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A dynamic model for level prediction in aerated tanks

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ABSTRACT

Stirred aerated tanks are a key unit operation in many industries, including froth flotation. Reliable and robust level control is of great importance in maintaining steady operation for successful implementation of higher level optimising control strategies, particularly when such tanks are arranged in series. When changes are made to the rate of aeration, there is a corresponding change in the pulp bubble size and gas holdup (the volume fraction of air in the tank), and consequently the pulp height. Stable operation of flotation tanks must, therefore, include the effect of air rate on pulp height in level control systems, especially if air rate is being actively controlled. In this paper, a model is developed from first principles to link the change in gas holdup with variation in air rate under dynamic conditions, accounting for the variability in gas holdup with height that results from differences in gas compressibility. This is validated experimentally.

In order to test the model, experiments were carried out using a 70 L laboratory tank comprising water and reagent systems. For both simple and complex changes in air rate, the model showed good agreement with the experimental results when predicting the change in pulp height at steady state. Under dynamic conditions, the experimental system exhibited a slightly slower response than is predicted by the model; this is likely to be due to the well mixed assumption not being adequately met.

This model provides a method to improve the operating stability of aerated tanks through better modelling of the dynamic pulp height changes that result from changes in air flowrate. In flotation tanks, this will enable greater control over froth height, which has been found to affect significantly mass pull, froth stability and flotation performance.

1. Introduction

Stirred tanks containing aerated slurry are found in numerous industrial operations, including froth flotation, tar sands recovery, waste water treatment and food production. Control of the flow into and out of such tanks is often by adjustment of the slurry outlet valve using cell level as the control variable. In froth flotation, one of the largest tonnage separation processes, the froth phase determines the separation performance of the desired minerals from the gangue. Control of the height of the pulp phase is critical, since it not only determines residence time, but also the depth of the overflowing froth phase, a key operating variable. There is a clear link between froth phase depth and flotation performance (e.g. Feteris et al., 1987; Hadler et al., 2012; Venkatesan et al., 2014), where, in general, deep froths result in higher grade concentrates but lower recoveries.

Flotation cells traditionally use proportional integral control (or PI control) to ensure cell levels remain at desired set-points (Kämpjärvi and Jämsä-Jounela, 2003; Carr et al., 2009; Shean and Cilliers, 2011). This is achieved by manipulating the out flow from the cell by adjustment of the slurry outlet valve. In aerated tanks, when changes are made to the gas addition rate, there is a corresponding effect on the pulp volume. This is associated with changes in both gas holdup (the volume fraction of air at a given tank height) (Vinnett et al., 2014) and bubble size (Gorain et al., 1995; Nesset et al., 2006). Vinnett et al. (2014) showed, for example, that both gas holdup (ε_g) and pulp bubble size (d_B) increased initially with increasing aeration after which further increases in aeration resulted in larger bubbles and lower ε_g. An increase in bubble size with increasing air rate was also shown in the extensive studies of Gorain et al. (1995) and Nesset et al. (2006). Gorain et al., (1995) additionally noted that the bubble sizes ranges also increased.

In order for pulp level to be well-controlled under changes in the rate of aeration, it is necessary to account for changes in pulp volume that occur as a result of the changes in both the air rate and slurry rate into the cell. Many modern advanced pulp level control systems do not account for the effects of varying air flowrate on ε_g and react retroactively. Kämpjärvi and Jämsä-Jounela (2003), for example, develop a model of flotation cells in series in order to test different control

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strategies. They do not, however, account for aeration in the tanks, stating that "the impact of the air feed on the pulp level is ignored". Stenlund and Medvedev (2002) discuss the merits of multivariable control strategies for level control in a flotation bank and consider the impact of feed rate but do not include any other operating variables such as air flowrate. Mintek's FloatStar control system, on the other hand, makes use of an "aggressive" PID controller for level control (Knights et al., 2012). The net result is that both traditional and modern advanced pulp level control systems do not offer as tight a control response as is possible. In addition, the dynamics of any change in cell level as a response to a change in air flowrate and ε_g are significantly faster than changes in cell level as a result of a change in pulp flows into or out of the flotation cell. As such, any control system solely controlling on the pulp flows into or out of the flotation cell may over compensate in its response actions when a rapid change in cell level is detected as a response to an air flowrate change.

