Design of a Mixing System for Simulated High-Level Nuclear Waste Melter Feed Slurries

**M. E. Peterson
D. McCarthy
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March 1986

**Prepared for the U.S. Department of Energy
under Contract DE-AC06-76RLO 1830**

**Pacific Northwest Laboratory
Operated for the U.S. Department of Energy
by Battelle Memorial Institute**

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PACIFIC NORTHWEST LABORATORY
operated by
BATTELLE
for the
UNITED STATES DEPARTMENT OF ENERGY
under Contract DE-AC06-76RLO 1830

Printed in the United States of America
Available from
National Technical Information Service
United States Department of Commerce
5285 Port Royal Road
Springfield, Virginia 22161

NTIS Price Codes
Microfiche A01

Printed Copy

Pages	Price Codes
001-025	A02
026-050	A03
051-075	A04
076-100	A05
101-125	A06
126-150	A07
151-175	A08
176-200	A09
201-225	A010
226-250	A011
251-275	A012
276-300	A013

DESIGN OF A MIXING SYSTEM FOR SIMULATED
HIGH-LEVEL NUCLEAR WASTE MELTER
FEED SLURRIES

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Pacific Northwest Laboratory
Richland, Washington 99352

SUMMARY

The United States is proceeding with plans to immobilize nuclear high-level liquid waste by conversion to borosilicate glass followed by burial in a deep geological repository. The vitrification process for converting the waste to glass involves mixing the waste with glass forming chemicals and processing this slurry through a high temperature furnace known as a liquid-fed ceramic melter (LFCM). Maintaining glass product quality control and safe efficient melter operation relies heavily upon the ability of the feed preparation system to produce a homogeneous mixture of waste and glass-former materials. Mixing tank and agitator designs exist in the literature but the applicability of these designs to complex, concentrated high-level waste and glass-former slurries has not been verified.

The purpose of this evaluation was to conduct bench-scale tests that would provide a quantitative description of the performance of a waste/glass-former mixing system in terms of measurable process variables. The results from these tests were used to predict the performance of a large-scale system. The specific objectives of the evaluation were to: 1) examine the effects of selected key variables on the distribution of solids in the mixing tank, and 2) select a reference prototype system for large-scale verification tests.

Bench-scale mixing tests were conducted in a 15-inch diameter clear plastic mixing tank. A statistical experimental design approach was used to assess the effects of impeller diameter, impeller type, impeller rotational speed, use of baffles, use of heating/cooling coils, and type of slurry on the distribution of solids in the mixing tank. The bench-scale mixing system was designed to be geometrically and kinematically similar to large-scale mixing tanks frequently used in solids suspension mixing applications.

This evaluation used nonradioactive melter feed slurries, tailored to simulate the physical and chemical properties of the radioactive slurries expected at two of the vitrification plants currently under design in the United States. One of the slurries represents an alkaline waste that has been treated with formic acid to which glass-formers are added as 1/3 glass frit and 2/3 glass-forming chemicals. This slurry is referred to as the Hanford Waste

(HW) slurry. The other slurry selected is a high nitrate waste to which inorganic ion exchange material and glass-forming chemicals have been added. This slurry is referred to as the West Valley (WV) waste slurry. The chemical, physical and rheological properties of the simulated melter feed slurries were evaluated in the laboratory. Based on chemical, physical, and rheological properties, the selected slurries were judged to be good candidates for evaluating mixing vessel designs.

The rheological properties displayed by the two simulated slurries allowed them to be classified as yield-pseudoplastic non-Newtonian fluids. Mathematical models have been developed to characterize this behavior. These models were applied to actual laboratory and bench-scale data and used to predict the level of turbulence that was present in the bench-scale mixing tank during each test. The level of turbulence was correlated with the degree of mixing achieved.

The primary overall trends revealed by the bench-scale experiments were that: 1) the distribution of solids in the mixing tank became more uniform as rotational speed, N , increased, 2) HW slurry is easier to mix than WV slurry at low values of N , 3) a high-efficiency turbine design and a standard 45-degree pitched-blade turbine achieved very similar mixing results, 4) impeller diameter was only significant at low values of N , 5) baffles improved mixing at low values of N but had little effect at higher rotational speeds, and 6) heating/cooling coils reduced the degree of mixing achieved at low values of N . Predictive equations were developed to quantify the magnitude of these trends.

The bench-scale predictive equations were scaled-up to predict the performance that is likely to occur in a large-scale mixing system. Scale-up of the bench-scale data included an assessment of the level of power needed to maintain solids in a uniform suspension, and an estimate of the power required to resuspend solids once they had settled. Detailed evaluations of large-scale mixing system designs for the full-scale vitrification plants at the Defense Waste Processing Facility (DWPF) located at the Savannah River (SR) Plant, and the Component Test Stand (CTS) located at the West Valley (WV) Plant were conducted using the HW and WV slurries. The mixing system for DWPF was designed for a Bingham Plastic slurry with a yield stress of 15 N/m^2 and an insoluble solids content of 48 wt%. The mixing system for CTS was designed for a slurry

very similar to the slurry used in these bench-scale tests. Mixing system designs used in mining and metallurgical industrial applications were also evaluated.

The shell of a large-scale mixing tank was selected for scale-up. This shell, a vertical cylinder with a flat bottom, had a diameter of 9 feet and a design slurry depth of 9 feet. A power level of less than one hp (which induces a bulk fluid velocity of 5 to 15 ft/min) will be sufficient to maintain the solids in the HW and WV melter feeds in a uniform suspension (± 1 percent of the mean value of total solids). This unusually low power requirement (and bulk velocity) are due to the very slow settling rates exhibited by the simulated HW and WV slurries (0.45 and 0.60 cm/hr, respectively). The yield-stress exhibited by each slurry is primarily responsible for the slow settling velocity of the solids in the slurry.

In addition to maintaining solids in a uniform suspension, the large-scale mixer must be able to resuspend solids if they settle out as a result of a power outage or other disturbance. The scale of agitation necessary to resuspend solids is higher than the value needed to maintain solids in suspension. Historically, bulk fluid velocities of approximately 50 ft/min are specified for solids suspension mixing applications where a uniform suspension of solids is required. Summarized below is the power that must be applied to HW and WV slurries for the large-scale mixing system to achieve a bulk velocity of 50 ft/min.

Mixing System Design	Impeller Diameter (feet)	Upper Impeller Design	Design Impeller Rotational Speed (rpm)	Shaft Power (hp)	
				HW	WV
DWPF (SR)	2.25	High Efficiency Turbine	190	5.5	5.1
CTS (WV)	2.70	High Efficiency Turbine	100	7.6	8.2
Industrial A ^(a)	4.5	45° pitched blade turbine	20	8.1	7.9
Industrial B ^(a)	2.70	45° pitched blade turbine	100	9.4	9.3

(a) Industrial A and B represent designs typically used in mining and metallurgical industries.

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1.0 INTRODUCTION AND OBJECTIVES

The liquid fed ceramic melter process is the reference process in the United States for treating both defense and civilian high-level liquid waste. Development of the current technology has been underway for approximately 10 years and as a consequence is nearly complete. The design and operating parameters for safe, efficient operation have been defined and successful design concepts identified. Documentation of past development and design work, resolution of a few remaining technical issues, and demonstration of remote process and product quality control/verification models are the key areas where work remains.

The Nuclear Waste Treatment Program (NWTP) at the Pacific Northwest Laboratory (PNL) focuses on generic research to permit the future commercialization of high level waste (HLW) treatment technologies and to support the research and development needs of the HLW demonstration program being conducted at the Department of Energy (DOE) facility at West Valley, New York. This facility was formerly operated by Nuclear Fuel Services Company (NFS) as a commercial nuclear fuels reprocessing plant. A primary objective of the NWTP is to complete development, testing, and process modeling of high-level waste treatment technology. This program will help to: 1) establish the acceptability of the selected waste forms for storage and disposal, 2) verify the reliability and licensability of processes for producing the waste forms, and 3) demonstrate the technology by solidifying actual radioactive wastes.

The NWTP development program consists of coordinated nonradioactive and radioactive testing combined with numerical modeling of the process to provide a complete basis for design and operation of a vitrification facility. The radioactive demonstration tests of equipment and processes are conducted before incorporation in radioactive pilot-scale melter systems for final demonstration. The mixing system evaluation described in this report was conducted as part of the nonradioactive testing.

The mixing tank is an integral part of the feed preparation system. The nuclear waste is blended with glass forming materials and other additives in this tank. The ability of the feed preparation system to produce and maintain

a homogeneous melter feed within compositional tolerances strongly affects process control and quality control of the LFCM process. The purpose of this mixing evaluation was to provide a quantitative description of the performance of a feed mixing system in terms of measurable process variables. The specific objectives of the evaluation were to: 1) examine the effects of impeller design, impeller diameter, impeller rotational speed, the presence of baffles, and of heating/cooling coils on the distribution of solids in the mixing tank, and 2) select a reference prototype system for large-scale verification tests.

The format of this report follows the sequence in which the design of a large-scale mixing system is determined. The initial program activity was concerned with gaining an understanding of the theoretical foundation of non-Newtonian mixing systems. Section 3 of this report describes the classical rheological models that are used to describe non-Newtonian mixing systems. Since the results obtained here are only valid for the slurries utilized, Section 4, Preparation of Simulated Hanford and West Valley Slurries, describes how the slurries were prepared. The laboratory-scale viscometric and physical property information is summarized in Section 5, Laboratory Rheological Evaluations. The bench-scale mixing evaluations conducted to define the effects of the independent variables described above on the degree of mixing achieved with each slurry are described in Section 6. Bench-scale results are scaled-up to establish engineering design requirements for the full-scale mixing system in Section 7.

2.0 CONCLUSIONS/RECOMMENDATIONS

The bench-scale mixing tests provided a quantitative description of the performance of the feed mixing system in terms of measurable process variables. The specific objectives of the evaluation were to: 1) examine the effects of impeller design, impeller diameter, impeller rotational speed, the presence of baffles and of heating/cooling coils on the distribution of solids in the mixing tank, and 2) select a reference prototype system for large-scale verification tests. The specific conclusions and recommendations are provided below.

This evaluation used two different simulated melter feed slurries: a typical Hanford Waste (HW) slurry and a West Valley (WV) slurry. The chemical, physical and rheological properties of the simulated HW and WV melter feed slurries were evaluated in the laboratory. Some significant physical properties of these slurries at 20°C are listed in Table 1.

The rheological properties displayed by the simulated HW and WV slurries allowed them to be classified as yield-pseudoplastic non-Newtonian fluids. Mathematical models have been developed to characterize this behavior. These models were applied to actual laboratory and bench-scale data and used to predict the level of turbulence that was present in the bench-scale data and mixing tank during each test. The level of turbulence was correlated with the degree of mixing achieved.

The primary overall trends revealed by the bench-scale experiments were that: 1) the distribution of solids in the mixing tank became more uniform as rotational speed, N , increased, 2) HW slurry was easier to mix than WV slurry at low values of N , 3) a high efficiency turbine design and a standard

TABLE 1. Physical Properties of HW and WV Slurries at 20°C

<u>Slurry</u>	<u>Density (g/cm³)</u>	<u>pH</u>	<u>Dissolved Solids (wt%)</u>	<u>Suspended Solids (wt%)</u>	<u>Total Solids (wt%)</u>	<u>Settling Rate (cm/hr)</u>
HW	1.29	8.3	20.9	20.8	41.7	0.45
WV	1.26	7.4	14.8	21.7	36.5	0.60

45-degree pitched blade turbine achieved very similar mixing results, 4) impeller diameter was only significant at low values of N , 5) baffles improved mixing at low values of N but had little effect at higher rotational speeds, and 6) heating/cooling coils reduced the degree of mixing achieved at low values of N . Predictive equations were developed to quantify the magnitude of these trends.

The bench-scale predictive equations were scaled-up to predict the performance that is likely to occur in a large-scale mixing system. Large-scale mixing system designs for the full-scale vitrification plants at the Defense Waste Processing Facility (DWPF) located at Savannah River (SR) Plant, and the Component Test Stand (CTS) located at the West Valley (WV) Plant and for mining and metallurgical industrial sites were evaluated in detail to determine their performance with HW and WV slurries. The mixing system for DWPF was designed for a Bingham Plastic slurry with a yield stress of 15 N/m^2 and an insoluble solids content of 48 wt%. The mixing system for CTS was designed for a slurry very similar to the slurry used in these bench-scale tests.

A large-scale mixing tank shell was selected for scale-up. This shell was a vertical cylinder with a flat bottom. It had a diameter of 9 feet and a design slurry depth of 9 feet.

Scale-up of the bench-scale data included an assessment of the level of power needed to maintain solids in a uniform suspension, and an estimate of the power required to resuspend solids once they had settled. A power level of less than one hp (which induces a bulk fluid velocity of 5 to 15 ft/min) will be sufficient to maintain HW and WV solids in a uniform (± 1 percent of the mean value of total solids) suspension. This unusually low power requirement (and bulk velocity) is due to the very slow settling rates exhibited by the simulated HW and WV slurries (0.45 and 0.60 cm/hr, respectively). The yield-stresses exhibited by each slurry is primarily responsible for the slow settling velocity of the solids in the slurry.

The large-scale mixer must be able to resuspend solids if they settle out as a result of a power outage or other disturbance. The scale of agitation necessary to resuspend solids is higher than the value needed to maintain solids in suspension. Historically, bulk fluid velocities of approximately

50 ft/min are specified for solids suspension mixing applications where a uniform suspension of solids is required. The power that must be applied to HW and WV slurries for the large-scale mixing system estimated to achieve a bulk velocity of 50 ft/min, is summarized in Table 2.

The motor used to operate the mixer will have a power rating higher than those listed in Table 2. This motor must overcome energy losses that will occur in the motor and the gearbox. It must also possess a start-up torque that can overcome the yield stress of the slurry in the tank. The yield stress will increase as the length of time the slurry remains motionless in the tank increases. The effect of time on this yield stress value was not evaluated. This evaluation should be performed before the motor size is specified. It is likely that this start-up torque requirement will be the critical factor that will fix the size of the motor needed for melter feed slurry mixing applications.

The large-scale impeller diameter and rotational speed can be specified once the need for and design of internal heating/cooling coils has been established. If coils are not required, the WV, Industry A or Industry B designs

TABLE 2. Calculated Large-Scale Mixer Shaft Power Required to Achieve a Bulk Velocity of 50 ft/min

Mixing System Design	Impeller Diameter (feet)	Upper Impeller Design	Design Impeller Rotational Speed (rpm)	Shaft Power (hp)	
				HW	WV
DWPF (SR)	2.25	High Efficiency Turbine	190	5.5	5.1
CTS (WV)	2.70	High Efficiency Turbine	100	7.6	8.2
Industrial A ^(a)	4.5	45° pitched blade turbine	20	8.1	7.9
Industrial B ^(a)	2.70	45° pitched blade turbine	100	9.4	9.3

(a) Industrial A and B represent designs typically used in mining and metallurgical industries.

could be specified. The WV or Industry B designs are preferred because they permit the specification of standard variable speed drive units. These units should be specified for each large-scale melter feed slurry application since the degree of agitation required is likely to vary with time as slurry composition and viscosities change.

The HW and WV melter feed slurries selected for this study do not necessarily represent the final waste glass composition to be processed at the vitrification plants. While the composition of the actual slurries may change, the data acquisition, data analysis, and mixing scale-up methods described in this report can be used to obtain estimates of the significance of compositional or other geometric or kinematic changes on the degree of mixing that can be achieved in full-scale melter feed mixing tanks. These methods provide a low-cost and rapid means of obtaining information that can be used to formulate efficient and cost effective designs for a wide variety of mixing systems.

3.0 MIXING OF NON-NEWTONIAN SLURRIES

Rheology is the field of study concerned with the deformation and flow of materials. The procedures used to design slurry mixing processes are based on classical rheological models for homogeneous and heterogeneous flow systems. These models were used to classify Hanford and West Valley melter feed slurries, and scale-up the bench-scale data collected to predict the behavior of the slurries in the full-scale mixing system. The accuracy of these predictions will depend on the degree of similarity that exists between the properties of the full-scale system, the bench-scale system and the underlying theoretical model. The theoretical models used to define the non-Newtonian behavior of Hanford and West Valley melter feed slurries are described in Sections 3.1 and 3.2. The design of solid-liquid mixing systems is discussed in Section 3.3.

3.1 CLASSIFICATION OF SLURRIES

A complex mixture can be defined as a single-phase, non-Newtonian fluid, a multiphase mixture of a fluid and a solid, or a mixture of two or more fluids. A classification of complex mixtures based on the effect of phase condition or flow behavior is depicted on Figure 1. A multiphase mixture can be classified as a fine dispersion, a coarse dispersion, a macro-mixed flow pattern, or as a stratified flow pattern.

Single Phase	Multi-Phase (Gas-Liquid, Liquid-Liquid, Gas-Solid, Liquid-Solid)			
	Fine Dispersion		Coarse Dispersion	Macro-Mixed Stratified
True Homogeneous	Pseudo-Homogeneous		Heterogeneous	
	Laminar Turbulent	Turbulent Only		
← Flow Behavior as Single Phase →		← Flow Behavior as Multi-Phase →		

FIGURE 1. Classification of Complex Mixtures

Macro-mixed and stratified heterogeneous flow behavior applies to gas-liquid and liquid-liquid mixtures. Examples of a macro-mixed flow are frothy and highly turbulent mixtures of gas and liquid or of immiscible liquids under conditions where neither phase is continuous. Stratified flow exists when each phase of a gas-liquid mixture or a mixture of two immiscible liquids is continuous.

For liquid-solid slurries a fine dispersion is a mixture of fine particles of solids that are uniformly dispersed in a continuous liquid phase. Solids with a diameter less than about 40 μm usually form fine dispersions and can be considered to be pseudohomogeneous mixtures (Bain and Bonnington 1970). Solid particles with a diameter greater than about 40 μm usually form coarse dispersions, have a tendency to settle, and can be considered to be heterogeneous mixtures. The techniques used to design mixing systems to achieve pseudohomogeneous and heterogeneous mixtures differ markedly. These procedures are described in Section 3.3.

