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OVERVIEW OF SLURRY STORAGE TANK DESIGN AND OPERATION FOR HYDROTRANSPORT SYSTEMS

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ABSTRACT: This paper provides a summary of slurry tank design for feeding and receiving slurry transported through pipelines. It considers the solids suspension mechanisms, the relevant physical properties of the solid and liquid phases, and the type, number and position of appropriate agitators. Slurry tank operation covers the degree of solid suspension required such as "just-suspension" and cloud height, including the prediction of "just-suspension" agitator speed. The effect of the partial discharge and complete draining of a slurry tank, and therefore a progressive decrease in the slurry level in the tank, on the extent of particle suspension is also covered.

KEY WORDS: slurry, tank, design, agitators, suspension levels

NOTATION

Ar	Archimedes number (-)
C	Agitator clearance (m)
C_v	Solid volume fraction (-)
C_{vo}	Bulk (average) solids volume fraction (-)
d_p	Particle size (m)
d_{xx}	Different particle sizes in size distribution, d_{10}, d_{32}, d_{43}, d_{50}, d_{90}, d_{max} (m)
D	Agitator diameter (m)
g	Acceleration due to gravity (m/s^2)
H	Height of slurry in tank (m)
H_c	Cloud height (m)
N	Agitator rotational speed (rps)
N_{JS}	Just off bottom suspension agitator speed (rps)
N_p	Power number (-)
P	Power input to slurry by agitator(s) (W)
Re	Agitator Reynolds number
S	Parameter in Zwietering equation for N_{JS}
T	Internal tank diameter (m)
X	Solids concentration in Zwietering equation (-)
X_v	Solids concentration in GMB correlation (%)
y	Vertical distance from tank base (m)
μ_s	Dynamic slurry viscosity (Pa s)
ν	Kinematic slurry viscosity (m^2/s)
$\Delta\rho$	Difference between solids and liquid densities (kg/m^3)
ρ_l	Liquid density (kg/m^3)

ρ_s Slurry density (kg/m^3)
z Parameter in GMB correlation for N_{JS}

1. INTRODUCTION

One or more slurry storage tanks are key components of a hydrotransport system. In previous BHR Group Hydrotransport conferences, Harrah (1974) and von Essen and Ricks (1999) discussed the selection and sizing of agitators for large slurry tanks. However, in the last 24 years slurry tank design and operation have been largely overlooked in Hydrotransport conferences despite significant technical advances in this area. For continuous slurry pipelining the pumps require an adequate reserve of slurry, ideally in a homogeneous or near homogeneous condition. Slurry surge storage tanks are installed to smooth out variations or stoppages in the feed rate (von Essen et al., 1999). The "live storage capacity" is defined as the tank volume between maximum and minimum operating levels which determine the corresponding slurry residence times. To determine this capacity, it has been suggested that the upstream and downstream facilities be considered in terms of their capacity and reliability, together with the pipeline system reliability and pipeline segment volumes, and that is normally about 8 to 24 hours of the pipeline design capacity. Slurry tanks are also useful to mix or blend slurries prior to pumping, to provide first stage pump suction pressure, and to provide an opportunity for entrained air to disengage from the slurry before entering the pumping system. Detrimental effects of solids on agitators can include wear of the blades (Fořt et al., 2001) and damage related to starting the agitator when submerged in a settled bed of solids.

In some systems, such as fine particle tailings systems, including mine backfilling applications, there isn't always a need to maintain the slurry in suspension in a tank. The primary consideration is that the tank should not choke or block. Here the use of an agitator is often avoided because it represents additional mechanical equipment to maintain and which is vulnerable to unplanned stoppages. In such circumstances, it may be necessary to drain the tank to restart an agitator if the agitator is partially or wholly buried in a settled bed of solids as a result of unplanned power outage.

2. SOLID AND LIQUID PHYSICAL PROPERTIES

Slurry storage includes many of the parameters encountered in pumping and pipeline flow such as particle size and size distribution, particle density, and wear problems. To characterise or design a slurry tank, the relevant physical properties of both the solid and liquid phases (and the resulting slurry) must first be established, although the degree to which correlations are available to characterise their impact varies significantly.

2.1. PARTICLE SIZE DISTRIBUTION AND SHAPE

Solid particles usually have a distribution of sizes, and those used for experiments in the laboratory frequently have quite wide size distributions. Caution should be used when measuring or interpreting particle size distributions. Many methods look at only a part of the distribution (for example, screens usually characterise all fine particles under about 45 or 74 μm as "fines"). This practice is common in the minerals industry. An example size distribution, with various characteristic dimensions indicated, is shown in Figure 1.

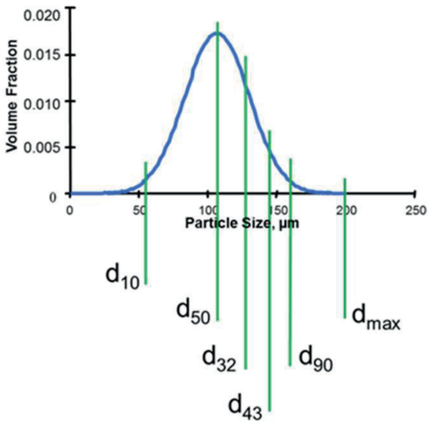

Figure 1 Typical particle size distribution

All the available design correlations employ a single characteristic particle size such as the Sauter mean diameter, d_{32}, rather than a distribution. Baldi et al. (1978) have shown that the characteristic moment of the distribution of solids in the measurement and correlation of the "just-suspension" agitator speed corresponds to d_{43}, but it should be emphasised that this is not the particle size that should be used for design purposes, which should usually be based on d_{max} as a minimum, and often with an appropriate confidence factor applied.

