In-line characterization of cement slurry - towards automated cementing operation

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ABSTRACT

Cement is a critical component for the integrity of oil and gas wells. As for many other industrial fluids, the quality and consistency of cement slurries depend on the equipment used for mixing and the conditions at which the mixing is performed. Optimization of slurry properties is commonly done at lab scale with small volumes and high shear mixing. Field-scale mixing conditions are significantly different so there is need for improved characterization of slurry properties when upscaling from the laboratory. Given the industry's goal of increased automation, the need for manual intervention through fluid sampling should be reduced. In this paper, we investigate the applicability of an in-line system for slurry characterization based on an instrumented standpipe principle for simultaneous determination of fluid density and viscosity. Characterization of non-Newtonian fluids is explored by systematically varying the imposed flow rate, and thereby the corresponding wall shear rate based on the fully developed non-Newtonian velocity profile. The feasibility of the system has been tested using both non-Newtonian model fluids and an oilfield cement slurry with comparable rheological profiles. Our results show that inline viscosity characterization of the test fluids agree well with independent viscometer measurements, especially for low shear rates. At greater shear rates, inline measurements result in a lower viscosity compared to the viscometer.

INTRODUCTION

Well cement is a critical barrier material that is placed behind casing strings in wells for geothermal energy recovery, geological CO_2 storage or hydrocarbon production. The cement slurry is mixed at the rig site and pumped down the well inside the casing that is to be cemented. At the bottom of the casing, the cement slurry flows into the annular space that is to be cemented, and displaces the in-situ fluids as it is pumped up toward the surface. The placement performance of the cement slurry and its ability to completely displace the original annular fluids are closely linked to its viscosity and density [1]. Thus, accurate and reliable measurement of these properties at the rig site can help ensure consistent large-scale mixing of the cement slurry, and that the actual slurry properties are in fact as planned before pumping the cement down the well.

Today, modern cementing units are equipped with monitoring devices for continuous measurement of volumetric flow, density, and pump pressure. However, they lack continuous in-line monitoring of the rheological properties of the mixed fluid, which could enable pro-active decision-making and improve quality control in the field. Conventionally, the practices recommended by American Petroleum Institute (API) standards are used for cement slurry properties measurements. However, these procedures are manual and prone to human error. To improve efficiency of fluid characterization during rig-site mixing, it would be advantageous to automate and do continuous measurements of slurry viscosity.

In this paper, a standpipe viscometer for automated cement slurry measurements is introduced. The design and operation is based on existing design concepts that have been developed primarily for automated drilling fluid characterization, see *e.g.* Refs. 2, 3. The aim of this work is to demonstrate the feasibility also for measurement of cement slurry viscosity and density.The proposed method may also facilitate automatic cement mixing operation where slurry properties can be measured in real-time and make it possible to adjust the slurry composition accordingly. We compare viscosity measurements acquired from the instrumented standpipe to off-line measurements from a conventional viscometer.

METHODS AND PROCEDURES

Experimental setup

The schematics of the instrumented standpipe used in this study is shown in Fig. 1. The current setup consists of a fluid mixing tank with agitator and ca. 80 liters capacity, a progressive cavity pump (screw pump) and two pipe measurement sections, each of 3 m length and with an inner diameter of 21 mm. As shown in Fig. 1, pairs of pressure sensors are placed along the horizontal and the vertical pipe sections for continuous measurement of friction pressure loss (horizontal section) and combined friction pressure loss and hydrostatic pressure gradient (vertical section). Thus, by subtracting the friction pressure component, as measured in the horizontal section, from the vertical pressure difference, the friction and hydrostatic pressure components can be identified separately, allowing continuous measurement of both viscosity and density. Differential pressure sensors with a measuring range from 0 to 1300 millibar are used in this study, and the pair of sensors is spaced 150 cm apart. The progressive cavity pump can operate up to pressures of 10 bar and at 175 revolutions per minute with no measurable pulsation effects. Finally, a temperature sensor is placed in the tank and a Coriolis flow meter for measurement of density and flow rate is placed between the mixing tank and the pump. The figure also shows how the right-hand side measurement loop can be integrated with a field batch mixer setup with a centrifugal pump for fluid re-circulation during mixing of cement slurry.