3.2 PSEUDOHOMOGENEOUS MIXTURES

Viscous, single-phase fluids and pseudohomogeneous multiphase fluid mixtures are classified according to their response to shearing stresses. A shear stress is defined as:

$$\tau = \frac{F}{A} \quad (1)$$

where τ = shear stress, dyne/cm^2

F = force, dyne

A = area, cm^2

For Newtonian fluids in laminar flow:

$$\tau = \mu \left(\frac{dv}{dy} \right) \quad (2)$$

where τ = shear stress, dyne/cm²

μ = viscosity, dyne sec/cm².

$\frac{dv}{dy}$ = shear rate in y direction, sec⁻¹

The viscosity, μ , is a proportionality constant that is independent of shear rate and affected only by pressure and temperature for a Newtonian fluid system. A plot of shear stress versus rate of shear in the laminar flow region is known as a rheogram (see Figure 2a). The slope of the curve is constant for a Newtonian fluid. All fluids that display rheograms that are not linear through the origin at a given temperature and pressure are considered to be non-Newtonian. Non-Newtonian fluids are usually classified as time-independent, time-dependent or viscoelastic fluids.

FIGURE 2. Rheograms for Time-Independent Non-Newtonian Fluids

3.2.1 Time-Independent Non-Newtonian Fluids

Time-independent non-Newtonian fluids are fluids for which rate of shear at a given point is solely dependent upon instantaneous shear stress at the point. Rheograms for classical time-independent non-Newtonian fluids are depicted on Figure 2a. The relationship between apparent viscosity and shear rate for these fluids is described on Figure 2b. The apparent viscosity is a function of shear rate for non-Newtonian fluids and is analogous to the Newtonian viscosity for Newtonian fluids. However, whereas the Newtonian viscosity does not vary with shear rate, the non-Newtonian apparent viscosity does.

3.2.1.1 Fluids Without a Yield Stress

Dilatant and pseudoplastic non-Newtonian slurries do not exhibit a shear stress at a shear rate of zero. Pseudoplastic and dilatant materials exhibit a continuous variation of shear stress with shear rate, the slope of the curve decreasing (pseudoplastic) or increasing (dilatant) as shear rate increases (see Figure 2a). Equations that relate shear stress and shear rate for these fluids are included in Table 3.

The majority of non-Newtonian slurries exhibit pseudoplastic behavior. The power law equation is widely used to describe the behavior of non-Newtonian fluids. This equation for different types of fluids is shown in Table 3. The parameter K is the fluid consistency index and n is the flow behavior index. For Newtonian fluids, $n=1$ and the power law equation reduces to Equation (2). The apparent viscosity for pseudoplastic fluids is defined as:

$$\mu_a = \frac{\tau}{\left(\frac{dv}{dy}\right)^n} = K \left(\frac{dv}{dy}\right)^{n-1} \quad (3)$$

Since the exponent n is less than unity for pseudoplastic fluids (see Table 3), the apparent viscosity μ , decreases with increasing shear rate. Pseudoplastic fluids include rubber solutions, adhesives, polymer solutions, starch suspensions, mayonnaise, soap, detergent slurries, Napalm, paints, biological sludges, and many other solid-liquid suspensions.

TABLE 3. Equations for Time-Independent Non-Newtonian Fluids

Type of Fluid	General Equation	Level of Exponent		
		τ_y	n	K
Yield-Dilatant	$\tau - \tau_y = K \left(\frac{dv}{dy} \right)^n$	>0	>1	---
Dilatant	$\tau = K \left(\frac{dv}{dy} \right)^n$	=0	>1	---
Yield-Pseudoplastic	$\tau - \tau_y = K \left(\frac{dv}{dy} \right)^n$	>0	<1	---
Pseudoplastic	$\tau = K \left(\frac{dv}{dy} \right)^n$	=0	<1	---
Bingham Plastic	$\tau - \tau_y = K \left(\frac{dv}{dy} \right)^n$	>0	=1	μ_p
Newtonian	$\tau = K \left(\frac{dv}{dy} \right)^n$	=0	=1	μ

Dilatant fluids are much less common than pseudoplastic fluids. Dilatancy is usually observed only at select solids concentrations in suspensions of irregularly shaped solids. The apparent viscosity of dilatant fluids increases with increasing shear rate since the exponent n (see Table 3) is greater than one. Quicksand, wet beach sand, and some aqueous suspensions of titanium dioxide are examples of dilatant fluids (Skelland 1967).

3.2.1.2 Fluids with a Yield Stress

Bingham plastic, yield-pseudoplastic and yield-dilatant pseudohomogeneous slurries exhibit a yield stress at a zero shear rate. This yield stress is caused by the internal structure created by the small particles in the fluid. The structure of particles prevents movement of the fluid for values of the shear stress, τ , less than the yield value, τ_y . When τ exceeds τ_y , the internal particle structure changes, and fluid movement begins. Empirical equations that describe these fluids are included in Table 3.

Bingham plastic fluids include thickened hydrocarbon greases, some asphalts and bitumens, and a variety of emulsions. The most common Bingham fluids include water suspensions of clays, fly ash, quartz, metallic oxides, sewage sludge and paint systems. Yield-pseudoplastic materials include

clay-water suspensions, solutions of carbonpolymethylene (carbopol), some grout formulations, rubber-benzene solutions, phosphate ore, and many coal-water slurries.

Yield-dilatant fluids are much less common than Bingham plastic or yield-pseudoplastic fluids. Yield-dilatant fluids will begin to flow when their yield stress is exceeded. Once motion begins, the rate of increase in shear stress with velocity gradient increases as the velocity gradient increases. Some solutions of titania (Metzner and Whitlock 1958), magnetite and galena (Govier et al. 1957) can exhibit this behavior.

3.2.2 Time-Dependent Non-Newtonian Fluids

The behavior of time-independent non-Newtonian fluids adjusts instantaneously as shear stresses are applied. The structural configuration and particle orientation of some slurries cannot adjust instantaneously to changes in shear stresses. These time-dependent non-Newtonian fluids are usually classified as thixotropic or rheopectic slurries. Rheograms for these slurries are depicted on Figure 3.

Thixotropic slurries exhibit a decrease in shear stress with time at a constant shear rate and temperature. A hysteresis loop (Figure 3) will result

FIGURE 3. Rheograms for Thixotropic and Rheopectic Non-Newtonian Fluids

as the shear rate is steadily increased from zero to a maximum value and then decreased steadily back to zero. The internal particle alignment mechanisms responsible for thixotropic behavior are similar to, but occur at a slower rate than, those causing pseudoplastic behavior. Some oil-well drilling muds, greases, margarines, shortenings, printing inks, and high molecular weight polymers exhibit thixotropic behavior (Jobling and Roberts 1959). Extensive surveys of thixotropic behavior have been prepared (Govier and Aziz 1982).

Rheopectic slurries exhibit an increase in shear stress with time at a constant shear rate and temperature. A hysteresis loop (Figure 3) will result as the shear rate is steadily increased from zero to a maximum value and then decreased steadily back to zero. The hysteresis loop produced is a reverse-image of the loop created by thixotropic materials. The internal particle alignment mechanisms responsible for rheopectic behavior are similar to, but occur at a slower rate than, those causing dilatant behavior. Some bentonite clays, vanadium pentoxide, gypsum, and ammonium oleate suspensions exhibit rheopectic behavior (Skelland 1967). Extensive surveys of rheopectic behavior have been prepared (Bauer and Collins 1967).

3.2.3 Viscoelastic Fluids

Viscoelastic fluids exhibit time-dependent non-Newtonian viscous behavior and elastic recovery properties. They flow when subjected to stress but part of their deformation is gradually recovered when the stress is removed. Viscoelastic fluids include bitumens, flour dough, Napalm, nylon, and nearly all molten polymers. Solid-liquid slurries rarely display significant degrees of viscoelastic behavior. Constitutive equations that describe viscoelastic behavior have been developed (Skelland 1967; Govier and Aziz 1982).

3.3 SOLID-LIQUID MIXING SYSTEMS

The design of solid-liquid mixing systems requires an understanding of the fundamental fluid dynamic relationships that exist in the mixing system. This understanding must be tempered with knowledge gained during experience with similar mixing applications. This experience usually permits the results of bench-scale mixing evaluations to be scaled-up to predict the performance that is likely to occur in the full-scale mixing tank. A brief summary of the

fundamental fluid dynamic relationships that occur in mixing tanks is presented in Section 3.3.1. Practical design considerations are discussed in Section 3.3.2. Scale-up methods that utilize these theoretical and practical relationships are discussed in Section 3.3.3.

3.3.1 Significant Theoretical Relationships

A mass and momentum balance around a mixing vessel for a constant density, Newtonian fluid yields the following relationship (Bird et al. 1960, Dickey and Fenic 1976, Bates et al. 1966):

$$N_p = f \left[N_{Re}, N_{Fr}, \frac{T}{D}, \frac{L}{D}, \frac{C}{D}, \frac{p}{D}, \frac{W}{D}, \frac{B}{B_0} \right] \quad (4)$$

where: $N_p = \text{Power Number} = \frac{P}{\rho N^3 D^5}$

$$N_{Re} = \text{Reynolds Number} = \frac{\rho N D^2}{\mu}$$

$$N_{Fr} = \text{Froude Number} = \frac{N^2 D}{g}$$

B = Number of blades

B_0 = Reference number of blades

C = Clearance between the impeller and tank bottom, m

D = Impeller diameter, m

L = Liquid level in tank, m

p = Impeller pitch, m

P = Power applied to fluid, w

N = Impeller rotational speed, rps

W = Impeller blade width, m

T = Diameter of tank, m

μ = Fluid viscosity, Pa · s

ρ = Fluid density, kg/m³

g = Gravitational constant, m/s²

Equation (4) implies that the power number, N_p , is a function of the Reynolds number, N_{Re} , the Froude number, N_{Fr} , and the geometric configuration of the

mixing system. The relationships among these parameters and among other parameters that arise from a heat balance around the mixing vessel (see Table 4) can be defined during pilot-scale tests and be scaled-up to predict the full-scale performance of a mixing system.

The relationship between the power number and the Reynolds number will depend on the configuration of the impeller and the mixing tank. The relationship that results for several impeller designs is depicted on Figure 4. The power correlations in Figure 4 indicate three ranges of liquid motion: turbulent, transition and viscous. Fully turbulent agitation occurs above a mixing Reynolds number of approximately 10,000. The viscous or laminar range occurs below a Reynolds number of about 20. The range between these limits can best be described as transition flow since flow patterns change depending on the Reynolds numbers. In agitation, transition flow occurs over a broader range of Reynolds numbers than is typical of flow in pipes.

The correlation in Figure 4 shows that the power number for any particular geometry becomes constant in baffled tanks. The turbulent range is important for agitator design, since nearly all low- and medium-viscosity applications of industrial equipment operate in this range (Dickey and Fenic 1976).

The blending time t_b^* , and pumping number N_Q , for a given agitator and mixing tank configuration will also depend on the mixing Reynolds number. These relationships are shown on Figure 5 for a turbine agitator system. The blending time is the time necessary to reach a specified concentration distribution

TABLE 4. Dimensionless Groups for Fluid Mixing Applications

Name	Formula	Relationship
Blend Time	$t_b^* = t_b N$	time
Froude Number	$N_{Fr} = DN^2/g$	inertial/gravitational forces
Peclet Number	$N_{Pe} = N_{Re} N_{Pr} = \rho C_p ND^2/k$	convection/conduction
Power Number	$N_p = Pg/\rho N^3 D^5$	imposed/inertial forces
Prandtl Number	$N_{Pr} = C_p \mu/k$	convection/conduction
Pumping Number	$N_Q = Q/ND^3$	turnovers
Reynolds Number	$N_{Re} = D^2 N \rho/\mu$	inertial/viscous forces
Schmidt Number	$N_{Sc} = \mu/\rho D_{AB}$	viscous/diffusive forces

Source: Dickey and Fenic 1976

FIGURE 4. Typical Power Number Versus Reynolds Number Curves for Fully-Baffled Tanks

in a tank. Blending time will be relatively constant in the viscous and turbulent flow regions (see Figure 5a), with more time required to achieve a mixed condition at low Reynolds numbers. The pumping number N_Q is an indicator of the pumping capacity of an agitator. It is relatively constant in the viscous and turbulent regions (see Figure 5b).

The values of N_p , N_Q , and t_b^* are relatively constant in the turbulent flow region. The identification of these values provides the basis for the design of solid-liquid agitation systems. One objective of the bench-scale tests conducted during this test program was to define these relationships for the agitator-tank configurations that are being considered for the West Valley and Hanford melter feed slurries.

FIGURE 5. Pumping Number and Blend Time Versus Reynolds Number

Each agitator-tank configuration will induce a unique flow pattern at a given Reynolds number. The turbulent flow patterns induced by pitched and flat blade turbines are shown on Figure 6. The pitched-blade turbine creates a primarily axial flow pattern. This flow pattern is common in solids suspension applications since it tends to prevent solids deposition on the bottom of the mixing tank. A flat-blade turbine induces a radial flow pattern. The flow pattern is often used in gas dispersion applications.

3.3.2 Practical Design Considerations

The selection of solid-liquid agitation systems requires an understanding of the principles discussed in Section 3.3.1 and of the practical design restraints that will impact each application. A logic diagram for the

a. Axial-Flow Pattern With Pitched Blade

b. Radial-Flow Pattern With Flat Blade

FIGURE 6. Flow Patterns--Turbine Agitators

selection of a solid-liquid agitator that incorporates these theoretical and practical requirements is shown on Figure 7. The first step in the selection process is to define the magnitude of the agitation problem. The size and difficulty of the mixing task, the process result, and the required dynamic response must also be specified (Gates et al. 1976).

The size and difficulty of the mixing problem depends primarily on the volume of slurry to be mixed, its viscosity and the settling velocity of the solids in the slurry. The desired process result could be to blend or dissolve solids added into the mixture, or to induce a desired heat or mass transfer. The required dynamic response will depend on the degree of solids suspension or bulk fluid velocity required in the vessel.

The required dynamic response will be related to the level of agitation that must exist in the tank. An agitation scale ranging from 1 to level 10 has

Source: Gates et al., 1976

FIGURE 7. Logic Diagram for the Selection of Solid-Liquid Agitation Systems

been prepared to define the degree of solids suspension that exists in a mixing tank (Gates et al. 1976). This scale of agitation is summarized in Table 5. Since a near homogeneous melter feed is required to assure HLW glass waste form qualification, a scale of agitation of 8 to 10 will be necessary for the HLW melter feed slurry application.

The design of an agitation system that will achieve the required scale of agitation can be based on previous experience with similar agitation problems and/or on bench-, pilot- or large-scale tests with an agitation system that is

TABLE 5. Scale of Agitation for Solids Suspension

Scale of Agitation	Description
1-2	<p>Agitation levels 1-2 characterize applications requiring minimal solids-suspension levels to achieve the process result.</p> <p>Agitators capable of scale levels of 1 will:</p> <ul style="list-style-type: none"> ● Produce motion of all of the solids of the design-settling velocity in the vessel. ● Permit moving fillets of solids on the tank bottom, which are periodically suspended.
3-5	<p>Agitation levels 3-5 characterize solids-suspension applications for most chemical-process industries. This scale range is typically used for dissolving solids.</p> <p>Agitators capable of scale levels of 3 will:</p> <ul style="list-style-type: none"> ● Suspend all of the solids of design-settling velocity completely off the vessel bottom. ● Provide slurry uniformity to at least one-third of fluid-batch height. ● Be suitable for slurry drawoff at low exit-nozzle elevations.
6-8	<p>Agitation levels 6-8 characterize applications where the solids-suspension level approaches uniformity.</p> <p>Agitators capable of scale level 6 will:</p> <ul style="list-style-type: none"> ● Provide concentration uniformity of solids to 96% of the fluid-batch height. ● Be suitable for slurry drawoff up to 80% of fluid-batch height.
9-10	<p>Agitation levels 9-10 characterize applications where the solids-suspension uniformity is the maximum practical.</p> <p>Agitators capable of scale level 9 will:</p> <ul style="list-style-type: none"> ● Provide slurry uniformity of solids to 98% of the fluid-batch height. ● Be suitable for slurry drawoff by means of overflow.

Source: Gates et al. 1976

geometrically, kinematically (velocity) and dynamically (power) similar to the design of the actual full-scale agitation system. The geometry of an agitation system designed to suspend solids is described in Figure 8. A summary of the geometric ratios recommended for this application by several investigators is provided in Table 6. The consensus of these recommendations is that the agitation system should consist of: 1) a cylindrical, baffled, vessel with a flat or dished bottom that contains slurry to a height roughly equal to the diameter of the vessel, and 2) one or two axial flow turbines with diameters equal to about 30 to 50% of the diameter of the vessel.

The fluid velocities required to suspend solids will depend on the geometry of the vessel and impellers, properties of the solids (size, density,

FIGURE 8. Typical Agitation System Geometry for the Suspension of Solids

TABLE 6. Geometric Ratios for Agitation Systems Designed to Suspend Solids

Ratio	Investigator					
	Garrison ^(a) 1983	Gates et al. 1976	Hicks et al. 1976	Lyons 1967	Nagata 1975	Oldshue 1983
L/T	1.3	<1.2, one Impeller <1.8, two Impellers	<1.2, one Impeller <1.8, two Impellers	0.6, one Impeller >0.6, two Impellers	<1, one Impeller >1, two Impellers	0.6-0.7, one Impeller >1.0, two Impellers
D/T	0.25-0.4	0.3-0.5	0.3-0.5	0.2-0.4	0.35-0.5	0.3-0.5
Type of Impeller	Axial Turbine	Axial Turbine	Axial Turbine	Various	Axial Turbine	Axial Turbine
No. of Impellers	1, L/T < 1.2 2, L/T > 1.2	1, L/T < 1.2 2, L/T > 1.2	1, L/T < 1.2 2, L/T > 1.2	1, L < 2D 2, L > 2D	1, L/T < 1 2, L/T > 1	1, L/T < 1 2, L/T > 1
W/D	0.15-0.25	0.15-0.25	0.15-0.25	--	0.1-0.2	0.2
C ₁ /T	0.4	<0.3, one Impeller <0.25, two Impellers	<0.3, one Impeller <0.25, two Impellers	0.2-0.4	0.1, two Impellers 0.3, one Impeller	--
C ₂ /T	~0.6	~0.6	0.6-0.8	0.2-0.4	--	--
B _w /T	0.083	0.083	0.083	--	0.1	0.083
B _g /T	0.014	0.014	0.014	--	--	0.042

(a) For 100 cP fluids.

shape, surface properties), properties of the liquid (depth, viscosity), and the desired process result. Bulk liquid velocities associated with the scale of agitation for blending applications have been established (Hicks et al. 1976). These velocities are summarized in Table 7.