2.2. SOLIDS CONCENTRATION

The volumetric or mass fraction of solids in slurries is easy to quantify. However, when calculating the power input from an agitator in the turbulent regime, the slurry density surrounding the agitator (and hence based on the local solids concentration in this region, rather than the vessel average) should be used. In some cases where the distribution is particularly poor, and hence where this issue is most significant, it may not be possible to know the slurry density surrounding the agitator. This effect can often be ignored and the average density used, but an estimate of the cloud height (see Section 4.3) may help determine if it needs to be considered.

2.3. LIQUID AND PARTICLE DENSITIES

Both solid and liquid phase densities are relatively simple to measure or define. Where the solids are flocs or agglomerates, or the particles are porous, the effective density of the "particles" should be applied (which factors in the mass of liquid contained within the particles) and if this information is not available the use of the primary particle density (and size) will be conservative for suspension processes.

2.4. PARTICLE SETTLING RATES AND PARTICLE DRAG COEFFICIENTS

Clift et al. (2005) discuss methods for calculating the free settling velocity of solids under a wide range of conditions. This velocity occurs from a balance of gravity against fluid drag in a quiescent liquid. The chief fluid dynamic function covering this is the particle drag coefficient which varies with the particle Reynolds number. Early researchers tried to correlate terminal settling rate with minimum suspension velocity in an agitated tank, but were unsuccessful in explaining large data sets. The process is one of lifting off the tank bottom and re-suspension, rather than preventing settling. The effects of a number of variables involved in both settling and suspension correlations are contradictory or inconsistent:

a) The effect of changes in density difference on the just-suspended agitator speed are much stronger than the effect of particle size, but in all models of settling velocity, the particle size has a stronger effect than density difference.

b) The effect of viscosity on settling rate is strong in the Stokes and intermediate regimes but is weak or non-existent in turbulent stirred tanks.

c) The effect of a solids concentration increase in one-dimensional sedimentation is to reduce the settling velocity (hindered settling), yet an increase in concentration will usually increase the required minimum agitator speed for solids suspension.

When solids are moved by fluid in an agitated tank a modified drag coefficient will come into play, but it is for a turbulent, recirculating velocity field - not a sphere in an infinite, unidirectional flow. It has been shown (Clift et al., 2005) that ambient turbulence and edge effects can result in drag coefficients very different from those measured in free-flowing systems.

3. AGITATION OF SETTLING SOLIDS

3.1. MECHANISMS OF SUSPENSION

Solid suspension requires the input of mechanical energy into the fluid-solid system by some mode of agitation. The input energy creates a turbulent flow field in which solid particles are lifted from the tank base and subsequently dispersed and distributed throughout the liquid (Nienow, 1985). Solids pick-up from the tank base in the turbulent and transitional regimes is achieved by a combination of the drag and lift forces on the solid particles caused by the mean flow and turbulent eddies originating from the bulk flow in the vessel. An illustration of the sudden pick-up by turbulent bursts is shown in Figure 2 (Cleaver and Yates, 1973). A turbulent burst is the sudden penetration of a high-energy turbulent eddy into the laminar sub-layer. Evidence of suspension through turbulent eddies in the inertial subrange in stirred tanks can be found in a number of published papers.

The distribution and magnitude of the mean fluid velocities and large anisotropic turbulent eddies generated by a given agitator determine to what degree solid suspension may be achieved. Thus, different agitator designs achieve different degrees of suspension at similar energy inputs. For the same reason, for any given agitator the degree of suspension will vary with both agitator-to-tank diameter and clearance-to-agitator diameter at constant specific power input.

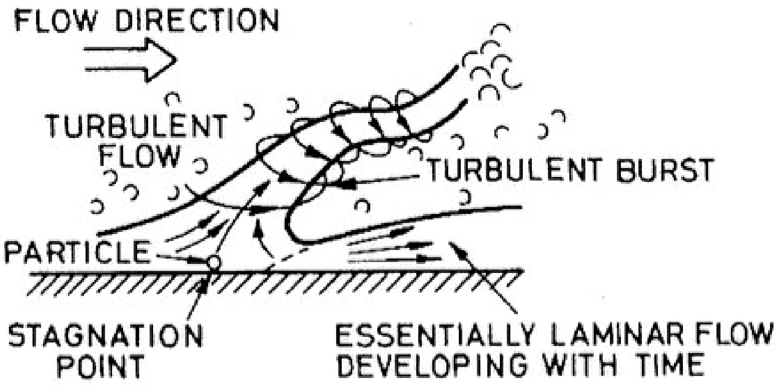

Figure 2 Sudden pick-up of solids by turbulent bursts (Cleaver and Yates, 1973)

3.2. DIMENSIONLESS GROUPS

Because no universally accepted mechanism for solids suspension exists empirical correlations based on experimental data and crude mechanistic models are employed. These correlations use dimensionless numbers.

The agitator Reynolds number indicates whether the flow at the agitator is laminar (Re < 10), turbulent (Re > 10,000) or transitional (10 < Re < 10,000):

$$Re = \frac{\rho_S N D^2}{\mu_S} \tag{1}$$

Traditionally the agitator swept diameter, D, is used as the length dimension in the relevant dimensionless number. In some cases, the more logical dimension would be the particle size. One would expect a particle Reynolds number based on particle diameter, d_p, and particle velocity, V_p, though this is rarely used as the velocity is unknown and maybe unknowable.

