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Mixing in SRS Closure Business Unit Applications

Michael R. Poirier

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INTRODUCTION

In addressing mixing problems in the Savannah River Site (SRS) Tank Farm, one must distinguish between different mixing objectives. These objectives include sludge mixing (e.g., Extended Sludge Processing[1,2]), sludge retrieval (e.g., sludge transfers between tanks[3]), heel retrieval (e.g., Tanks 18F and 19F[4,5,6]), chemical reactions (e.g., oxalic acid neutralization[7]) and salt dissolution[8,9]. For example, one should not apply sludge mixing guidelines to heel removal applications.

Mixing effectiveness is a function of both the mixing device (e.g., slurry pump, agitator, air sparger) and the properties of the material to be mixed (e.g., yield stress, viscosity, density, and particle size). The objective of this document is to provide background mixing knowledge for the SRS Closure Business Unit personnel and to provide general recommendations for mixing in SRS applications.

MIXING EQUIPMENT

The following equipment is commonly used to mix fluids: mechanical agitators, jets (pumps), shrouded axial impeller mixers (Flygt mixers), spargers, pulsed jet mixers, boiling, static mixers, falling films, liquid sprays, and thermal convection. This discussion will focus on mechanical agitators, jets, shrouded axial impeller mixers, spargers, and pulsed jet mixers, as these devices are most likely to be employed in SRS Closure Business applications.

Mechanical Agitators

Mechanical agitators consist of a rotating impeller around a cylindrical shaft immersed in the fluid to be mixed. The vessels are usually cylindrical with a liquid height approximately equal to the tank diameter. If the liquid height is much less than the tank diameter, the power applied to ensure fluid motion and mixing throughout the tank, can cause excessive fluid motion and splashing at the liquid surface. If the liquid height is much greater than the tank diameter, multiple impellers are commonly used to ensure good mixing throughout the tank.

The bases of these tanks can be flat, round, or conical. Flat bottom tanks are generally the least expensive, but the corners are regions of reduced turbulence and stagnant flow. Solids particles can accumulate in the corners. Round bottom tanks eliminate the regions of low turbulence and improve solids suspension, but cost more. Conical bottom tanks with a bottom draw off are useful when heavy particles need to be removed from the tank. If the cone is long and narrow, a second impeller may be required to ensure the material in the cone is well mixed. Contoured bottoms, such as a Bourne bottom, can improve solids suspension.[10]

Baffles are frequently added to the tanks to improve vertical mixing, ensure the entire tank contents pass through the impeller, produce more even power draw, produce more uniform shaft loads, and prevent vortexing with low viscosity fluids.[11] A typical baffle design employs four baffles 90 degrees apart, at or near the tank wall. Typical baffle width is 1/10 -1/12 of the tank diameter. Baffles are generally not needed for high viscosity fluids (e.g., Bingham plastics). When the tank's contents can vary from low viscosity to high viscosity, triangular baffles can be employed.[12] A method to avoid the need for baffles with low viscosity fluids is to mount the agitator off center or at an angle.

Numerous types of impellers are available for mechanical agitators. One type is the propeller. Propellers typically have three blades, are angled, and produce axial flow parallel to the impeller shaft. Propellers include high efficiency impellers and marine impellers. Propellers typically require the least amount of power, and produce the most flow. These impellers are typically used for blending low viscosity liquids, dispersing gases in liquids, and liquid-liquid contacting.[13] Radial flow impellers include Rushton impellers, flat blade impeller, and disk turbines. They produce flow, which moves radially away from the impeller shaft. They typically require the most power to operate and produce the most shear. Radial flow impellers are typically used for applications in which shear is important, gas dispersion, and viscous fluids.[14] Pitched bladed impellers are compromise between the propellers and radial flow impellers. They produce axial flow and are commonly used for solids suspension applications.[10] For optimum performance the impeller should produce turbulent flow.

The Reynolds number for an agitated tank is calculated from equation [1]

$$Re = N\,D^2/\nu \qquad [1]$$

where N is the impeller speed (sec^{-1}), D is the impeller diameter (cm), and ν is the kinematic viscosity (cm^2/sec). Typically a Reynolds number greater than 5,000 – 10,000 is required to produce turbulent flow.

For high viscosity laminar flow applications, devices such as anchor impellers, helical screws, and helical ribbons are used.[10] These impellers have diameters that are 85 – 95 % of the tank diameter and are close to the tank walls.

Draft tubes are sometimes placed around impellers to increase circulation and improve mixing.

The power requirements for an axial flow impeller can be calculated with equation [2]

$$P = N_p\,\rho\,D^5\,N^3 \qquad [2]$$

where P is power (Watts), N_p is power number, ρ is density (kg/m^3), D is impeller diameter (m), and N is impeller speed (sec^{-1}). The impeller number is a function of Reynolds number for laminar and transition regime flow, and a constant for turbulent flow. Table 1 shows turbulent power numbers for some common impellers.