Developments in process control, particularly with regards to predictive control and artificial intelligence, continue apace (Bergh and Yianatos, 2011; Jovanovic and Miljanovic, 2015). Furthermore new flotation control and optimisation systems are seeking to optimise air flowrate changes using machine vision to measure flotation response (Shean et al., 2017). A predictive model that can rapidly and robustly determine the effects of air flowrate changes on gas holdup will allow more accurate and precise dynamic level prediction and consequently improved operating stability and performance of flotation banks.

This paper describes the development and validation of a model, based on first principles, that predicts the dynamic level response due to changes in both pulp flowrates and air addition rates.

2. Model development

For an aerated tank, the total system volume (V_{system}) comprises contributions from both the gas (V_{gas}) and pulp (V_{pulp}), as shown in Eq. (1). Additionally, V_{system} is equal to the product of the height of the aerated pulp (h) and the cross sectional area of the cell (A).

$$V_{system} = hA = V_{gas} + V_{pulp} \tag{1}$$

The change in the gas volume in the cell can be expressed in terms of a mass balance where Q_{in} and Q_{out} are the volumetric flowrates of gas into and out of the cell respectively. Both of these flowrates will assume that the gas is at STP. This will be close to true for the out flowing gas, while the flowmeters used to measure the inflowing gas are typically calibrated to STP (a hydrostatic pressure for the incoming gas is also not appropriate due to the suction effect of the impeller). In this mass balance, the ratio between the air density at STP (assumed to be the same as the surface density), ρ_{air0}, and the average density of the air in the pulp, ρ_{airAve}, needs to be considered, especially in industrial cells where the hydrostatic pressure head provides an appreciable pressure contribution in addition to atmospheric pressure:

$$\frac{d\left(V_{gas}\frac{\rho_{airAve}}{\rho_{air0}}\right)}{dt} = Q_{in} - Q_{out} \tag{2}$$

The ratio of the average to surface air density is a function of the height of the aerated pulp and the gas holdup near the surface of the pulp, ε_0, as well as the slurry density, ρ_{slurry}, and the atmospheric pressure, P_0 (see Appendix A for the derivation of this equation):

$$\frac{\rho_{gAve}}{\rho_{g0}} = X_\rho \approx \frac{\dfrac{\rho_{slurry}g(1-\varepsilon_0)h}{P_0 ln\left(\frac{\rho_{slurry}g(1-\varepsilon_0)h}{P_0}+1\right)} - \varepsilon_0}{1-\varepsilon_0} \tag{3}$$

Furthermore, Q_{out} can be expressed as the product of upward velocity of the gas at the surface of the pulp zone, v_{gas}, A and ε_0; where v_{gas} is dependent on both ε_0 and the bubble size (d_B). Note that $J_g = v_{gas}\varepsilon_0$ under steady state conditions.

$$\frac{d(V_{gas}X_\rho)}{dt} = Q_{in} - Av_{gas}\varepsilon_0 \tag{4}$$

In terms of the depth of the cell, it is the total system volume rather than the gas volume that is important:

$$\frac{dV_{gas}}{dt} = \frac{d(V_{system}\varepsilon_{Ave})}{dt} = A\frac{d(h\varepsilon_{Ave})}{dt}. \tag{5}$$