Maximum flow rate and locating the pressure sensors

In this work, the pump rates were selected to ensure laminar flow during the measurements. With the first pressure sensor located 70 cm downstream of the pump, the flow should be fully developed over a length no greater than about 33.3 hydraulic diameters of the pipe. Using the development length correlation developed by Poole and Ridley [4] for pipe flow of power law fluids, we take $X_D/D \approx 0.0567\mathrm{Re}_g$ as a reasonable approximation for the ratio of development length, X_D, to inner pipe diameter D at Reynolds numbers $\mathrm{Re}_g \gtrapprox 100$. Here, $\mathrm{Re}_g = 8\rho V^{2-n} D^n (n/(6n+2))^n / K$ is the Metzner-Reed Reynolds number for power law fluids, n is the flow index and V is the imposed bulk velocity of the

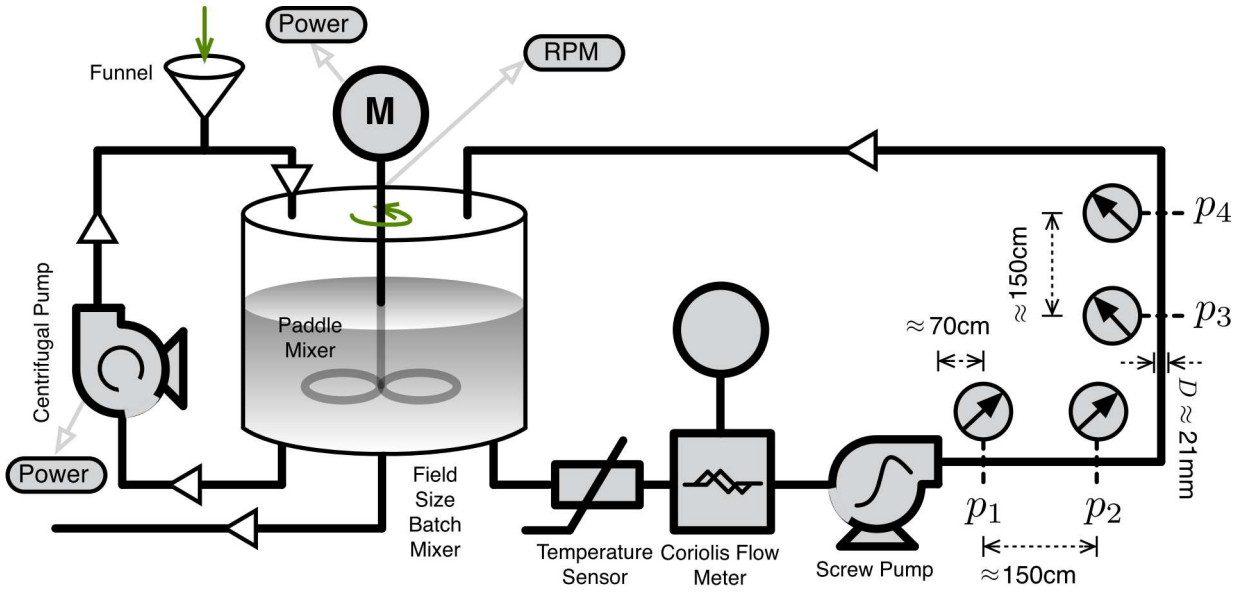


FIGURE 1: Schematic of the instrumented standpipe setup for continuous viscosity measurement of the cement slurry. In current tests, only the right-hand side measurement flow loop has been implemented. In future use, the measurement system will be integrated with a field size batch mixer with a centrifugal pump for circulation and mixing.

flow [4]. Since $X_D = 70$ cm, we require $\mathrm{Re}_g \lessapprox 588$ to have fully developed flow throughout the measurement section.

Test fluids

We studied the instrumented standpipe concept using both non-hardening fluids and a conventional oil well cement slurry. Glycerol and a xanthan gum solution were used as the two non-hardening fluids. While glycerol has close to Newtonian behaviour, xanthan gum solutions are shear thinning fluids. As cement slurry, we used cement class G with silica to gain water/cement ratio of 0.52 and water/dry material ratio of 0.39 (including silica). In order to compare and verify the results from the standpipe measurement, a conventional rotational viscometer (Fann 35) was used as reference. The procedure was chosen to obtain the rheological profile of the tested fluids at the same time and the same temperature through both standpipe and Fann 35.

