An Experimental Study of Gas Lift Efficiency in Polymer Systems

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ABSTRACT

This paper describes initial experiments that are done at the University of Stavanger in a new flow rig for study of gas lift under static and flowing conditions. Lift efficiency is determined from lift of a liquid column under various experimental conditions, while bubble rise velocity and bubble sizes are determined using high speed video recordings. For the experiments water, and also PAC (polyanionic cellulose) are used as fluids. Both static lift and air-circulation experiments involving polymers is found to be more time dependent than with water, due to different accumulation of gas bubbles.

INTRODUCTION

Gas-lift techniques are based on injection of gas into fluids to be transported in vertical upward flows in pipes and circulation systems. Gas reduces the mixture density and lowers the hydrostatic pressure gradient. It is used as a pumping technique often also for enhancement of chemical reactions via interfacial mass transfer and turbulence. It is used in a variety of engineering applications, as well as environmental cleaning and bioreactor systems, which generally involves non-Newtonian fluids [1,3,4].

Gas lift is also very important in oil production. A substantial fraction of drilled oil wells worldwide are “helped” by gas lift, and may increase production rates very much. Basically two applications are well known for petroleum. One is “gas lift production”, where usually produced natural gas is injected through gas-lift valves into the production tubing at the bottom of the well. The bottom-hole pressure (BHP) is reduced and thereby the inflow from the producing reservoir is increased. In this case the liquid is usually oil which is predominantly Newtonian, or an oil-water mixture which has a complex rheological nature not easily described by common non-Newtonian models. Another application is the addition of gas into drilling mud for so-called “underbalanced drilling”. The aim is to reduce the pressure at the bottom of the well to slightly below the reservoir pressure. The purpose is to avoid reservoir fracture and loss of lubrication drilling fluid into the reservoir formation. Drilling mud is usually non-Newtonian.

The gas-lift efficiency depends on the in-situ gas fraction for a given amount of injected gas. The lift process leads to increase in liquid flow-rate and thus the total flow-rate and mixture velocity will increase. However a number of parameters play a role in the overall process, where the rise velocity or “slip” is one of the most important. Additional factors involve solubility of the gas into the liquid, and also the total mixture velocity. Solubility is difficult to know in petroleum systems since the produced natural gas is most often used as gas source, and this aspect is not covered here since air and water based systems are used, with only minor interfacial mass transfer.

Rheology of the liquid is likely to have a major impact on the lift efficiency, in particular in conjunction with the suspended bubbles. The power-law behavior of the polymer solutions will not necessarily be the same in a gas-liquid system, like with shear rate versus shear stress in a vertical pipe. The rise velocity is closely connected to the bubble size but in non-Newtonian flows this is not a simple connection. It depends very much on the rheological parameters of the fluid. Furthermore it is found[1] that even for the same fluid the rise velocity depend in a more non-linear way on bubble size than in Newtonian flows. Finally, the bubble size in multiphase flow is in general a "self-organized" quantity. It is influenced by upstream injection or mixing history, and also by the turbulence intensity. The latter is a challenge to determine in polymer liquids with dispersed gas bubbles.

THEORY

The gas-lift efficiency is basically determined by the in-situ gas fraction which controls the hydrostatic pressure gradient. The gas fraction is for a given flow rate of gas directly connected to the slip velocity or the slip ratio $S = u_G/u_L$. Here u_G and u_L are the gas and liquid phase velocities respectively.

For low mixture velocity systems like in a bubbling column, it is more convenient to use the bubble rise velocity instead of the slip ratio, since u_L is practically zero and S would become very large. This is a common parameter to use also in drift-flux models for numerical simulation. In the drift-flux formulation the gas fraction can be determined theoretically by equating the two equations for the "drift flux" (*Zuber and Findlay, 1965 and Wallis, 1969*):

$$j_{GM} = \varepsilon_G (1-\varepsilon_G)^n \cdot U_T \qquad (1)$$

and

$$j_{GM} = (1-\varepsilon_G) \cdot U_{GS} - \varepsilon_G \cdot U_{LS} \qquad (2)$$