If the cross-sectional area of the pulp is assumed to be constant with respect to height, then this can be further simplified to:

$$\frac{dV_{gas}}{dt} = A\frac{d(h\varepsilon_{Ave})}{dt} \tag{6}$$

As the bubble rise velocity is a function of the surface rather than the average gas holdup it is convenient to introduce the ratio of the average to surface gas holdup (again derived in Appendix A):

$$\frac{\varepsilon_{Ave}}{\varepsilon_0} = X_\varepsilon \approx \frac{P_0}{\rho_{slurry}g(1-\varepsilon_0)h} ln\left(\frac{\rho_{slurry}g(1-\varepsilon_0)\text{h}}{P_0}+1\right) \tag{7}$$

This means that:

$$\frac{dV_{gas}}{dt} = A\frac{d\left(\frac{h\varepsilon_0}{X_\varepsilon}\right)}{dt} \tag{8}$$

In addition, the height of the system can be expressed in terms of the 'gas free' height of the pulp (h_0) and ε_0:

$$h = \frac{h_0}{1-\varepsilon_{Ave}} = \frac{h_0}{1-\frac{\varepsilon_0}{X_\varepsilon}} \tag{9}$$

Combining Eqs. (4), (8) and (9) results in Eq. (10); an expression is derived for the change of ε_0 with time as a function of Q_{in}:

$$\frac{d\left(\frac{X_\rho\varepsilon_0}{X_\varepsilon}\right)}{dt} = \frac{\left(\frac{\varepsilon_0}{X_\varepsilon}-1\right)^2}{h_0}\left[\frac{Q_{in}}{A} - v_{gas}(\varepsilon_0, d_B)\varepsilon_0\right] - \frac{dh_0}{dt}\frac{1}{h_o}\frac{\varepsilon_0}{X_\varepsilon}\left(1-\frac{\varepsilon_0}{X_\varepsilon}\right) \tag{10}$$

Furthermore, h_0 can be related to the mass balance of the pulp phase, whereby the rate of change of V_{pulp} is equal to the difference in the volumetric flow of pulp into ($Q_{pulp,in}$) and out of ($Q_{pulp,out}$) the flotation cell (Eq. (11)), where $Q_{pulp,out}$ includes both the tailings and concentrate flowrates. This expression assumes the density of pulp into and out of the system is constant, and should be corrected if the density difference between the streams is significant.

$$\frac{dh_0}{dt} = \frac{Q_{pulp,in}}{A} - \frac{Q_{pulp,out}}{A} \tag{11}$$

Substitution of Eq. (11) into Eq. (10) allows for further simplification:

$$\frac{d\left(\frac{X_\rho\varepsilon_0}{X_\varepsilon}\right)}{dt} = \frac{\left(\frac{\varepsilon_0}{X_\varepsilon}-1\right)^2}{h_0}\left[\frac{Q_{in}}{A} - v_{gas}(\varepsilon_0, d_B)\varepsilon_0\right] - \frac{1}{h_o}\frac{\varepsilon_0}{X_\varepsilon}\left(1-\frac{\varepsilon_0}{X_\varepsilon}\right)\left(\frac{Q_{pulp,in}}{A} - \frac{Q_{pulp,out}}{A}\right) \tag{12}$$

The upward gas velocity of a bubble in a flotation cell can be expressed in terms of ε_0 and the terminal rise velocity (Richardson and Zaki, 1954; Pal and Masliyah, 1989) with the terminal rise velocity, in turn, being expressed in terms of d_B, the gravitational force (g), the pulp density (ρ_{pulp}) and viscosity (μ_{pulp}) (Eq. (13)). The density of the air is assumed to be insignificant compared to that of the slurry. Importantly, v_{gas} is strongly dependent on the bubble size, with $v_{gas} \propto d_B^2$.

$$v_{gas} = \frac{gd_B^2\rho_{pulp}}{18\mu_{pulp}}[1-\varepsilon_0]^{1.39} \tag{13}$$

Since the dependency of gas velocity on bubble size is not linear, the velocity at the average bubble size will not be the same as the average rise velocity. Furthermore, even though perfect mixing is assumed, the