-of-thumb in the oil industry are that one would like a gas velocity of 5 m/s or better, or a liquid velocity of 1 m/s or better for good operation of a wet gas pipeline. These conditions are desirable for a number of reasons:

- It assures relatively complete mixing between the organic and aqueous phases;
- It keeps any small particulates (scale or corrosion products or production fines) entrained and moving;
- It assures reasonable production of droplets and effective inhibition of the top of the line.

The rules of thumb above correspond quite nicely with the Froude number, as we will demonstrate below. The Froude number can be used not only to predict droplet formation, but also to provide guidance as to the extent of mixing between the oil and

water phases. Also, the Froude number can provide quantitative as well as qualitative guidance for a corrosion inhibition program, as this section will demonstrate.

Consider Table 2, which gives the wet gas rate, liquid hold-up, gas velocity, liquid velocity, Froude number, and droplet fraction – as predicted by OLGA2000. First, note that the hold-up in the pipeline is almost independent of flow rate. This is typical of wet-gas pipelines; for Froude numbers over 1, the liquid hold-up is quite insensitive to flow rate. For Froude numbers below 1, however, there is a large increase in liquid hold-up, resulting in stagnant water "pooling" in low spots, dropout of solids, and under-deposit corrosion.

Flow rate (mmscfd)	Hold-up (-)	UG (m/s)	UL (m/s)	Froude (-)	Drop Fraction (-)
20	0.048	2.5	0.41	0.92	0.01
30	0.048	3.6	0.61	1.35	0.03
40	0.047	4.2	0.80	1.68	0.07
50	0.046	5.2	1.00	2.12	0.11
60	0.044	5.7	1.18	2.56	0.14
70	0.043	6.2	1.43	3.19	0.20
80	0.041	6.5	1.67	3.72	0.25
90	0.040	6.7	1.97	4.43	0.31
100	0.041	7.2	2.18	6.00	0.40
200	0.030	7.4	4.93	16.00	0.84

Table 2 – OLGA2000 results for a wet gas pipeline

For Fr < 1 (see, for example, White[1]), the gas-liquid interface is stratified smooth, with complete stratification of the oil and water layers, and potential for stagnant water build-up in the low portions of the line. This corresponds in the wet-gas case above to a liquid velocity of about 0.4 m/s, which occurs at 20 mmscfd. A Froude number below 1 also indicates that the pipeline has a potential for terrain slugging (i.e., periodic accretion and expulsion of slugs of liquid, which is very undesirable from a production standpoint), which could definitely be a concern for any flow rate at or below 30 mmscfd.

Often, in situations where Fr < 1, the multiphase flow is stratified in downward sections, and slug in upward sections, that is, slugs are created at all dips in the pipeline. Based on ConocoPhillips' experience, this is the worst possible environment for corrosion – particularly when there are solids present. The process of slug creation is very energetic, and any solids present in the pipe scour away the protective corrosion inhibitor film, leading to high rates of localized corrosion.[2] An examination of the OLGA output indicates that this is precisely what is happening in the 20 mmscfd case. Thus, the Froude number can be a very effective indicator of potential corrosion problems, even with the addition of corrosion inhibitor chemical.

For 1 < Fr < 2, the gas-liquid interface is in the stratified-wavy regime. Here we begin to have mixing between the oil and water phases, but there is still significant stratification pf the oil and water, and droplet production is minimal. For this case, at 30 mmscfd, the liquid velocity is 0.6 m/s, the gas velocity is 3.6 m/s, and the Froude number is about 1.4.

For Fr = 2, we begin to have significant droplet formation (according to the OLGA model), and good mixing of the oil-water layers. This corresponds in Brodgar to a gas velocity of 5.2 m/s, a liquid velocity of about 1 m/s, which occurs at 50 mmscfd. Note that a gas velocity of 5 m/s, a liquid velocity of 1 m/s, and a Froude number of 2 all occur at very close to the same point – supporting the traditional "rules of thumb" in pipeline design.

The Froude number is, in fact, preferable to the other two rules of thumb; due to the fact that it is a dimensionless group, it should have much better scaling characteristics to larger pipeline diameters. It is ConocoPhillips practice to design pipelines with – at a minimum – Froude number of 2 or better. If this is not possible, then a program of routine pigging will be required for effective corrosion inhibition.

Operations at 100-200 mmscfd exhibit quite high Froude numbers (6.0 and 16.0 respectively) and droplet fractions (40% and 84% respectively), which will result in very turbulent and well-mixed liquid layers, and quite high droplet fractions. Both of these effects will act to inhibit both the top and bottom of the pipeline quite effectively.

Figure 3 gives the OLGA2000-predicted droplet fraction plotted as a function of Froude number. Note that there is almost a linear relationship between the Froude number and the percentage of liquid being carried as droplets, indicating that the Froude number could be a very effective dimensionless group for quantifying droplet production and corrosion inhibitor effectiveness.