Table 1. Turbulent Flow Power Numbers

Impeller Type	Power Number
Fluid foil	0.3
Propeller	0.6
Pitched blade turbine	1.27
Flat blade impeller	4.0
Flat blade disk impeller	5.0

When mixing Bingham plastic fluids, one can apply the cavern model. The basis of this model is that close to the impeller the shear stresses generated by the impeller are greater than the yield stress of the slurry, and the slurry is well mixed. At large distances from the impeller, the slurry yield stress is larger than the shear stress generated by the impeller, and the slurry is not well mixed. Equations [3] – [5] describe the cavern model.[15,16,17]

$$N_c = \frac{\pi}{D_I}\sqrt{\frac{\left(\frac{H_c}{D_c}+\frac{1}{3}\right)\left(\frac{T}{D_I}\right)^3 \tau_y}{\rho N_p}} \quad [3]$$

$$N_c = \sqrt{\frac{4V_c\pi\left(\frac{H_c}{D_c}+\frac{1}{3}\right)\tau_y}{\left(\frac{H_c}{D_c}\right)\rho N_p D_I^{5}}} \quad [4]$$

$$N_c = \sqrt{\frac{\left(\frac{T}{D_I}\right)^3 \pi^2 \tau_y}{1.36\,\rho N_p D_I^{2}}} \quad [5]$$

In equations [3] – [5], N_c is the impeller speed (sec^{-1}) required to mix the entire tank, V_c is the cavern volume (cm^3), H_c is the cavern height (cm), D_c is the cavern diameter (cm), T is the tank diameter (cm), D_I is the impeller diameter (cm), τ_y is the slurry yield stress ($dynes/cm^2$), ρ is the slurry density (g/cm^3), and N_p is the impeller power number. For a well mixed tank, the cavern volume is equal to the slurry volume, the cavern height is equal to the slurry height in the tank, and the cavern diameter is equal to the tank diameter.

A number of researchers have investigated the requirements for suspending insoluble solids in tanks mixed with agitators. In these studies, they measured the agitator speed required to "just suspend" all of the particles (i.e., no particle remains stationary on the tank bottom for more than one second). Table 2 shows the correlations developed by these researchers.

In Table 2, ρ_s is the particle density (kg/m^3), ρ_f is the fluid density (kg/m^3), g is the gravitational acceleration (9.8 m/s^2), d_p is the particle size (m), C_v is particle volume concentration (fraction), T is tank diameter (m), C_D is the drag coefficient, N_p is the power number, D is the impeller diameter (m), Z is the liquid height (m), ν is the kinematic viscosity (m^2/sec), X is the particle weight fraction, and S is a constant. Typical values of S are between 5 and 7.

Table 2. Agitator Speed Required to Suspend Insoluble Particles

Reference	Correlation
18	$N_{js} = \dfrac{3.5\left(\dfrac{\rho_s - \rho_f}{\rho_f}\right)^{1/2} g^{1/2} d_p^{\,1/6} C_v^{\,1/3} T}{C_D^{\,1/6} N_p^{\,1/3} D^{5/3}}$
19	$N_{js} = \left(\dfrac{g(\rho_s - \rho_f)}{\rho_f}\right)^{1/2} \dfrac{1}{N_p^{\,1/3}} \left(\dfrac{T}{D}\right) \dfrac{d_p^{\,1/6}}{D^{2/3}} \dfrac{1}{Z}$
20	$N_{js} = \dfrac{S\nu^{0.1} d_p^{\,0.2} \left(\dfrac{g(\rho_s - \rho_f)}{\rho_f}\right)^{0.45} X^{0.13}}{D^{0.85}}$
21	$N_{js} = \dfrac{S\nu^{0.1} d_p^{\,0.15} \left(\dfrac{g(\rho_s - \rho_f)}{\rho_f}\right)^{0.4} X^{0.12}}{D^{0.76}}$
22	$N_{js} = \dfrac{\mu^{0.17} d_p^{\,0.14} \left[g(\rho_s - \rho_f)\right]^{0.42} T}{\rho_f^{\,0.58} D^{1.89} N^{\,0.28}}$

A number of researchers have investigated the influence of operating parameters on mixing time. Mersmann suggest a general correlation described by equation [6]

$$\theta_m = 6.7\,(T/D)^{5/3} N_p/N \qquad [6]$$

where θ_m is the mixing time (sec), T is the tank diameter (m), D is the impeller diameter (m), Np is the impeller power number, and N is the impeller speed (sec^{-1}).[23] Additional mixing time correlations are provided in other references.[24,25]

Peter Holman, a mixing consultant who formerly worked for Lightnin, provided the following guidelines for mixing liquids.[26] Table 3 shows the recommended power per unit volume and impeller tip speed for various mixing applications.

Table 3. Guidelines for Liquid Mixing with Agitators

Operation	HP/1000 gallons	Tip Speed (ft/s)
Blending	0.2 – 0.5	N/A
Homogeneous reaction	0.5 – 1.5	7.5 – 10
Reaction with heat transfer	1.5 – 5.0	10 – 15
Liquid-liquid mixtures	5	15 – 20
Liquid-gas mixtures	5 – 10	15 – 20
Slurries	10	N/A

Jet Mixers

Turbulent jet mixing is common in the chemical processing industry. In this process, a fast moving stream of liquid (jet) is injected into a slow moving or stationary liquid (bulk liquid). The velocity difference between the jet and the bulk liquid creates a mixing layer at the jet boundary. The mixing layer grows in the direction of the jet flow, entraining and mixing the bulk liquid into the jet. Equation [7] describes the jet velocity

$$V = 6\ Dj\ Vj/x \qquad [7]$$

where V is the jet centerline velocity (ft/sec), D_j is the jet nozzle diameter (ft), V_j is the jet nozzle velocity (ft/sec), and x is the distance from the jet nozzle (ft). The velocity falls to ~ 5% of its initial value after 100 – 120 nozzle diameters. Jet mixing is considered insignificant after 400 nozzle diameters. Maximizing the product V_jD_j will slow the jet decay and improve fluid mixing far from the jet.

