Plant design for slurry handling

by P. Slatter*

Journal Paper

Synopsis

The key issue when designing plants for slurry handling, is understanding the slurry environment. Pressure to use less water and operate at higher concentrations directly affects slurry flow behaviour. Using the Bingham plastic rheological model, the impact that slurry rheology has on transitional pipe flow, particle settling in laminar shear flow, losses in valves and fittings, centrifugal pump derating and launder flow are presented.

Introduction

The minerals processing industry is under continuous pressure from environmental, legal and financial quarters to use less water, and designers are obliged to consider the option of operating at higher concentration. As the concentration of fine particle mineral processing slurries increases, viscous stresses also increase, and inevitably become non-Newtonian in nature. For some years, the Flow Process Research Centre (FPRC) at the Cape Peninsula University of Technology has been researching the behaviour of high concentration non-Newtonian slurries in pipes, valves and fittings, pumps and launders, and these will form the focus of the paper.

The aim of this paper is to highlight the important practical aspects of these fundamental issues, and their implications for design for slurry handling. In particular, the objective is to present the principal conceptual issues that underpin sound design. The full detail of the design process for handling slurries is an extensive topic, beyond the scope of this paper, and the reader is referred to sources dealing directly with these for more detail[1].

The important point to make at the outset is that the foundation of sound design for slurry handling does not revolve around the task of choosing or producing special materials and plant—although these are often required. The foundation lies rather in having a good understanding of the slurry environment[1,2], which is the basis of this paper and our work at the FPRC.

Rheological characterization

Slurry rheology (viscous character) is a dynamic property of microstructure[3]. When the slurry is stationary, the attractive forces between particles or agglomerates form a three-dimensional structure, which extends to the walls of the container. The shear stress required to rupture this structure and initiate flow, is called the yield stress. Below this stress, the material behaves like an elastic solid. As shear stresses and shear rates increase, the agglomerates gradually re-orientate and disintegrate, resulting in a decrease in the viscosity of the material. This process is known as shear thinning. At very high shear stresses and shear rates, the re-orientation and disintegration process reaches equilibrium, and the viscosity becomes constant.

Although this portrayal of the relationship between viscosity and microstructure is idealized, it is useful for engineering purposes. The simplest steady state, time independent rheological model, which can accommodate the behaviour described, above is the Bingham plastic model. This model can be formulated in terms of shear stress τ;

$$\tau = \tau_y + K\dot{\gamma}, \qquad [1]$$

or viscosity η;

$$\eta = \frac{\tau}{\dot{\gamma}} = \frac{\tau_y}{\dot{\gamma}} + K, \qquad [2]$$

where τ_y is the yield stress, K is the plastic viscosity and $\dot{\gamma}$ is the shear rate or velocity gradient. The two terms on the right-hand side of Equation 2 will be equal when[3].

$$\dot{\gamma}_b = \frac{\tau_y}{K}. \qquad [3]$$

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* *Cape Peninsula University of Technology.*

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The importance of the boundary shear rate $\dot{\gamma}_b$ is that it marks the boundary between yield stress and plastic viscosity domination of viscosity. This is shown graphically in Figure 1.

The two terms on the right-hand side of Equation 2 each represent the asymptotes shown in Figure 1. It is significant to note from Figure 1 that the ordinate intercept (at a shear rate of unity) of the oblique asymptote is in fact the yield stress value τ_y. This illustrates the fact that—for practical purposes—the viscosity values for shear rates less than the boundary shear rate $\dot{\gamma}_b$ are directly proportional to the yield stress. This serves to emphasize the importance of the yield stress value when operating in the region to the left of the boundary shear rate. The yield stress τ_y is a strong function of particle properties, concentration and solution chemistry, with the plastic viscosity *K* usually being a weaker function of these.

Clearly, knowledge of the slurry rheology is of prime importance to understanding the slurry environment, and its determination would be one of the first steps in the design process.

Transitional pipe flow

The laminar/turbulent transition is of extreme importance for plant design, because at this point the behaviour of the fluid changes fundamentally. The manner in which slurry rheology affects transitional flow behaviour is critically important to understanding the slurry environment. For Newtonian fluids, such as water and oil, the location of the transition is well established and the calculation is simple. For systems conveying non-Newtonian slurries, however, this is not the case. Although an accurate method is available[4], it is computationally somewhat cumbersome. However, for large, industrial sized pipes, the estimation of the laminar/turbulent transition velocity, V_c, using this approach, resolves to a surprisingly simple relationship dominated by the yield stress, and which excludes pipe diameter and plastic viscosity:

$$V_c = 26\sqrt{\frac{\tau_y}{\rho}}, \qquad [4]$$

where ρ is slurry density.