Fluid property measurement

Viscosity

The rheological behavior of well construction fluids such as drilling fluids, spacer fluids and cement slurries is often characterized by shear thinning and yield stress behavior. Consequently, a Herschel-Bulkley constitutive model is often used for representing the steady state viscosity, μ, in sheared regions of such fluids, *i.e.* $\mu = \tau/\dot{\gamma} = \tau_y/\dot{\gamma} + K\dot{\gamma}^{n-1}$. Here, τ and $\dot{\gamma}$ denote the shear stress and shear rate, while τ_y, K and n are the yield stress, the consistency index and the flow index, respectively. In regions where the shear stress is

less or equal to the yield stress, the model treats the fluid as a rigid body with $\dot{\gamma} = 0$. The three model parameters can be estimated from either viscometer measurements of the fluid or by the measurement of fully developed laminar friction pressure gradients in a conduit of known dimensions at different imposed volumetric flow rates. For fully developed flow in a pipe of inner diameter D, the axial momentum equation becomes $\tau_w = D\Delta P_f/(4L)$, where ΔP_f is the friction pressure loss over a length L of the pipe. As shown by Magnon and Cayeux [5], the wall shear rate in laminar pipe flow of a Herschel-Bulkley fluid is

$$\dot{\gamma}_w = \frac{3}{4}\dot{\gamma}_{N,w} + \frac{1}{4}\dot{\gamma}_{N,w}\frac{d\ln(\dot{\gamma}_{N,w})}{d\ln(\tau_w)} \tag{1}$$

where $\dot{\gamma}_{N,w} = 8v/D$ is the Newtonian wall shear rate, and where

$$\frac{d\ln(\dot{\gamma}_{N,w})}{d\ln(\tau_w)} = \frac{n+1}{n}\frac{\tau_w}{\tau_w - \tau_y} - \frac{C_1\tau_w + 2C_0}{C_2\tau_w^2 + C_1\tau_w + C_0} - 1, \tag{2}$$

with $C_0 = (6n - 4n^2)\tau_y^2$, $C_1 = 2n(1+n)\tau_y$ and $C_2 = (1+n)(1+2n)$. For measurements of the wall shear stress τ_w, or equivalently the friction pressure drop ΔP_f in pipe flow, Eqs. (1) and (2) can be used to evaluate the corresponding wall shear rates for given model parameters τ_y, K and n. However, as the model parameters are generally unknown at the time of measurement, we apply an iterative procedure where we solve for the least squares estimates of the model parameters by minimizing the residual sum of squares of the measured and modelled wall shear stress at each wall shear rate from the experiments.

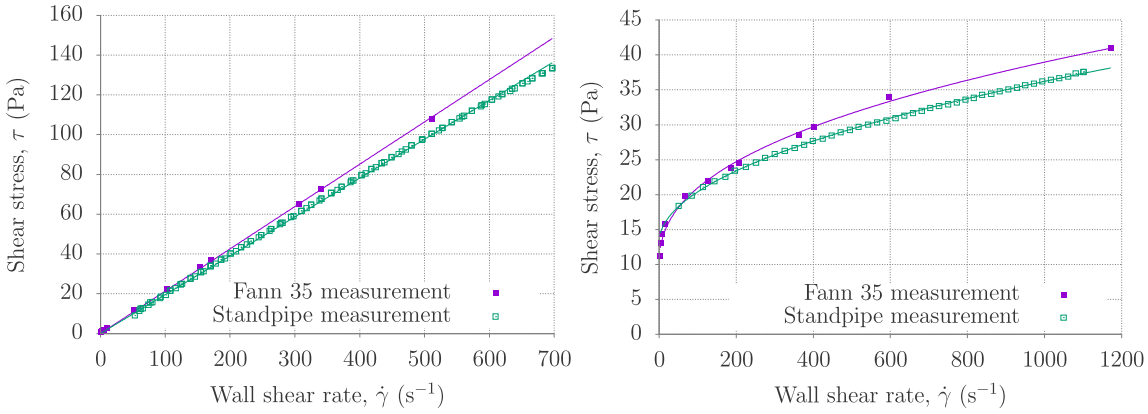
In this study, we compare the model parameter estimation from measurements of friction pressure in a pipe, as explained above, to flow curve measurements using a Fann 35 viscometer fitted with the standard R1-B1 measurement geometry. As explained by *e.g.* Lima *et al.* [6], the shear stress and shear rate at the wall of the stator in the viscometer is normally calculated from measurement of the torque applied to the stator and from the rotational speed of the rotor. While conventional conversion factors are often based on an assumption of a Newtonian fluid, for shear thinning and yield stress fluids these conversion factors will generally depend on the unknown viscosity. Consequently, an iterative approach is required to identify wall shear stresses and shear rates that are consistent with Herschel-Bulkley measurement fluids. As per above and Ref. 6, we solve for the least squares model parameters by minimizing the modelled and measured torque at each rotational speed, using the non-Newtonian end effect correlations developed by Lac and Parry [7] and correcting the wall shear rate as per Skadsem and Saasen [8].