Jet mixing time has been measured experimentally with tracers. In these tests, t_m is the time until the maximum absolute deviation of the mixture is (100 – m)%. Most studies have been conducted to determine t_{95}. Equation [8] estimates the variation in mixing time with m.[27]

$$t_m/t_{95} = -1/3\ \ln[(100-m)/m] \qquad [8]$$

The following equations can be used to calculate the mixing time in a tank. Van de Vusse[28] measured mixing of liquids of different density in a 36 meter diameter storage tank. He correlated his measurements of tank mixing time (in seconds) with equation [9].

$$t = 3.7\ \frac{D_{tank}^2}{V_j\ D_j} \qquad [9]$$

In equation [8], D_{tank} is the tank diameter (ft), V_j is the jet nozzle velocity (ft/sec), and D_j is the jet nozzle diameter (ft). The correlation is limited since it is based on three tests in which only the height of the liquid and the jet velocity were varied. The degree of mixing was not quantified.

Fossett and Prosser[29] injected a small amount of tracer into a 1.52 meter diameter tank and measured the mixing time. They correlated their data by equation [10]

$$t = 4.5\ \frac{D_{tank}^2}{V_j\ D_j} \qquad [10]$$

Okita and Oyama[30] measured the mixing time of a tracer in 0.4 and 1.0 meter diameter tanks. They correlated their data with this expression

$$t = \frac{5.5\ D_{tank}^{1.5}\ H^{0.5}}{V_j\ D_j} \qquad [11]$$

where H is the liquid height (ft). Fox and Gex[31] measured mixing times for side entry jets in tanks 0.29m, 1.52m, and 4.27m in diameter. They correlated their data with equations [12], [13], and [14].

$$t = \frac{A\ D_{tank}\ H}{V_j\ D_j}\ (D_j/H)^{0.5} \qquad [12]$$

$$A = \frac{120 Fr^{0.17}}{Re_j^{\ 0.17}} \qquad [13]$$

$$Fr_j = \frac{V_j^2}{g\ D_j} \qquad [14]$$

In equation [14], g is the gravitational acceleration constant (32.2 ft/sec^2).

Hiby and Modigell[32] used a tracer technique to measure mixing in tanks with axial jets. They correlated their data with equation [15].

$$t = 3.2\ \frac{D_{tank}^2}{V_j\ D_j} \qquad [15]$$

The author recommends calculating the mixing time with all five equations. The user can choose the maximum time for conservatism or the average time. Koh et al. measured the mixing of liquids with multiple rotating nozzles in a cylindrical vessel.[33] They found the mixing time to be inversely proportional to the square root of the number of nozzles.

Revill provides the following guidelines for jet mixing.[27] Single axial jets should be used when the tank height to diameter ratio is between 0.75 and 3.0. Single horizontal jets should be used when the height to diameter ratio is 0.25 – 1.5. Multiple horizontal jets should be used if the height to diameter ratio is less than 0.25 or greater than 3.0.

Mixing of Bingham plastic fluids can be modeled with the effective cleaning radius model (which is analogous to the cavern model discussed earlier). Churnetski[1] developed an effective cleaning radius model to describe the suspension and mixing of slurries. The model predicts the volume of a tank in which the pump supplies sufficient energy to suspend and mix the slurry as a function of pump operating parameters and fluid properties.

When a turbulent jet impinges on a surface, the force of the jet (dynes) is

$$F_j = \rho v^2 A/2g_c \qquad [16]$$

where ρ is the density (g/cm^3), v is the jet velocity (cm/sec), A is the surface area (cm^2), and g_c is a conversion factor (1 dyne sec^2/g cm). If the jet force is greater than the surface force, the jet will break or move the surface. For slurries in SRS waste tanks, the surface force is the yield stress of the slurry (τ_y) in dynes/cm^2, and the point at which the jet shear stress equals the slurry yield stress is the effective cleaning radius.

$$\tau_y = \frac{F}{A} = \rho v^2/2g_c \qquad [17]$$

The velocity of a turbulent jet is a function of nozzle diameter, jet nozzle velocity, and axial distance, and is described by

$$v = C_1 \frac{D_j \ V_j}{x} \qquad [18]$$

where D_j is the nozzle diameter (cm), V_j is the jet velocity at the nozzle (cm/sec), C_1 is a constant, and x is the axial distance (cm). Substituting equation [18] into equation [17] yields

$$\tau_y = C_2 \rho D_j^2 V_j^2/x^2 g_c \qquad [19]$$

Solving equation [19] for axial distance provides the following equation for the effective cleaning radius

$$x = C_3 \ D_j \ V_j \,(\rho/\tau_y)^{1/2} \qquad [20]$$

where x is the effective cleaning radius (ECR) in cm.