Yield stress domination of the transition behaviour is in fact to be expected, since it has been shown[3] that for large pipes, the transition point always occurs to the left of the boundary shear rate $\dot{\gamma}_b$ shown in Figure 1, in the yield stress dominated region.

Settling in laminar shear flow

Pipelines conveying non-Newtonian slurries in laminar flow will undergo laminar shear flow settling of the coarser particles. In the absence of any effective re-suspending mechanism, a bed will form, which could eventually block the pipe[6]. The AMIRA P599/A project 'High Concentration Suspension Pumping' was initiated precisely to address this problem. These flows can be modelled using a generic two-layer model[7], which is usually applied to heterogeneous, settling slurries, as shown in Figure 2.

Figure 2, a plot of wall shear stress vs bulk shear rate ($8V/D$), illustrates the obscure nature of these flows. Although strongly heterogeneous, settled bed flow was observed for all four pipe sizes, but only the largest pipe shows evidence of the settled bed behaviour in Figure 2. In fact, in the two smaller diameter pipes, the experimental data show convincing evidence of only homogeneous behaviour. If pressure gradient prediction were to be done for the larger pipe on this basis, Figure 2 clearly shows that under-prediction errors of 100 per cent or more will result. It is important to note that this behaviour is true even for slurries with a yield stress that is high enough to support the largest settleable solid particles[7].

Losses in valves and fittings

Losses in valves and fittings can form a significant portion of the total energy losses in the relatively short pipe run lengths typically found in minerals processing plants[8]. This point is well illustrated by recent texts on the subject[9].

An exacerbating factor is that these losses become much more significant in laminar flow. Very little design information is available on loss coefficients in laminar flow, and is urgently needed[10]. The FPRC has already begun producing such data for practical design purposes[11].

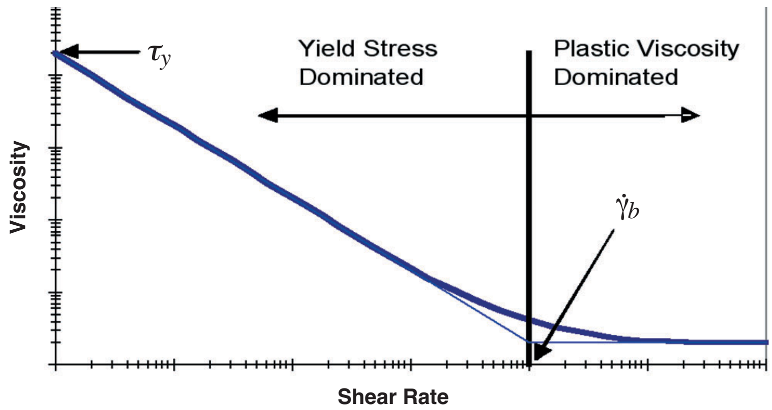

Figure 1—Graphical presentation of Equation 2 showing the boundary shear rate $\dot{\gamma}_b$[3]

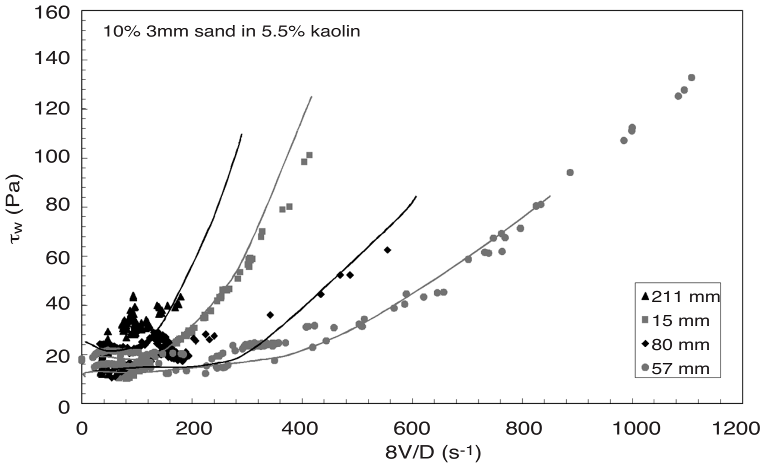

Figure 2—Laminar settling pipe data and two-layer model predictions for sand/clay suspensions in various pipe sizes[7]