The ratio of agitator diameter to tank diameter (D/T) is important as it affects the flow patterns. Sometimes the particle diameter is made non-dimensional with the tank diameter directly, though by doing this any mechanism is likely to be ignored, and the agitator diameter must still be considered.

The Archimedes number, sometimes called the Galileo number, is a grouping of solid and liquid fluid properties without any velocity or agitator diameter:

$$\boldsymbol{Ar} = \left(\frac{g\Delta\rho}{\rho_l}\right)\frac{{d_p}^3}{\nu^2} \tag{2}$$

Unlike most dimensionless numbers which are ratios of competing forces or phenomena, the Archimedes number is a purely physical property group, and has been useful in correlations for deposit velocity in pipelines and for the "just-suspended" agitator speed. In quiescent settling, it is found that there are critical Archimedes numbers corresponding to the flow transitions observed. In addition to these dimensionless numbers there are the various dimensionless ratios describing tank geometry such as the agitator clearance to the agitator diameter (C/D) and the relative fill level (H/T) in the tank. The agitator power number, N_p, is defined by

$$N_p = \frac{P}{\rho_s N^3 D^5} \tag{3}$$

and this number is useful for determining the agitator power input, P.

4. SOLIDS DISTRIBUTION IN TANKS

4.1. DEGREE OF SUSPENSION

In agitated slurry tanks, the degree of solids suspension is often classified into three levels: on-bottom motion, complete off-bottom suspension and uniform suspension (Fig. 3). On bottom motion (Fig. 3a) is characterised by the visual observation of the complete motion of all particles around the bottom of the vessel. It excludes the formation of fillets, a loose aggregation of particles in corners or other parts of the tank bottom. There is little literature on the prediction of agitation conditions to achieve just on bottom motion.

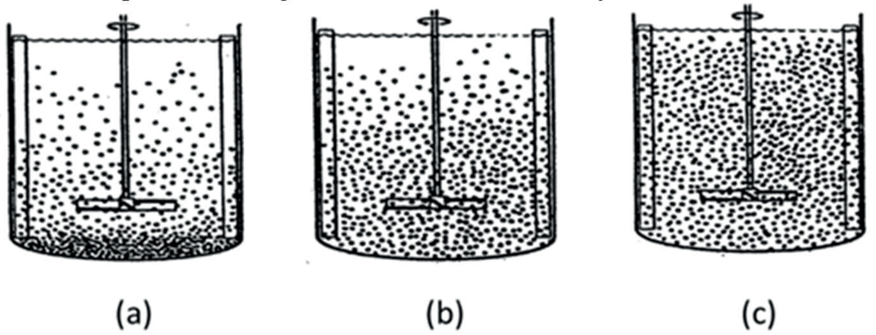

Figure 3 Degrees of Suspension. (a) partial suspension: some solids rest on the tank bottom for short periods (b) complete suspension: all solids are just off the tank bottom; minimum desired condition for most solid-liquid systems. (c) uniform suspension: solids suspended uniformly throughout the tank

Complete off-bottom suspension (Figure 3b), or the “just-suspended” speed, N_{JS}, is characterised by the complete motion of all particles, with no particle remaining on the tank base for more than 1 to 2 seconds (the “Zwietering criterion”, 1958). While solids may all be moving on the base, the distribution of solids throughout the tank is usually non-uniform and often with a clear upper layer. The height of this layer is referred to as the cloud height. This is discussed in Section 4.3.

Uniform distribution (Figure 3c) corresponds to the state of suspension at which solids concentration and particle size distribution are practically uniform throughout the tank, so that any further increase in agitation speed or power does not appreciably enhance the solids distribution in the fluid. In many cases, the achievement of a completely uniform distribution of solids throughout the tank is impractical or simply unachievable because of agitator speed or power limitations, but close to completely uniform conditions must be achieved in some cases. Examples include the need to allow representative sampling, or to feed material at a constant concentration to a slurry pipeline or to a downstream process such as a centrifuge or a hydrocyclone.

There are three commonly employed measures of solid particle distribution in an agitated slurry tank: the just-suspended agitator speed, the cloud height and the uniformity of the distribution of solids throughout the vessel. In designing an agitation system for a given slurry, the first design criterion is the solids suspension level that needs to be achieved, particularly during tank discharge, when the outflow solids concentration and particle size distribution normally needs to match closely the bulk values for the slurry in the tank. Low levels of agitation provide the "just-suspended" criterion, whereas high intensity agitation will give a high degree of homogeneity in the tank but often at a high power consumption.

4.2. JUST-OFF-BOTTOM SUSPENSION

The standard metric with regard to the "just-suspended" speed, N_{JS}, comes from Zwietering (1958), where it is defined as follows: "when no deposits remained on the bottom (of the tank) for more than one second the suspension was considered complete". This condition is usually considered to be the minimum design requirement, but there are applications (e.g., the minerals industry) where process economics dictate that operation at agitator speeds a degree below N_{JS} is required, and solids are allowed to build up in the tank corners. The Zwietering equation has been widely used to predict N_{JS}:

$$N_{JS} = S\frac{\nu^{0.10} d^{0.20} X^{0.13} (\frac{g\Delta\rho}{\rho_l})^{0.45}}{D^{0.85}} \tag{4}$$

where ν is the liquid kinematic viscosity, d the particle size, X a solids concentration parameter (volume of solids/volume of liquid times 100%), **Δ**ρ the density difference between the solids and the liquid, D the agitator diameter, and S is a parameter, assumed to be scale independent, which depends on the agitator geometry. Framatome Ltd's FMP consortium has measured the parameter "S" over a very wide range of conditions.