The bulk fluid velocity in a vessel is created by the rotation of the impeller(s). The fluid velocity at the tip of the impeller will exceed the bulk fluid velocity. Impeller tip speeds associated with gas, liquid, and solid dispersion applications have been identified (Garrison 1983). These tip speeds are summarized in Table 8.

The power needed to achieve the desired process response will depend on the process characteristics of the agitator and the geometry of the vessel in which it is located. Calculation procedures to define the power required for blending (Hicks et al. 1976) and solids suspension (Gates et al. 1976) have

TABLE 7. Bulk Velocities Created at Various Scales of Agitation

<u>Scale of Agitation</u>	<u>Bulk Velocity (ft/min)</u>
1	6
2	12
3	18
4	24
5	30
6	36
7	42
8	48
9	54
10	60

TABLE 8. Impeller Tip Speeds (ft/min) for Dispersion Applications

<u>Application</u>	<u>Course Dispersion (>100 μm)</u>	<u>Fine Dispersion (<100 μm)</u>
Gas Dispersion	600-800	1000-1200
Liquid-Dispersion	600-800	1000-3000
Solid Dispersion	600-800	1000-5000

been established. The power must be delivered by commercially available electric motors through standard American Gear Manufacturers Association (AGMA) gear reducers. A chart of the power and axial turbine shaft speed needed to suspend spherical particles with a settling velocity of 10 ft per minute has been prepared (Gates et al. 1976). This chart is included as Table 9. The chart identifies the power and impeller speed combinations required to achieve different scales of agitation in vessels of varying sizes. This chart can be used to approximate ($\pm 30\%$) mixer requirements for many industrially significant solid suspension applications.

3.3.3 Scale-Up of Bench- and Pilot-Scale Data

Bench and/or pilot-scale tests are usually conducted when: 1) the mixing application is new, 2) the fluids agitated are expensive or hazardous, 3) the fluid properties change significantly during the agitation sequence, 4) several components are added to the agitation tank during the agitation sequence, and 5) when a clear understanding of the mixing process is required. Bench or pilot-scale tests usually follow the path outlined in Figure 9 (Garrison 1983). The first step is to define the goal of the mixing process. Once the goal is established, the class of mixing required to achieve the goal can be identified. Several classes of mixing have been established (Garrison 1983). These classes are summarized in Table 10. The process then continues along the path described on Figure 9.

Bench and pilot-scale units are designed to be as geometrically, kinematically and dynamically similar to full-scale equipment as practical. Geometric similarity requires that the ratio of corresponding dimensions must be the same in bench-, pilot-, and full-scale equipment. Emphasis is placed on maintaining L/T , D/T , C/D , W/D , and B_w/T ratios.

Kinematic similarity is defined in terms of corresponding times or velocities. The bulk velocity in an agitation system is induced by the rotation of an impeller. The bulk velocity is related to the scale of agitation in the mixing vessel (Tables 5 and 7). This scale of agitation can be related to t_b and N_Q (Bowen 1985). These ratios become constant in the turbulent flow region (Figure 5). As a result, kinematic similarity occurs with geometric similarity in turbulent agitation (Rautzen et al. 1976). A specific level of bulk

TABLE 9. Power and Shaft Speed (hp/rpm) Needed to Suspend Solids with a Settling Velocity of 10 ft/min

Scale of Agitation	Equivalent Volume, gal							
	500	1,000	2,000	5,000	15,000	30,000	75,000	100,000
1	1/350	1/190	2/190	5/125	10/84	20/100	50/68	60/84
			1/100	3/84	7.5/68	15/68	40/84	50/68
				3/68	5/45	10/45	40/56	40/56
				2/45	3/37	7.5/37	20/37	30/37
2	1/230	1/100	2/125	7.5/125	20/100	40/84	100/100	125/68
			1.5/84	5/100	15/68	30/68	75/68	100/56
				5/84	10/45	25/56	60/56	75/45
				3/56	7.5/37	20/37	50/45	75/37
3	1/190	2/190	2/84	3/37	25/100	60/125	100/68	75/30
			1.5/56		20/68	50/100	100/56	60/20
					15/56	50/84	75/45	
					10/37	30/45	60/30	
4	1/155	2/155	5/155	7.5/84	30/100	60/84	150/84	200/68
		1.5/100		5/56	25/84	50/38	125/68	150/56
					15/45	40/56	75/37	125/45
						30/37		100/30
5	1/125	1.5/84	3/84	15/155	40/100	75/100	75/30	300/100
		2/125		10/100		60/68	60/20	250/84
				7.5/38		50/56		150/45
				5/45		30/30		125/37
6	1/100	2/100	5/125	10/84	40/84	75/68	250/84	300/68
		1.5/68	3/68		30/68	60/56	200/68	250/56
			3/56		25/56	50/45	150/45	200/45
			2/45		20/37	40/37	125/37	150/37
7	2/190	2/84	7.5/155	15/84	60/125	100/68	350/84	200/30
		1.5/56	7.5/125	10/56	50/100		200/45	150/30
			5/84	7.5/45	40/56		150/37	150/25
				7.5/37	30/45		100/20	
8	1.5/84	3/84	7.5/84	25/125	75/100	125/68	300/68	400/56
	2/125		5/56	20/100	60/84	100/56	250/56	350/45
				15/68	50/68	75/45	150/30	300/45
				10/45	30/37	75/37	125/25	250/37
9	2/84	7.5/155	15/155	40/155	75/68	75/30	400/56	
		5/125	10/100	30/100	60/56		300/45	
		5/100	7.5/68	25/84	50/45		250/37	
		3/68		20/68	40/37		200/30	
10	5/125	7.5/125	20/100	50/100	150/84	250/84	600/84	
		5/84	15/84	40/84	125/68	200/68	500/68	
			10/84	30/68	100/56	150/45	350/45	
				25/56	75/45	125/37

FIGURE 9. Logic Diagram for Pilot-Scale Evaluations

TABLE 10. Basic Mixing Classes

Dominant Impeller Effect	Mixing Class
Flow Controlling	Fluid Blending Solid Suspension Heat Transfer
Combination	Gas Dispersion Mass Transfer
Shear Rate Controlling	Liquid Dispersion Solid Dispersion Chemical Reaction

velocity may be required to achieve a given process result. In this case the impeller tip speed, V_T , must be held constant.

Dynamic similarity requires that fixed ratios of corresponding forces (inertial, viscous, gravitational) be held constant. The mixing parameters that relate to these forces are listed in Table 11. All of these parameters are a function of both N and D . It is not possible to hold both N and D

TABLE 11. Parameters Associated with Dynamic Similarity

Parameter	Formula
Power/Volume	$\rho N^3 D^5 / V$
Torque/Volume	$N^2 D^5 / V$
Reynolds Number	$\rho N D^2 / \mu$
Froude Number	$N^2 D / g$
Pumping Number	$N D^3$
Weber Number	$\rho N^2 D^3 / \sigma$
Impeller Tip Speed	$\pi D N$

during scale-up. In addition, the size distribution of particles in solid-liquid agitation systems will not change as the size of the agitation system is scaled-up or scaled-down.

An empirical approach must be taken to scale-up bench and pilot-scale mixing results. Bench and pilot-scale mixing tanks are fabricated to assure dimensional similarity. The range of impeller rotation speeds evaluated is chosen to induce tip speeds that bracket the speeds often used in full-scale equipment (Table 8). Finally, a scale-up rule is chosen that accounts for the dynamic dissimilarity that will exist between the pilot- and full-scale equipment.

Geometric similarity fixes the level of D for each of the dynamic similarity parameters listed in Table 11. As a result, the scale-up problem becomes one of determining the appropriate large-scale impeller speed. The selection of the appropriate scale-up rule to establish this rotational speed is based on experience. An experience-based selection procedure has been documented (Rautzen et al. 1976). This procedure identifies a scale ratio, R as:

$$R = \left(\frac{V_2}{V_1} \right)^{1/3} \quad (5)$$

where,

the subscript 2 refers to large-scale equipment and the subscript 1 refers to bench- and pilot-scale equipment.

The scale ratio R is used to fix the physical dimensions of the full-scale equipment as follows:

$$R = \frac{L_2}{L_1} = \frac{T_2}{T_1} = \frac{D_2}{D_1} = \frac{W_2}{W_1} = \frac{BW_2}{BW_1} \text{ etc.} \quad (6)$$

The large-scale impeller speed is then related to bench- and pilot-scale impeller speed. This relationship adjusts the scale ratio, R , by raising it to an exponent, m :

$$\frac{N_2}{N_1} = \left(\frac{1}{R}\right)^m \quad (7)$$

The selection of the exponent, m , is based on a knowledge of the controlling mixing class in the agitation system (Table 10) and on experience. The levels of m associated with common mixing applications are indicated in Table 12.

TABLE 12. Scale-Up Exponents for Agitation

Mixing Application	Dominant Impeller Effect	Description	Exponent, m
Equal Blending Time	--	Reaction Time is controlling	0
Equal Surface Motion	Shear Rate (Fr, Tip Speed)	Vortex formation	1/2
Equal Mass Transfer	Shear Rate/Flow (Power/Volume, Tip Speed)	Liquid Motion/ Turbulence/ Dispersion of Solids	2/3
Equal Solids Suspension	Flow (N_Q)	Suspension of Solids	3/4
Equal Liquid Motion	Flow (Torque/Volume)	Corresponding Velocities are equal	1

Some mixing problems entail a combination of the applications listed in Table 12. In these instances, a variable speed control system is often specified. The range of speeds available is selected to bracket the speeds identified for each potentially controlling impeller effect (Table 12).

Some of the most important results of the study are as follows: (1) The results of the study indicate that the use of the proposed method is effective in reducing the error rate of the classification. (2) The results of the study indicate that the use of the proposed method is effective in reducing the error rate of the classification. (3) The results of the study indicate that the use of the proposed method is effective in reducing the error rate of the classification.

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4.0 PREPARATION OF SIMULATED WV AND HW SLURRIES

The slurry metered to the ceramic melter for processing consists of two fractions, HLW and glass formers. Each of these fractions is a slurry. The waste fraction consists of the unwanted HLW byproducts produced during nuclear fuel reprocessing. Glass formers are mixed with the raw waste to create a melter feed stream that will produce a glass product with acceptable properties. The glass formers can be added as individual glass forming chemicals or as a glass frit. Glass forming chemicals are usually in the form of metal hydroxides, nitrates, or oxides. Glass frit is ground glass that was produced by premelting glass forming chemicals, cooling the resulting glass product, and milling the product to the required particle size.

The simulated WV and HW melter feed slurries have been tailored to produce an acceptable glass product based on site-specific HLW chemical and physical properties. The composition of these feed slurries is described in Section 4.1. The procedures used to prepare the slurries during this evaluation are discussed in Section 4.2. The means used to acquire the specific aliquots of slurry utilized are discussed in Section 4.3.

4.1 COMPOSITION OF WV AND HW MELTER FEED SLURRIES

The WV and HW melter feed slurries have been tailored to create a glass with properties that exceed all applicable regulatory requirements. The design of the slurry begins once the HLW composition has been specified. An empirical model based on many years of HLW glass development is used to define a first approximation of a glass composition for initial testing. Next a simulant of this HLW is prepared. Glass formers are added to this simulated waste, the combined slurry is melted and the properties of the resulting glass are analyzed. This procedure is repeated until an acceptable working glass is defined. This working glass which becomes the reference for that waste composition is then subjected to extensive testing on both the laboratory and large-scale melter equipment with changes being made to optimize the

composition and define its tolerance to variations in waste composition. This procedure has been used to define the current reference HW and WV melter feed slurries.

The simulated WV slurry composition shown in Table 13 is such a working glass composition. This slurry contains glass forming chemicals and an ion exchange material. The ion exchange material, unique to the WV slurry, is used to remove radioactive cesium from the supernate solution found in HLW storage tanks. Nitric acid is added to the waste, glass formers and ion exchange material to account for recycle from the off-gas treatment system and to help improve the rheological characteristics of the slurry. Sugar, a reductant, is added to control the reduction/oxidation state of the glass in the melter.

The simulated HW melter feed slurry composition used for this study is shown in Table 14. The waste composition shown is the composition prior to the addition of formic acid to the feed. The final simulated slurry consists of the formatted waste fraction and glass formers added as 1/3 glass frit and 2/3 glass-forming chemicals. The waste is treated with formic acid to improve the rheological properties and melting behavior of the slurry. During this process, the mono- and di-valent elements, Na, Ni, Mn, Mg, and Ca, are converted to formates. Sugar provided additional control of the reduction/oxidation state of the glass in the melter.

4.2 SLURRY PREPARATION PROCEDURES

A properly agitated mixing tank is required to prepare simulated WV and HW melter feed slurries. The first step in the preparation of WV feed is to add chemicals procured as solutions to the tank. Once the agitator is started, dry chemicals (except for the SiO_2) and the ion exchange resin are added to the tank and mixed with the liquid solutions. Approximately 2/3 of the water required is then added. An in-line homogenizer is located in a 2-in. recirculation line. Slurry is drawn from the tank bottom through the homogenizer and then pumped back into the tank. The slurry in the tank is homogenized for 8 to 12 hours to reduce the particle size of the dry chemicals to less than about 200 mesh and to speed the dissolution of these chemicals. The SiO_2 required in the reference feed compositions is very abrasive. This material is purchased

TABLE 13. Simulated WV Slurry Composition

Chemical Source	(g/L)
Waste	
$\text{Fe}(\text{NO}_3)_3 \cdot 9 \text{H}_2\text{O}$	11.80
$\text{Fe}(\text{OH})_3$	36.75
$\text{FePO}_4 \cdot 2 \text{H}_2\text{O}$	22.98
NaNO_3	6.22
NaOH	0.16
$\text{Cr}(\text{NO}_3)_3 \cdot 9 \text{H}_2\text{O}$	0.90
$\text{Cr}(\text{OH})_3$	0.79
CsNO_3	0.46
$\text{ZrO}(\text{NO}_3)_2 \cdot 2 \text{H}_2\text{O}$	23.42
$\text{Mn}(\text{NO}_3)_2 \cdot 4 \text{H}_2\text{O}$	17.23
$\text{Ni}(\text{NO}_3)_2 \cdot 6 \text{H}_2\text{O}$	3.11
$\text{Ni}(\text{OH})_2$	1.99
$\text{La}(\text{OH})_3$	0.12
$\text{Sm}(\text{OH})_3$	0.12
$\text{Pr}(\text{OH})_3$	0.12
$\text{Nd}(\text{OH})_3$	0.12
$\text{Y}(\text{OH})_3$	0.04
$\text{Al}(\text{NO}_3)_3 \cdot 9 \text{H}_2\text{O}$	15.65
$\text{Al}(\text{OH})_3$	5.14
KNO_3	0.54
$\text{Ba}(\text{OH})_2$	0.21
$\text{Sr}(\text{OH})_2$	0.11
$\text{Ce}(\text{OH})_3$	0.63
$\text{Ca}(\text{OH})_2$	2.23
Na_2SO_4	2.49
NaCl	0.11
Ion Exchange Material	
IE-96®	39.18
Glass Formers	
$\text{Mg}(\text{NO}_3)_2 \cdot 6 \text{H}_2\text{O}$	25.69
NaCOOH	36.98
$\text{Na}_2\text{B}_4\text{O}_7 \cdot 10 \text{H}_2\text{O}$	100.13
Li_2CO_3	26.39
TiO_2	3.38
SiO_2	134.76
K_2CO_3	17.63
Miscellaneous	
HNO_3	8.50
Sugar	24.60

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TABLE 14. Simulated HW Slurry Composition

Chemical Source	(g/L)
Waste	
Fe(OH) ₃	59.50
Al(OH) ₃	26.30
NaNO ₃	13.40
Cr(OH) ₃	7.21
NaOH	3.58
Na ₂ C ₂ O ₄	3.37
Zr(OH) ₄	3.11
SiO ₂	3.00
Ni(OH) ₂	3.00
Na ₂ SO ₄	2.32
LaF ₃	2.16
Na ₂ MoO ₄ • 2 H ₂ O	2.00
Nd(OH) ₃	1.84
CsOH	1.05
Ce(OH) ₃	0.59
Cu(OH) ₁	0.74
Mn(OH) ₂	0.74
BaSO ₄	0.63
NdF ₃	0.58
Sr(OH) ₂	0.47
Pr(OH) ₃	0.47
La(OH) ₃	0.47
NaF	0.47
Mg(OH) ₂	0.42
CaF ₂	0.42
Y(OH) ₃	0.26
Sm(OH) ₃	0.21
Formic Acid	9.20
Glass Formers	
Glass Frit -80 +200 mesh	100.0
SiO ₂	134.4
NaCOOH	20.0
Na ₂ B ₄ O ₇ • 10 H ₂ O	69.9
LiCOOH • H ₂ O	46.9
Mg(NO ₃) ₂ • 6 H ₂ O	12.8
Ca(NO ₃) ₂ • 4 H ₂ O	31.6
Sugar	4.5

as particles with diameters less than 0.0017 in. (325 mesh) and added to the homogenized mixture after the homogenizer has been stopped. Finally, the remaining volume of water is added to the tank to create the reference melter feed slurry.

The preparation of simulated HW feed proceeds in a similar fashion. Once the simulated waste is prepared, it is heated to 90°C, and formic acid (0.2 M) is added to convert the mono- and di-valent elements (Na, Ni, Mn, Mg, and Ca) in the waste to formates. Once formed, the simulated waste is transferred to a larger mixing tank where glass formers, frit and sugar are added and thoroughly mixed with the formed waste. Finally the remaining volume of water is added and the slurry is transferred to a melter feed tank.