Density

As shown in Fig. 1, measurements of pressure difference are performed both in a horizontal and a vertical section of the pipe. Since the length L between the pairs of pressure transmitters is the same in the two sections, the friction pressure drop over L will be the same, ΔP_f. In the vertical section, the height difference between the two pressure transmitters will give rise to a hydrostatic pressure difference $\Delta P_g = \rho g L$, where ρ is the fluid density and g is the gravitational acceleration. Consequently, the fluid density can in principle be measured from the vertical pipe section by subtracting the friction pressure loss, as measured in the horizontal pipe section. This measurement principle has been investigated by *e.g.* Carlsen *et al.* [2] and Sui and Vidaur [3].

RESULTS AND DATA ANALYSIS

Flow curve measurements for the non-hardening test fluids, *i.e.* glycerol and the xanthan gum solution, are shown in Fig. 2. In both cases, the flow curves have been measured with the Fann 35 viscometer using the standard R1-B1 measurement geometry, and these results are compared to the equivalent flow curves from the instrumented stand pipe. As explained above, measurements of flow rate and friction pressure gradient in the instrumented standpipe have been converted to equivalent wall shear stress and wall shear rates, and plotted in Fig. 2.



(a) Glycerol (b) Xanthan gum

FIGURE 2: Comparison of flow curve measurements for glycerol (a) and a xanthan gum solution (b) acquired with the instrumented standpipe and using a Fann 35 viscometer.

As expected, the flow curve for glycerol suggests a Newtonian fluid with constant viscosity at all shear rates. We observe from Fig. 2a that the flow curve based on Fann 35 measurements suggests a larger viscosity (approximately 212.9 cP) compared to the flow curve from the instrumented stand pipe (approximately 195.4 cP). A similar trend is also observed when comparing flow curves for the xanthan gum solution in Fig. 2b. The Fann 35 measurements suggest highly shear thinning behavior, and also yield stress behavior. The solid line in Fig. 2b is a Herschel-Bulkley parametrization of the viscometer measurements corresponding to $\tau_y = 9.8$ Pa, $K = 1.65$ Pa·s^n and $n = 0.42$. We note that these measurements have been corrected for non-Newtonian end effects, as described above. The flow curve based on the instrumented standpipe measurements is also now suggesting a lower effective viscosity compared to the viscometer measurements, with Herschel-Bulkley parameters of $\tau_y = 13.3$ Pa, $K = 0.69$ Pa·s^n and $n = 0.51$.

To test the standpipe measurement with a cement slurry, class G cement combined with silica were used and mixed with water to obtain 54 liters of slurry with mass density of 1920 kg/m^3. The cement slurry flow curve was measured four times in the instrumented stand pipe, each starting at a high flow rate of approximately 38 l/min. The flow rate was next ramped down according to a pre-programmed sequence, allowing ample time for pressure and flow stabilization during the ramp down. After completing the first sequence, the second, third and fourth flow curve measurement were performed 19 minutes, 33 minutes and 52 minutes after the first sequence. At the beginning of each of the four test sequences, cement slurry was sampled from the mixing tank and the flow curve was measured using a Fann 35 viscometer for comparison with results from the instrumented standpipe.