Grenville et al. (2015) took a very large data set, in part obtained within FMP, with much larger scale data for 45-degree, pitched-blade turbines and hydrofoils in standard baffled vessels of 0.31, 0.61 and 1.00m diameter in which the "just-suspended" speed has been measured visually using Zwietering's criterion and developed the GMB correlation, which indicates that N_{JS} is independent of liquid viscosity:

$$N_{JS} = \frac{z}{N_p^{0.333} D^{0.667}} \left(\frac{\Delta\rho}{\rho_l}\right)^{0.5} d_p^{0.167} X_v^{0.154} \left(\frac{C}{D}\right)^{0.1} \tag{5}$$

Here, N_P is the agitator Power number, defined by Eqn (3), X_v is a solids concentration parameter (volume of solids / volume of slurry times 100%), C is the agitator clearance, and where the parameter, z, for a 45-degree pitched blade turbine is 1.53, and for a hydrofoil such as the SPX Flow's A310, Chemineer's HE3, etc. is 1.21. While the Zwietering (1958) correlation is purely empirical, the Grenville et al. (2015) correlation has a sound mechanistic basis. Both have significantly different scale-up implications. Giacomelli et al. (2022) using large and dense solid particles in tests found that the range over which the GMB correlation could be applied increased significantly, and Giacomelli et al. (2023) investigated and evaluated several improvements to the GMB correlation to generalise it. They developed two geometry-specific as well as generalised suspension correlations for N_{JS} based on experiments made with 225 µm sand in water in four vessel geometries: cylindrical flat-based, cylindrical torispherical dish-based, square flat-based, and horizontally oriented cylindrical vessels. They defined a vessel sphericity parameter to generalise the various tanks into a single correlation permitting interpolation to other vessel types and sizes not tested.

Great care should be taken when considering systems where the tank fill level, H, is not equal to the tank diameter, T. This will be the case when a slurry tank is feeding a pipeline but the slurry is being replenished in the tank at a lower rate. Most of the published work on solids suspension has been based on systems where H = T, while it is possible that H is greater or less than T due to filling and/or draining operations. With axial flow agitators, as the fill level is reduced below H = T, N_{JS} initially reduces slightly with the fill level, but once the fill level reduces to about H = 0.5T the agitator suction is throttled to such an extent that the solids potentially cannot be suspended at any speed.

When there is a mixture of solids of two or more different densities to be suspended the simple equations for predicting solids suspension need to be altered. Garcia-Barberena (2009) and Etchells et al. (2010) suggested using a combination of the minimum suspension speed of each component by itself, so that the combined N_{JS} is given by:

$$N_{JS\,Mix} = \left(N_{JS\,a}^{A} + N_{JS\,b}^{B}\right)^{1/a} \tag{6}$$

where N_{JSa} and N_{JSb} are the "just-suspended" agitator speeds for the various individual components and A , B and a are fitting constants. Both Ayranci and Kresta (2014) and Meyer et al. (2007) found that a value of 3 for all three constants fit the data very well for a wide range of materials and never under predicted the mixture minimum suspension speed. The implication of Eqn (6) calculations cannot be based on just the density of the heaviest particles. The agitator speed required for the "just-suspended" conditions is increased by the presence of the lower density particles.

4.3. CLOUD HEIGHT

At solids concentrations greater than about 5% by volume there is often a clear demarcation between a lower region rich in solids and an upper region almost empty of solids. This maximum height that the agitated solids achieves in the tank is referred to as the cloud height, H_c, which is a simple and useful measure of performance largely because it can be relatively easy and unambiguous to measure, compared with the three-dimensional solids concentration distribution.

FMP consortium research has shown (Brown, 2018) that when the agitator speed exceeds N_{JS} the dimensionless cloud height, H_c/T, increases with liquid viscosity. In addition, for the same solid/liquid system and scale at constant agitator power input per unit mass of slurry (P/M) a single pitched-blade turbine or hydrofoil gives the same cloud height. The published correlation for H_c for single agitators by Bittorf and Kresta (2003) is not recommended because it assumes that cloud height can be based on N_{JS} which is not the case. The use of multiple agitators is often the only way to get a good solids distribution.

4.4. SOLIDS DISTRIBUTION

A complete picture of the quality of a solids-liquid distribution only can be obtained by measuring the solids concentration at a number of locations throughout the tank. These measurements can be undertaken using a variety of techniques including grabbing samples, local concentration measurements (using conductivity probes or a range of other methods), or tomographic methods. The distribution can be quantified in three dimensions, though the tangential component is usually ignored as an assumption of axi-symmetry is made.

Axial concentration distribution plots present the concentration relative to the mean on the x- axis, as a function of the axial height normalised with respect to the tank diameter (at a fixed radial and tangential location). Hence a perfect distribution would be represented by a vertical line at $C_v/C_{vo} = 1$ and a relative standard deviation (RSD) of 0. A poor distribution would show concentrations significantly above average in the bottom parts of the tank, and lower than average values at the top. An example is presented in Figure 4. Detailed one-dimensional solids concentration data for fine and coarse sand slurries at high concentrations when using separately five different agitators are available (Heywood et al., 1985).