4.3 SLURRY ACQUISITION PROCEDURES

The WV and HW slurries used during the bench-scale mixing tests were prepared for use during a nonradioactive pilot-scale ceramic melter (PSCM) experiment and a High Bay Ceramic Melter (HBCM) experiment. Both slurries were prepared in large batches; the WV slurry batch volume was 12,000 L and the HW slurry batch volume was 1000 L. Approximately 50 L of each slurry was taken from the melter feed holding tank for use during the mixing evaluation. The WV slurry was drained to a 50 L carboy through the tank drain valve. The agitation in holding tank for the HW slurry was not sufficient to prevent the formation of a gel at the outer edges of the tank. Prior to transferring this slurry, the entire tank was air sparged to completely mix the slurry so that a representative sample could be obtained. A 50 L sample of this well-mixed HW slurry was gravity-drained from this tank. The carboys containing the 50 L samples were agitated thoroughly before the samples were transferred to the bench-scale mixing tanks.

5.0 LABORATORY RHEOLOGICAL EVALUATIONS

Laboratory rheological evaluations were conducted with samples of simulated Hanford and West Valley melter feed slurries. The slurry samples analyzed were aliquots of the batches prepared according to the recipe described in Section 4.

The laboratory evaluations were conducted to: 1) measure the physical properties of the slurries, 2) classify the rheological behavior of each slurry, 3) determine the effects of shear rate and temperature on the apparent viscosity of each slurry, and 4) provide background information needed to design and operate the bench-scale mixing equipment.

5.1 EXPERIMENTAL EQUIPMENT AND PROCEDURES

Laboratory rheological data were obtained with a Haake Rotovisco® viscometer equipped with an M150 measuring/drive head and the MV1P sensor system. The measurement of the rheological properties with this instrument (see Figure 10) requires that the sample be placed into the gap between two coaxial cylinders. A motor drives the inner cylinder. A viscosity-related torque, caused by the resistance of the sample to shearing, acts on the inner cylinder. This torque deflects a measuring spring placed between the motor and the inner cylinder. The magnitude of the spring deflection correlates linearly with the torque. The spring deflection is transformed into an electrical signal. The torque and tachometer signals are recorded.

The speed signal required for the determination of the shear rate does not come from the rotor directly but from the tacho-generator (TG) which sits above the motor and is mechanically linked to it. For slurries that do not exhibit yield stresses, the speeds of the tacho-generator, motor, and the rotor, are identical so that the speed signal n_{tg} also defines the rotor speed n_r . A slurry with a yield stress will "clamp" the rotor to the cup, until the torque τ , applied is greater than the yield stress τ_y .

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FIGURE 10. Schematic of a Haake Viscometer

While the rotor is still clamped and remains at standstill, the motor rotation will cause some spring deflection and consequently a torque signal τ , on the ordinate. For a slurry with a yield stress, a straight line, leaning off the vertical with an angle β , is recorded initially (see Figure 11). When the shear stress surpasses the yield value τ_y the rotor breaks loose from its "clamp", the rotor picks up speed, accelerating almost instantaneously to the value n_{tg} . The change in the behavior of the sample becomes apparent in the rapid turn of the flow curve. The ordinate value of this turning point defines the yield value τ_y . On the flow curve two regions can be distinguished: the first region is a segment of the "flow curve of a solid," while the second region describes the flow of the material after it had been transformed into a liquid.

Depending on the test conditions selected one may also get a more or less gradual change of the flow curve from one region to the other. The transition

FIGURE 11. Flow Curve (Rheogram) for a Yield-Pseudoplastic Fluid

from a solid gel to an easy-flowing liquid may go through a phase of creep, which makes the turning point less clear. Under these conditions the yield point is determined by the crossing point of the tangents on both branches.

Standard laboratory techniques were used to measure the pH, density, dissolved solids, suspended solids and total solids content of each sample.

5.2 RHEOLOGICAL PROPERTIES

Laboratory evaluations were conducted to classify the rheological behavior of HW and WV melter feed slurries and to determine the effects of shear rate and temperature on the apparent viscosity of each slurry. These evaluations provided the background information needed to design and operate the bench-scale mixing system. The bench-scale data were scaled-up to predict the performance of the full-scale mixing tank.

5.2.1 Classification of Slurries

An assumption implicit in this evaluation was that the melter feed slurries would be mixed at velocities sufficient to prevent particles from settling in the mixing vessel. This assumption implies that heterogeneous flow behavior (solids concentration at any point deviates from mean value by greater than ± 5 percent) cannot be tolerated. The objective was to create a pseudohomogeneous or nearly pseudohomogeneous flow regime in the mixing vessel.

A wide variety of pseudohomogeneous flow regimes are possible (see Section 3.2). The majority of slurries can be treated as if they were pseudoplastic or yield-pseudoplastic (Y-P) materials. The behavior of these materials can be defined by the following equations (Herschel and Bulkley 1926):

$$\tau = K \left(\frac{dv}{dy} \right)^n \quad \text{when } \tau_y = 0 \quad (8)$$

$$\tau = \tau_y + K \left(\frac{dv}{dy} \right)^n \quad \text{when } \tau_y > 0 \quad (9)$$

where:

τ = shear stress, N/m^2 (Pascal)

τ_y = yield stress, N/m^2

K = consistency index, Ns/m^2

$\frac{dv}{dy}$ = shear rate in y direction, sec^{-1}

n = flow behavior index.

The yield-pseudoplastic model defined by Equation (9) reduces to the Bingham Plastic model when $n=1$, the pseudoplastic model, Equation (8), when $\tau_y = 0$, and to the Newtonian model when $\tau_y = 0$ and $n=1$.

A typical rheogram for the simulated HW melter feed slurry is depicted on Figure 12. This rheogram is similar to the theoretical rheogram for a yield-pseudoplastic material shown in Figure 11. The data of Figure 12 fit the

FIGURE 12. Typical Rheogram for HW Melter Feed Slurry

yield-pseudoplastic model of Equation (6) with a correlation coefficient of 0.999. The Y-P equation for this sample of HW melter feed slurry that results, is

$$\tau = 4.27 + 0.103 \left(\frac{dv}{dy} \right)^{0.59} \quad (10)$$

The three samples of HW melter feed slurry that were evaluated exhibited similar Y-P behavior.

A typical rheogram for WV melter feed slurry is shown on Figure 13. This rheogram displays a much lower yield stress than was exhibited by simulated HW feed. The slope of the rheogram decreased as the shear rate increased. This behavior is typical of a pseudoplastic material. The data of Figure 13 were fit to the yield-pseudoplastic model described by Equation 9. The data fit this model with a correlation coefficient of 0.999. The yield-pseudoplastic equation for this sample of WV melter feed is:

$$\tau = 0.164 + 0.0187 \left(\frac{dv}{dy} \right)^{0.786} \quad (11)$$

FIGURE 13. Typical Rheogram for WV Melter Feed Slurry

The three samples of WV melter feed slurry that were evaluated exhibited similar behavior. The slurry is a yield-pseudoplastic fluid with a very low yield stress. It is likely that both the pseudoplastic and yield-pseudoplastic rheological models will accurately represent this simulation of WV melter feed slurry.

5.2.2 Evaluation of Rheograms

The yield-pseudoplastic rheological model (Equation 9) was used to evaluate the rheograms developed for HW and WV melter feed. This evaluation was designed to: 1) develop the relationship between the apparent viscosity of each sample and shear rate, and 2) determine the effects of temperature on viscosity. Glycerol, corn syrup, and water were used to develop the relationship between N_p and N_{Re} for each bench-scale mixing configuration evaluated. The viscosity of these Newtonian fluids will also change with temperature.

5.2.2.1 Newtonian Fluids

The Haake Rotovisco viscometer was used to assess the effects of temperature on the viscosity of the corn syrup and glycerol samples used during the mixing evaluation. Rheograms for corn syrup prepared at 20, 25, 35, and 40°C are plotted on Figure 14. The slope of each rheogram is constant ($n=1$), and each rheogram begins at the origin ($\tau_y = 0$), confirming the fact that corn syrup is a Newtonian fluid. The viscosity of corn syrup at each temperature was calculated by: 1) establishing 10 equidistant shear rate values, 2) determining the shear stress at these shear rates, and 3) dividing the computed value of τ by the shear rate (see Equation 2). The mean values and standard deviation of the viscosity of corn syrup that were determined for each temperature are listed in Table 15.

FIGURE 14. Rheogram for Corn Syrup as a Function of Temperature

TABLE 15. Viscosity of Corn Syrup as a Function of Temperature

Temperature (°C)	Viscosity	
	Mean Value (cP)	Standard Deviation (cP)
20	4898	44
25	2976	24
35	1372	19
40	1032	31

Rheograms for glycerol prepared at 20, 25, 35, and 40°C are plotted on Figure 15. These rheograms confirm the fact that glycerol is a Newtonian fluid. The calculated mean value and standard deviation of the viscosity of glycerol as a function of temperature are listed in Table 16.

FIGURE 15. Rheograms for Glycerol as a Function of Temperature

FIGURE 16. Rheograms for HW Melter Feed Slurry as a Function of Temperature

TABLE 17. Yield-Pseudoplastic Coefficients for HW Melter Feed Slurry as a Function of Temperature

Coefficient	Temperature, °C				Mean Value
	20	25	30	35	
τ_y , Yield Stress, Pa	4.3	4.0	4.4	4.4	4.3
K, Consistency Index, Pa·s	0.077	0.018	0.076	0.101	0.103
n, Flow Behavior Index	0.646	0.515	0.641	0.581	0.593

The mean value coefficients in Table 17 define the relationship between shear stress and shear rate for simulated HW melter feed slurries. The viscosity of HW was determined by: 1) evaluating the value of τ at shear rates of interest (Equation 9) and 2) dividing this τ by the shear rate, to yield the apparent viscosity of HW melter feed slurry at this shear rate. A plot of apparent viscosity versus shear rate for typical mixing system shear rates is contained on Figure 17.

5.2.2.3 WV Melter Feed Slurry

The Haake Rotovisco viscometer was used to assess the effects of temperature on the viscosity of the WV melter feed slurry used during the mixing

TABLE 16. Viscosity of Glycerol as a Function of Temperature

Temperature (°C)	Viscosity	
	Mean Value (cP)	Standard Deviation (cP)
20	994	5
25	648	3
35	316	6
40	212	1

One aliquot of glycerol was diluted with water to create a mixture that contained 74 volume percent glycerol. This mixture had a pH of 6.9 and a density of 1.2 g/cm^3 at a temperature of 22°C . A rheogram for the mixture confirmed that the mixture was a Newtonian fluid with a viscosity of 34 cP (standard deviation = 0.28 cP).

Rheograms for tap water were not prepared. Values of viscosity as a function of temperature were obtained from standard reference texts.

5.2.2.2 Simulated HW Melter Feed Slurry

The Haake Rotovisco viscometer was used to assess the effects of temperature on the viscosity of the simulated HW melter feed slurry used during the mixing evaluation. The rotational speed of the viscometer was fixed to achieve a shear rate of 234 sec^{-1} . The temperature of a sample of simulated HW melter feed was gradually increased from 20 to 35°C over an 18-min time interval. The rheogram produced is included as Figure 16. The measured value of shear stress declined from 8.2 to 7.8 Pa during the test interval. This slight decline was not considered significant.

Rheograms for simulated HW melter feed slurry were prepared at temperatures of 20 , 25 , 30 and 35°C . The yield-pseudoplastic coefficients associated with each rheogram are listed in Table 17. The mean values in Table 17 were calculated by: 1) averaging the values of τ obtained at 20 , 25 , 30 and 35°C , at each of 10 shear rates, and 2) 'fitting' the yield-pseudoplastic model to the average shear stress values. A rheogram based on these mean values was included as Figure 12.

FIGURE 17. Apparent Viscosity of HW Melter Feed Slurry as a Function of Shear Rate

evaluation. Rheograms for WV melter feed slurry prepared at 20, 25, 30, and 35°C are plotted on Figure 18. These rheograms are typical of those expected for a yield-pseudoplastic fluid with a low yield stress. The rheograms at 25, 30 and 35°C do not exhibit significant (>±5%) differences in shear stress at a given shear rate. The yield-pseudoplastic coefficients associated with each rheogram are listed in Table 18.

The coefficients in Table 18 were used to calculate the apparent viscosity of WV melter feed slurry as a function of temperature (Equation 9). Because the rheograms at 25, 30, and 35°C do not exhibit significant differences in shear stress at a given shear rate, the apparent viscosities calculated for

FIGURE 18. Rheograms for WV Melter Feed Slurry as a Function of Temperature

TABLE 18. Yield-Pseudoplastic Coefficients for WV Melter Feed Slurry as a Function of Temperature

Coefficient	Temperature, °C			
	20°C	25°C	30°C	35°C
τ_y , Yield Stress, Pa	0.16	0.22	0.10	0.24
K, Consistency Index, Pa·S	0.0187	0.0051	0.020	0.022
n, Flow Behavior Index	0.786	0.979	0.754	0.722

these temperatures were averaged. Plots of apparent viscosity for simulated WV melter feed slurry at 20°C and at 25-35°C are given on Figure 19.

5.3 PHYSICAL PROPERTIES

The parameters used to evaluate solid-liquid mixing systems (see Tables 4, 11) require information about the density of the fluids being mixed. The settling velocity of the solids in a suspension will affect the design of the mixing system selected to assure a homogeneous suspension of these solids. The dependent variable in this mixing evaluation was the degree of solids

FIGURE 19. Apparent Viscosity of WV Melter Feed Slurry as a Function of Temperature

distribution homogeneity achieved in the mixing system that resulted from the combination of independent variables that were evaluated. The density, settling rate, and total solids content of the fluids used during this mixing evaluation are discussed in this section.

5.3.1 Newtonian Fluids

The corn syrup, glycerol, and tap water used during the evaluation did not contain appreciable quantities of suspended solids. Standard laboratory techniques revealed the density information summarized in Table 19.

TABLE 19. Density of Corn Syrup, Glycerol and Tap Water as a Function of Temperature

Fluid	20°C	Density, g/cm ³		
		25°C	30°C	35°C
Corn Syrup	1.374	1.371	1.370	1.366
Glycerol	1.256	1.254	1.251	1.248
Tap Water	0.998	0.997	0.996	0.994

5.3.2 Simulated Hanford Melter Feed Slurry

Standard laboratory procedures were used to measure the macro-scale physical properties of this slurry. The density, pH, and solids weight distribution results obtained for a representative sample of simulated HW melter feed slurry are summarized in Table 20.

Samples of simulated HW melter feed slurry were shaken, placed in a 100 ml graduated cylinder, and allowed to settle. The level of the solids-liquid interface was recorded as it descended in the cylinder. Plots of interface height versus time were prepared at temperatures of 25 and 40°C (Figure 20). The settling rate for the linear portion of the curve was 0.45 and 0.19 cm/hr, respectively. The yield-stress displayed by simulated HW melter feed is the likely cause of the very low settling rates that were observed.

Bench-scale mixing tests were conducted during a six-week period. The sample of simulated HW slurry was transferred between containers frequently. As a result of these transfers and of the bench-scale mixing tank sample acquisition procedures utilized, the total solids content of the HW sample varied during the test period. The total solids content varied between 24 and 31 wt%.

An attempt was made to evaluate the size distribution of the solids in the simulated HW melter feed sample. A PA-720 Automatic Particle Size Analyzer from Pacific Scientific, HIAC Instrument Division, was used for this analysis.

TABLE 20. Physical Properties of Simulated HW Melter Feed Slurry

Property	Temperature °C			
	20	25	35	40
Density, g/cm ³	1.286	1.281	1.282	1.279
pH	8.34	8.34	8.35	8.38
Dissolved Solids, wt%	20.9	n.e. (a)	n.e.	n.e.
Suspended Solids, wt%	20.8	n.e.	n.e.	n.e.
Total Solids, wt%	41.7	n.e.	n.e.	n.e.

(a) n.e. = not evaluated.

FIGURE 20. Solids Settling Rate of Simulated HW Melter Feed Slurry

The simulated HW sample was diluted substantially before it was inserted in the device. The results obtained are listed in Table 21.

The 9 μm mean particle size indicated by Table 21 seems unreasonably small, since the smallest particles added to the simulated HW mixture were about 325 mesh or 44 μm in diameter. Additional work will be required to quantify the size distribution of this slurry.

TABLE 21. Approximate ($\pm 100\%$) Particle Size Distribution of Simulated HW Melter Feed Slurry

Particle Size (μm)	Cumulative Amount Greater than (vol %)
22	10
15	25
9	50
6	75
3	99

5.3.3 Simulated WV Melter Feed Slurry

The density, pH, and solids weight distribution results obtained for a representative sample of simulated WV melter feed slurry are summarized in Table 22.

Samples of simulated WV melter feed slurry yielded the settling rate information (at 25 and 40°C) plotted on Figure 21. The settling rates for the linear portion of the curves was 0.6 and 0.8 cm/hr at 25 and 40°C, respectively.

The total solids content of this slurry varied between 35 and 39 wt%.

The size distribution data obtained with the Pacific Scientific Automatic Particle Size Analyzer for the simulated WV melter feed slurry is summarized in Table 23.

The dilution procedures described in Section 5.3.2 may have contributed to the unreasonably small particle size values listed in Table 23. Additional work will be required to quantify the size distribution of this slurry.

TABLE 22. Physical Properties of Simulated WV Melter Feed Slurry

Property	Temperature °C			
	20	25	35	40
Density, g/cm ³	1.263	1.261	1.259	1.257
pH	7.44	7.44	7.44	7.47
Dissolved Solids, wt%	14.8	n.e. ^(a)	n.e.	n.e.
Suspended Solids, wt%	21.7	n.e.	n.e.	n.e.
Total Solids, wt%	36.5	n.e.	n.e.	n.e.

(a) n.e. = not evaluated.

FIGURE 21. Solids Settling Rate of Simulated WV Melter Feed Slurry

TABLE 23. Approximate ($\pm 100\%$) Particle Size Distribution of Simulated WV Melter Feed Slurry

Particle Size (μm)	Cumulative Amount (wt%)
20	10
15	25
10	50
6	75
3	99

FIGURE 51. Sedimentation rate of isolated W-100 test slurry.

TABLE 52. Approximate (40°C) Particle Size Distribution of Slurry at Initial Test Slurry

Particle Size (microns)	Cumulative Amount (%)
10	10
15	25
20	40
25	55
30	70
35	85
40	95

6.0 BENCH-SCALE MIXING EVALUATIONS

The preparation of melter feed slurries involves a simultaneous mixing and heat transfer process. The mixing process must provide enough agitation to maintain a homogeneous distribution of solids in the tank. An external jacket or internal heat transfer coils will provide the heat transfer surface necessary to concentrate the slurry and remove decay heat generated by the radioactive waste slurries.