There is very little droplet fraction data, but one fairly good source of such data in the open literature comes from the work of Tayebi[3], reproduced here in Table 3. Tayebi used Exxol D80 (820 kg/m^3 density, $1.7x10^{-3}$ kg/m-s viscosity) as the liquid and SF_6 (45.8 kg/m^3 density, $1.5x10^{-5}$ kg/m-s viscosity) as the gas. SF_6 has the advantage that, because of its high molecular weight (146 gm/gm-mole, or roughly 5 times the molecular weight of air), very high gas densities can be achieved at moderate pressures.

Figure 4 gives the Tayebi droplet fraction data plotted against the experimental Froude number. Note, again, the linear relationship between the Tayebi data and the Froude number. Also included in Figure 4 is a simple model for droplet entrainment fraction, based on the Tayebi data. The model works as follows:

droplet fraction = Minimum(1, Maximum((Fr-2)/10, 0))

This simplified droplet model has been applied to this pipeline for various flow rates, and a comparison of this model to the OLGA2000-predicted droplet fractions at 100 mmscfd is given in Figure 5. Note that there is reasonable agreement between the OLGA2000 droplet fraction prediction and those of the simple model above.

USG (m/s)	USL (m/s)	Gas density (kg/m3)	Liquid Density (kg/m3)	Hold-up (-)	Film thickness (m)	Drop percentage (%)
4.33	0.25	23.4	824	0.237	0.0287	2.789
5.38	0.25	23.4	824	0.192	0.0247	5.071
6.89	0.25	23.4	824	0.157	0.0214	8.914
7.03	0.26	24.1	824	0.158	0.0215	9.851
3.35	0.26	32.5	824	0.259	0.0306	1.651
4.51	0.26	32.5	824	0.200	0.0254	4.246
5.53	0.26	32.5	824	0.163	0.022	9.105
7.00	0.26	32.5	824	0.126	0.0184	13.025
4.50	0.26	46.5	824	0.183	0.0239	6.727
5.50	0.26	46.5	824	0.151	0.0208	12.369
7.00	0.26	46.5	824	0.109	0.0168	22.966

Table 3 – Experimental results for droplet fraction, taken from Tayebi.

Droplet Size Calculation using the Weber Number

Droplet size in the production header and pipeline is governed by the Weber number, given by

$$We = \rho_g U^2 d/\sigma$$

where

ρ_g = gas density, (kg/m^3);
U = relative velocity of the continuous phase (basically the gas velocity), (m/s);
d = droplet diameter, (m);
σ = surface tension (kg/s^2).

Droplet breakup occurs at a critical Weber number, denoted We_c. The critical, or largest, droplet diameter possible is found by inverting the Weber number equation to give,

$$d_c = We_c\sigma/\rho_g U^2$$

For small viscosity liquids, as we have here, the critical Weber number is around 13.[4] Using this equation, the drop sizes predicted are:

- 20 mmscfd: 1200 microns
- 100 mmscfd: 114 microns
- 200 mmscfd: 13 microns

The 1200-micron droplets size predicted at 20 mmscfd would clearly not be able to be suspended by the gas, and therefore no droplets are expected for this flow rate. This corresponds well with the prediction of no droplets present based on the Froude number criterion outlined in the previous section.

Drop-Drop and Drop-Wall Collisions

Drop-drop and drop-wall collisions both need to be considered as a mode of mixing. The governing equation for turbulent collisions of small droplets is given by the equation[5]

$$L_{ij} = (54/5\pi^3)^{1/2}\theta_i\theta_j((d_i+d_j)3/(d_id_j)^3(\varepsilon/\nu)^{1/2}$$

Where

$\theta_{i,j}$ = volume fraction of droplets i, j (-);
$d_{i,j}$ = diameter of droplets i, j (m).
ε = turbulence dissipation rate (m^2/s^3);
ν = kinematic viscosity, (m^2/s).

Here, the $(\varepsilon/\nu)^{1/2}$ term gives the turbulence shear rate in the flow.

The turbulence quantities can be obtained from the following set of equations,

I = turbulence intensity = $0.16/ Re^{1/8}$
k = turbulent kinetic energy per unit mass = $3/2 (U_g I)^2$
L = largest turbulent length scale = 0.07D
ε = turbulent dissipation rate = $0.16k^{3/2} / L$

In order to obtain the drop-drop mass transfer rate, the collision rate is multiplied by the mass of a typical droplet. This gives a mass transfer rate of

$$M_{ij} = L_{ij}\ \rho_L\pi d_i^3/6$$

Where

M_{ij} = drop-drop mass transfer rate, (kg/m^3-s).

The results of the drop-drop and drop-wall calculations are given in Table 4 (the drop-wall collision rates have obtained from OLGA, where they are available as an output variable). As can be seen from the table, the drop-drop collisions are not an important factor for mixing, and the vast majority of mixing of the corrosion inhibitor with the process stream occurs at the wall, in the bulk liquid stream.

Flow rate (mmscfd)	UG (m/s)	UL (m/s)	Θbulk (-)	Θdroplet (-)	Regime	Drop-Drop (kg/m3-s)	Drop-Wall (kg/m3-s)
200	8.0	2.84	0.0360	0.0470	ST/MIST	0.0020	14.60
100	7.80	1.29	0.0092	0.0020	ST/MIST	0.0002	0.73
20	3.16	0.42	0.0260	0.0001	ST/WAVY	0.0000	0.02

Table 4 – Calculation of drop-drop and drop-wall collisions

Mixing Efficiency

The corrosion inhibitor is injected into the subsea production header, which is constructed out of corrosion resistant material. A critical question that must be answered is whether the corrosion inhibitor chemical is thoroughly mixed before the multiphase mixture reaches the carbon-steel heating medium spool and pipeline.