Eq. (1) is referred to as the "dynamical equation" since fluid flow dynamics controls the rise velocity u_T while n is a system dependent parameter. The equation also includes effective rise velocity in a system of bubbles with gas fraction ε_G as compared to terminal velocity for the bubbles. U_T is the stabilized rise velocity for one single bubble in a large tank. Equation (2) called the "kinematic equation" is theoretically just a connection between the superficial velocities in a fixed reference frame to the so-called "drift–flux" j_{GM}. This is the superficial velocity in a system following the mixture velocity, which again is the sum of the superficial velocities

$$U_{mix} = U_{GS} + U_{LS} \qquad (3)$$

For pure bubble column mode with liquid circulation U_{LS}=0, and the equations simplify to solving

$$\varepsilon_G (1-\varepsilon_G)^{n-1} = \frac{U_{GS}}{U_T} \qquad (4)$$

Alternatively if the gas fraction is known, it possible to derive the average bubble terminal velocity U_T using

$$U_T = \frac{U_{GS}}{\varepsilon_G (1-\varepsilon_G)^{n-1}} \qquad (5)$$

However the rise velocity may be very different for single bubbles in a big tank compared to in pipes with high gas fraction and complex velocity profiles. One objective of this study is to investigate also the effect of non-Newtonian liquids on the bubble flow pattern and net bubble rise velocity.

Good survey articles of gas-lift systems can be found[6,9]. The first focuses on experimental issues, while the other has more focus on mass transfer and computational models.

EXPERIMENTAL SETUP

The particular setup was originally motivated by need to determine gas-lift dynamics for petroleum engineering applications. Gas-lift assisted flow in wells "production" will normally be a one-way transport from "bottom-hole" up to the surface "wellhead". For underbalanced drilling it may involve also downwards gas transport. Gas is then injected from top into the drill pipe.

It was decided to build the laboratory flow rig with two 5 m high vertical legs; a "Riser" and "Downcomer", connected by two horizontal crossover pipes 1.5 m long, as shown in Fig. 1. In this way it would be also a very general rig for many different applications.

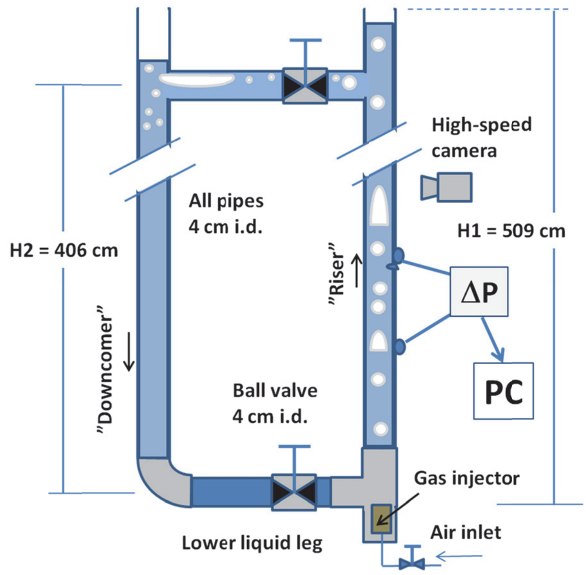


Figure 1. Sketch of the gas-lift loop, with definitions of "Riser", "Downcomer" and crossover "Legs".

It is quite similar to what is used also for chemical reactors or bioreactors where recirculation flow is desirable. The two crossover pipes can be opened or closed by means of valves.

Gas lifting is obtained by injection of gas (air) bubbles at the bottom of the right leg in Fig. 1. A close up of the injector is shown in the insert of Fig. 2. The bubble size is only partly controllable with different types of injectors or "spargers". In this work was used a porous injector made from small sintered metal particles, as shown in the electron microscope picture in Fig. 2.

A problem with this type and most other "passive" spargers is that bubbles tend to hang on to the pore openings during the inflation stages and may become several diameters larger than the openings of the gas outlet.

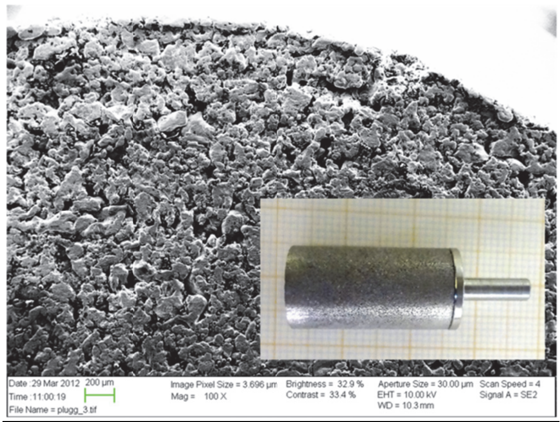


Figure 2. Picture of the gas injector inserted into an electron microscope image of the surface. Magnification is 100X, and the image width is approximately 3.8 mm.