Churnetski[1] performed tests to determine the applicability of the ECR model to suspension of SRS sludge and to determine the constant C_3. Based on these results, the effective cleaning radius (ECR) can be described by

$$ECR = 0.97 \ D_j \ V_j \,(\rho/\tau_y)^{1/2} \qquad [21]$$

The constant 0.97 is less than the constant predicted by turbulence theory ($C_3 = 4.2$). The reason for the difference is probably the friction caused by the tank wall.

Churnetski noticed a slight decrease in the effective cleaning radius when the pump rotation rate was increased.

Another model which can predict shear stresses in turbulent jets is the impinging jet model. Consider a circular jet impinging on a flat surface (Figure 1).

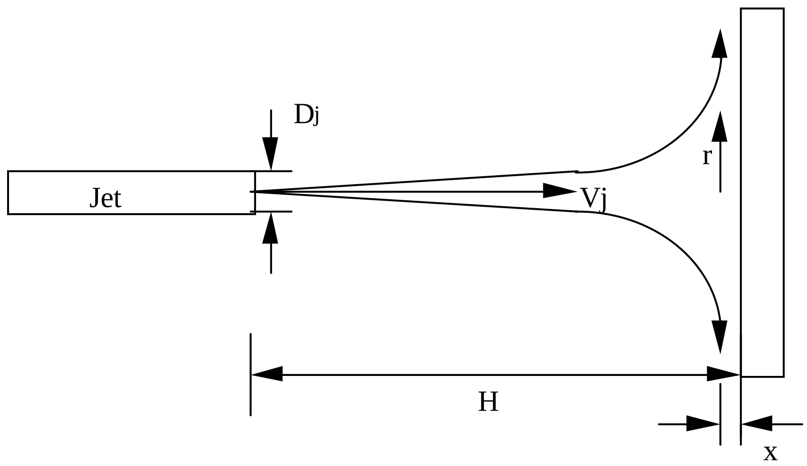

Figure 1. Impinging Jet

The maximum shear stress of the jet (dynes/cm^2) is described by[34]

$$\tau_m = \frac{0.16 \rho V_j^2 D_j^2}{H^2} \qquad [22]$$

where ρ is the fluid density (g/cm^3), V_j is the jet nozzle velocity (cm/sec), D_j is the nozzle diameter (cm), x is the distance from the solid surface (cm), and H is the distance between the jet and the surface (cm). The shear stress variation as a function of radial distance from the jet centerline (r) is described by[35]

$$\frac{\tau}{\tau_m} = 0.18 \left\{ \frac{1 - \exp[-114(r/H)^2]}{(r/H)} \right\} - 9.43 \frac{r}{H} \exp[-114 (r/H)^2] \qquad [23]$$

According to Rajaratnam[34,35], the maximum shear stress occurs at r/H = 0.14 rather than at the jet centerline. Since the maximum shear stress occurs at r/H = 0.14, the best mixing and suspension may occur at an angle of 8• rather than at the jet centerline. If the shear stress at a point is greater than the fluid yield stress, that point is inside the cavern and good fluid mixing will occur. This model can be applied to SRS waste tanks to determine the cavern size or effective cleaning radius of the slurry pumps.

A third model is the shear stress generated by a wall jet (Figure 2).

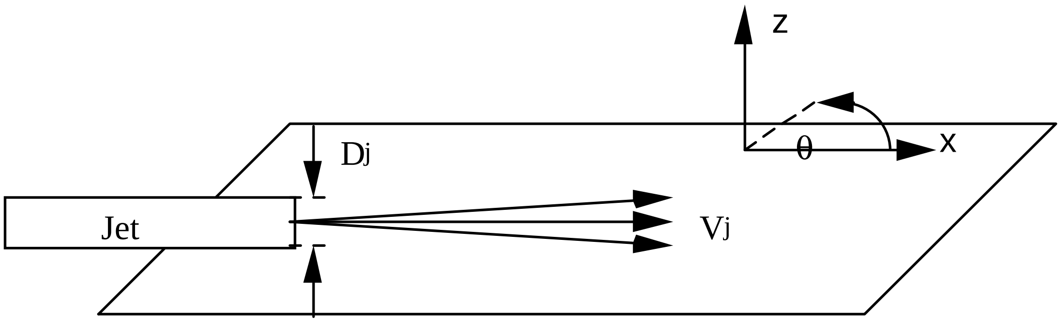

Figure 2. Turbulent Wall Jet

The maximum shear stress of a wall jet is described by[36]

$$\tau_W(z{=}0) \;=\; 0.02782\,\rho\;(V_j\;D_j/x)^2\;\{8.5\,[1 - \exp(\frac{1.1284\,\frac{x}{D_j} + 12}{-17})] - 2\} \qquad [24]$$

The shear stress as a function of position is

$$\tau_W(z) \;=\; \tau_W(z{=}0)\,\exp[-63.63\,(z/x)^2] \qquad [25]$$

In equations [24] and [25], ρ is the fluid density (g/cm^3), V_j is the jet nozzle velocity (cm/sec), D_j is the nozzle diameter (cm), z is the distance between the jet and the slurry surface (cm), and x is the axial distance from the jet nozzle (cm). If the shear stress at a point is greater than the fluid yield stress, that point is inside the cavern and good fluid mixing will occur. This model can be applied to SRS waste tanks to determine the cavern size or effective cleaning radius of the slurry pumps. This model can also be used to determine the height of the cavern.