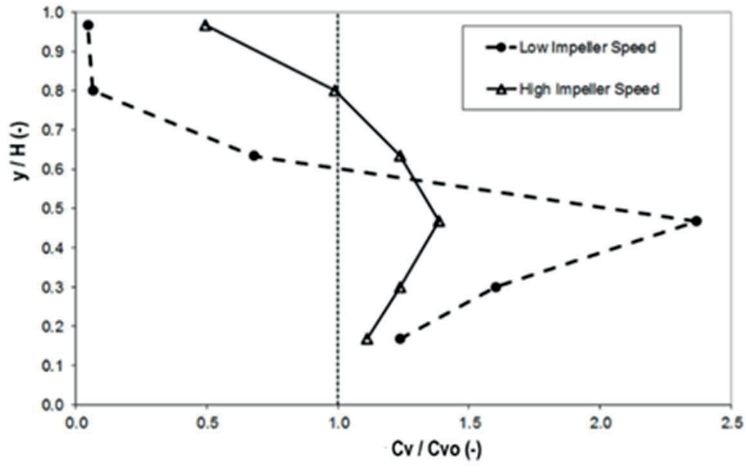

Figure 4 Typical axial solids concentration distribution profiles

Ideally (assuming perfect radial and tangential distribution, and a full range of axial measurements), the area under any normalised axial concentration distribution curve should be equal to the tank aspect ratio, H/T (or 1 if the y axis is normalised by the H/T rather than by T only). However, as measurements are often not made very close to the base, where the concentration is highest, and variations in radial and axial homogeneity can exist, this is not always the case.

4.5. EFFECT OF MATERIAL PROPERTIES ON SUSPENSION AND DISTRIBUTION

The individual effects of particle size, density and viscosity are well understood for suspension processes, but their impact on distribution performance is more complicated and interrelated. Faster settling particles (large diameter, large density difference, low viscosity) are significantly more difficult to disperse than slower settling particles. A formal link to the particle terminal settling velocity has not been found, and terminal velocity correlations relate to the motion of a particle in a quiescent (or at least laminar) bulk. Even when the particle drag may be in the turbulent regime, this drag effect relates to particle (rather than liquid) Reynolds number, and so not directly to particle motion in a turbulent liquid.

4.6. EFFECTS OF AGITATOR AND TANK GEOMETRY

All the proceeding discussion and correlations relate to relatively "standard" geometries, here defined as systems with a fully baffled cylindrical tank containing a single agitator and a fill level equal to the tank diameter, operating in the turbulent regime. Limited aspects of mixer geometry have been incorporated into correlations for N_{JS} with relatively standard geometries; the GMB correlation includes the effect of both D/T and C/T within sensible ranges, and by changing a single constant, z, can be used for all narrow blade hydrofoils and pitched-blade turbines.

The Zwietering correlation incorporates all geometric variables in its constant, 'S', which must be either measured or obtained from tables, though modified versions of the correlation do exist to allow for changes in D/T and C/T. However, in most cases simple predictive correlations (either for N_{JS} or distribution quality) cannot be adapted and applied to systems with significantly varied geometries. This difficulty arises because of the complex interactions between the geometry (which is difficult to quantify in simple numerical terms) and specific aspects of the system's performance (such as distribution quality, which is also difficult to quantify in simple numerical terms).

Slurries are frequently agitated in tanks that do not meet the criteria outlined above. The impact of changing agitator/tank geometry on all aspects of solid-liquid mixing performance can be profound. In some cases, the change in geometry may have a significant effect on the measured N_{JS}, but the amount of solids on the bottom can be relatively small. A lot of agitator power may be required to lift the last few kg. Often, a level of engineering pragmatism is required; achieving N_{JS} may not be possible, but a failure to reach N_{JS} does not mean a poorly-performing slurry tank.

Tanks Containing Multiple Agitators

Slurry tanks with multiple agitators mounted on the same centrally-positioned shaft are common, as this design is generally the best method of ensuring a relatively uniform distribution of solids in a tank with a H/T ratio equal to or greater than one. The best geometry to use to suspend and distribute solids in a large aspect ratio tank is one employing multiple axial flow hydrofoil agitators. Both cloud height and distribution quality can be significantly improved over either single agitator or multiple agitator mixed or radial flow systems, often at significantly lower specific agitator power inputs. If pitched blade turbines are to be used, it is recommended that they have a D/T ratio less than 0.4. For hydrofoil designs, a D/T of less than 0.45 is recommended.

Much academic work has been performed using multiple radial flow agitators such as Rushton turbines, but it is not a geometry recommended for industrial implementation as multiple Rushton turbines tend to form distinct axial zones, which usually are unwanted. The application of "tickler" agitators, designed as additional agitators mounted towards the tank bottom to aid in solids draining is discussed in Section 6.2.

Effect of Tank Base Shape

The shape of the tank base in part dictates the flow patterns in the vessel (in particular that in the agitator discharge), and hence can affect the Power number. In general, this effect is small at higher agitator clearances (> T/3) for most agitator types and base shapes, but at smaller clearances (or with more radical base shapes further removed from the standard dished or torispherical base, such as a cone), the effect can be significant. Changes in Power number are usually related to changes in agitator discharge flow patterns. The effect of the tank base shape on Power number can depend strongly on the agitator type and baffling configuration. In general, the effects of changing the base shape alone are less significant than changes in baffle design.