This bench-scale mixing evaluation focused on: 1) determining the agitation level necessary to maintain a homogeneous slurry concentration, and 2) determining the effect of a number of independent variables on the homogeneity in the tank. This evaluation did not examine heat transfer mechanisms. However, heating/cooling coils were placed in the bench-scale tank to determine their effect on mixing. These coils were sized to simulate the coil arrangement now installed in the melter feed tank undergoing evaluation at the Savannah River Laboratory (SRL).

Bench-scale mixing tests were conducted to: 1) examine the effects of significant independent variables (see Section 5.2) on the distribution of solids in the mixing tank, and 2) provide a rational basis for the design of the full-scale mixing system that will be built during FY-86. The results obtained with the bench-scale mixing system are described in this section.

6.1 EXPERIMENTAL EQUIPMENT AND PROCEDURES

The bench-scale mixing system was installed in the high-bay annex of Building 324 on the Hanford site. The system was composed of an 11-gal clear plastic cylindrical tank, motor, gear reducers, impeller shaft, impellers, coils and baffles (see Figure 22). Power was supplied by a 0.25 hp, 0-1100 rpm, AC motor. The agitator assembly (shaft plus impellers) was supported on "frictionless" bearings to facilitate the measurement of the torque applied to the slurry in the tank. A load cell connected to a 10-in. lever arm measured this applied torque.

The load cell consisted primarily of a Wheatstone Bridge circuit. An excitation voltage was applied to the circuit. The voltage induced in the

FIGURE 22. Bench-Scale Mixing System Arrangement

circuit was measured. The induced voltage is directly related to the applied torque. The load cell output was 0 to 10 V, which corresponded to 0 to 100 in.-lb_f of torque applied. The load cell had an accuracy of $\pm 0.02\%$ of full-scale. A signal conditioner provided the excitation voltage to the load cell and amplified the signal from the load cell. This output signal was

recorded on a chart recorder. The accuracy of the chart recorder was $\pm 1.5\%$ of full scale or ± 0.08 in.-lb_f. The accuracy of the entire torque measurement system was ± 0.1 in.-lb_f.

The power applied to the impeller shaft was calculated as:

$$P = (\text{Shaft Torque})(N) \quad (12)$$

The power calculated by Equation (12) includes the power applied to the slurry and the power dissipated by friction in the system. Each impeller configuration was operated in air at various rotational speeds. The resulting torque was determined. This torque was subtracted from the torque measured when the impeller configuration was placed in the Newtonian or non-Newtonian fluids to obtain the effective torque applied to the fluid. The torque applied to the fluid alone was the torque used to calculate the power actually applied to the fluid by each impeller configuration tested.

The shaft rotational speed, N , was measured with a digital tachometer when speeds were in a range of 30 to 1000 rpm. Speeds less than 30 rpm were verified with a stop watch--visual count of the rotations procedure. These procedures yielded rotational speeds that correlated well with the factory calibration of the AC motor variable speed drive assembly.

All impellers, coils and baffles used during the experiment were fabricated of 304 L or 316 L stainless steel. The impellers utilized are depicted on Figure 23. The coils placed in the mixing tank during some tests are shown on Figure 24. The rationale behind the selection of the impellers and coils shown on Figures 23 and 24 is discussed in Section 6.2.

The homogeneity of solids distribution in the mixing tank was evaluated by extracting samples of the tank contents at six locations in the tank. The degree of homogeneity was determined by measuring the total solids content of each sample and comparing this concentration to the solids content of a homogeneously dispersed sample of the slurry.

Prior to each test, the tank contents were gently stirred by a flat paddle for a specified time to resuspend any solids that may have settled before the

FIGURE 23. Bench-Scale Impeller Arrangements

test began. The impeller configuration installed in the tank was then operated for the period of time necessary to 'turnover' the contents of the tank approximately 40 times, to assure that all samples were collected on a consistent basis.

The time required to accomplish one 'turnover' of tank contents is given by:

$$\text{time} = \frac{V}{Q} = \frac{V}{N_Q N D^3} \quad (13)$$

The pumping number N_Q was assumed to be equal to the sum of the N_Q for a single 4-blade, 45-degree pitched turbine and the N_Q of a 4-blade, flat turbine with

FIGURE 24. Bench-Scale Coil Arrangements

dimensions similar to those of the bench-scale units. The N_Q for a 4-blade, 45° pitched turbine as a function of Re and D/T is shown in Figure 25. This N_Q was added to published values of N_Q for flat blade turbines (Oldshue 1983, Nagata 1975) to yield an overall N_Q value of 1.58 for the combined axial

FIGURE 25. Impeller Pumping Number N_Q for Axial Flow Turbines

flow and radial flow dual impeller system. The time required to achieve 40 tank turnovers for each impeller system evaluated is summarized in Table 24.

A stainless steel pipet (see Figure 26) was used to withdraw samples from the tank at the times indicated in Table 24. The suction required to draw the slurry into the pipet was provided by a pipet pump. The sample volume was approximately 10 ml. The samples collected were placed in aluminum dishes, weighed and placed in a drying oven at 103°C for at least 12 h. After the samples were dried, they were placed in a desiccator to cool for 20 min. The dried solids were weighed to determine the total solids concentration.

6.2 SELECTION OF INDEPENDENT AND DEPENDENT VARIABLES

Bench-scale mixing tests were conducted: 1) to examine the effects of significant independent variables on the distribution of solids in the mixing tank, and 2) to provide a rational basis for the design of the large-scale

TABLE 24. Mixing Time Required to Obtain 40 Tank Turnovers

Impeller Diameter (in.)	Rotational Speed (rpm)	Mixing Time (min:sec)
3.75	50	25:27
	140	9:26
	230	5:25
	320	3:48
4.50	50	14:44
	140	5:28
	230	3:08
	320	2:12
7.50	10	15:55
	23	7:14
	36	4:33
	50	3:19

FIGURE 26. Stainless Steel Pipet Used to Sample Slurry

mixing tank. The design of the bench-scale test program followed the logic outlined in Figure 9. A key step was to identify an approximate full-scale mixing system design before the bench-scale system was designed.

An extensive literature review was conducted to develop a basis for the selection of the significant variables that were most likely to effect the mixing of melter feed slurries. The literature review focused on current industrial practices for the mixing of slurries (see Section 3.3) and the mixing tank designs for the Defense Waste Processing Facility's (DWPF) Slurry Mix Evaporator (SME) and the West Valley (WV) concentrator/melter feed tank. The DWPF tank shown on Figure 27 and WV tank shown on Figure 28 were specifically designed to mix and concentrate a slurry of HL waste and glass formers. The mixing system for DWPF was designed for a Bingham Plastic slurry with a

FIGURE 27. Schematic of DWPF Slurry Mix Evaporator
(DWPF/SR Drawing #W752193, Rev D. 6/24/83;
#W752280, Rev. 8, 3/11/83; #W752197, Rev A, 6/24/83)

FIGURE 28. Schematic of WV Concentrator/Melter Feed Tank

yield stress of 15 N/m^2 and an insoluble solids content of 48 wt%. The mixing system for CTS was designed for a slurry very similar to the slurry used in these bench-scale tests.

As a result of the literature review, six independent variables were selected for testing. The remaining geometric ratios were held constant at values commonly used in industrial slurry mixing applications (see Table 6).

The six independent variables selected for testing include the type of impeller, impeller diameter, impeller rotational speed, use/absence of wall baffles, use/absence of internal heat transfer coils, and the type of melter feed slurry evaluated.

A key geometric consideration that was not evaluated was the design of the mixing tank shell. Industrial practice is to use an L/T of approximately 1 (see Table 6) for solids suspension applications. The L/T of the DWPF tank is 1.08, while the L/T of the WV tank is 0.80. An L/T of 1.0 was specified for the bench-scale tank. The maximum bench-scale tank diameter that could be evaluated with the available mixing test stand was 15 inches. A vertical, cylindrical, flat-bottomed clear plastic tank with this diameter and a height of 18 inches (3 inches freeboard) was used during the bench scale tests. Approximately 11 gallons of fluid were added to the tank to achieve a liquid height of 15 inches.

A dual impeller design was selected to permit the bench-scale system to be consistent with the DWPF and WV designs. Impeller clearances were selected to be geometrically consistent with these designs ($C_2/T = 0.3$, $C_1/T = 0.06$).

The distribution of solids in the mixing tank is affected to a large extent by the selection and location of the impellers installed in the tank. The location of the impellers was fixed by the existing DWPF and WV designs and current industrial practice (see Table 6). The lower impeller chosen was a 4-blade flat turbine that can act as a sweeper to keep solids from depositing on the tank bottom. Two upper impeller designs, a 45° pitch 4-blade turbine impeller and a 3-blade high efficiency turbine impeller, were evaluated. Both of these impellers provide an axial flow pattern. This flow pattern is preferred for solids suspension applications (see Table 6). Normally, the location of the upper impeller is determined by the different liquid operating levels that will occur in the tank. These operating levels have not been defined for the HW or WV melter feed tanks. Therefore, the location of the upper impeller was fixed at $C_2/T = 0.3$, which is the location used in the DWPF, SME tank.

Since the impeller diameter (D) has a large effect on the mixing in the tank (see Tables 4 and 11) three impeller diameters were selected for

testing. A 3.75 in. diameter impeller with the D/T ratio of 0.25, used in the DWPF tank, was selected. A second impeller with a diameter of 7.5 in. was selected to yield a D/T ratio of 0.50. This ratio is commonly used in industry. A third impeller with a diameter of 4.5 in., yielded a D/T ratio of 0.30. This D/T ratio is used in the WV tank and will produce a tank turnover rate mid-way between the turnover rates achieved by the 3.75 in. and 7.5 in. diameter impellers.

The third independent variable evaluated was the use/absence of baffles. Wall baffles are commonly used in mixing tanks to prevent vortex formation and to direct the flow. Tests were conducted with and without wall baffles to determine their effect on the solids distribution in the tank. Four vertical and flat standard baffles were used. The baffle width was 0.08 times the tank diameter or 1.2 in. The gap between the baffles and the tank wall was 0.2 times the baffle width or 0.25 in. The baffles were equally spaced 90° apart.

The use/absence of heating/cooling coils may affect the distribution of solids in the mixing tank. The coil design evaluated was the bayonet assemble used in the DWPF SME tank. This 3-wrap coil design, extends from the lower to the upper impeller and is in close proximity to each impeller. The coils of the second and third wraps are centered in the open areas of the previous wrap (see Figure 24). The pipe diameter for the coils and the overall coil length were scaled-down according to the ratio of these dimensions to the SME tank diameter. The horizontal distance between the impeller and the first coil was scaled-down by the ratio of this distance to the SME impeller diameter. The gap distance between coil wraps was scaled to maintain the shear rate through this gap that results from the DWPF design. This scaling procedure was applied to size coils for the 3.75- and 4.5-in. impellers. The procedures could not be practically applied to the 7.5-in. impeller.

The rotational speed of the impeller is the most critical parameter in the scale-up of bench-scale data. Preliminary scoping tests were conducted to bracket speeds that produced well-mixed and not-mixed conditions in the bench-scale system. Four equally spaced rotational speeds were selected for each impeller. The speeds selected were based on a 'best judgment' decision prior

to the test sequence. The 3.75- and 4.5-in. impellers were operated at one set of 4 rotational speeds. The 7.5-in. impeller was operated at a different, lower set of four speeds to maintain reasonably constant power levels in the tank during tests with each impeller. A discussion of the effects of impeller diameter on power was provided in Section 3.3.1.

Two slurries with different physical and rheological properties were selected to determine the effect of slurry constituents on the solids distribution in the mixing tank. One of the slurries was a WV zeolite with unreacted glass-forming chemicals and the second slurry was a HW formatted waste with glass formers added as 1/3 glass frit and 2/3 unreacted glass-forming chemicals. The preparation of these slurries was discussed in Section 4.0.

The dependent variable in this evaluation was the distribution of solids in the mixing tank. Samples were withdrawn from six locations in the mixing tank during each run. The total solids content of these samples was evaluated. This total solids content was compared with the total solids content of a homogeneously mixed batch of slurry. The mixing tank was considered well-mixed when the total solids content of each sample varied by less than ± 5 percent from the solids content of the homogeneously mixed batch of slurry. The location of the sample points and the sampling procedures are discussed in Section 6.4.1.

6.3 EXPERIMENTS WITH NEWTONIAN FLUIDS

Newtonian fluids with known rheological properties were evaluated in the bench-scale mixing system to: 1) demonstrate the mechanical integrity of the system, 2) permit a determination of the accuracy and precision of data acquisition methods, and 3) identify the baseline relationship between power number and Reynolds number for each mixing configuration evaluated.

6.3.1 Selection of Newtonian Fluids

The power number versus Reynolds number relationship is an integral part of the evaluation and scale-up of mixing systems. This relationship is a function of the configuration of the impeller(s) and the mixing tank. The

power number versus Reynolds number relationship was defined for all of the bench-scale mixing configurations evaluated during the experiment.

The power number-Reynolds number relationship is similar to traditional friction factor-Reynolds number relationships encountered in pipe flow. The relationship evolves as the flow regime progresses from a laminar to a transition and finally to a well developed turbulent flow regime. The definition of the power number-Reynolds number relationship required that Reynolds numbers of from 1 to 10^5 be developed in the bench-scale mixing tank. Three Newtonian fluids: corn syrup, glycerol and water, were used to define this power number-Reynolds number relationship. The viscosity of these fluids was discussed in Section 5.2.2.1.

Corn syrup was used to obtain laminar Reynolds numbers ($Re < 15$); glycerol was used to obtain Reynolds numbers in the transition region ($15 < Re < 10^3$); a glycerol-water solution (74 vol% glycerol) was used to obtain Reynolds numbers in the near turbulent range ($10^3 < Re < 10^4$); and tap water was used to obtain Reynolds numbers in the fully developed turbulent flow region ($Re > 10^4$).

6.3.2 Torque-in-Air Measurements for Each Impeller Configuration

The definition of the power number-Reynolds number relationship required the determination of the power applied to the fluid. The overall power applied to the impeller shaft was determined by the method described in Section 6.1. The power measured was the sum of the power applied to the fluid and the power dissipated by friction. The power dissipated by friction was determined in the following way. Each impeller configuration was operated in air at various rotational speeds. The resulting torque was measured. These torque measurements are shown in Table 25. This torque value was subtracted from the torque measured when a given impeller configuration was tested in either a Newtonian or a non-Newtonian fluid, to obtain the effective torque applied to the fluid. The torque applied to the fluid alone, was the torque used to calculate the power applied to the fluid.

The torque-in-air measurements for the 3.75-in. and 4.5-in. diameter impellers were determined at rotational speeds ranging from 50 to 320 rpm. The

TABLE 25. Torque-in-Air Measurements for All Impeller Configurations

Impeller Diameter (in.)	Impeller Type	Torque (in.-lb _f)	
		Mean	Standard Deviation
3.75	High Efficiency Turbine	0.248	0.005
	Pitch Blade Turbine	0.271	0.005
4.5	High Efficiency Turbine	0.258	0.005
	Pitch Blade Turbine	0.243	0.005
7.5	High Efficiency Turbine	0.316	0.009
	Pitch Blade Turbine	0.294	0.005

torque-in-air measurements for the 7.5-in. diameter impeller were determined at rotational speeds ranging from 10 to 50 rpm. The torque measurements did not vary significantly over the range of rotational speeds evaluated.

6.3.3 Power Number Versus Reynolds Number Relationship for Newtonian Fluids

The power number versus Reynolds number relationship was determined for each test configuration evaluated during the bench-scale experiment. The eight configurations shown in Table 26 were evaluated for the 3.75 and 4.5 in. impeller diameter and each fluid type. The 7.5 in. diameter impeller was evaluated using configurations 2, 4, 6 and 8 with each fluid type. Coils could not be evaluated with this diameter.

The power number versus Reynolds number relationship generated for configurations 5 to 8 with a 4.5-in. diameter lower impeller and a pitched blade turbine upper impeller is shown on Figure 29. Note the effect of baffles, coils, and baffles and coils on the power consumed by the mixing system. The shape of this curve is typical of the shape of the curves that were obtained with other bench-scale test configurations (see Appendix A), and that were obtained by other investigators (see Section 3.3.1).

Four fluids: corn syrup, glycerol, glycerol/water, and water were used to generate Figure 29. The laminar region of the curve has a declining linear slope. In this region, portions of the fluid in the vessel may remain stagnant. The end of the laminar region is the point where the linear slope

TABLE 26. Eight Basic Bench-Scale Mixing Configurations

Configuration	Upper Impeller	Coils	Baffles
1	High Efficiency Turbine	Yes	No
2	High Efficiency Turbine	No	Yes
3	High Efficiency Turbine	Yes	Yes
4	High Efficiency Turbine	No	No
5	Pitch Blade Turbine	Yes	No
6	Pitch Blade Turbine	No	Yes
7	Pitch Blade Turbine	Yes	Yes
8	Pitch Blade Turbine	No	No

FIGURE 29. Power Number-Reynolds Number Relationship for Bench-Scale Mixing Configuration with Newtonian Fluids

changes drastically. This point occurs at a Reynolds number of approximately 15. The turbulent region occurs at Reynolds numbers greater than 10^4 . In this region the power number remains constant as the Reynolds number increases. Most mixing applications are designed to operate in the turbulent region. The transition region occurs at Reynolds numbers ranging from 15 to 10^4 . Flow near the impeller is turbulent, while motion away from the impeller is laminar, in the transition region.

The effect of baffles, coils and the combination of baffles and coils on the power consumed by the mixing system becomes apparent in the turbulent range. In the turbulent region, the viscous forces dissipate and the inertial forces dominate. The inertial forces are strongly affected by baffles and coils. The baseline case is the configuration where baffles and coils are absent. The data points associated with this configuration slope downward between Reynolds numbers of 10^2 and 3×10^4 . This reduction in power number was due to the vortexing that was developing as the Reynolds number increased. The vortexing or swirling of the slurry, requires less power than would be needed in a fully-baffled tank. Approximately 2.5 times more power was required when coils were placed in the tank. The coil and baffle combination required 4.4 times more power, while the baffles alone required 5 times more power than the base case. The baffles create more fluid motion in the tank than the coils and would be expected to draw more power. The power consumed by baffles and the combination of baffles and coils was similar. With the baffle and coil combination, a vortex was present inside the coil assembly. This reduced the power applied to slurry.