However even if the pore openings are microscopic, 10-100 µm, it is difficult to obtain smaller bubbles than 1-2 mm, due to bubbles clinging to the pore outlets during the inflation, or by coalescence with other bubbles before they are released.

Alternative spargers are planned in order to generate smaller bubbles. These could be based on applying a shearing force to the bubbles either by a strong cross flow, or by using a fast mechanical scraper mechanism to wipe the bubbles off the surface.

EXPERIMENTS

Experiments were planned and carried out to determine the following issues: 1) flow regimes in the riser, crossovers and downcomer pipe, 2) gas-liquid fractions as a function of superficial gas velocity, 3) flooding limit for gas bubbles in the downcomer pipe in terms injected gas, 4) bubble size for given gas sparging unit as a function of gas flow rate, 5) pressure gradients versus flowrates for different configurations of valve openings, and finally 6) velocity profiles of liquid and gas in riser and downcomer. Determination of the velocity profile in the downcomer could also enable calculation of the liquid flow-rate.

There are already quite a lot of data generated from these initial experiments. They have been acquired during a very busy experimental campaign the last month. This has left less time to carry out all the planned analysis, so only some few examples will be presented and discussed here. More will be presented at the conference and in later papers.

Experimental program

The different tests were grouped as

i) *Static column bubbling*

ii) *Gas-lifted circulation*

Static bubbling was used to determine liquid height as a function of gas flow rate. The goal was to determine gas fraction. This was done both for water and PAC. Flow regime determination and a feasibility study to find bubble sizes gas fraction distributions was also part of this study.

Circulation with fully opened valves was carried out to study dynamic lift, both in Newtonian and non-Newtonian flow. As described later a direct comparison between the experimental modes was influenced by the gradual entrainment of bubbles especially in the non-Newtonian case.

The lift efficiency can be measured from the pressure gradient in the riser, but it is representative only in the initial period when gas bubbles still have not reached the downcomer pipe. On the measurement program was also the flooding limit for gas bubbles in the downcomer, both for water and PAC

Flow aspects regarding the tests

The “static 1” experiments were carried out with the crossover pipes closed These tests reveal the potential lift properties of a full circulating the gas-lift system. It is also possible to close just the upper valve to aloe inflow from the bottom into the riser. These tests mainly shift the column level to a slightly higher level, but do in principle not add much more information.

The circulating fully dynamic tests also involve flow friction. Both in two-phase flow, and in single-phase non-Newtonian system this is important to determine. A comment on the static versus circulating test regards the analogous equivalent when measuring the voltage difference between the poles of a battery when not loaded. The circulation tests give the corresponding voltage when current flows in an external circuit. In his case the voltage is reduced due to internal losses in the battery. The friction in the left downcomer leg corresponds to the internal loss.

Instrumentation

The flow loop has been set up to give a best possible access to vital information to determine lift dynamics and efficiency. Transparent pipes give excellent information about flow regimes, bubble size and velocities, and also liquid rise levels. Pressure is measured at the gas flowmeter with a Honeywell sensor. Also the pressure difference ΔP between the two locations shown in Fig. 1 was measured, using a Rosemount transmitter 3051C. It has a range of 62 mbar and time resolution 100ms. Finally the gauge pressure is measured at the lowest position of the two used for the ΔP sensor. A "Gauge Digital Testing" sensor (Crystal Engineering, XP2I) was used. Pressures are recorded to PC using a fast datalogging card (brand NIDAQ 6024E, 12 bit). Phase distributions and velocity profiles are measured by means of a high speed video camera (SpeedCam MiniVis e2) that records up to 2500 fps using full resolution 512x512 pixels. It may even record up to 120.000 fps at reduced resolution. The camera has onboard memory for 8223 full frames at full resolution. Images are downloaded to computer via a GigaBit Ethernet cable by means of a dedicated communication program ("MotionBlitz", by Mikrotron).

Fluids

Basically three different fluids were used; tap water, and two different concentrations of polyanionic cellulose (PAC) with 2g and 4g dry powder per Liter of distilled water. These are referred to as PAC 200 and PAC 400. The viscosity of the PAC solutions was measured using a rheometer of type Physica UDS 200 with cone plate configuration.