SRTC compared the predictions of the models described by equations [21], [22], and [25] with operating experience at the Extended Sludge Processing (ESP) facility and found good agreement.[2]

The radial wall jet model can be used to predict shear stresses and velocities at the liquid-vapor interface in a tank (see Figure 3).

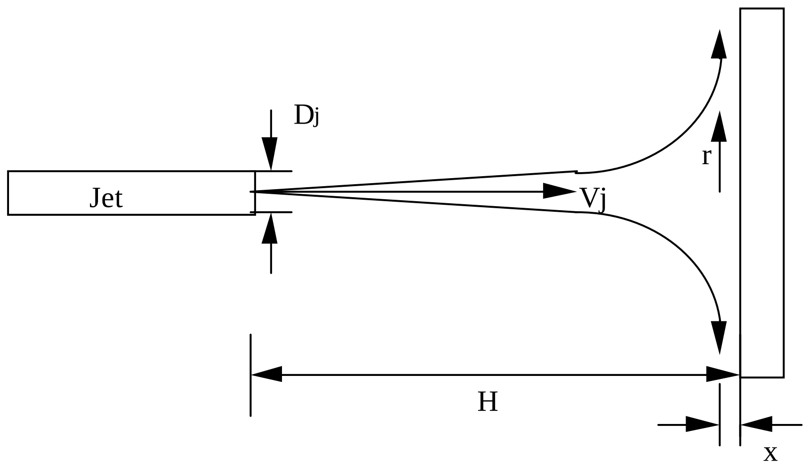

Figure 3. Radial Wall Jet

The model is valid if r/H > 0.1. The shear stress is described by[37]

$$\frac{\tau_W\, x^2}{\rho\, K} = 0.3\,(Re)^{-0.3}\ \ (r/x)^{-2.3} \quad [26]$$

where

$$K = 0.153\,\pi\, D_j^2\, V_j^2 \quad [27]$$

$$Re = \frac{D_j\, V_j}{\nu} \quad [28]$$

The velocity is described by[28]

$$V = 0.9\,\frac{H^{.1}\ D_j\ V_j}{r^{1.1}} \quad [29]$$

In equations [26] – [29], τ_w is the wall shear stress (dynes/cm^2), x is the distance from the wall (cm), ρ is density (g/cm^3), D_j is the nozzle diameter (cm), V_j is the jet nozzle velocity (cm/s), H is the distance between the jet and the wall (cm), and r is the distance from the jet centerline (cm).

If the shear stress at the liquid-vapor interface is greater than the yield stress of the fluid or the velocity is greater than 1.0 feet/second, good mixing should occur at the surface. This type of model can be used to predict the effect of tank level on mixing.

Tatterson examined the Tank 16 sludge removal demonstration data and developed a model which calculates the cleaning radius as a function of time.[38] Since the pump operating parameters and sludge properties did not change, this model does not include their effect. Equation [30] shows this model

$$x = 4.1\, t^{0.36} \qquad [30]$$

where x is the cleaning radius in feet and t is the mixing time in hours.

This observation suggests that two mechanisms contribute to sludge mobilization by jets: overcoming the slurry yield stress and erosion. If the turbulent jet produces a shear stress that exceeds the slurry yield stress, the fluid will be well mixed. This process will occur quickly. At longer times, slow changes are observed in the cleaning radius. The phenomenon seems similar to erosion. As the jet continues to impact the sludge, it will slowly cause the sludge to erode and be mixed. This phenomenon has been observed in other testing.[6,54,59]

Shrouded Axial Impeller Mixers

Shrouded axial impeller mixers (ITT Flygt) consist of an electrically powered 3-bladed propeller with a close-fitting shroud (see Figure 4). Depending on the size of the tank, different-sized propellers or propeller numbers can be used. A propeller spinning rapidly will create a turbulent jet with an average exit velocity of 3 m/s for a 4 hp mixer.

Figure 4. ITT Flygt Mixer

Spargers/Spouting Beds

Gas sparging has been used to mix liquids in tanks. The mixing is performed by injected gas through two mechanisms: gas expansion and kinetic energy transfer during injection, and gas bubbles rising through the liquid. The contribution from the first mechanism is generally small and neglected. The power input for gas sparging processes is given by equation [31]

$$P = Q\, \rho\, g\, H \qquad [31]$$

where P is power (W), Q is gas flow rate (m^3/s) , ρ is density (kg/m^3), g is gravitational acceleration (9.8 m/sec^2), and H is height (m).

David Dickey (a mixing consultant) recommended the following operating parameters for spargers and bubble columns.[39] Typical superficial velocities for mixing by gas sparging are 0.005 - 0.35 ft/s. A superficial gas velocity of 0.05 ft/s is the minimum velocity to stir the vessel. A superficial velocity of 0.1 ft/s is recommended to stir the vessel. A superficial gas velocity of 0.25 ft/s is the maximum that should be used, and at that velocity, splashing will occur.