A mixing efficiency for corrosion inhibitor in the production header can be established as follows:

- The mass exchange rate given in OLGA between the wall and the droplets (the most important mode of mixing) has the units of mass/time/volume;
- Multiply the volume of the pipe as a function of distance through the header/spool, to obtain a mass rate, in units of mass/time;
- The total droplet mass in this same volume can be obtained by (liquid density in drops)*(drop volume fraction)*(volume);
- The total droplet mass divided by the mass exchange rate with the wall gives a characteristic time of mixing;
- This characteristic time can then compare this with the residence time in that same volume, which is given by the length divided by the gas velocity;
- The residence time divided by the mixing time gives the number of characteristic mixing times we spend in the volume;
- Based on this, a mixing parameter can be established.

Consider a volume filled with N droplets, colliding randomly with the wall at a rate of n collisions per unit time. During one unit time, there have been n collisions; however, half of these droplets have already collided with the wall once before, and therefore do not advance mixing. Likewise, half of the half that has collided twice has collided three times, and so on,

$$\text{Total Collisions } n = \text{droplets colliding once} + \text{droplets colliding twice} + \ \ldots$$

or

$$\text{Total Collisions } n = n(1/2 + 1/4 \ + 1/8 + ,\ldots)$$

Thus, half of the total number of collisions are droplets colliding with the wall for the first time. Therefore, the mixing half-life of any population of droplets is

$$T_{\text{half-life}} = N/n = M/m$$

where M is the mass of droplets per unit volume and m is the drop-wall mass flux per unit volume. The number of ‘half lives’ experienced by a population of droplets is given by

$$n_{half\text{-}life} = T_{residence} / T_{half\text{-}life}$$

A mixing parameter is can be defined as

$$\beta = 1 - (1/2)^{n\text{-}half\text{-}life}$$

A value of $\beta = 0$ indicates no mixing, and β approaches 1 for complete mixing.

Figure 7 gives the critical droplet size, droplet fraction, and mixing parameter (evaluated at the start of the spool) as a function of wet gas rate for the wet gas case under consideration here. Note that the droplets are predicted to be very large for 20 mmscfd. No droplet this large could be suspended in the gas phase – therefore, this indicates that there are no droplets to speak of at 20 mmscfd, and the corrosion inhibitor drains into the bulk liquid pool, and mixing is not a concern. On the other hand, the value of the mixing parameter for the higher rate flows (100 and 200 mmscfd) clearly indicate that there is incomplete mixing in the production header prior to entry into the heating medium spool.

Figure 8 gives the value of the mixing parameter as a function of distance from the last well inlet into the production header. Note that for 200 mmscfd, it takes 30 meters past the last well inlet into the production header – well into the pipeline spool - to achieve a better than 80% mixing of corrosion inhibitor with the bulk produced fluid. For this reason, a decision was made to construct the production heating medium spool out of a corrosion-resistant alloy.

Relationship between Droplet Fraction and Pressure Gradient

Droplet production is directly related to the pressure gradient in multiphase flow. If the droplet fraction is over-predicted, then the pressure drop will also be over-predicted, and vise versa. In fact, a problem in an earlier commercial version of OLGA was that the droplet fraction was greatly over-predicted (predicting 98% droplets at very low rates, for example), which resulted in very high pressure drop predictions (as much as 40-50% higher than field measurements). The droplet entrainment fraction was reworked in OLGA2000 (the current commercial version), but still exhibits inadequacies.

The single-phase Reynolds number is defined as

$$Re = \rho U D / \mu$$

Where

ρ = fluid density
U = fluid velocity
D = pipe diameter
μ = fluid viscosity

The pressure drop in a single-phase line is given by the following relation

$$\delta P/\delta z = f \rho U^2 / D / 2$$

where f is the Darcy friction factor, which can be modeled (smooth wall pipe) with[1]:

$$f = 1.02 / \log(Re)^{2.5}$$

By analogy, a multiphase Reynolds number can be defined as follows:

$$Re_M = \rho_M U_M D / \mu_M$$

Where

ρ_M = mixture density = $\rho_L H_L + \rho_G (1 - H_L)$
μ_M = mixture viscosity = $\mu_L^{HL} \mu_G^{(1-HL)}$
U_M = mixture velocity = $U_{SL} + U_{SG}$
D = pipe diameter
ρ_L = liquid density
ρ_G = gas density
μ_L = liquid viscosity
μ_G = gas viscosity
H_L = liquid hold-up

A simple multiphase friction factor can be proposed as a direct analogy to single-phase flow, as follows:

$$f_{simple} = 1.02/\log(Re_M)^{2.5}$$
$$(\delta P/\delta z)_{simple} = f_{simple} \rho_M U_M^2 / D / 2$$

The simple model prediction has been compared to actual experimental data taken at the SINTEF Tiller flow loop, in Trondheim Norway. The Tiller loop is a 800 meter long, 0.159 m ID loop with a 55 meter riser section. Fluids used are naphtha, diesel, and lube oil, with nitrogen as the gas phase.