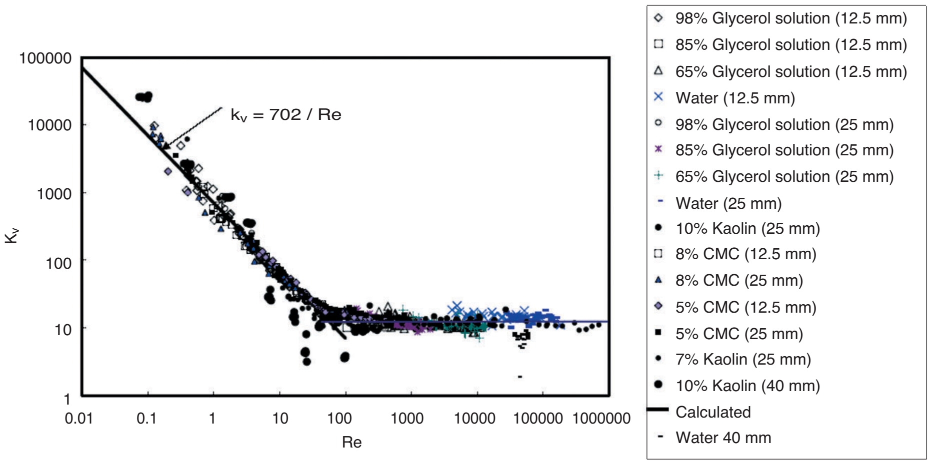

Figure 3—Experimental data from the FPRC for 12.5 mm, 25 mm and 40 mm globe valves[12]

The loss coefficient k_v of a valve (or fitting) defines the head loss across the valve H_{Lv} as a portion of the velocity energy head of the slurry:

$$H_{Lv} = k_v \frac{V^2}{2g}. \qquad [5]$$

Figure 3 shows that the work at the FPRC provided experimental data over eight orders of magnitude of Reynolds numbers for the loss coefficient for three differently sized (12.5 mm, 25 mm and 40 mm) geometrically similar globe valves. This work was carried out using nine different Newtonian and non-Newtonian fluids on two different test rigs, demonstrating that dynamic similarity—in the broadest possible sense—can be achieved for valves and fittings. This data can therefore be used with confidence for accurate and efficient design purposes.

Centrifugal pump derating

Pumping head characteristics must be based on the pump performance for the fluid that must be pumped. Conventionally, manufacturer's pump catalogue curves are presented for clear water performance only. When designing for slurry handling, these curves must be derated, and the usual approach is to determine the head and efficiency ratios (as slurry:water performance), referred to as HR and ER respectively.

For settling slurries, standard methods are available[1], which can usually be used successfully. For non-Newtonian slurries, there is less certainty, and methods based on the slurry rheology are presently under investigation.

The Walker and Goulas[13] pump Reynolds number is an effective method to predict pump performance for non-Newtonian slurries[14]. However, it is based either on the plastic viscosity only, or on an apparent viscosity, both of which present fundamental rheological problems. These problems arise immediately from examination of Figure 1—unless an accurate estimate of the shear rate within the pump can be defined, these values are meaningless. Sery and Slatter[15] present a new analysis, NRe_{p2} based on the pump geometry and the more comprehensive Herschel-Bulkley rheological model. Experimental data from centrifugal pump tests performed at the FPRC have been used to compare these two approaches, as shown in Figure 4.

It is of particular importance to be able to predict the performance of centrifugal pumps, since at high concentration, total cost comparison with PD pump systems need to be made over the life of the plant. Since the main issue in

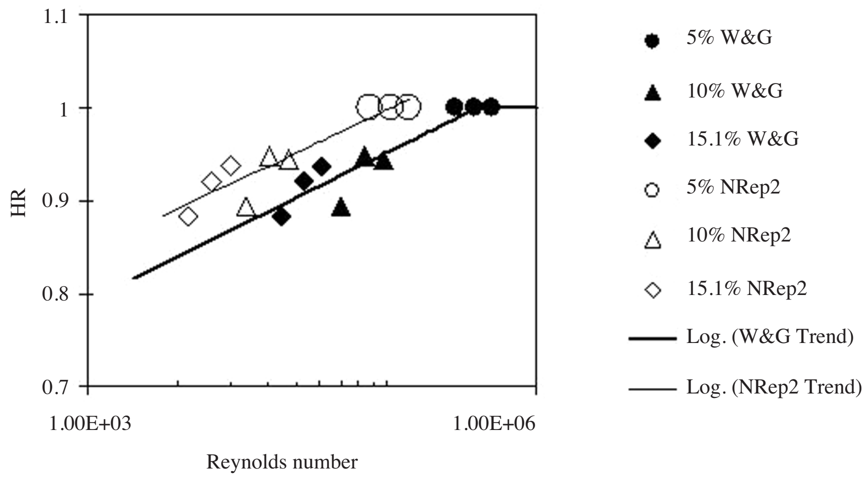

Figure 4—Comparison of the two Reynolds number approaches, using kaolin clay slurry experimental data[15]

these comparisons is the capital cost of the PD pump system versus the running cost of the centrifugal pump, accurate derating of both head and efficiency is of prime importance.