In Fig. 3, we show the flow curves obtained from the first test sequence. The Fann 35 measurements suggest a relatively viscous cement slurry that is nearly viscoplastic. The overlaid solid curve is the Herschel-Bulkley parametrization of the viscometer measurements, $\tau_y = 5.4$ Pa, $K = 0.41$ Pa·s^n and $n = 0.83$. As per the xanthan gum measurements discussed above, the viscometer readings have been corrected for non-Newtonian end effects and wall shear rate. As pointed out above, the flow curve for the cement slurry is

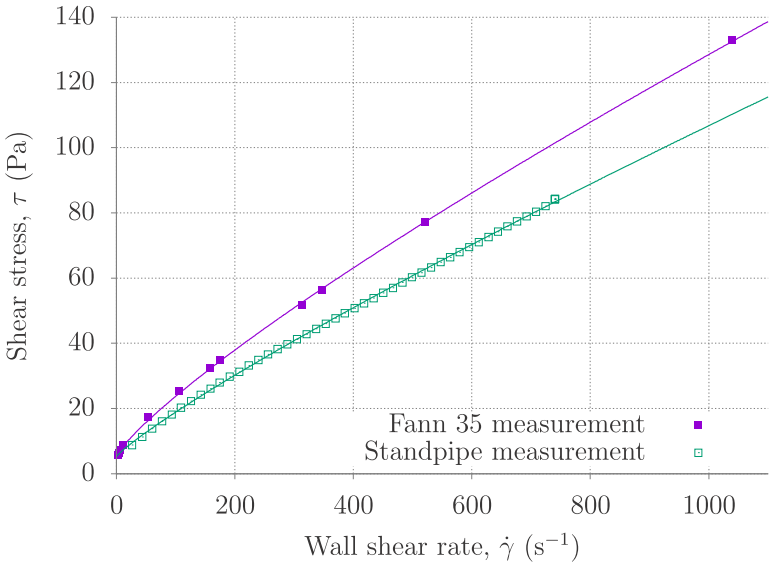


FIGURE 3: Flow curve measurements using viscometer and standpipe for the cement slurry from the first test interval.

acquired by ramping the flow rate down from approximately 38 l/min. At the highest flow rate, we evaluate the wall shear rate to approximately 740 s^{-1}. The final (lowest) wall shear rate from the instrumented stand pipe is estimated to approximately 17 s^{-1}, corresponding to a flow rate of approximately 0.4 l/min. The Herschel-Bulkley model parameters obtained from the instrumented standpipe measurements are $\tau_y = 5.2$ Pa, $K = 0.25$ Pa·s^n and $n = 0.87$, with the resulting model fitting shown in Fig. 3.

As per Fig. 2, for measurements of glycerol and the xanthan gum solution, the flow curves obtained from the viscometer and from the instrumented standpipe show qualitatively similar flow curves, but the flow curves based on viscometer measurements correspond in all cases to effectively larger viscosity than that from the standpipe. In Fig. 3 this is reflected primarily in the parameter estimate for the consistency index, *i.e.* $K = 0.41$ Pa·s^n for viscometer flow curve and $K = 0.25$ Pa·s^n from the stand pipe. Similar results were obtained also for the second, third and fourth test sequence. Flow curves from all four test sequences are presented in Fig. 4. As can be observed when comparing the four flow curves, a slightly decreasing trend was found for the effective cement slurry viscosity, both for the viscometer and the standpipe measurements. This trend may be attributed to a gradual evolution of the slurry viscosity over time due to continued mixing as the slurry is circulated through the flow loop. A minor temperature increase of approximately 5°C was recorded between the first and the last measurement, and this will also contribute to a slight drop in slurry viscosity.

The results compiled in Figs. 2, 3 and 4 consistently show that the flow curves and the corresponding effective viscosity are higher when measured using the viscometer. The cement slurry flow curves are very similar at the lowest shear rates, and the yield stress