When the two frictional pressure gradients were compared, it was found that the simple model was as much as 20% high to as much as 50% low compared to the actual, experimentally-determined frictional pressure gradient. However, when the ratio of the two friction factors are plotted vs. the Froude number, there is a very strong trend in the data, as can be seen from Figure 9.

Note that for Fr < 2, the pressure drop of the simple model is higher than that for the experiments. As Fr = 2, the two are equal. As the Froude number increases, the difference between the two increases, until it reaches a maximum at around Fr = 10-12. This data trend clearly shows a regularity that implies a physical process of some kind. It

seems plausible that the reason that ratio of the experimental pressure gradient over the simple model pressure gradient mirrors the production of droplets in the flow.

In fact, this view is supported by recent data taken by ConocoPhillips, using an optical test section installed in the Tiller 8-inch flow loop.[6]

Conclusions

The Froude number has been shown to be a very powerful method for evaluating the performance of corrosion inhibitor chemical. Because it is dimensionless, the Froude number enjoys several advantages over other methods based upon either the gas or liquid velocity. The Froude number can give a great deal of guidance over and above the information generally available from multiphase flow simulators.

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7. Figures

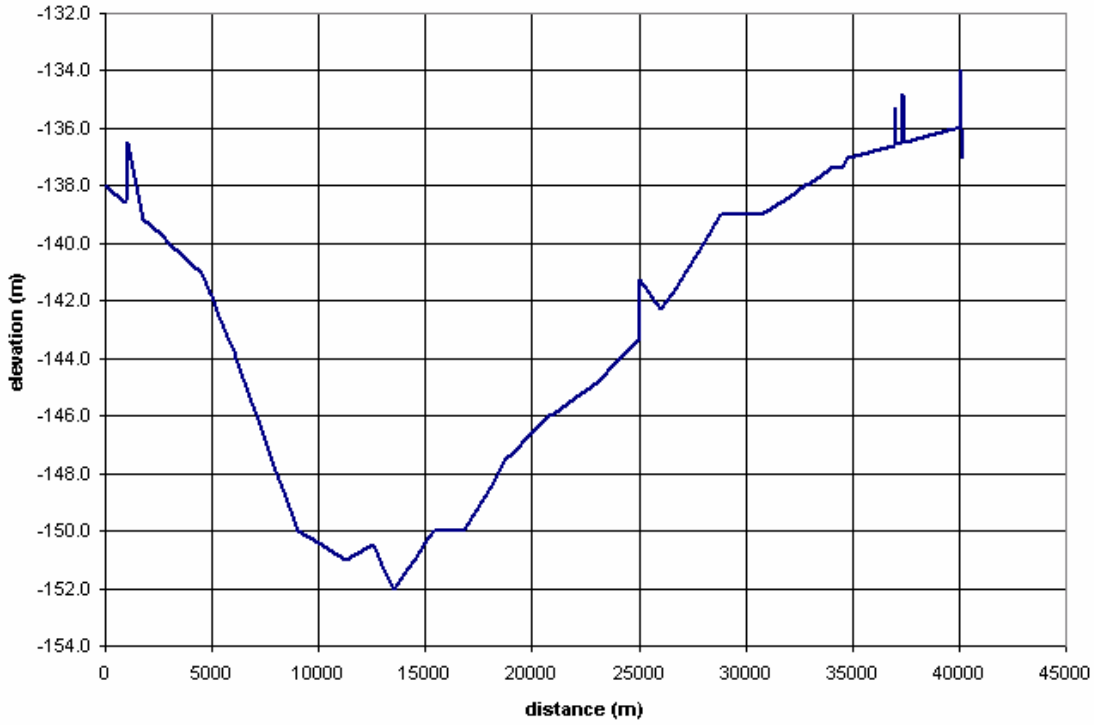

Figure 1

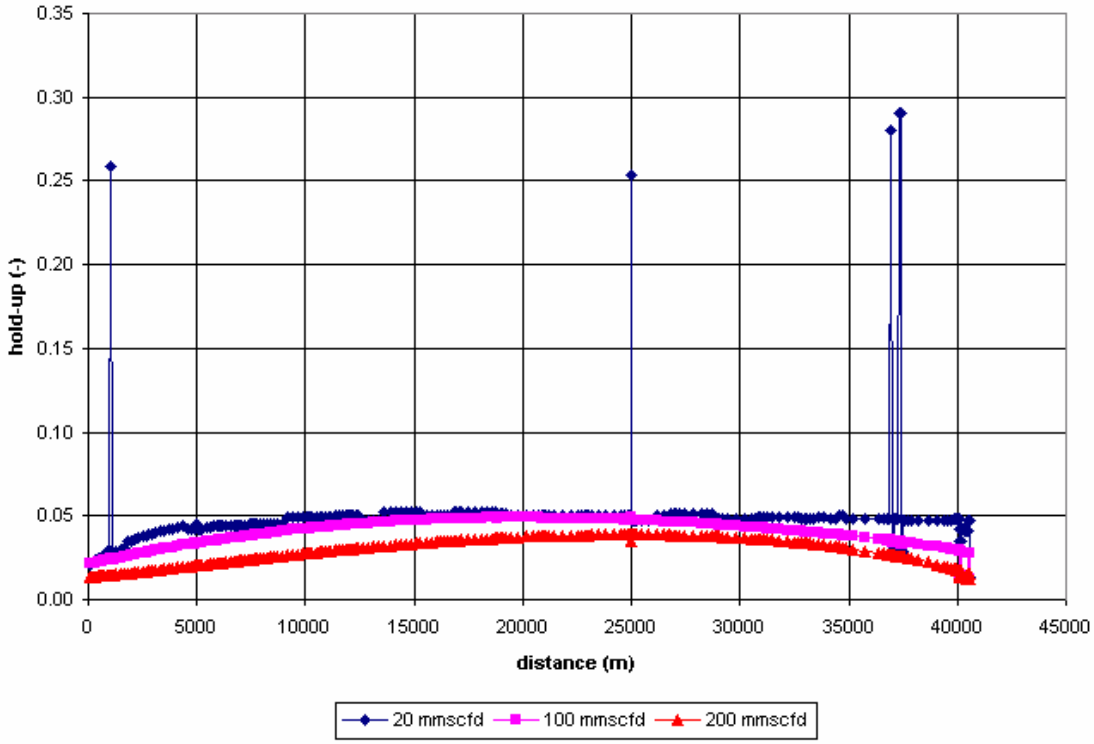

Figure 2

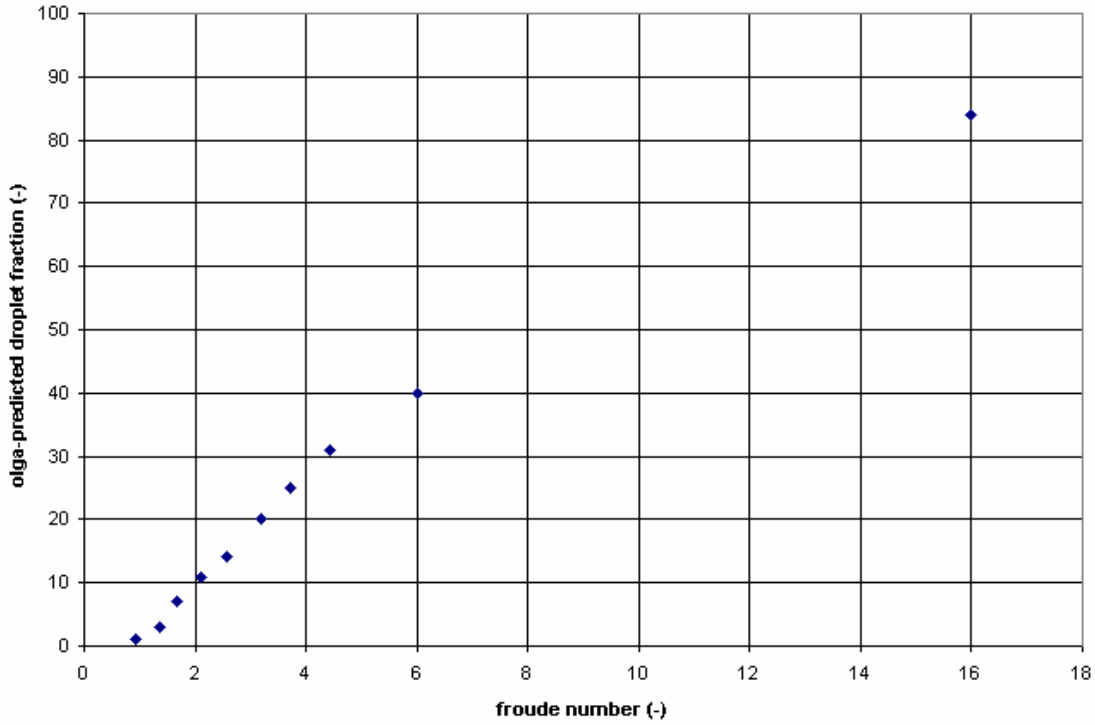

Figure 3

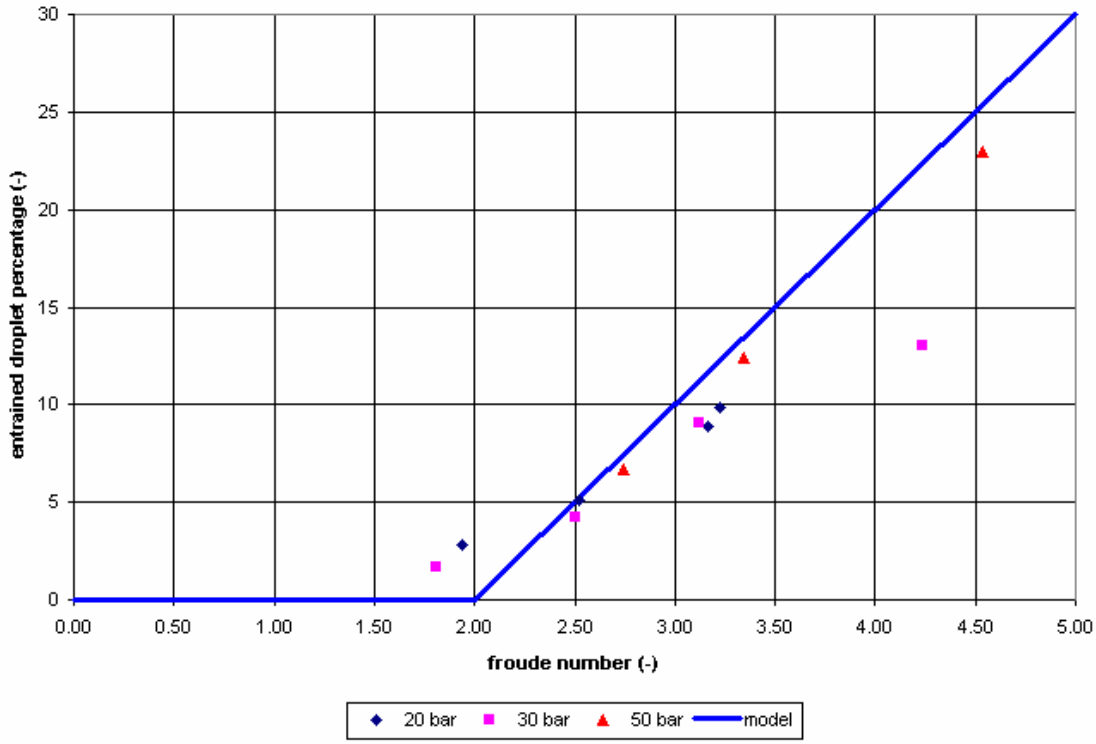