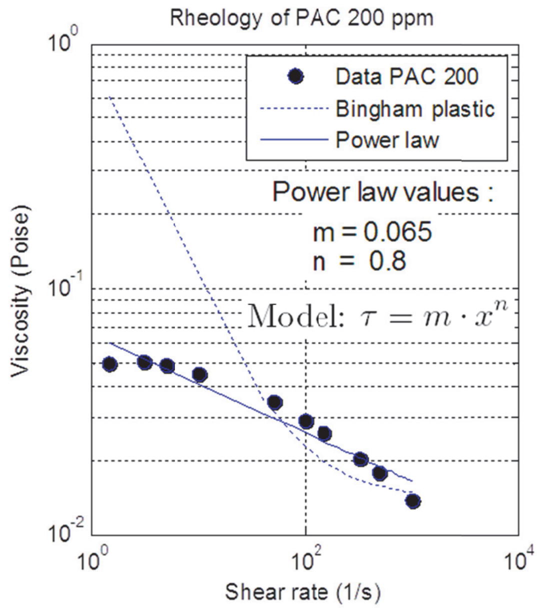
PAC is chosen because of good optical transparency. The PAC solutions behave as power-law fluids with parameters as given in Fig. 3.

RESULTS AND DISCUSSION

Video recordings allowed determination of flow regime, gas bubble rise velocities and gas-liquid distribution, and this serves as a reference technique. This paper focuses on what can be obtained from overall measurements like pressure gradient and liquid lift heights.

Static experiments

For static experiments with both valves closed, the flow regime becomes more and more inhomogeneous at the same time as the bubble size increases with the gas flow



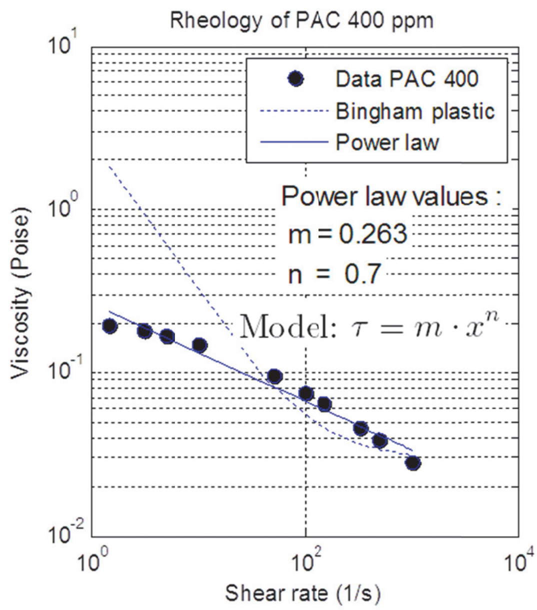


Figure 3. Rheology of PAC 200 and 400.

rate. Transition to slug flow is shown in Fig. 4.

For Newtonian flows the transition from dispersed bubble (DB) to slug flow is sometimes described with a "maximum packing" model. This is one part of what is often referred to as the mechanistic "Taitel Barne and Dukler model"[12]. Below a critical gas fraction the bubbles remain distinct and dispersed. Exceeding this fraction the dense bubble cloud causes coalescence into larger bubbles. For Newtonian liquids, at low flow rates the maximum gas fraction for low viscosity fluids is approximately 25%, while at highly turbulent flows it may stay bubbly up to over 50%.

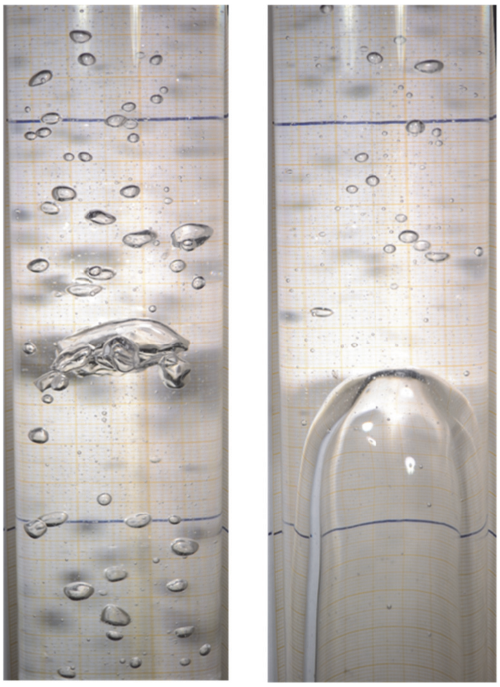


Figure 4. Flow regime in water-filled riser at moderate gas influx varies between bubbly and slug flow. The two images are taken with 2 seconds interval.