The tank base shape can have a significant effect on the solids suspension (and distribution) performance, and the impact of this effect should be considered where data are available only with other geometries. Not only can the agitator speed required to just suspend the solids change with the geometry, but the location of that last suspended material can also change. As the mechanism of suspension can change with extreme changes of base shape (or baffling), it can become difficult to measure N_{JS} in a consistent fashion. For some bases it can be difficult to observe the whole base when making N_{JS} measurements.

The effect of base shape described in the following text focuses on the use of axial and mixed flow agitators. Effects are broadly similar for radial flow agitators. For the "standard" DIN or ASME torispherical or dished bottom, the last point of suspension is the centre of the tank base for all sensibly positioned turbine agitators (with or without baffles, and generally irrespective of agitator type), as both velocities, and turbulence levels are lowest in this region. Suspension and distribution performance in dished based tanks, where there is no knuckle radius between the vessel walls and base, can generally be assumed to be identical to that in a torispherical vessel, but the last point of suspension can vary with agitator type, clearance and diameter. In most cases, N_{JS} is not significantly different in dished or torispherical based tanks.

N_{JS} for flat based tanks, particularly common in very large tanks, can be significantly higher than that required for dished or torispherical based tanks. An increase of 25% is not uncommon. This is because the solids tend to collect in and be hard to remove from the tank corners. The absolute necessity to suspend all the solids (which will depend significantly on the process) should be considered here, as the power required to do so may be prohibitive (an increase of about 100% is typical). Sometimes a small gap is left between the bottom of the baffles and the base to help prevent the collection of particles in this region.

Conical or wedge-shaped bases are sometimes an option. They are often used with reduced baffling and are non-optimal for solids suspension and distribution. Typical cone angles are about 45°, but shallower 30° cones are also common. Slopes are sometimes specified to prevent the hang-up of sticky slurry and to allow the slurry to flow to the tank outlet as slumping may occur if material is allowed to accumulate in dead zones. Very shallow cones can behave like dished based tanks, but with steeper cones the relative agitator speed required to suspend solids is considerably increased. The last point of suspension in these geometries (irrespective of agitator type and baffle geometry) is usually at the bottom of the cone (usually on top of the drain valve). As conical-based tanks are usually poorly baffled, and therefore exhibit significant tangential swirling flow and very poor axial flow/mixing, the distribution of solids in these vessels can be highly non-uniform, with much higher than average concentrations towards the bottom of the vessel. This can also lead to poor uniformity of concentration during draining.

It is generally assumed that there is no significant difference between the cloud height achieved with dished, torispherical, and elliptical based tanks, providing similar agitators and baffles are employed. Cloud height may be compromised slightly by the use of a flat base. Cloud height and distribution can be significantly compromised by the use of conical bases (with their associated poor baffling).

Effect of Slurry Fill Level

Great care should be taken when considering systems where the fill level is not equal to the tank diameter. Most of the published work on solids suspension has been based on systems where H = T, while many slurry tanks operate with a fill level greater than or less than the tank diameter due to other process requirements, tank availability, and tank diameter. Von Essen and Ricks (1999) state that tank cost is minimised for a fixed live volume at H/T =1.05, and so this is frequently selected but that agitator and power costs are optimised at lower H/T, and a total project saving of about 15% can be realised at an optimum H/T = 0.63.

The Power number of most single agitators mounted at standard clearances is not significantly affected by the slurry fill level when it is greater than the tank diameter. At lower fill levels (H < T) where the slurry surface approaches the agitator, the flow patterns can change significantly as the agitator suction and discharge flows are affected. With axial flow agitators (smaller diameter pitched blade turbines and hydrofoils), the agitator suction can become "throttled" and, significant quantities of air are entrained and so the agitator does not work effectively and the Power number (and hence power draw) can be significantly reduced. This throttling can occur when the fill level drops below about one agitator diameter above the agitator and severe splashing can occur, which further compromises performance and also affects tank draining (see Section 6.2).

With axial flow agitators, as the fill level is reduced below H = T, N_{JS} initially reduces slightly with the fill level, but once the fill level reduces to about H = 0.5T the agitator suction is throttled to such an extent that the solids cannot be suspended at any speed. This transition is consistent with the drop in Power number (Motamedvaziri and Armenante, 2012). However, von Essen and Ricks (1999) have suggested that the absolute minimum level for an axial flow agitator occurs at H/T = 0.15 with a small bottom heel agitator, or tickler (see Section 6.2). They also maintain that a single very large agitator with D/T = 0.4 cannot suspend solids when H/T < 0.25.

Care should be taken with the definition of mass fraction employed in correlations to predict N_{JS} when the fill level is not equal to the tank diameter. All correlations have been developed for systems where H = T. When H < T it is reasonable to base calculations on the real system mass fraction, but where H > T, however, it is safer to assume that the mass fraction is based on the total mass of solids present and the volume of the tank up to a fill level equal to the tank diameter. Measurements show that N_{JS} increases (at a constant total mass fraction) as the fill level increases above H = T, as the "local" concentration in the bottom part of the vessel is higher owing to the poor distribution. Power calculations should also consider the local concentration (and hence slurry density) in the region of the agitator.