Figure 4 shows that both approaches appear to have similar merit, and work in this area is ongoing. However, what is clear from Figure 4 is that pump performance decreases significantly as the Reynolds number decreases below $1\text{x}10^6$.

Launder flow

Launders are used extensively in the minerals processing industry. However, very little is known of the way in which the rheology of the slurry affects the behaviour of these flows, especially as the slurry becomes more viscous, and laminar flows predominate. Clearly, knowledge of this interaction is vitally important to the understanding of the slurry environment for plant design.

Our approach at the FPRC is to adopt the time-honoured approaches of engineering hydrodynamics, and develop a Reynolds number, which we could use to establish dynamic similarity[16].

For launder flow, the Fanning friction factor f is given by

$$f = \frac{2R_h g \sin\theta}{V^2} \qquad [6]$$

where R_h is the hydraulic radius and θ is the launder slope. An appropriate Reynolds number[16,17] which accounts for the non-Newtonian viscous stresses in launder flow is

$$\mathrm{Re}_{2(YPP)} = \frac{8\rho V^2}{\tau_y + K\left(\frac{2V}{R_h}\right)} \qquad [7]$$

Using these as the basis for our approach, laminar flow data can be compared with

$$f = \frac{16}{\mathrm{Re}}, \qquad [8]$$

and turbulent flow data with the Blasius equation

$$f = \frac{0.079}{\mathrm{Re}^{0.25}}. \qquad [9]$$

Using data from the experimental programme at the FPRC, this approach can be evaluated[17] as presented in Figure 5.

Figure 5 shows that this approach has merit, and adequately models the interaction between slurry rheology and laminar launder flow behaviour. Figure 5 also shows that the transitional behaviour can be expected to be more complex than for pipe flow—and research continues in this direction.

Discussion

One of the hallmarks of research is that it asks more questions than it answers, and this discipline is no exception. This paper started by making definitive statements describing slurry rheology. While these statements are often adequate for engineering purposes, the truth is that slurry rheology depends on physical and chemical state, as well as the flow conditions themselves. There is much research still to be done. However, for those who need to make effective designs that work in the present, it is believed that the issues raised here will provide a basis for understanding the slurry environment—a vital precursor for sound design.

Conclusions

Our point of departure for this paper has been that sound design for slurry handling rests squarely on an understanding of the slurry environment. The manner in which slurry rheology affects flow behaviour in a number of typical plant design hydrodynamic contexts has been presented. Slurry rheology itself is, however, a complex topic, and research in this area is ongoing.

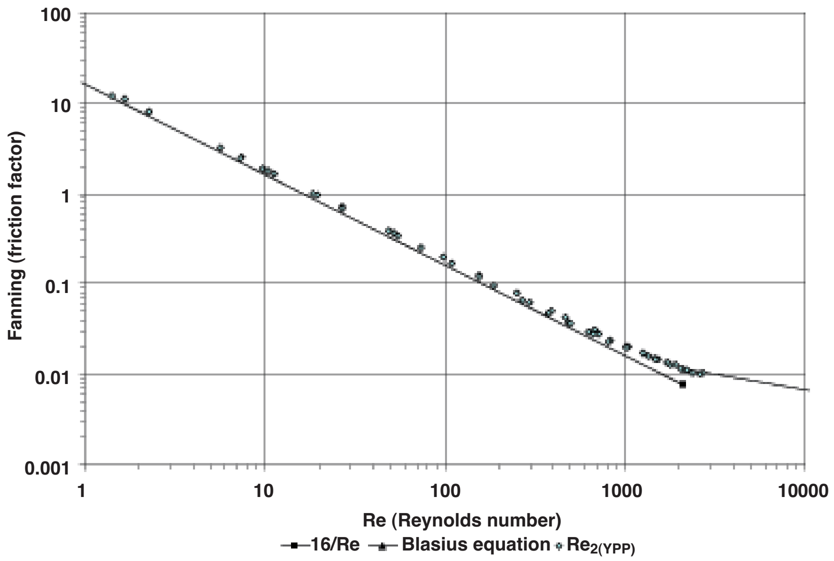

Figure 5—A 6% bentonite slurry flowing in a 300 mm rectangular flume[17]

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References

1. PATERSON AND COOKE Annual Course, *The Design of Slurry Pipeline Design*, Paterson and Cooke Consulting Engineers, Claremont, Cape Town.