The model[12] is validated only for co-current upwards Newtonian flows. In the present static experiments ("bubble column" mode) the liquid flow is stagnant. The experiments reveal a very complicated bubble flow with recirculation cells that change continuously.

This enhances the bubble collision rate and would for a non-Newtonian system lead to slugging at lower gas fractions.

However it is also seen other cases where the bubbles set up collective pathways resembling transport veines.

In Fig. 5 the gas fraction in the riser is calculated directly from the liquid level increase.

However, in the static case the gas fraction can also be calculated from the pressure gradient in the pipe using the relation

$$\varepsilon_G = \frac{\Delta P}{g \cdot (\rho_L - \rho_G) \cdot h_{12}} \qquad (6)$$

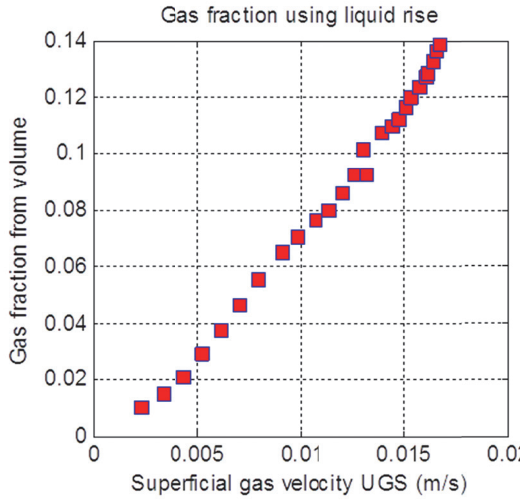


Figure 5. Gas fraction with PAC 400 as a function of superficial gas velocity, based on volumetric increase of liquid in riser is used.

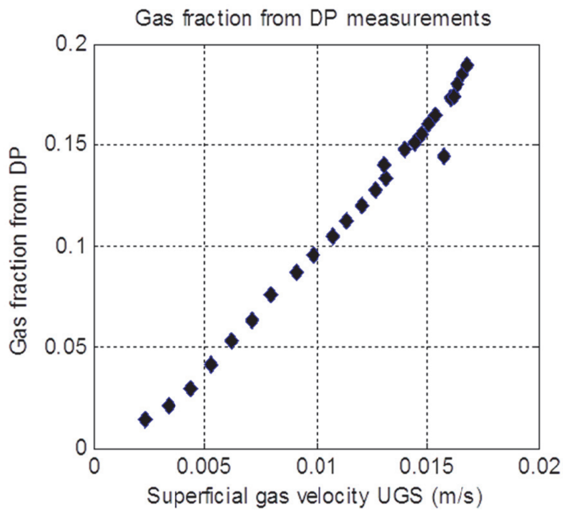


Figure 6. Gas fraction with PAC 400 as a function of superficial gas velocity, based on pressure gradient in riser.

since the difference between the two inlets 1 and 2 to the DP sensor then is due to the gas fraction. Using the densities of PAC as for water and air at the ambient pressure in the pipe (1.5 bar), we get the relation shown in Fig. 7. Ideally the two methods should have given the same result, following a 45^0 line. It is quite obvious from the plot that although there is a good linearity, the pressure method overestimates the gas fraction compared to when using the volumetric, which must be assumed closer to the real value since volumes are directly connected to the fractions.

This comparison shows that the DP pressure sensors indicate more gas from a higher DP, equivalent to saying that the pressure drop upwards is higher than expected. This could be explained in terms of a friction component in the multiphase mixture. This seems plausible since also the pipeline pressure increases. This assumption is also supported by the variation in pressure (gauge) versus gas flow rate as shown in Fig. 8.

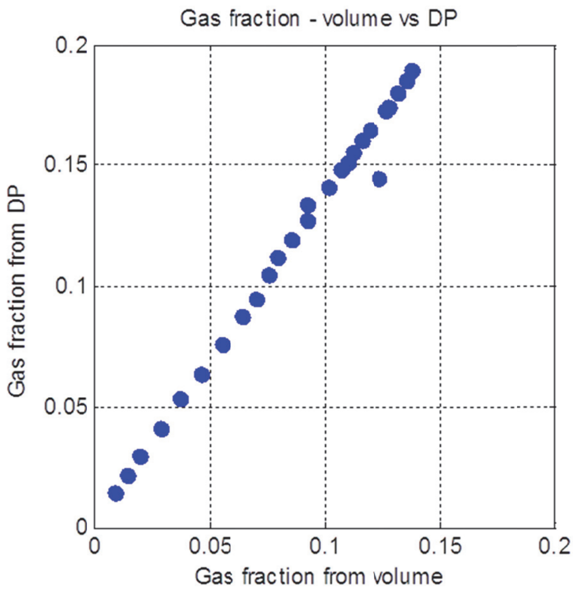


Figure 7. Gas fraction with PAC 400 – volumetric method versus pressure.