The data summarized on Figure 29 was compared with published information (Metzner et al. 1961) for similar mixing configurations. This published data is shown as a solid line on Figure 29. The published data was collected when a single PBT impeller was installed in the mixing tank. This data was adjusted to account for the additive effect in the laminar region of the lower turbine that was installed in the bench-scale tank. The very good correlation of the bench-scale data with the published information is an indication that the power and rotational speed data acquisition and analysis methods were accurate and precise.

The power number versus Reynolds number relationship is dependent only upon the tank configuration (i.e., baffles, coils, impeller configuration) and is independent of the type of fluid utilized (Metzner and Otto 1957). For a specific test configuration, the power number versus Reynolds number relationship for a Newtonian and non-Newtonian fluid will be identical. This fact permits the scale-up of bench-scale data to predict full-scale performance. The optimal impeller-tank configuration-rotational speed design point is

defined during bench-scale tests. The bench-scale rotational speed is scaled-up (Section 3.3.2) to a full-scale value. This full-scale value is used to compute a full-scale operational Reynolds number. The power number-Reynolds number relationship is used to identify the power that will be consumed by the full-scale impeller in the full-scale tank configuration. The power number-Reynolds number plots obtained with each bench-scale mixing configuration evaluated are included in Appendix A.

6.4 EXPERIMENTS WITH SIMULATED HW AND WV SLURRIES

One goal of this evaluation was to assess the effects of selected independent mixing system variables (Section 6.2) on the distribution of solids in simulated batches of HW and WV melter feeds. The design of the experiment utilized to gather the data necessary to assess these effects is discussed in Section 6.4.1.

The analysis of the results obtained during the experiment requires a determination of the shear rate in the mixing tank, and the apparent viscosity of the slurry being mixed. The methods used to determine these quantities are described in Section 6.4.2.

The statistical procedures used to quantify the effects of selected independent variables on the degree of mixing achieved, the result of this analysis, and the predictive equations developed to model the affects of the independent variables on mixing of HW and WV slurries are presented in Section 6.4.3.

6.4.1 Design of Bench-Scale Experiments

The large number of tests required to assess the effects of significant independent variables on the solids distribution in the mixing tank (see Section 6.2) required a detailed and organized approach to data collection and analysis. A statistical experimental design approach was selected. This approach is a proven technique that provided four major benefits to the test program. First, it provided more information per experiment than other approaches. Time and cost constraints limited the number of experiments which could be performed, making this benefit important. Second, it provided a very

organized approach toward the collection and analysis of the data. Third, an assessment of the data reliability in light of experimental and analytical variation was obtained. This assessment adds more credibility to the experimental results and conclusions. Fourth, the statistical design provided information about the interactions among experimental variables, leading to more-reliable predictions of performance for mixing configurations not directly covered by experimentation.

The experimental design contains six basic features: 1) statement of the objectives, 2) list of response variables, 3) list of factor variables, 4) choice of factor levels, 5) size of the design, and 6) order of the experimentation. The overall objectives of this design were: 1) to determine the effects of the factors and factor interactions on the response or dependent variable, and 2) to generate predictive equations based on the bench-scale data that could be used to predict the performance that is likely to occur in a full-scale mixing tank.

The distribution of solids in the mixing tank was chosen as the response variable and used to assess the degree of homogeneity in the mixing tank. In a homogeneously-mixed tank the solids are distributed uniformly throughout the tank. A judgment had to be made as to the number of samples and sample locations needed to evaluate the homogeneity of the solids distribution in the tank. Time and cost constraints limited the total number of samples to a total of six per test. Samples were taken at two locations 180° apart, at three different fluid levels. The samples were taken 0.5 in. from the tank wall at heights of 0.5 in., 6.5 in., and 13 in. from the tank bottom. These locations were known to be sensitive to changes in total solids content. The experiment was designed to determine the differences in the total solids content due to the sample location. Of primary interest was the determination of the deviation in the total solid content of the samples from the total solids content of a homogeneously mixed tank. Large deviations indicate a not-mixed condition. A standard procedure was prepared for obtaining each slurry sample and measuring the total solids content (see Section 6.1).

The selection of significant factors or independent variables and the levels of these variables was discussed in Section 6.2. The six factors

evaluated and their corresponding levels are summarized in Table 27. Three impeller diameters (3.75 in., 4.5 in., and 7.5 in.) were evaluated. The evaluation of the 7.5 in. diameter impeller revealed some problems in the experimental design. The use of heating/cooling coils was not possible with the 7.5 in. impeller. Coils scaled to this impeller diameter would not fit in the bench-scale mixing tank. The 7.5 in. diameter impeller had to be rotated at very low rotational speeds to obtain a power level in the bench-scale tank that was equivalent to the levels obtained with the other smaller impellers. To overcome this problem a two stage experiment was designed. The first experiment focused on testing the 7.5 in. diameter impeller at lower rotational speeds without heating/cooling coils. The second experiment evaluated the effects of the remaining two impeller diameters, and use of heating/cooling coils on the solids distribution in the mixing tank.

In order to evaluate the diameter effect, data from both experimental designs had to be combined. The data combined was data obtained using the same

TABLE 27. Factor Variables and Levels Used in the Experimental Design

<u>Factor</u>	<u>Number of Levels</u>	<u>Actual Level</u>
<u>Experiment 1</u>		
Impeller Design	2	High Efficiency Turbine
Impeller Diameter	1	7.5 in.
Impeller Rotational Speed	4	10, 23, 36, 50, rpm
Wall Baffles	2	With Baffles, Without Baffles
Heating/Cooling Coils	1	Without Coils
Slurry Type	2	WV, HW
<u>Experiment 2</u>		
Impeller Design	2	High Efficiency Turbine
Impeller Diameter	2	3.75 in., 4.5 in.
Impeller Rotational Speed	4	50, 140, 230, 320, rpm
Wall Baffles	2	With Baffles, Without Baffles
Heating/Cooling Coils	2	With Coils, Without Coils
Slurry Type	2	WV, HW

impeller design, rotational speed, the baffle level, slurry type, and heating/cooling coil arrangement. The rotational speed common to each experiment was 50 rpm.

The size of the experimental design was limited by time and cost constraints. The total number of tests for both experiments was set at 120. The experimental design selected permitted an evaluation of all the main factor effects, the two-way factor interactions and many three-way factor-interaction effects on the solids distribution in the bench-scale tank.

The experimental design for the first experiment called for a full-factorial experiment. All possible combinations of the four impeller rotational speeds, two slurry types, two impeller types, and two baffle configurations were tested, for a total of 32 tests. This experimental design permitted an examination of all main factor effects, all two-way factor-interaction effects and all three-way factor-interaction effects for the 7.5 in. diameter impeller configuration. The experimental design for the second experiment used the remaining 88 tests. A computer-aided design approach was required. This experimental design permitted an examination of all main factor effects, all two-way factor-interaction effects, and all three-way factor-interaction effects. The data from both experiments, for common coil configurations and speeds, was combined to determine the overall main effects due to impeller diameter and all two-way and three-way interactions involving the impeller diameter, slurry type, impeller type, and baffle configuration.

Both experimental designs were combined. The entire experiment was randomized to guard against systematic trends in uncontrolled or unknown variables. In addition, the order in which the six samples were taken with each test was randomized. The final experimental test array is shown in Table 28.

Each experimental design was based on a number of assumptions. These assumptions are listed below:

1. The replication variability is constant at all factor-level combinations (homogeneity of variance)
2. All four-way and higher-order interactions are not significant

TABLE 28. Experimental Test Array

Test #	RPM	Impeller Diameter	Coil	Slurry	Impeller Type ⁽¹⁾	Baffle	Sample Location and Order					
							#1	#2	#3	#4	#5	#6
1	230	4.5	No	WV	PBT	No	6.5 L	0.5 R	0.5 L	6.5 R	13 R	13 L
2	320	4.5	Yes	WV	A310	Yes	6.5 R	13 R	6.5 R	13 L	0.5 R	0.5 L
3	140	3.75	Yes	WV	PBT	Yes	0.5 R	13 L	6.5 L	13 R	6.5 R	0.5 L
4	50	7.5	--	HW	PBT	--	6.5 R	0.5 R	13 L	13 R	0.5 L	6.5 L
5	50	4.5	--	WV	PBT	Yes	13 L	6.5 L	13 R	6.5 R	0.5 R	0.5 L
6	50	4.5	Yes	HW	A310	--	0.5 R	13 R	6.5 L	0.5 L	13 L	6.5 R
7	23	7.5	--	HW	PBT	Yes	0.5 L	0.5 R	6.5 R	13 L	6.5 L	13 R
8	140	3.75	Yes	WV	A310	--	0.5 R	6.5 R	13 L	13 R	6.5 L	0.5 L
9	230	4.5	Yes	HW	A310	Yes	6.5 R	0.5 R	13 R	13 L	0.5 L	6.5 L
10	320	3.75	--	HW	A310	--	13 R	6.5 L	13 L	0.5 R	0.5 L	6.5 R
11	230	3.75	--	WV	PBT	--	6.5 R	13 L	6.5 L	0.5 R	0.5 L	13 R
12	50	4.5	Yes	WV	PBT	--	0.5 R	0.5 L	13 L	6.5 R	6.5 L	13 R
13	320	3.75	--	WV	PBT	--	0.5 L	6.5 L	0.5 R	13 R	13 L	6.5 R
14	230	4.5	--	HW	PBT	Yes	6.5 L	0.5 R	0.5 L	13 L	13 R	6.5 R
15	10	7.5	--	HW	A310	--	0.5 L	13 R	6.5 R	13 L	6.5 L	0.5 R
16	10	7.5	--	WV	A310	Yes	6.5 L	6.5 R	13 R	13 L	0.5 L	0.5 R
17	50	3.75	Yes	WV	A310	--	0.5 R	6.5 R	13 R	0.5 L	6.5 L	13 L
18	50	4.5	--	WV	A310	Yes	13 L	0.5 R	6.5 L	13 R	6.5 R	0.5 L
19	50	7.5	--	WV	A310	Yes	6.5 L	0.5 L	6.5 L	13 L	0.5 R	13 R
20	320	4.5	Yes	WV	A310	--	13 L	0.5 R	6.5 L	6.5 Y	0.5 L	13 R
21	140	3.75	--	WV	PBT	--	0.5 R	0.5 L	13 L	13 R	6.5 L	6.5 R
22	50	3.75	--	HW	PBT	--	13 L	0.5 L	13 R	6.5 R	0.5 R	6.5 L
23	10	7.5	--	HW	PBT	--	6.5 L	0.5 R	13 R	6.5 R	0.5 L	13 L
24	140	4.5	Yes	WV	A310	Yes	0.5 R	0.5 L	13 L	6.5 R	6.5 L	13 R
25	320	4.5	--	HW	A310	Yes	6.5	0.5 L	13 R	6.5 R	13 L	0.5 R
26	230	3.75	--	WV	A310	Yes	0.5 R	13 R	6.5 R	13 L	0.5 L	6.5 L
27	50	3.75	Yes	HW	PBT	Yes	0.5 R	13 L	6.5 L	0.5 L	13 R	6.5 R
28	50	4.5	Yes	HW	PBT	Yes	13 R	13 L	0.5 R	6.5 R	6.5 L	0.5 L
29	320	4.5	--	HW	PBT	--	0.5 L	6.5 L	6.5 R	0.5 R	13 L	13 R
30	50	4.5	--	HW	PBT	Yes	6.5 L	0.5 L	13 R	13 L	6.5 R	0.5 R
31	50	3.75	Yes	WV	PBT	Yes	0.5 R	0.5 L	13 R	13 L	6.5 L	6.5 R
32	230	3.75	--	HW	A310	--	0.5 R	6.5 L	13 R	13 L	6.5 R	0.5 L
33	50	4.5	Yes	WV	A310	Yes	6.5 L	13 R	0.5 L	6.5 R	13 R	0.5 L
34	50	3.75	--	WV	A310	--	6.5 L	0.5 R	13 R	6.5 R	13 L	0.5 L
35	230	3.75	--	WV	PBT	Yes	0.5 R	0.5 L	6.5 R	6.5 L	13 R	13 L
36	36	7.5	--	WV	PBT	Yes	6.5 R	13 R	13 L	6.5 L	0.5 R	0.5 L
37	140	4.5	Yes	HW	A310	--	13 R	6.5 L	6.5 R	0.5 L	0.5 R	13 L
38	230	4.5	Yes	WV	PBT	Yes	0.5 L	13 R	0.5 R	13 L	6.5 L	6.5 R
39	50	4.5	--	HW	PBT	--	0.5 R	13 L	6.5 R	0.5 L	13 R	6.5 L
40	230	4.5	--	HW	A310	Yes	0.5 R	6.5 R	6.5	13 L	13 R	0.5 L

TABLE 28. (contd)

Test #	RPM	Impeller Diameter	Coil	Slurry	Impeller Type	Baffle	Sample Location and Order					
							#1	#2	#3	#4	#5	#6
41	23	7.5	--	WV	PBT	Yes	0.5 L	6.5 R	13 L	0.5 R	6.5 L	13 R
42	320	4.5	Yes	HW	A310	--	13 R	13 L	6.5 R	0.5 L	6.5 L	0.5 R
43	10	7.5	--	HW	PBT	Yes	6.5 R	13 L	6.5 L	0.5 L	0.5 R	13 R
44	230	3.75	--	HW	PBT	--	6.5 L	13 L	0.5 L	6.5 R	0.5 R	13 R
45	140	3.75	--	HW	PBT	--	6.5 L	13 R	0.5 R	0.5 L	6.5 R	13 L
46	140	4.5	--	WV	A310	--	13 L	6.5 L	6.5 R	0.5 R	13 R	0.5 L
47	230	4.5	--	WV	A310	Yes	0.5 L	13 L	0.5 R	6.5 R	6.5 L	13 R
48	230	3.75	--	HW	PBT	Yes	0.5 R	0.5 L	13 L	6.5 R	13 R	6.5 L
49	10	7.5	--	WV	PBT	Yes	13 L	6.5 L	13 R	6.5 R	0.5 L	0.5 R
50	23	7.5	--	HW	A310	--	13 R	13 L	6.5 R	0.8 R	0.5 L	6.5 L
51	36	7.5	--	HW	PBT	Yes	6.5 L	6.5 R	0.5 R	13 L	13 R	0.5 L
52	50	3.75	--	HW	A310	Yes	6.5 R	6.5 L	0.5 R	0.5 L	13 L	13 R
53	50	3.75	--	WV	PBT	Yes	0.5 L	13 L	6.5 L	0.5 R	13 R	6.5 R
54	320	3.75	Yes	HW	A310	Yes	0.5 L	0.5 R	6.5 R	6.5 L	13 L	13 R
55	140	3.75	Yes	HW	A310	Yes	6.5 L	6.5 R	0.5 R	0.5 L	13 L	13 R
56	230	3.75	Yes	WV	A310	--	6.5 L	13 L	0.5 L	13 R	0.5 R	6.5 R
57	50	3.75	Yes	HW	PBT	--	6.5 R	0.5 R	0.5 L	13 L	6.5 L	13 R
58	50	4.5	--	WV	A310	--	13 R	6.5 R	0.5 R	6.5 L	0.5 L	13 L
59	36	7.5	--	WV	A310	Yes	13 L	13 R	0.5 L	6.5 L	6.5 R	0.5 R
60	23	7.5	--	WV	A310	Yes	6.5 R	0.5 R	13 L	6.5 L	13 R	0.5 L
61	36	7.5	--	HW	A310	--	6.5 L	13 R	13 L	6.5 R	0.5 L	0.5 R
62	140	4.5	--	HW	PBT	--	13 R	13 L	6.5 L	0.5 L	0.5 R	6.5 R
63	10	7.5	--	WV	A310	--	0.5 R	0.5 L	13 R	6.5 L	13 L	6.5 R
64	320	4.5	--	WV	PBT	--	6.5 R	6.5 L	13 R	13 L	0.5 R	0.5 L
65	140	4.5	Yes	HW	PBT	Yes	0.5 R	13 R	13 L	6.5 R	6.5 L	0.5 L
66	140	3.75	--	HW	A310	--	6.5 L	13 L	6.5 R	13 R	0.5 R	0.5 L
67	140	3.75	Yes	HW	PBT	--	13 R	0.5 R	6.5 R	0.5 L	6.5 L	13 L
68	50	7.5	--	HW	A310	Yes	13 L	0.5 L	6.5 R	0.5 R	13 R	6.5 L
69	230	3.75	Yes	HW	PBT	Yes	13 L	6.5 L	0.5 L	6.5 R	0.5 R	13 R
70	50	4.5	--	HW	A310	--	13 R	0.5 L	0.5 R	6.5 L	13 L	6.5 R
71	23	7.5	--	WV	A310	--	13 L	13 R	0.5 L	6.5 R	0.5 R	6.5 L
72	10	7.5	--	HW	A310	Yes	0.5 L	13 L	6.5 L	0.5 R	13 R	6.5 R
73	320	4.5	--	WV	PBT	Yes	6.5 R	0.5 R	6.5 L	0.5 L	13 R	13 L
74	320	3.75	Yes	WV	PBT	--	0.5 R	13 L	0.5 L	13 R	6.5 R	6.5 R
75	50	7.5	--	HW	A310	--	13 L	6.5 L	0.5 L	0.5 R	6.5 R	13 R
76	50	3.75	Yes	HW	A310	--	13 R	13 L	6.5 L	0.5 L	0.5 R	6.5 R
77	320	4.5	Yes	WV	PBT	--	0.5 L	13 L	13 R	6.5 R	0.5 R	6.5 L
78	50	3.75	--	WV	PBT	--	13 L	6.5 R	13 R	0.5 R	6.5 L	0.5 L
79	320	4.5	Yes	HW	PBT	Yes	0.5 R	6.5 R	13 L	13 R	0.5 L	6.5 L
80	320	3.75	--	HW	A310	Yes	13 R	13 R	0.5 L	0.5 R	6.5 R	6.5 L

TABLE 28. (contd)