Peter Holman, a mixing consultant, has provided mixing support to the Alternative Salt Disposition program. Mr. Holman recommended mixing guidelines for agitated tanks (see Table 3).[26] Those guidelines can be considered for this type of mixing, but should be used with caution.

An approach to predicting mixing time for spargers is to calculate the bulk circulation time in the tank and multiply by 5.[25] The circulation time is described by equations [32] - [33]

$$Q_{liq}^3 = 4 \times 10^{-4} Q_{gas}\, g\, H^5 = Q_{circ}^3 \qquad [32]$$

$$t_{circ} = V/Q_{circ} \qquad [33]$$

where V is the liquid volume (cm^3), H is the liquid depth (cm), g is gravitational acceleration (980 cm/s^2), Q_{circ} is the bulk circulation rate (cm^3/sec), Q_{liq} is the liquid circulation rate (cm^3/sec), and Q_{gas} is the gas flow rate (cm^3/sec). The bulk circulation rate is assumed equal to the liquid circulation rate.

To determine the ability of spargers or bubblers to suspend solid particles in a tank, one can draw an analogy to spouting beds. A spouting bed is a process for fluid-solid contacting which is similar to fluidization.[40] The fluid enters the flat or conical base of a cylinder through a small orifice, entrains the solids, carries the solids to the top of the bed, followed by the solids falling back into the annular core. Spouting beds are used with coarse particles and have a single air stream.

Figure 5 shows the flow regimes that exist with spouting beds. At very low gas flow rates, the solid particles remain settled and the bed behaves as a packed bed (a). As the flow rate increases, the bed shows internal spouting (b), and then good spouting (c).[40] To suspend solid particles, the sparger or bubbler should operate at a high enough flow rate to have good spouting.

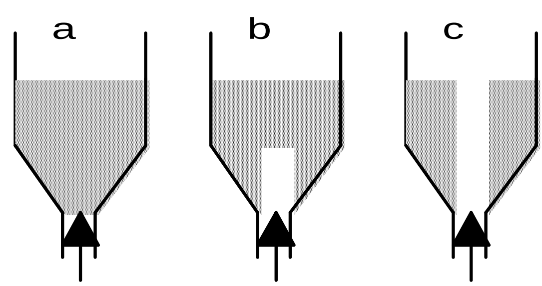

Figure 5. Spouting Bed Flow Regimes

Mathur and Gishler developed a correlation to predict the minimum spouting velocity for uniform-sized particles in conical bottom vessels.[40,41] Equation [34] describes the correlation

$$U_{MS} = \left(\frac{d_p}{D_c}\right)\left(\frac{D_i}{D_c}\right)^{1/3}\left[2gH\left(\frac{\Delta\rho}{\rho_l}\right)\right]^{1/2} \qquad [34]$$

where U_{MS} is the minimum spouting velocity (cm/sec), d_p is the particle diameter (cm), D_c is the evaporator pot diameter (cm), D_i is the spout diameter (cm), g is gravitational acceleration (980 cm/sec^2), H is the bed height (cm), $\Delta\rho$ is the density difference between the fluid and the solid particles (g/cm^3), and ρ_l is the fluid density (g/cm^3). In the case of non-uniform particle size, the correlation is generally accurate to within 25%.

Smith and Reddy developed a correlation to predict the minimum spouting velocity (cm/sec) for non-uniform size particles in conical bottom vessels.[42] Equation [35] describes the correlation

$$U_{MS} = d_{p-avg}\left[\frac{g\Delta\rho}{\rho_l D_c}\right]^{1/2}\left[0.64 + 26.7\left(\frac{D_i}{D_c}\right)^2\right]\left(\frac{H}{D_c}\right)^{0.5-1.76(Di/Dc)} \qquad [35]$$

where d_{p-avg} is the average particle size (cm).

Uemaki et al. developed a correlation to predict the minimum spouting velocity for non-uniform size particles in conical bottom vessels.[40,43] Equation [36] describes the correlation

$$U_{MS} = 0.977\left(\frac{d_{p-avg}}{D_c}\right)^{0.615}\left(\frac{D_i}{D_c}\right)^{0.274}\left[2gH\Delta\rho/\rho_l\right]^{0.324} \qquad [36]$$

where U_{MS} is the minimum spouting velocity (m/sec), d_{p-avg} is the average particle diameter (m), D_c is the column diameter (m), D_i is the spout diameter (m), g is gravitational acceleration (9.8 m/sec^2), H is the bed height (m), $\Delta\rho$ is the density difference between the fluid and the solid particles (kg/m^3), and ρ_l is the fluid density (kg/m^3).

Lamont investigated air agitation with Pachuca tanks, which are similar to spouting beds and have conical bottoms.[44] He found a minimum spout velocity of 0.33 cm/sec was needed for moderate agitation.

The work developing correlations was performed with small diameter vessels. Fane and Mitchell developed a correction to apply the Mathur-Gishler correlation to larger diameter tanks.[45] The correction is described by equations [37] and [38]

$$U_{MS} = U_{MG}\,2\left(\frac{D_c}{D_{c-ref}}\right)^n \qquad [37]$$

$$n = (1 - \exp\{-7(D_c/D_{c\text{-}ref})^2\}) \quad [38]$$

where $D_{c\text{-}ref}$ is the reference column diameter and U_{MS} is the minimum spouting velocity.