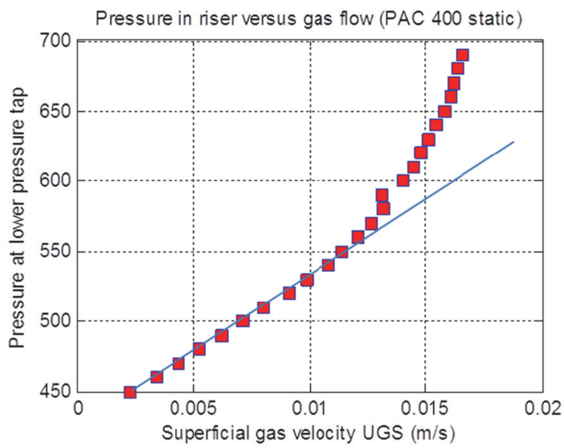


Figure 8. Gauge pressure (in mbar) at bottom of riser as a function of gas inflow rate for PAC 400.

Fig. 8 shows a linear increase up to 0.012 m/s and then a sharper rise which could be attributed to increased turbulence. Since only the gas flowrate is measured for Fig. 8, a detailed calculation of Reynolds number must await the analysis using the high-speed video system. A fullbore liquid flowmeter could be used, but very few if any of the conventional techniques are feasible; neither electromagnetic nor acoustic for gas rich flows.

Finally, the bubble rise velocities may be calculated using the drift-flux model presented earlier. This is shown in Fig. 9. It may be seen that for superficial gas velocities over 0.008 m/s the bubble rise velocity is fairly constant around 0.15 m/s. This is a reasonable value, and somewhat expected since the bubble interaction is quite complex for the higher gas flowrates.

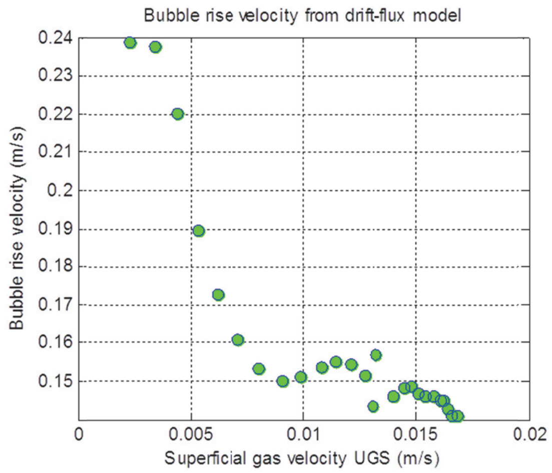


Figure 9. Average bubble rise velocity for PAC 400, based on drift-flux model (Eq.5).

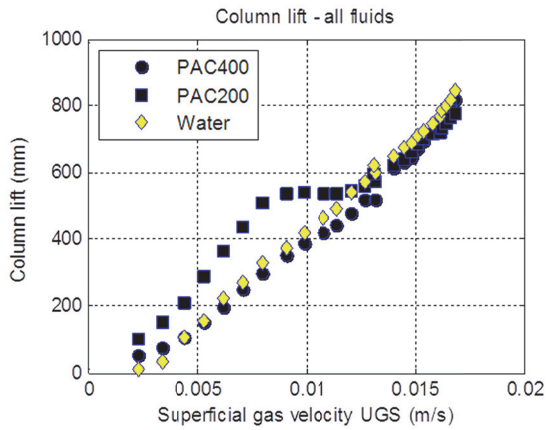


Figure 10. Column lift for all the fluids, static bubbling - without circulation.

At gas flow lower than 0.007 m/s the bubble rise velocities are much higher, because the bubbles move without collisions and in fairly straight paths. In some tests trains of bubbles were developed along the pipe wall, where the bubble speed was visibly much higher than in the central part of the pipe. This is very likely a kind of Coanda effect.

For the shear thinning PAC system this kind of dynamics is modified due to higher viscous damping of the liquid which decreases the collision strength. However shear thinning also increases bubble deformation, and thus makes the bubble more susceptible to coalescence.