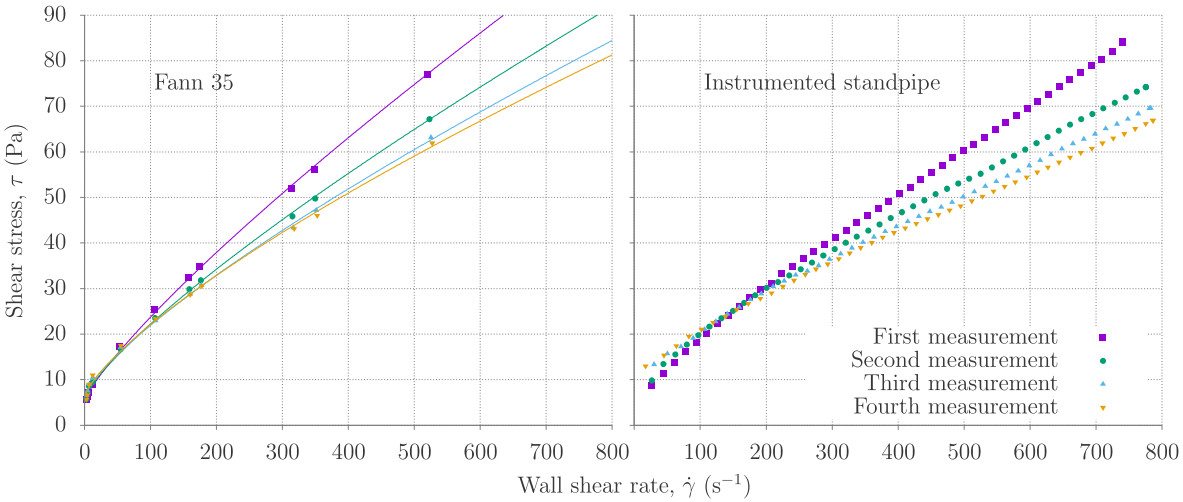


FIGURE 4: Development of cement slurry flow curves during mixing.

parameter ranges from about 4.2 Pa up to 10.6 Pa based on the standpipe, and ranges from 5.4 Pa to about 6.6 Pa for the viscometer measurements. At the lowest flow rates, the wall shear rate in the instrumented standpipe is estimated to approximately 10 s^{-1}, while the corresponding (corrected) lowest shear rate in the viscometer was about 3 s^{-1} (corresponding to 0.9 revolutions per minute). As such, the viscometer provides a slightly better sampling of the low end of the flow curve. At greater shear rates, the difference between the viscometer and standpipe flow curves are expressed primarily through the consistency index, with the viscometer producing the larger values.

Possible reasons for the observed deviations in Figs. 2, 3 and 4 may be due to different concentrations of air entrained in the fluids, imprecise calibration of differential pressure sensors or viscometer, and/or measurement errors due to the pressure tap geometry. The cement slurry measurements may also be affected by slight variations in cement particle concentrations across the sampled volumes and in the instrumented standpipe. Evaluation of these possible explanations is planned as part of future work.

As pointed out above, the vertical pipe segment in the instrumented standpipe concept shown in Fig. 1 was introduced to allow the separate measurement of friction pressure and hydrostatic pressure gradients. This approach offered continuous density measurements that were deviating from lab sample values by 2-3 %, which was poorer quality than provided with the Coriolis flow meter. Further, during testing with cement slurry, the hydrostatic pressure measurement was unfortunately affected by particle plugging of the pressure ports, which rendered this measurement unreliable. Consequently, future studies will rely on the density measurement provided by the Coriolis sensor.

SUMMARY AND CONCLUSIONS

This study has focused on investigating the feasibility of using the instrumented standpipe concept for automatic and continuous measurement of the viscosity of industry fluids, such as oil well cement slurry. Measurements of the flow curve of glycerol, a xanthan gum suspension and a conventional cement slurry with silica addition have shown qualitative agreement between viscometer measurements and flow curves acquired from friction pressure measurements in the standpipe. Future work will aim to improve the correspon-

dence between the two independent measurement methods, and to retrofit the concept to an existing, large-scale batch mixer for mixing of cement slurry and alternative barrier materials.

Acknowledgements

The authors acknowledge the Research Council of Norway (RCN) for financing the Centre for Research-based Innovation "SWIPA – Centre for Subsurface Well Integrity, Plugging and Abandonment", RCN project number 309646, for which the work has been carried out. The centre is also financed by the operating companies AkerBP, Equinor ASA and Wintershall Dea Norway, and includes in addition more than 20 in-kind contributing industry partners. The R&D partners in SWIPA are SINTEF, NORCE, IFE, NTNU and the University of Stavanger. Ullrigg Test Centre is acknowledged for providing technical support that made the experiments possible.

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