Pulsed Jet Mixers

Pulsed Jet Mixers contain a large reservoir with a nozzle located at the bottom and an air supply connected to the top. The air stream pulls a vacuum on the reservoir that draws fluid into it. The air also pressurizes the reservoir to discharge the fluid. The discharging fluid produces a turbulent jet. Because the discharge pulse is short and not continuous, the jet does not have time to become fully developed. The Hanford River Protection Program is currently conducting extensive testing with this technology to evaluate its ability to mix concentrated sludge slurries.

SRS MIXING EXPERIENCE

The Savannah River Site has employed water jets and long shaft vertical centrifugal pumps (i.e., slurry pumps) to remove sludge from waste tanks. In addition, they have used mechanical agitators to mix and suspend particles in smaller tanks. Those experiences will be described in this section.

Actual Waste

Hill investigated the use of high pressure water jets to remove sludge from Savannah River Site High Level Waste Tanks.[46] The jets tested had nozzle diameters of 1/8", 3/16", and 1/4". The jet pressures were 500 – 3000 psig. They discharged horizontally ½ - ¾ inches above the tank bottom. The 1/8" nozzle operating at 500 psig produced a cleaning radius of ~ 6 ft. The ¼" nozzle operating at 3000 psig produced a cleaning radius of ~ 30 ft. No material or rheological properties were provided in the documentation. During tests in which the nozzles rotated, the best mixing occurred with rotation rates less than ½ rpm. The slower rotation rate provides more time for the jet to develop. He did observe that changing the direction of rotation increased the amount of sludge that was removed, with the greatest improvement observed behind obstructions.

Four sets of two ¼" nozzles were installed close to the bottom in Tank 2. The nozzles operated at 500 – 2000 psig and rotated 15 degrees every 5 minutes. Approximately 90% of the original sludge was removed and transferred to Tank 7. The process required 5.3 gallons of water for each gallon of sludge removed.

High pressure water jets were also installed in Tank 9. Nearly all of the sludge was removed from the tank and the water to sludge removed ratio was 5.7.

Bradley et al. reported that high pressure water jets were employed to remove sludge from seven waste tanks between 1966 and 1969.[47] The process removed approximately 95% of the sludge, but it required approximately 5 gallons of water for each gallon of sludge removed.

Hill and Parsons prepared a technical data summary for sludge removal in Tank 16.[48] The target supernate sludge ratio for slurry pumps is 1:1. The number of pumps required for sludge removal depends on the properties of the sludge.

SRS placed slurry pumps in Tank 16 and demonstrated their ability to remove sludge.[49,50] Table 4 shows the results. During the demonstration, 98% of the sludge was removed. During the first transfer, the pump cleaning radius was ~ 20 ft. During the second transfer, it was ~ 30 ft. The demonstration used 1.4 volumes of water per volume of sludge.

Table 4. Tank 16 Sludge Removal Demonstration

Transfer	Pumps	Time (hr)	Cummulative Liquid Volume added (L)	Sludge Transferred (L)	Fraction Sludge Remaining
1	1	86	64,000	44,000	85 %
2	1	208	160,000	59,000	65 %
3	2	75	248,000	139,000	17 %
	3	94			
4	3	76	254,000	44,000	2 %

SRS conducted an in-tank sludge processing demonstration in 1982.[51] Periscopic inspections of the tank showed good mixing and ~ 75% of the aluminum dissolved with four slurry pumps.

SRS performed a salt removal demonstration in 1983 in Tank 19F.[52] During the demonstration, they removed over one million gallons of salt, using one and two slurry pumps. In 1994, SRTC developed a model of the salt dissolution process to help optimize slurry pump operating parameters for salt dissolution.[53] They compared the model predictions with the Tank 19 demonstration and found good agreement.

For salt dissolution, one needs water and time. Mechanical agitation can speed up the dissolution process, by increasing the rate of transport of saturated salt solution away from the salt-supernate interface. The recommendations for salt dissolution with slurry pumps are the following.[53] The supernate should be concentrated to no more than 90% of saturation (~ 1.34 g/ml density, 6.7 - 7.7 molar sodium). The pump flow rate should be full speed. Reducing the pump flow rate will increase dissolution time. The slurry pumps should be lowered in 2.5 – 5 foot increments. Smaller increments are desirable, but the time, cost, and exposure required to lower the pumps must also be considered. Larger increments would significantly increase the salt dissolution time. The recommended volume of dissolution water for one batch is 105,000 - 210,000 gallons. The salt dissolution process should be performed at the highest temperature possible. Increasing the operating temperature will increase salt solubilitity and diffusivity which will increase the salt dissolution rate.

Simulant

Bradley et al. developed a low pressure slurrying technique to improve the sludge removal process.[47] In this process, a pump is immersed in the sludge so that a recirculating mixture of sludge and supernate is the feed to the pump rather than fresh water. The V_jD_j of this technique is the same as the V_jD_j of the water jet. However, because the slurry pump contains a larger diameter nozzle, the jet decays more slowly and has a larger cleaning radius. During simulant testing with kaolin, personnel found no significant changes in cleaning radius or kaolin suspension after 50 hours. They also observed that changing the direction of pump rotation improved the rate of sludge removal from behind large obstacles.