Finally considering the column lift in the static cases, all the tests are summarized in Fig. 10. From the graph there is clear trends for the superficial gas velocities below 0.012 m/s. For higher flow rates, the lift seems to be slightly higher for water.

Experiments with open valves - circulation

When the valves are open, liquid is allowed to circulate until a new equilibrium levels is reached. At first sight this may seem to the same as the static case. However in this new situation the liquid may transport gas to the crossover pipe and the downcomer. Important new regimes occur when the liquid start to flood the gas down as in the example Fig. 11.

The different flow regimes that arise influence the gas fraction and circulation rate. The equilibrium liquid levels in the two pipes reflect the difference in pressure gradient. Gas bubbles in the downcomer reduce the mixture density and much of the gas-lift effect is lost. This in turn reduces the circulation velocity and allows more of the gas to ventilate out at the top of the downcomer. It is only the “excess gas” (difference in gas fraction in the two pipes) in the riser which sets up a lift effect.

Density variations with time

With PAC, over time the gas accumulation in the liquid complicates the analysis somewhat, in particular since also larger bubbles with unpredictable rise behavior may continue into the downcomer pipe. This pipe is supposed to be a reference for the liquid density, which is then gradually reduced.

To some extent this effect can be determined and accounted for using closed valves and comparing liquid level in the downcomer with the level in a third thinner pipe which contains only liquid without bubbles.

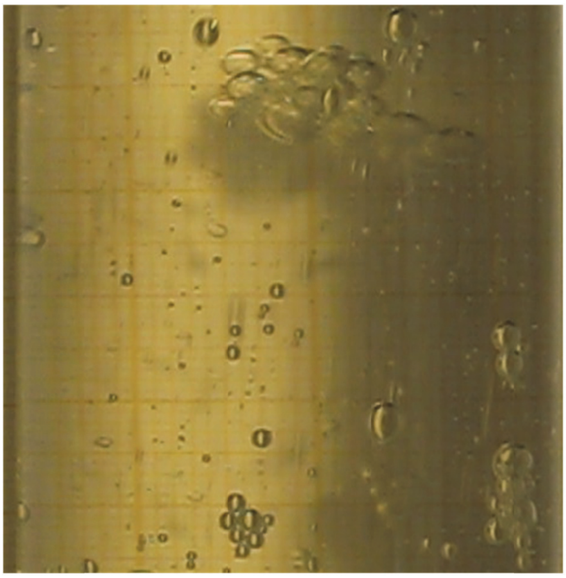


Figure 11. Flooding of gas bubbles into the downcomer with PAC 400 liquid. The bubbles agglomerate, rotate and rearrange in constantly varying patterns. Image snapshot from video.

In circulation mode also bubble size separation effects takes place in the downcomer pipe. This is different from flow in the riser. The immediate reason is the change of flow regimes (single and two-phase) and flow dynamics in concurrent and countercurrent flows. This again sets up specific and different effects in water and polymer flows. In the riser both large and small bubbles flow side by sides, and the large bubbles causes a type of “piggyback riding” effect - the small bubbles follow in the wake of the larger bubbles. The variations in Fig. 12 mainly show the large scale impact of the pressure gradients and flow conditions in the riser and downcomer.

Flooding dynamics in downcomer

In the downcomer pipe there are two effects present; 1) bubbles in the vicinity of the wall decrease the near wall liquid velocity. For non-Newtonian flows this forces a higher downward liquid flow in the middle (“core”) part of the pipe. 2) This