Test #	RPM	Impeller Diameter	Coil	Slurry	Impeller Type	Baffle	Sample Location and Order											
							#1		#2		#3		#4		#5		#6	
81	230	4.5	Yes	HW	PBT	--	13	R	13	L	0.5	L	0.5	R	6.5	R	6.5	L
82	320	3.75	--	WV	A310	Yes	13	R	6.5	R	0.5	L	6.5	L	0.5	R	13	L
83	320	3.75	--	HW	PBT	Yes	13	R	6.5	R	6.5	L	0.5	R	0.5	L	13	L
84	140	3.75	--	HW	PBT	Yes	13	R	13	L	6.5	L	0.5	R	0.5	L	6.5	R
85	36	7.5	--	WV	PBT	--	6.5	R	0.5	L	13	L	6.5	L	13	R	0.5	R
86	230	3.75	Yes	WV	PBT	--	6.5	L	6.5	R	0.5	L	0.5	R	13	L	13	R
87	36	7.5	--	HW	A310	Yes	13	R	0.5	L	6.5	R	13	L	6.5	L	0.5	R
88	50	7.5	--	HW	PBT	Yes	13	R	6.5	R	13	L	6.5	L	0.5	L	6.5	R
89	23	7.5	--	HW	A310	Yes	6.5	R	0.5	L	0.5	R	6.5	L	13	L	13	R
90	230	4.5	--	WV	A310	--	13	L	0.5	L	13	R	6.5	L	0.5	R	6.5	R
91	230	3.75	Yes	HW	A310	Yes	6.5	R	0.5	R	6.5	L	13	R	13	L	0.5	L
92	230	3.75	Yes	WV	A310	Yes	6.5	L	0.5	R	0.5	L	13	R	13	L	6.5	R
93	36	7.5	--	HW	PBT	--	13	R	6.5	L	0.5	L	6.5	R	13	L	0.5	R
94	320	4.5	--	HW	A310	--	0.5	L	6.5	L	13	R	0.5	R	6.5	R	13	L
95	230	3.75	Yes	HW	A310	--	6.5	R	13	R	13	L	6.5	L	0.5	R	0.5	L
96	320	4.5	--	WV	A310	Yes	6.5	R	0.5	L	6.5	L	13	L	0.5	R	13	R
97	23	7.5	--	HW	PBT	--	13	R	6.5	L	0.5	L	6.5	R	13	L	0.5	R
98	320	3.75	Yes	HW	A310	--	6.5	R	0.5	R	0.5	L	6.5	L	13	L	13	R
99	230	4.5	Yes	WV	A310	--	0.5	L	0.5	R	6.5	L	6.5	R	13	R	13	L
100	320	4.5	Yes	WV	PBT	Yes	0.5	R	13	R	13	L	6.5	R	6.5	L	0.5	L
101	140	4.5	--	HW	A310	Yes	13	L	0.5	L	6.5	L	6.5	R	13	R	0.5	R
102	320	3.75	Yes	WV	PBT	Yes	13	L	6.5	R	13	R	6.5	L	0.5	R	0.5	R
103	140	3.75	--	WV	A310	Yes	0.5	R	13	R	6.5	L	0.5	L	6.5	R	13	L
104	140	4.5	Yes	WV	PBT	--	6.5	L	6.5	R	13	R	0.5	L	0.5	R	13	L
105	23	7.5	--	WV	PBT	--	13	L	0.5	L	13	L	6.5	R	6.5	L	0.5	R
106	10	7.5	--	WV	PBT	--	6.5	R	13	R	0.5	L	0.5	R	6.5	L	13	L
107	50	7.5	--	WV	PBT	--	13	R	6.5	R	13	L	0.5	L	0.5	R	6.5	L
108	50	7.5	--	WV	PBT	Yes	0.5	L	6.5	R	0.5	R	13	L	6.5	L	13	R
109	320	4.5	Yes	HW	PBT	--	0.5	R	6.5	L	0.5	L	13	R	6.5	R	13	L
110	50	4.5	Yes	HW	A310	Yes	13	R	0.5	L	13	L	0.5	R	6.5	L	6.5	R
111	230	3.75	--	HW	A310	Yes	6.5	R	13	L	13	R	0.5	R	6.5	L	0.5	L
112	320	3.75	--	WV	A310	--	13	L	6.5	R	13	R	0.5	R	6.5	L	0.5	L
113	320	3.75	--	WV	PBT	Yes	6.5	R	13	R	0.5	L	6.5	L	0.5	R	13	L
114	230	4.5	--	HW	A310	--	0.5	L	6.5	R	13	R	6.5	L	13	L	0.5	R
115	50	7.5	--	WV	A310	--	6.5	L	13	R	0.5	R	6.5	R	0.5	L	13	L
116	320	3.75	Yes	HW	PBT	--	6.5	R	0.5	L	0.5	L	6.5	L	13	R	13	L
117	320	3.75	Yes	WV	A310	Yes	0.5	L	6.5	L	6.5	R	13	R	13	L	0.5	R
118	50	3.75	Yes	WV	A310	Yes	13	R	6.5	R	6.5	L	0.5	R	0.5	L	13	L
119	140	4.5	--	WV	PBT	Yes	13	L	0.5	R	6.5	R	0.5	L	13	R	6.5	L
120	36	7.5	--	WV	A310	--	6.5	R	13	L	6.5	L	0.5	L	13	R	0.5	R
							6.5	R	13	L	0.5	R	0.5	L	13	R	6.5	L

(1) PBT--Pitched Blade Turbine Impeller.
A310-- High Efficiency Turbine

3. All three-way interactions associated with the quadratic portion of the impeller diameter factor are not significant
4. The three and four factor levels (diameter, rpm) are equally spaced across some scale
5. The error structure is normally distributed.

6.4.2 Determination of the Shear Rate and Apparent Fluid Viscosity in the Mixing Tank

The simulated HW and WV melter feed slurries exhibited yield-pseudoplastic rheological behavior (see Section 5.2.1). This non-Newtonian behavior will cause the apparent viscosity of the slurries to decline as the shear rate in the mixing tank increases. This relationship between viscosity and shear rate must be specified before the bench-scale and full-scale mixing system design procedures described in Section 3.3 can be implemented. Fortunately, theoretical models for the mixing of non-Newtonian slurries are available. A model that describes the mixing behavior of pseudoplastic slurries is described in Section 6.4.2.1. The application of this model to yield-pseudoplastic slurries is discussed in Section 6.4.2.2.

6.4.2.1 Pseudoplastic Fluids

Equations that describe the behavior of time-independent non-Newtonian fluids were listed in Table 3. The behavior of pseudoplastic fluids was defined by:

$$\tau = K \left(\frac{dv}{dy} \right)^n \quad (8)$$

Laboratory evaluation procedures (see Section 5.1) are used to determine the values of the consistency index, K, and flow behavior index, n, for pseudoplastic fluids. The apparent viscosity of pseudoplastic fluids decreases as the rate of shear, dv/dy , increases. The shear rate in the mixing system must be specified before the apparent viscosity in the mixing tank can be determined. The apparent viscosity in the mixing tank must be known before the Reynolds number can be calculated.

The velocity (and shear rate) of fluids in a mixing tank decrease as they

leave the vicinity of the impeller. As the velocity decreases, the apparent viscosity of a pseudoplastic fluid will increase. The average shear rate in a mixing system is assumed to be related to the speed of the impeller utilized (Metzner and Otto 1957), or:

$$\left(\frac{dv}{dy}\right)_{\text{average}} = kN \quad (14)$$

Values of the mixing shear rate proportionality constant, k , have been established for common impeller configurations. Some of these values are summarized in Table 29. The addition of a second impeller to a mixing tank does not significantly effect the shear rate profile in the tank (Metzner et al. 1961). The average value of k , for single and double impeller configurations, listed in Table 29 is approximately 11.5. This value of k was used to determine the shear rate present in the bench-scale mixing tank during each test.

The simulated WV melter feed slurry exhibited yield-pseudoplastic properties (see Section 5.2.1). However, the value of τ_y was very small. It is likely that the pseudoplastic model can be accurately applied to this slurry. The apparent viscosity of WV melter feed slurry as a function of shear rate and temperature was plotted on Figure 19. The shear present during each bench-scale run was assumed to be equal to 11.5 times the impeller rotational speed during the run. The viscosity at this shear rate was then determined by inspection of Figure 19.

TABLE 29. Values of Mixing Shear Rate Constant, k

Impeller Configuration		k	Source
Type	D/T		
Flat Blade	0.33	13 $\pm$ 20%	Metzner and Otto 1957
Turbine		11.6 $\pm$ 8%	Metzner et al. 1961
	0.33	10 $\pm$ 11%	Calderbank and Moo-Young 1961
6-Blade, 45° Pitch Axial Flow Turbine	0.3 - 0.7	13 $\pm$ 15%	Metzner et al. 1961

6.4.2.2 Yield-Pseudoplastic Fluids

The simulated HW melter feed slurry exhibited yield-pseudoplastic behavior (see Section 5.2.1). Yield-pseudoplastic materials develop a yield stress that must be overcome before fluid motion can begin. Once this yield stress is overcome, the fluid behaves as a pseudoplastic material. Non-Newtonian mixing systems designed to suspend solids operate in the turbulent flow regime (see Section 3.3). The yield stress must be overcome before turbulent motion can begin. It was assumed (for turbulent flow) that the shear rate of a yield-pseudoplastic material could be estimated by the same procedure used to estimate the shear rate of pseudoplastic materials. This shear rate was estimated to be equal to 11.5 times the impeller rotational speed during each run with simulated HW melter feed slurry.

The apparent viscosity of simulated HW melter feed slurry as a function of shear rate and temperature was plotted on Figure 17. The apparent viscosity of the slurry was determined by inspection of this figure at the shear rate that was present during each bench-scale run.

6.4.3 Experimental Results

The data collected during the experimental phase of this project was analyzed by standard statistical techniques. These techniques will be described briefly in Section 6.4.3.1. These statistical techniques identified the significance of factor and factor-interaction effects on the solids distribution attained in the bench-scale mixing tank. Significant factor and factor-interaction effects were included in polynomial equations that were used to model the behavior of the bench-scale tank. These equations are described in Section 6.4.3.2 and 6.4.3.3.

6.4.3.1 Data Analysis Procedures

An analysis of variance (ANOVA) technique was used to identify the significance of factor and factor-interaction effects on the solids distribution in the bench-scale mixing tank. The ANOVA computational technique has been fully described (Hicks 1982).

The experimental design (see Section 6.4.1) specified a three-phase approach to the analysis of the data collected. The first phase of the

analysis focused on data collected with the 7.5 inch diameter impeller (D/T = 0.5). The factors evaluated during this experiment are listed in Table 30. The number of combinations of factors and levels possible during this experiment was 32. A full factorial experiment, evaluating each of these combinations was conducted. This methodical approach permitted the calculation of the analysis of variance (ANOVA) for the experimental data. The ANOVA calculation provided insight into the significance of main factor effects, factor-interaction effects and random error effects on measured values of the distribution of solids in the mixing tank. The standard deviation of the six values of the solids concentration in the tank measured during each run was used as the dependent variable during this evaluation.

The functional equation used during the ANOVA computation was:

$$\begin{aligned} \sigma_s = & \mu + P_1 B_i + P_2 S_j + P_3 NS_{ij} + P_4 I_k + P_5 NI_{ik} \\ & + P_6 SI_{jk} + P_7 NSI_{ijk} + P_8 B_l + P_9 NB_{il} \\ & + P_{10} SB_{jl} + P_{11} NSB_{ijl} + P_{12} IB_{kl} + P_{13} NIR_{ikl} \\ & + P_{14} SIB_{jkl} + P_{15} BSU_{B_{ijkl}} + E_{ijkl} \end{aligned} \quad (15)$$

where σ_s = standard deviation of solids distribution in the bench-scale mixing tank, (wt%)

N = rotational speed, rpm

S = slurry type

I = impeller type

B = use of baffles

i,j,k,l = levels of factors

μ = overall mean effect

E_{ijkl} = random error effect

P_1 - P_{15} = polynomial coefficients

Equation (15) states that the measured solids distribution in the bench-scale mixing tank is a function of some overall mean effect, plus the singular effects of each factor, plus the combined effects of the various factor interactions, plus some random experimental error effect. If none of the

TABLE 30. Factors Evaluated During Experiments With a 7.5 Inch Diameter Impeller

Factor	Level
Rotational Speed	10, 23, 36, 50 rpm
Type of Slurry	HW, WV
Type of Impeller	High Efficiency Turbine, Pitched Blade Turbine
Use of Baffles	Yes, No

factors or combination of factors had any effect on σ_s , σ_s would only be a function of the constant overall mean effect and random experimental error. The ANOVA technique was used to quantify the significance of each factor and factor-interaction effect.

Equation (15) is valid when the random error associated with the experimental data is normally distributed about zero with a variance equal to σ_s^2 . Equation (15) was used during this analysis because a statistical test for normality (Kolmogorov D test) indicated that the σ_s data collected was normally distributed.

The random error term E_{ijk1} , must be evaluated before "F statistics" can be computed to determine the significance of each of the factor and factor-interaction effects. Several statistical techniques were employed to determine which factor-interaction effects were nothing more than additional estimates of the error variance E_{ijk1} . For the 7.5 inch diameter impeller experiment, the variance estimates associated with the four-way interaction, and most three-way interaction effects were combined to provide an estimate of E_{ijk1} . This combined estimate of E_{ijk1} may result in more conservative tests of the significance of factor and factor-interaction effects.

The random error term E_{ijk1} is the random variability associated with estimates of σ_s . Another source of variability, the experimental variability, is the random variability of total solids measurements. This experimental variability was estimated in a separate analysis of variance where the effect of sample acquisition height was included as a factor such that differences in total solids measurements due to the height at which the measurement was taken was accounted for and not included in the random variability estimate. There were two sources of experimental variability: variations due to sampling

methods and analysis procedures, and variations due to equipment set-up and instrument accuracy changes. The contributions of each source of variability in total solids measurements are listed in Table 31.

The statistical approach described above was applied to the data collected during each phase of the experiment. The approach was tailored to each data set, since the functional equation for each phase of the experiment was different. The results obtained by the ANOVA of each phase of the experiment are summarized in Section 6.4.3.2.

6.4.3.2 Solids Distribution ANOVA--Experiment With 7.5 Inch Diameter Impeller

The ANOVA for the 7.5 inch diameter impeller experiment is summarized in Table 32. The calculated F statistics listed are an indication of significance of each factor and factor-interaction effect. The higher the F statistic the more significant the factor or factor interaction becomes. The level of significance column provides an indication of the confidence level associated with each F statistic. For instance, an effect is significant at the 95% confidence level if the level of significance is less than 0.05.

Inspection of Table 32 reveals that only the overall mean effect of rotational speed, N, is significant at a level of 0.05. However, the effects of slurry type, S, presence or absence of baffles, B, and interactions between N and S, and N and B become significant if the confidence level criteria is reduced from 95 to 90 percent.

The effects of the interaction between N and S on the standard deviation of the solids distribution in the bench-scale mixing tank is described on Figure 30. There was no significant difference between the HW and WV slurries at rotational speeds of 10, 36 and 50 rpm. A significant difference was observed

TABLE 31. Sources of Experimental Error

Experiment	Total Variability of σ_s	Source of Error	
		Sampling Data Analysis	Set-Up Variance Instrument Accuracy
7.5 Inch Impeller	0.904	82%	18%
3.75, 4.5 Inch Impeller	0.933	100%	0%

TABLE 32. ANOVA for 7.5 Inch Diameter Impeller Experiment

Source of Variation	Degrees of Freedom	"F" Value	Level of Significance
N	3	5.65	0.028
S	1	4.73	0.066
I	1	2.32	0.172
B	1	4.92	0.062
NS	3	3.81	0.066
NI	3	1.61	0.271
NB	3	3.89	0.063
SI	1	1.72	0.231
SB	1	0.00	0.975
IB	1	0.03	0.859
NSI	3	2.51	0.142
NSB	3	2.29	0.166
Residual	<u>7</u>		
	31

FIGURE 30. Experiment with 7.5 Inch Diameter Impeller: Interaction Between Rotational Speed and Slurry Type

at 23 rpm. For the HW slurry alone, rotational speed did not significantly alter the value of σ_s observed. For WV slurry there was a significant difference between the results achieved at 23 and 36 rpm and the results achieved at 50 rpm. The increase of speed from 10 to 50 rpm did not substantially impact the solids distribution in the bench-scale tank.

The effects of the interaction between N and B on σ_s is described on Figure 31. The effects of baffles is negligible at 23, 36, and 50 rpm. At 10 rpm, baffles significantly enhanced the solids distribution in the bench-scale tank. The optimal mixing configuration for the 7.5 in. diameter impeller experiment occurred when the impeller was installed in a baffled tank and rotated at 10 rpm. A rotational speed of 10 rpm is appropriate because rotational speeds did not significantly impact the solids distribution in the bench-scale tank (Figure 30). In this case, the lowest acceptable speed is selected to reduce equipment cost. Baffles are significant at 10 rpm (Figure 31) so they should be part of the optimal mixing configuration.

FIGURE 31. Experiment With 7.5 Inch Diameter Impeller: Interaction Between Rotational Speed and Presence of Baffles

The specific linear, quadratic and cubic components of the significant factor and factor-interaction effects listed in Table 32 were identified. These components are listed in Table 33, which contains a larger number of significant factor and factor-interaction effects than Table 32. This situation arises from the fact that the most significant factor identified in Table 32, the rotational speed N , was broken down into its linear, quadratic, and cubic components in Table 33.