SRTC personnel investigated the ability of the mixing equipment in DWPF to suspend sludge and MST slurries that settle for extended times at elevated temperatures.[54] As the sludge and MST settle and sit undisturbed in the process tanks, they can develop Bingham plastic properties, such as a yield stress. Heating the sludge and MST slurry that sits for extended times at elevated temperature significantly increases the yield stress (1.8 – 16X). Cavern model calculations predicted that the pilot-scale agitation equipment could not suspend slurries in that test, but the equipment did suspend the slurries. The reason the equipment was able to suspend the sludge could be the erosion mechanism. In addition, sludge mobilization work conducted for SRS and Hanford applications showed sludge could be mobilized when the applied shear stress was two orders of magnitude less than the sludge yield stress.[6]

DOE SITE MIXING EXPERIENCE

SRTC, PNNL, and ITT Flygt personnel conducted scaled tests with shrouded axial impeller mixers to determine mixer requirements for suspending sludge heels.[4,5,6,55] The tests were performed with zeolite in scaled tanks which have diameters of 1.5, 6.0, and 18.75 ft. The mixer speeds required to suspend zeolite particles were measured at each scale. The data were analyzed with various scaling methods to compare their ability to describe the suspension of insoluble solids with the mixers and to apply the data to a full-scale waste tank.

Scaling of the suspension of fast settling zeolite particles was best described by the constant power per unit volume method. Increasing the zeolite particle concentration increased the required mixer power needed to suspend the particles. Decreasing the zeolite particle size from 0.7 mm – 0.3 mm decreased the required mixer power needed to suspend the particles. Increasing the number of mixers in the tank decreased the required mixer power needed to suspend the particles by 25%, which is close to $1/n^{1/2}$ from the work of Koh et al.[23] A velocity of 1.6 ft/sec two inches above the tank bottom is needed to suspend zeolite particles.

Powell et al. reviewed work conducted at DOE sites to describe the mixing of sludge in waste tanks.[56] They described their results with equation [39]

$$\text{ECR (cm)} = 4.0\ \text{VjDj}\ \tau_s^{-0.46} \qquad [39]$$

where Vj is the jet discharge velocity (cm/sec), Dj is the jet diameter (cm), and τ_s is the sludge shear strength (dynes/cm^2). They observed that 50 – 500 hours were required before the cleaning radius growth rate became insignificant. They also discussed applying the extensive literature on river bed erosion to sludge mobilization.

MIXING RECOMMENDATIONS

In addressing mixing problems, Closure Business Unit personnel should apply the following approach. First, identify the type of mixing required. The mixing could be for blending miscible fluids, salt dissolution, bulk sludge mixing, or heel removal. If the application contains a liquid phase, determine the density and viscosity of the liquid stream. If the application contains a solid phase, determine the particle size, particle density, and rheology of the solid phase. If the process contains a solid-liquid slurry, determine whether it is fast settling or slow settling. Fast-settling slurries typically have low concentrations of large (> 100 micron), heavy particles. Slow-settling slurries typically have high concentrations of small (< 10 – 20 micron), slow settling, cohesive particles. SRS sludge typically behaves as slow settling solids. Zeolite particles typically behave as fast settling solids.

When trying to suspend fast-settling particles with mechanical agitators, one should use correlations such as the Zwietering correlation (Table 2). If trying to suspend them with bubblers or spargers, apply the spouting bed models described in equations [34] – [38]. If trying to suspend fast-settling slurries with jets, testing is probably required.

When trying to suspend slow-settling particles with mechanical agitators, one should use the cavern model. If trying to suspend slow-settling slurries with jets, one should use the effective cleaning radius model. These models may overpredict the amount of mixing required because of erosion. Modeling the sludge mixing by erosion requires testing

When trying to determine the mixing requirements with mechanical agitators for blending miscible fluids or chemical reactions with miscible fluids, one should use the mixing guidelines provided by Peter Holman (Table 3) and the mixing time equations (equation [6]). With jets, one should use the mixing time calculated with equations [9] – [15]. With spargers or bublers, one should target having a superficial gas velocity of 0.05 – 0.25 ft/s.

For salt dissolution, one needs water and time. Mechanical agitation can speed up the dissolution process. The recommendations for salt dissolution with slurry pumps are the following. The supernate should be concentrated to no more than 90% of saturation (~ 1.34 g/ml density, 6.7 - 7.7 molar sodium). The pump flow rate should be full speed. The slurry pumps should be lowered in 2.5 - 5 foot increments. The recommended volume of dissolution water for one batch is 105,000 - 210,000 gallons. The salt dissolution process should be performed at the highest temperature possible.

FINAL THOUGHTS

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The following mixing consultants could be contracted to help the Closure Business Unit solve mixing problems.

- Art Etchells, retired DuPont mixing consultant
- David Dickey, independent mixing consultant, MixTech, Inc., formerly with Chemineer
- Alvin Nienow, School of Chemical Engineering, University of Birmingham
- Richard Calabrese, Department of Chemical Engineering, University of Maryland
- Peter Holman, independent mixing consultant, formerly with Lightnin

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