There are few or no data available for the effect of cloud height with single agitators where H ≠ T. It is to be expected that when the fill level is significantly greater than the tank diameter the distribution will be poor, with few solids being transported to the upper regions of the tank. Distribution with single agitators may be reasonable down to an H/T of about 0.7, and below that the drop in Power number and poor suspension discussed previously mean the distribution will be very poor. In the former case, where large H/T values are used, it is strongly recommended that multiple agitators should be used. The agitators should be axial so a large overall flow pattern can be established. The lower agitator is for off-bottom solids suspension and the upper agitator(s) distribute solids over the height. The use of radial agitator(s) as the upper agitator(s) will cause the formation of undesirable segregated zones.

5. AGITATOR SELECTION AND OPERATION

The optimum agitator choice to minimise the required power input for a solids-suspension process in a standard baffled tank is generally a narrow blade hydrofoil, pumping towards the base. The D/T ratio should be between 0.33 and 0.45, and the clearance to tank diameter ratio between 0.2 and 0.33. The optimum clearance depends on the minimum expected fill level and the distribution quality required.

Mechanical agitation need not always be continuous, particularly if there is intermittent draw-off from the tank feeding a pipeline, and any settled bed formed on the tank bottom is not cohesive and can be readily resuspended. Such an approach saves agitator power consumption, and minimises particle attrition if the solids are friable, as well as reducing abrasion of the agitator and tank internals if the slurry is abrasive,

Agitator erosion or wear is related to the material and design of the blades, as well as the size, density, hardness and concentration of the solids. Particular care should be taken when hard solids are suspended but poorly distributed, because the concentration in the region of the agitator can be particularly high.

5.1. AGITATOR START-UP IN A SETTLED BED

The start-up of an agitator that is either partially or wholly submerged in a settled bed can cause damage not only to the agitator but also the drive/gearbox and cause shaft damage when these components are not designed to work at the high torque levels required to exceed the shear strength of the settled bed. When the settling of solids so they cover the agitator is a significant risk, liquid or gas jets, can be used to aid the initial re-suspension.

Two simple laboratory tests are available to assess whether, firstly an agitator can be partially or wholly submerged by solids as a result of a planned or unplanned power outage, and, secondly, whether the agitator shaft thickness, motor and gearbox can cope with the high start-up torque. In the first lab test, the settled bed height over time, and final settled bed height, can be measured using a 2-litre graduated cylinder and this will determine if the lower (or only agitator) will be partially or fully immersed in the bed. The ratio of bed height to slurry height in the graduated cylinder can be directly scaled up to tank operation (with an adjustment for different tank bottom profiles) provided that the diameter of the graduated cylinder is large enough with respect to the slurry particle size so that there are no significant wall effects. In the second laboratory test, settled beds can also be formed in standard laboratory glass beakers, and their shear strength measured as a function of time using a vane attached to a suitably designed rheometer. While not necessarily providing a direct way to predict the start-up torque of the agitator in a settled bed, and how this may increase with time, it does provide a useful guide.

6. CONTINUOUS FLOW AND SEMI-BATCH OPERATION

It is believed that the various process results such as off-bottom suspension, distribution, attrition, etc. are not affected by whether the system is batch or continuous (flow through), and so there is no effect on the specific, batch-developed correlations. This effect is minor because in most cases the through flow through and the through momentum are orders of magnitude less than the flow and momentum generated by the agitator. If this condition were not the case, then through flow might have some effect but there are practically no studies or effects reported so the assumption is made. However, there are several instances where the through flow can modify the *in situ* and exit concentrations. Some hydrotransport systems may allow slurry to drain back into the tank after a pump shutdown. This must be taken into account in fixing the tank volume so that the tank does not overfill.

6.1. EFFECT OF SLURRY OUTLET LOCATION IN A CONTINUOUS TANK

If the tank contents are completely homogeneous so that the solids concentration is the same at all locations then the outlet location should have little effect on the concentration downstream. However, it has been shown that the power requirements to achieve this degree of homogeneity are significant and excessively high, and, in most cases, there exists a vertical concentration gradient. In such cases, the height of the outlet will affect the system performance. At steady-state, the concentration of solids leaving the vessel must equal the mean concentration in the tank.

If the outlet is low in the tank the solids concentration there would be higher than average, and so the average concentration in the tank would be lower than the composition leaving the tank. If the outlet location is higher, where the local concentration is lower than the average, then the average concentration in the tank will increase with time. This situation can cause problems. The higher concentration may cause the minimum suspension speed to be higher than that for the design condition. The higher speed can promote attrition and increase the residence time of the solids over that of the liquids. It may even be possible to stall the agitator by running at an unmixable concentration many times higher than what is being pumped. Because of this phenomenon and for robust tank drainage the slurry outlet is usually near the bottom.

6.2. DRAINING OR WITHDRAWAL FROM A TANK

During vessel discharge a constant concentration in the outlet line that is representative of the mean tank concentration is desirable for feeding a slurry pipeline, but this can be difficult to achieve. The draining profile depends heavily on the concentration distribution at the onset of discharge. If a significant concentration gradient exists in the tank before discharge starts, the initial concentration withdrawn will be representative of that at the bottom, and higher than the average. Thus, the concentration withdrawn will reduce over time if the slurry level in the tank is allowed to fall.

To maintain a more uniform discharge profile a constant concentration distribution is required. This can be achieved by increasing the number of agitators and/or optimising their separation. While the cloud height and axial distribution of solids can, to a degree, be maximised by increasing the agitator separation, on discharge this spacing recommendation means that the upper agitator will become clear of the slurry level earlier during discharge, and so achieving this optimum is difficult.