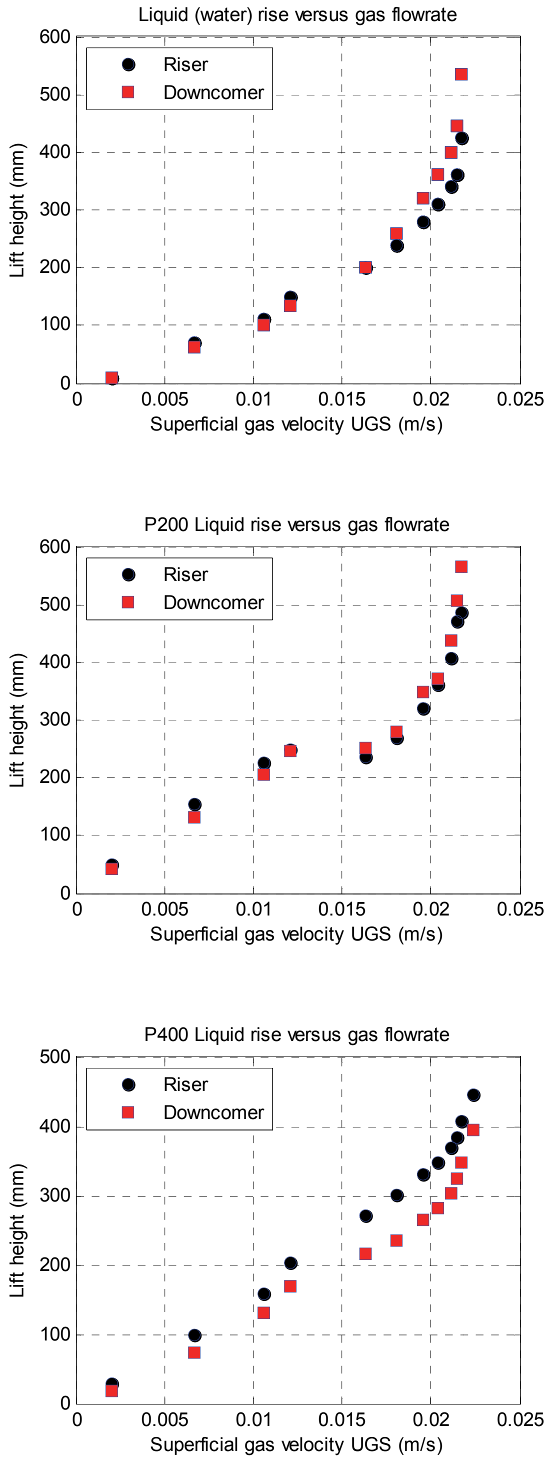


Figure 12. Difference in liquid height in riser and downcomer as a function of gas flow rate. Circulation experiments with fully opened valves using water (top), PAC200 (middle) and PAC400 (bottom).

causes a "pinch" effect so that there is higher shear further away from the wall than for water. Since PAC is shear thinning this causes a reduction of apparent viscosity in the main flow of the pipe. So even though the downflow is stronger in the core, the reduced liquid viscosity in some cases may cause enhanced counter current upwards bubble flow compared to with water circulation.

Similar phenomena are seen in the riser, but with opposite effect. Bubbles rise faster close to the wall causing a positive feedback with increased wall shear stress and reduced apparent viscosity. As an example, this is as reflected in Fig. 9.

The high speed camera is particularly useful for this study. It enables both calculation of the liquid phase velocity with the addition of small tracer particles. It also visualizes the velocity profile, bubble size and bubble fractions profile, as well as bubble shape.

CONCLUSIONS

Two different modes of operation were used, with and without external circulation, often referred in literatures to as bubble column and gas-lift mode. In this work the bubble column mode is mainly used as an indicator of the gas rise velocity as reflected in the gas fraction.

It was found that gas-lift with Newtonian and non-Newtonian liquids differ in several ways. Fig. 10 and 12 illustrate the overall behavior in various ways. These are not conclusive until a more detailed analysis from high-speed recordings is done. The lift in the non-Newtonian PAC 200 case is slightly higher for low gas flow rates than the corresponding Newtonian case, even though the overall viscosity of water is lower.

The interpretation is therefore not straightforward, since the lift process dynamics involves both the liquid viscosity and the density. The main mechanisms seems to be the reduced rise velocity of the gas bubbles which increases gas fraction as seen in the "static" experiments where both of the crossovers valves were closed. Reduced density should in the case of equal viscosity lead to higher flow. However for PAC the viscosity is higher and friction increases. The flowing case (circulation mode) is in accordance with this conclusion, but is complicated by the gradual saturation of small and microscopic bubbles over time.

The effect of gas bubble size is for the present initial study difficult to estimate, since bubble size was not easily controllable. Similar conclusions are given in most publications too; that diameter of gas sparger orifices has small effect on the overall gas fraction and pressure drop. We conclude this from a theoretical point of view; the bubble rise velocity approaches zero for small bubbles, and thus the gas fraction would increase with smaller diameter. This would be even more pronounced with shear thinning fluids like PAC.

For the bubble column mode this would lead to accumulation of time until the gas creates a foamy mixture at the sparger, eventually leading to slugging when the foam breaks. For the gas-lift mode a similar effect is expected, although taking much longer time and leading to a more even gas distribution.

Friction is modified by the bubbly flow as compared to single phase flow for both the Newtonian and the non-Newtonian case, in accordance with the general conclusion in most published works[8].

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