2. WASP, KENNY AND GANDHI *Solid-Liquid Flow: Slurry Pipeline Transportation*, Trans Tech Publications, New York. 1977.

3. SLATTER, P.T. Tailings Transport—Back to Basics!; Invited Keynote Address, Paste 2005, *International Seminar on Paste and Thickened Tailings,* Santiago, Chile, 20-22 April, 2005. pp. 165–176.

4. SLATTER, P.T. The role of rheology in the pipelining of mineral slurries, *Min. Pro. Ext. Met. Rev*., vol. 20, 1999. pp. 281–300.

5. SLATTER P.T. and WASP. E.J. The laminar/turbulent transition in large pipes; *10th International Conference on Transport and Sedimentation of Solid Particles*—Wrocław: 4-7 September, 2000. pp. 389–399.

6. COOKE, R. Laminar flow settling: the potential for unexpected problems, *British Hydromechanics Research Group 15th International Conference on Slurry Handling and Pipeline Transport HYDROTRANSPORT 15*; Banff, June, 2002. pp. 121–133.

7. PULLUM, L., GRAHAM, L.J.W., AND SLATTER, P.T. A non-Newtonian two-layer model and its application to high density hydrotransport, *British Hydromechanics Research Group 16th International Conference on Hydrotransport;* Santiago, April, 2004. pp. 579–594.

8. SLATTER, P.T., PIENAAR, V.G., and PETERSEN, F.W. Non-Newtonian fittings losses; *9th International Conference on Transport and Sedimentation of Solid Particles*, Cracow: 2–5 September, 1997. pp. 585–596.

9. CHHABRA, R.P. and RICHARDSON, J.F. *Non-Newtonian Flow in the Process Industry,* Oxford, Butterworth-Heinemann 1999.

10. PIENAAR, V,G., SLATTER, P.T., ALDERMAN, N.J. and HEYWOOD, N.I. A Review of Frictional Pressure Losses for Flow of both settling and non-settling Non-Newtonian Fluids through Pipe Bends, *British Hydromechanics Research Group 15th International Conference on Slurry Handling and Pipeline Transport HYDROTRANSPORT 15;* Banff, June, 2002. pp. 511–527.

11. PIENAAR, V.G., NTLALETŠENG, S., and SLATTER, P.T. Resistance coefficients of mineral tailings flowing in half open diaphragm valves, *MINPROC '05*, SAIMM Western Cape, 3 August 2005.

12. PIENAAR, V.G., SLATTER, P.T., ALDERMAN, N.J. and HEYWOOD, N.I. A Review of Frictional Pressure Losses for Flow of Newtonian and non-Newtonian Slurries Through Valves, *British Hydromechanics Research Group 16th International Conference on Hydrotransport,* Santiago, April, 2004. pp. 219–230.

13. WALKER, C.I. and GOULAS, A. Performance characteristics of centrifugal pumps when handling non-Newtonian slurries, *Proc. Inst. Mech. Engrs.* 1984.

14. SERY, G.A. and SLATTER, P.T. Centrifugal pump derating for non-Newtonian slurries, *British Hydromechanics Research Group 15th International Conference on Slurry Handling and Pipeline Transport HYDROTRANSPORT 15;* Banff, June, 2002. pp. 679–692.

15. SERY, G,A, and SLATTER, P.T. Centrifugal pump performance Reynolds number for non-Newtonian slurries, *12th International Conference on Transport & Sedimentation of Solid Particles,* Prague, Czech Republic, 20–24 September, 2004. pp. 601–609.

16. HALDENWANG, R., CHHABRA, R.P., and SLATTER, P.T. Open channel flow of non-Newtonian fluids, *10th International Conference on Transport and Sedimentation of Solid Particles*, Wrocław: 4–7 September, 2000. pp. 269–280.

17. HALDENWANG, R.H. and SLATTER, P.T. Experimental procedure and database for non-Newtonian open channel flow, *Journal of Hydraulic Research,* vol. 44, no. 2, 2006. ◆