Figure 4

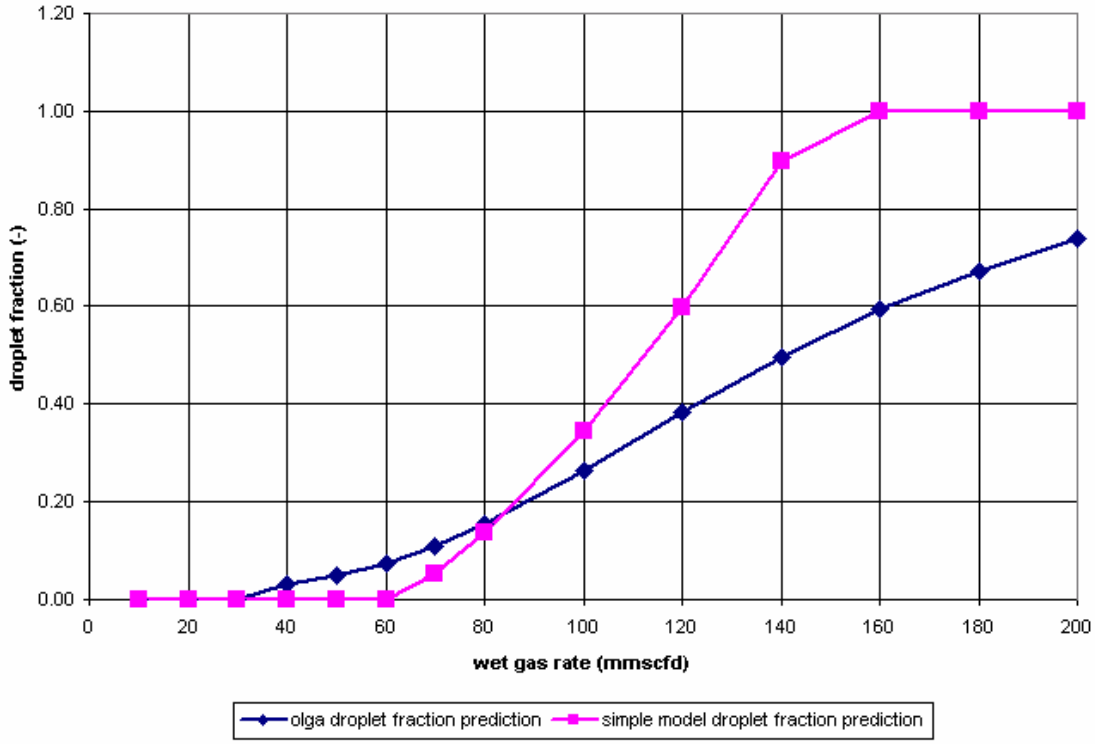

Figure 5

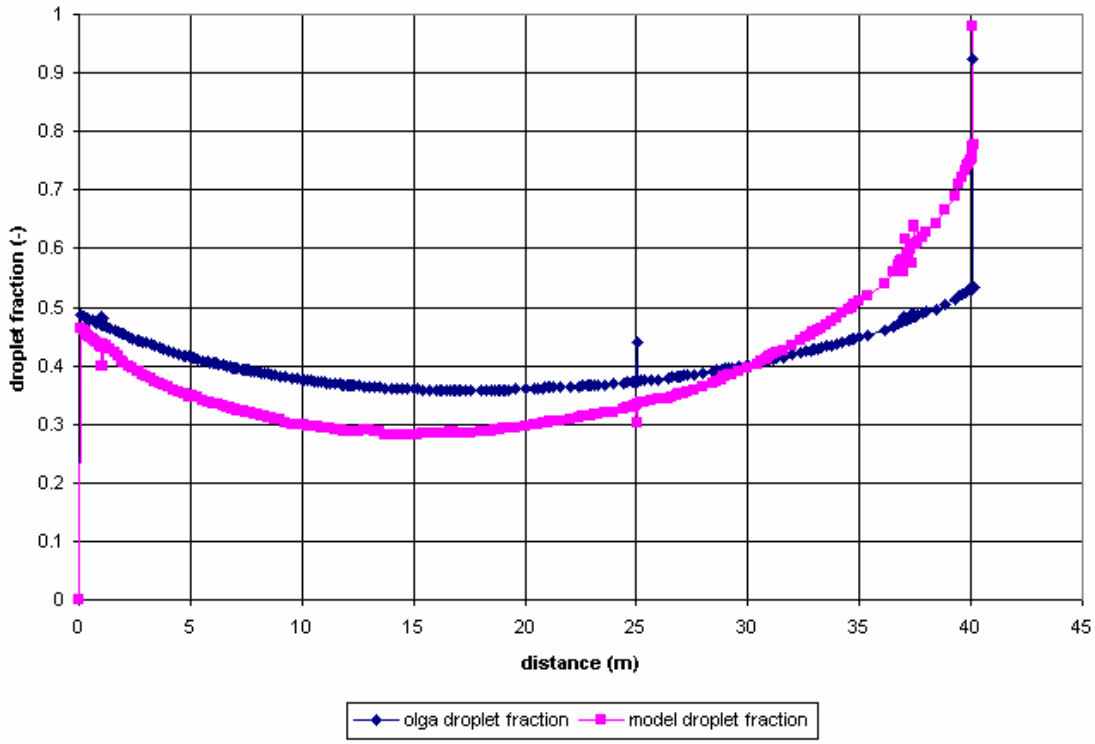

Figure 6

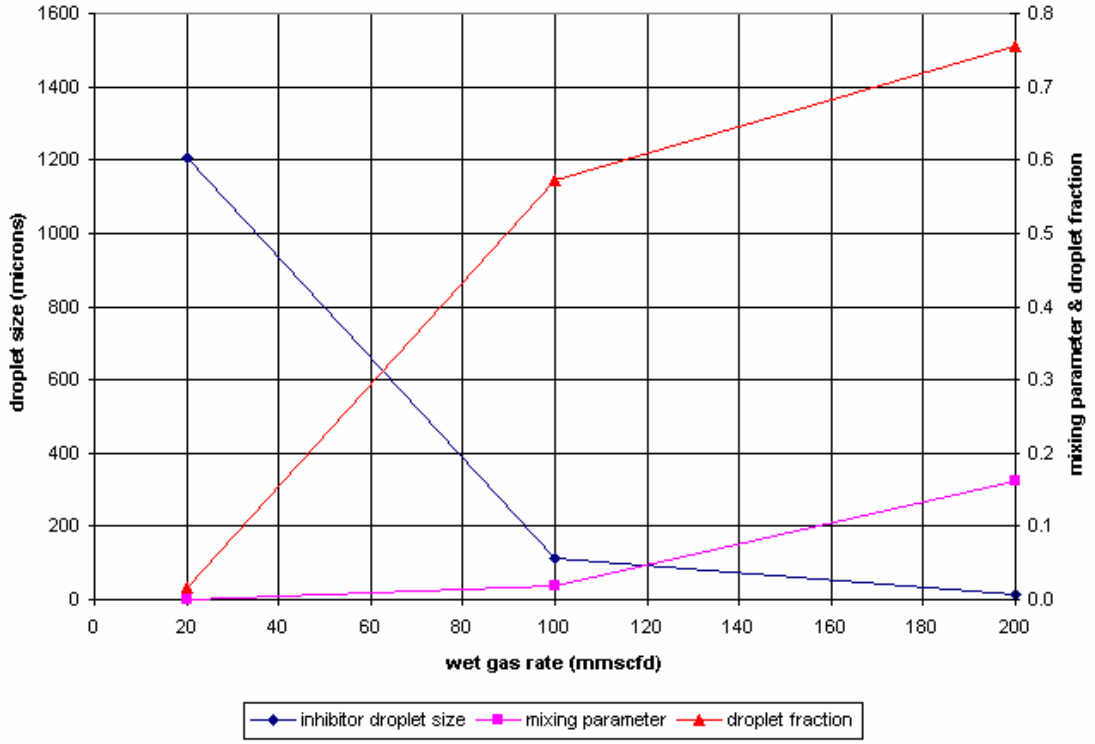

Figure 7

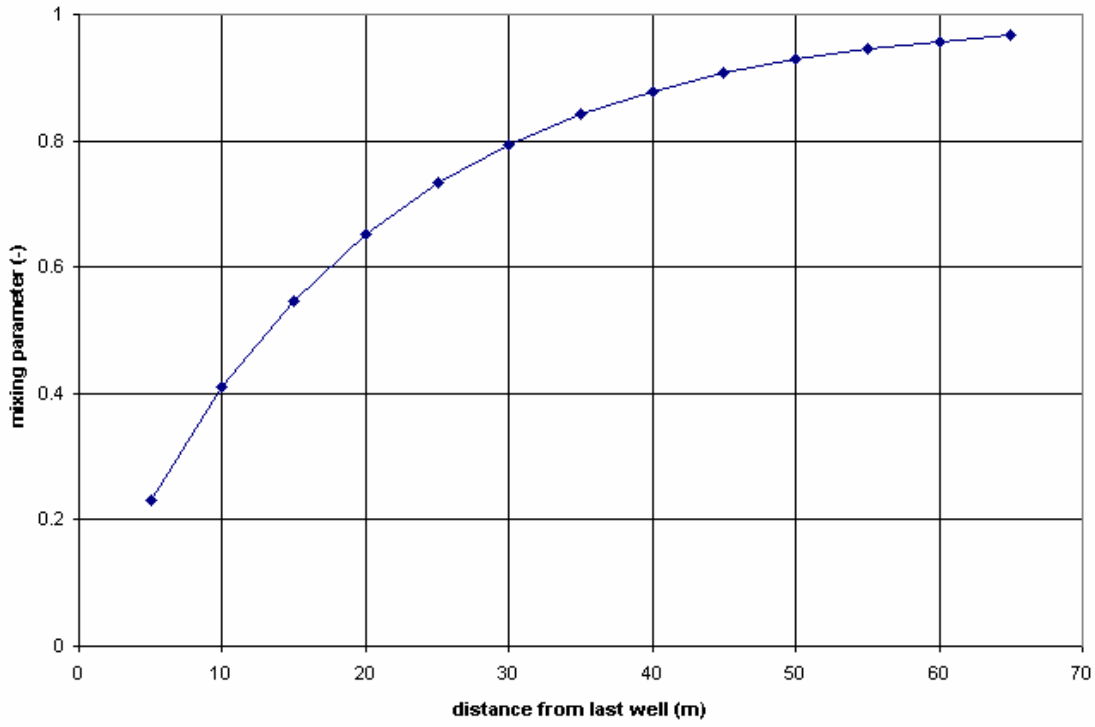

Figure 8

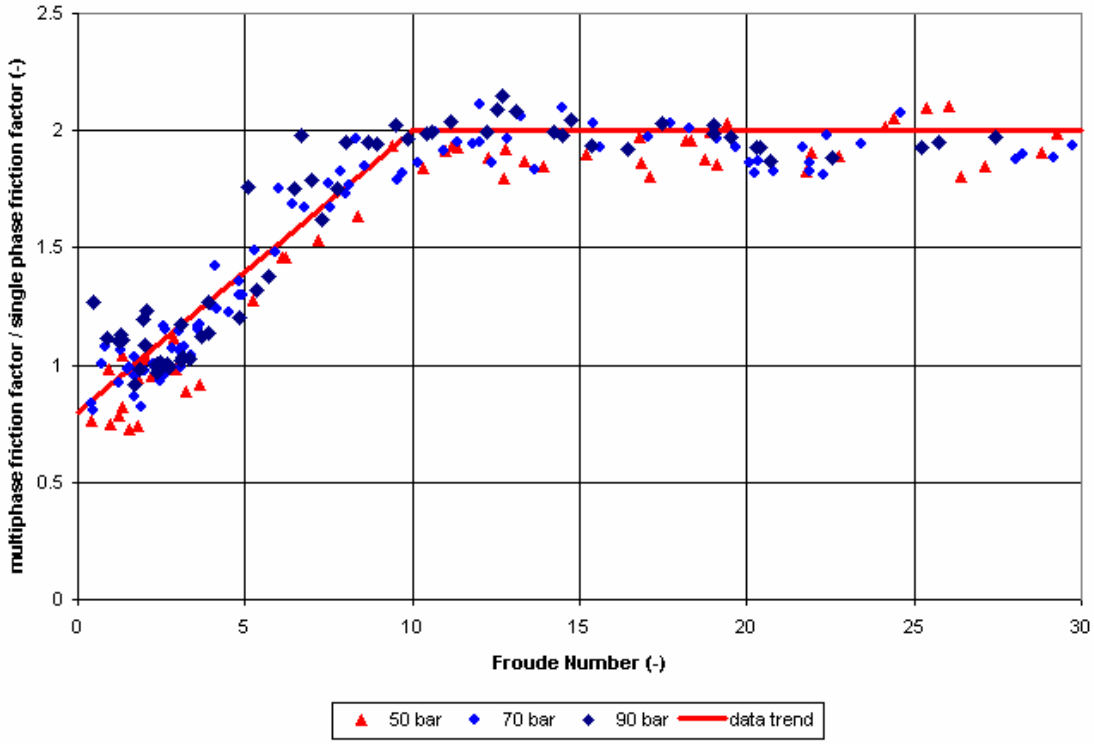
multiphase friction factor / single phase friction factor (-)
2.5
2
1.5
1
0.5
0
0
5
10
15
20
25
30
Froude Number (-)
50 bar
70 bar
90 bar
data trend

Figure 9