When no agitation is provided in a slurry tank, settled solids can accumulate in dead zones, particularly when flat-bottom, rectangular tanks are used. This can be a problem when settled solids accumulate over a non-active outlet nozzle, or slump within the tank to cover the outlet. Slurry feed lines and outlet nozzles are often arranged to minimize dead zones. Water injection points are sometimes provided around the outlet nozzles to move settled solids. Another approach to minimising dead zones is to minimise the tank size, so minimising the inventory of solids that could cause blockage on shutdown.

Tickler Agitators

Often, a lower “tickler” agitator is put very low in the tank to maintain solids suspension at low fill levels during draining. In such cases, many designers simply use a four-blade, radial flow agitator, sometimes with swept blades and rotation to push solids outward to minimise plugging. Axial flow agitators running close to the bottom are sometimes employed but these act like radial agitators at very low clearances and the Power number is often higher than at higher positions off the tank bottom. Tickler agitators are often placed at or just below the tangent line, or in the case of large, flat-bottom storage tanks, at a fixed distance off bottom such as 250 mm.

Tickler agitators generally have a significantly smaller diameter than the main agitator(s) designed to mix the tank's contents, and hence at the same shaft speed draw very little additional power. They also have limited effectiveness from a blending and solids distribution point of view. Tickler agitators should be sized effectively to control the volume of the tank below the main agitator, not just be arbitrarily small. In these cases, the agitator is generally employed simply to aid in the withdrawal of the final solids from the "heel" of the tank. Few studies have been done on optimising this design. An exception is the work by Dow (Motamedvaziri and Armanante, 2012) whose design is called the KT-3 (Figure 5) and is marketed by SPX Flow. They chose for practical reasons to optimise the design that left the fewest solids behind, while an additional criterion was to minimise splashing and gas entrainment. This is a frequent problem with running at low slurry levels that can affect the operation of a transfer pump.

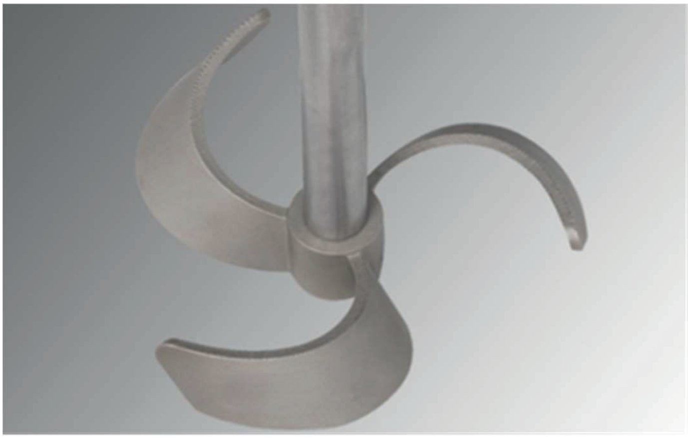

Figure 5 KT-3 "Tickler" agitator

7. DESIGN RECOMMENDATIONS AND SCALE-UP

For slurry tank applications where minimising the specific agitator power input (P/M, agitator power input per unit mass of material in tank) for suspension/distribution in the transitional or turbulent regimes is important, it is recommended that axial flow hydrofoil agitators pumping towards the tank bottom are employed, along with three or four standard flat baffles and a dished or torispherical base, if possible. However, when a tank has a large diameter, the base often has to be flat. Hydrofoil agitators typically suspend solids at about half the power input (and torque) of equivalent diameter and clearance pitched-blade agitators, significantly reducing both capital and operating costs, though at a higher agitator speed. Hydrofoils are significantly more power efficient at suspending solids than radial flow agitators such as the Rushton disc turbine. Use of radial flow agitators is not recommended. A hydrofoil agitator-to-tank diameter, D/T between 0.33 and 0.45 is recommended, with 0.4 a good standard.

An agitator clearance of T/6 is recommended if achieving N_{JS} is the priority, though higher agitator clearances may be required if a tickler impeller is installed. Clearances of less than T/6 are not recommended. Where multiple agitator systems are employed using a common shaft, again hydrofoil agitators are recommended as they produce optimum axial flow patterns and exchange flows between agitators, maximising the transport of solids throughout the vessel. Agitator separation must be carefully considered, with recommended separations no greater than two agitator diameters. The use of multiple agitators will increase the power per unit mass required to just suspend solids, with the addition of a second agitator having little effect on N_{JS}. The use of a tickler agitator is recommended where draining of the solids from the tank is required. This use of an additional agitator will usually not affect either the power draw or suspension performance of the slurry tank, provided that the tickler diameter is significantly less than that of the main agitator(s).

For design purposes, it is recommended that the GMB correlation (Grenville et al., 2015) be used to predict the agitator speed for just-suspended conditions where axial flow or pitched-blade turbine agitators are employed. Scaling on an equal power per unit mass is recommended for both solids suspension and cloud height/distribution limited processes.

8. CONCLUDING REMARKS

There exists much design information for slurry storage tanks. It is important to be clear what degree of slurry agitation is required to ensure that the slurry concentration in the slurry discharging from the tank to the pipeline or some other downstream process is representative of the bulk concentration in the tank. Otherwise, for a continuous system, this slurry parameter will not be maintained constant and so could impact negatively on the pipeline hydraulics. Thus, correct tank design, agitator selection, sizing and positioning within the tank together with the specification of the agitator speed and mixer drive system are all vital. In addition, the position of the point of slurry discharge from the tank needs to be carefully considered.

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