TABLE 33. Non-Linear ANOVA for the 7.5 Inch Diameter Experiment

<u>Source of Variation</u>	<u>Degrees of Freedom</u>	<u>Level of Significance</u>	<u>Polynomial Coefficient</u>
N	1	0.027	0.5473
N^2	1	0.068	-0.0163
N^3	1	0.162	0.0001
S	1	0.012	6.7105
I	1	0.133	3.5664
B	1	0.022	1.8407
NS	1	0.006	-1.0121
N^2S	1	0.010	0.3555
N^3S	1	0.017	-0.0004
NI	1	0.063	-0.6221
N^2I	1	0.067	0.0231
N^3I	1	0.083	-0.0002
NB	1	0.020	-0.1325
N^2B	1	0.017	0.0022
SI	1	0.086	-5.8577
SB	1	0.090	1.1572
NSI	1	0.066	0.8709
N^2SI	1	0.084	-0.0307
N^3SI	1	0.012	0.0003
NSB	1	0.064	-0.0386
Intercept	-	0.057	-3.3921

The information in Table 33 permits the construction of a polynomial equation that can be used to model the results obtained during the experiment. This polynomial equation can be described by:

$$\begin{aligned} \sigma_s = & -3.3921 + P_1N + P_2N^2 + P_3N^3 + P_4S + P_5I + P_6B + P_7NS + P_8N^2S + \\ & P_9N^3S + P_{10}NI + P_{11}N^2I + P_{12}N^3I + P_{13}NB + P_{14}N^2B + P_{15}SI + \\ & P_{16}SB + P_{17}NSI + P_{18}N^2SI + P_{19}N^3SI + P_{20}NSB \end{aligned} \quad (16)$$

where P_1 - P_{20} = polynomial coefficients in Table 33

S = slurry type (0 = WV, 1 = HW)

B = baffles (1 = no, 0 = yes)

I = impeller (1 = PBT, 0 = High Efficiency Impeller)

N = rotational speed

The solution of Equation (16) is plotted on Figure 32. Inspection of this figure indicates that the optimal rotational speed (minimal standard deviation) for the no baffles case occurs at about 20 rpm, while speeds as low as 10 rpm will be adequate in a baffled tank. This inference is consistent with the information included on Figures 30 and 31.

6.4.3.3 Solids Distribution ANOVA--Experiments With 3.75 and 4.5 Inch Diameter Impellers

The factors and levels evaluated during experiments with the 3.75 and 4.5 in. diameter impellers are listed in Table 34. The ANOVA technique described in Sections 6.4.3.1 and 6.4.3.2 was used to examine the significance of each factor, two-way factor interactions, and three-way factor interactions. The significance of higher order interaction effects were considered likely to be so small as to provide only an estimate of the error variance, E_{ijklmn} .

The ANOVA for these experiments is summarized in Table 35. Inspection of Table 35 reveals that the overall mean effects of rotational speed and type of slurry are significant at a level of 0.05. There are two significant three-factor interactions that involve factors slurry type, S, presence of coils, C, diameter, D, and Impeller Type, I. To understand these interactions, the

FIGURE 32. Experiment With 7.5 Inch Diameter Impeller: Solution to Predictive Equation

effects due to each of the four factors must be examined at the various levels of the other two factors involved in the three-way interaction. The most significant effects are listed in Table 36.

TABLE 34. Factors Evaluated During Experiments With 3.75 and 4.5 Inch Diameter Impellers

Factor	Level
Rotational Speed, N	50, 140, 230, 320 rpm
Type of Slurry, S	HW, WV
Type of Impeller, I	High Efficiency Impeller, Pitched Blade Turbine
Impeller Diameter, D	3.75, 4.5 inch
Use of Baffles, B	Yes, No
Use of Coils, C	Yes, No

TABLE 35. ANOVA for Experiments With 3.75 and 4.5 Inch Diameter Impellers

Source of Variation	Degrees of Freedom	"F" Value	Level of Significance
N	3	8.30	0.0001
S	1	9.17	0.003
I	1	0.00	0.980
B	1	0.31	0.581
C	1	0.00	0.954
D	1	2.92	0.092
NS	3	3.24	0.028
NI	3	3.15	0.031
NB	3	2.12	0.106
ND	3	3.10	0.033
SC	1	0.27	0.608
SD	1	0.06	0.799
IC	1	1.69	0.198
ID	1	1.68	0.199
CD	1	0.10	0.754
SCD	1	7.09	0.010
ICS	<u>1</u>	6.03	0.016
	27		

TABLE 36. Experiments with 3.75 and 4.5 Inch Impellers: Factor and Factor-Interaction Effects

<u>Factors</u>	<u>Trend</u>	<u>Effect</u>
N	N ↑	σ_s ↓
N,S	HW or WV at N > 50	σ_s ↔
	HW or WV at N = 50	HW σ_s < WV σ_s
N,I	PBT or A310	σ_s ↔
N,D	D ↑ at N > 140	σ_s ↔
	D ↑ at N < 140	σ_s ↓
S, C, D	Various	See Figures 33-36
I, C, D	Various	See Figures 33-36

The primary overall trends indicated in Table 36 are that: 1) σ_s decreases as N increases, 2) HW slurry is easier to mix than WV slurry at N = 50 rpm, 3) the high efficiency turbine and pitched blade turbine achieved very similar mixing results, 4) impeller diameter was only significant at lower rpms, 5) baffles improved mixing at 50 rpm but had little effect at higher rotational speeds, and 6) coils reduced the degree of mixing achieved at low rpm values.

The effects listed in Table 36 were masked by the low values of σ_s that were measured. The maximum value of σ_s that was encountered during the test program was a σ_s of approximately 4. This value of σ_s corresponded to a solids distribution in the mixing tank that was homogeneous to the extent that the six samples obtained were within ±10 to 15 percent of the mean value. The majority of σ_s values recorded had values less than 1.0. A σ_s value of 1 corresponds to a solids mixing distribution within approximately ±5 percent of the mean value.

The specific linear, quadratic, and cubic components of the significant factor and factor-interaction effects are listed in Table 37. This information was used to construct a polynomial equation that included the significant effects of each component. Equation (17) has the same general form as

Table 37. Non-Linear ANOVA for the 3.75 and 4.5 Inch Diameter Experiments

<u>Source of Variation</u>	<u>Level of Significance</u>	<u>Polynomial Coefficient</u>
N	0.019	-0.0281
N ²	0.026	0.0002
S	0.0045	-1.4421
I	0.0024	0.7470
B	0.0303	-1.0480
C	0.2270	0.1791
D	0.1790	-0.6649
NS	0.0126	0.0295
N ² S	0.0213	-0.0002
NI	0.0107	-0.0073
NB	0.0182	0.0278
N ² B	0.0163	-0.0002
ND	0.0499	0.0229
N ² D	0.1024	-0.0001
SC	0.1442	-0.2534
SD	0.1027	-0.2839
IC	0.4281	-0.1379
ID	0.0130	-0.4424
CD	0.0091	-0.5627
SCD	0.0121	0.6237
ICD	0.0189	0.5873
Intercept	0.0014	1.6483

Equation (16) but it contains 21 polynomial coefficients. The solution of this equation for each mixing configuration (see Table 26) evaluated, is plotted on Figures 33 through 36. The trends indicated on Figures 33 through 36 parallel those listed in Table 36.

$$\sigma_s = 1.6483 + P_1N + P_2N^2 + P_3S + P_4I + P_5B + P_6C + P_7D + P_8NS + P_9N^2S + P_{10}NI^2 + P_{11}NB + P_{12}N^2B + P_{13}ND + P_{14}N^2D + P_{15}SC + P_{16}SD + P_{17}IC + P_{18}ID + P_{19}CD + P_{20}SCD + P_{21}ICD \quad (17)$$

where: P_1 - P_{21} = Polynomial coefficients in Table 37

N = Rotational speed

S = Slurry type (0 = WV, 1 = HW)

I = Impeller type (0 = A310, 1 = PBT)

B = Baffles (0 = yes, 1 = no)

C = Coils (0 = yes, 1 = no)

D = Diameter (0 = 4.5 in., 1 = 3.75 in.)

An attempt was made to compare the effects of the 3.75, 4.5 and 7.5 in. diameter impellers at a rotational speed of 50 rpm in a tank with baffles and without coils. It was only possible to identify the single factor and two-way factor-interaction effects on the solids distribution in the mixing tank. Unfortunately, the generally good mixing ($\sigma_s < 1$) achieved by each impeller at 50 rpms, masked the diameter effect that was anticipated (see Section 3). None of the effects evaluated was significant at the 0.05 level. Several two-way interactions were significant at the 0.10 level.

FIGURE 33. Experiments with 3.75 and 4.5 Inch Diameter Impellers: Solution to Predictive Equation for Mixing Configurations 1,5

FIGURE 34. Experiments With 3.75 and 4.5 Inch Diameter Impellers: Solution to Predictive Equation for Mixing Configurations 2,6

FIGURE 35. Experiments with 3.75 and 4.5 Inch Diameter Impellers: Solution to Predictive Equation for Mixing Configurations 3,7

FIGURE 36. Experiments with 3.75 and 4.5 Inch Diameter Impellers: Solution to Predictive Equation for Mixing Configurations 4,8

7.0 DESIGN REQUIREMENTS FOR THE FULL-SCALE SYSTEM

The data collected during the bench-scale mixing experiments (see Section 6) established the significance of factor and factor-interaction effects on the degree of mixing achieved in the bench-scale mixing tank. This information is: 1) scaled-up to predict the full-scale performance that is likely to occur in mixing systems similar to those installed at the CTS (WV) and DWPF (SR), and 2) used to design full-scale mixing systems that can achieve a solids distribution in the mixing tank that falls within ± 1 , 3, and 5 percent of the mean value, in Section 7.1.

One goal of this evaluation was to prepare fabrication drawings for a mixing tank shell that incorporates the geometric features necessary to properly mix simulated HW and WV melter feeds on a large-scale. The selection of the tank shell configuration and the design details of the tank shell are discussed in Section 7.2.

7.1 SCALE-UP OF BENCH-SCALE MIXING DATA

The bench-scale mixing results obtained during this evaluation were discussed in Section 6.4.3. The procedures used to scale-up bench-scale mixing data were discussed in Section 3. An important step in the scale-up procedure is to establish the configuration of the large-scale mixing system. The selection of the large-scale mixing system configurations that are likely to be effective in mixing HW and WV melter feed slurries is discussed in Section 7.1.1. The procedures used to size the mixing equipment that would be installed in each large-scale configuration are summarized in Section 7.1.2.

7.1.1 Selection of Large-Scale Mixing System Configurations

The geometry of the bench-scale mixing system was based on existing melter feed slurry mixing tank designs that are being evaluated at the DWPF (SR) and at CTS (WV), and on the design of slurry mixing tanks frequently used in the mining and metallurgical industry (see Section 3.3.2). The bench-scale information can only be scaled-up to predict the performance of geometrically similar large-scale equipment. A total of eight geometric configurations were evaluated during the bench-scale tests (see Table 26). The bench-scale data

collected for four of these configurations was scaled-up to provide an indication of full-scale performance.

The four bench-scale configurations that were considered most applicable to the melter feed slurry mixing application are listed in Table 38. The DWPF and CTS configurations were selected because they are currently being evaluated for this application. Mixing system configuration 6 was selected since it closely approximates the configuration that is used to suspend mining and metallurgical slurries. The D/T ratio used in the mining and metallurgical industry ranges from 0.33 to 0.50. The bench-scale impeller diameters corresponding to these D/T ratios are 4.5 and 7.5 inches, respectively.

The experimental data associated with the bench-scale mixing system configurations listed in Table 38 were scaled-up to provide an indication of full-scale performance. The scale-up procedures employed are discussed in Section 7.1.2.

7.1.2 Sizing of the Full-Scale Mixer

The first step in the scale-up procedure is to define the desired process result (see Figure 9). The dependent variable evaluated during the bench-scale experiments was the degree of mixing achieved. The standard deviation of the six total solids content values measured during each run was the specific dependent variable utilized. This dependent variable, σ_s , is related to the absolute value of the average percent total solids variation, $\bar{\Delta}\%$, that existed in the bench-scale tank during each run. The $\bar{\Delta}\%$ value is the index usually associated with the solids distribution in a mixing tank. A $\bar{\Delta}\%$ value of ≤ 5 indicates that all six total solids content values measured during a bench-scale run were within $\pm 5\%$ of the mean total solids content in the tank during that run.

The relationship between σ_s and $\bar{\Delta}\%$ for the bench-scale configurations listed in Table 38 is plotted on Figure 37. The slurry utilized had a significant effect on the σ_s - $\bar{\Delta}\%$ relationship. For a given value of $\bar{\Delta}\%$, the simulated WV slurry produced a higher value of σ_s than the simulated HW slurry. The σ_s values associated with $\bar{\Delta}\%$ levels of ± 1 , 3 and 5% are listed in Table 39. The

TABLE 38. Bench-Scale Mixing System Configurations Selected for Scale-Up

Configuration Description	Number	Upper Impeller		Bayonet Coils	Baffles
		Type	Size (inches)		
DWPF (SR)	1	High Efficiency Turbine	3.75	Yes	No
CTS (WV)	2	High Efficiency Turbine	4.5	No	Yes
Industrial A(a)	6	Pitch Blade Turbine	7.5	No	Yes
Industrial B(a)	6	Pitch Blade Turbine	4.5	No	Yes

(a) Industrial A and B represent designs typically used in the mining and metallurgical industries.

FIGURE 37. Relationship Between σ_s and $\bar{\Delta}\%$

TABLE 39. Scale-Up Decision Criteria

Configuration		Decision Value of σ_s					
Description	Number	Simulated HW Slurry			Simulated WV Slurry		
		$\pm 1\%$	$\pm 3\%$	$\pm 5\%$	$\pm 1\%$	$\pm 3\%$	$\pm 5\%$
DWPF (SR)	1	0.30	0.95	1.70	0.45	1.30	2.20
CTS (WV)	2	0.30	0.95	1.70	0.45	1.30	2.20
Industrial A (7.5)	6	0.30	0.95	1.70	0.45	1.30	2.20
Industrial B (4.5)	6	0.30	0.95	1.70	0.45	1.30	2.20

bench-scale conditions associated with these $\bar{\Delta\%}$ values were scaled-up to provide an indication of full-scale performance.

The next step in the scale-up procedure is to identify the rotational speed at which each mixing configuration achieved a level of homogeneity equal to ± 1 , ± 3 and ± 5 wt% solids. These rotational speeds were identified by inspecting the bench-scale predictive equation plots on Figures 32, 33, and 34. They are listed in Table 40.

The scale-up procedure described in Section 3.3.3 is now used to identify the full-scale rotational speed that will achieve the same degree of mixing achieved in the bench-scale unit. Recall that the scale ratio, R , was defined as:

$$R = \left(\frac{V_2}{V_1} \right)^{1/3} \quad (5)$$

The large-scale mixing tank volume, V_2 , was assigned a value of 16,200 liters. This value approximates the fluid volume that will be contained in the large-scale mixing tank shell described in Section 7.2. The lab-scale mixing tank volume was 43 liters. The calculated scale ratio, R , becomes 7.2 for this scale-up application.

The full-scale rotational speed was related to the bench-scale rotational speed by Equation (7).

$$\frac{N_2}{N_1} = \left(\frac{1}{R} \right)^m \quad (7)$$

TABLE 40. Bench-Scale Rotational Speeds Needed To Achieve ± 1 , ± 3 and ± 5 Percent Solids Homogeneity in Selected Mixing Configurations

Configuration		Simulated Melter Feed Slurry					
Description	Number	HW			WV		
		$\pm 1\%$	$\pm 3\%$	$\pm 5\%$	$\pm 1\%$	$\pm 3\%$	$\pm 5\%$
DWPF (SR)	1	150	125	50	100	50	<50
CST (WV)	2	50	<50	<50	80	50	<50
Industrial A (7.5)	6	20	15	10	>10	10	<10
Industrial B (4.5)	6	100	50	<50	95	50	<50

The selection of exponent, m , is based on a knowledge of the controlling mixing class in the agitation system (Table 10) and on experience (Table 12). The calculated full-scale rotational speeds associated with the bench-scale rotational speeds listed in Table 40, are summarized in Table 41, for values of exponent, m , equal to 0.5, 0.67, 0.75 and 1.0. The calculated full-scale rotational speed increases substantially as m decreases from 1.0 (for equal fluid motion) to 0.5 (for equal shear rates). The value of m usually associated with solids suspension applications is $m = 0.75$.

What level of agitation will the rotational speeds listed in Table 41 induce in the full-scale tank? The scale of agitation in a mixing tank (Table 5) can be related to the bulk velocity in the tank (Table 7). The bulk velocity, V_b , in a mixing tank depends on the fluid motion induced by the impeller. The volumetric flow induced by an impeller equals (see Section 6.1):

$$q = N_Q ND^3 \quad (18)$$

$$v_b = \frac{q}{A} \quad (19)$$

where: q = volumetric flow rate, ft^3/min

N = rotational speed, rpm

D = Impeller diameter, ft

N_Q = Pumping Number (1.58)

v_b = Bulk Velocity, ft/min

A = cross sectional area of mixing tank, ft^2

TABLE 41. Calculated Large-Scale Rotational Speeds Needed to Achieve ± 1 , ± 3 and $\pm 5\%$ Solids Homogeneity in Selected Mixing Configurations

Configuration		Decision Criteria ($\pm\%$)	Full-Scale Rotation Speed (rpm)							
			m = 0.5		m = 0.67		m = 0.75		m = 1.0	
			Slurry		Slurry		Slurry		Slurry	
Description	Number		HW	WV	HW	WV	HW	WV	HW	WV
DWPF (SR)	1	1	56	37	40	27	34	23	21	14
		3	47	19	33	13	28	11	17	7
		5(a)	19	13	13	9	11	8	7	5
CTS (WV)	2	1	17	30	13	21	11	18	7	11
		3(b)	13	19	9	13	8	11	5	7
		5(b)	13	13	9	9	8	8	5	5
Industrial A (7.5)	6	1(c)	8	8	5	5	5	5	3	3
		3	6	4	4	3	3	2	2	1
		5(d)	4	3	3	2	2	2	1	1
Industrial B (4.5)	6	1	37	35	27	25	23	22	14	13
		3	19	19	13	13	11	11	7	7
		5(b)	13	13	9	9	8	8	5	5

(a) Assumes bench-scale $N = 35$ rpm for WV

(b) Assumes bench-scale $N = 35$ rpm for HW and WV

(c) Assumes bench-scale $N = 20$ rpm for WV

(d) Assumes bench-scale $N = 7$ rpm for WV

The scale of agitation N_I was defined as (Section 3.3.2):

$$N_I = \frac{v_b}{6} \quad (20)$$

The full-scale impeller diameters were established by maintaining the same D/T ratios that were used during bench-scale tests. These impeller diameters are listed in Table 42. They were used in conjunction with the rotational speeds in Table 41, and an N_Q value of 1.58, to establish the predicted scale of agitation values listed in Table 43. These scale-of-agitation values are unusually low for a solids suspension application (see Table 5).

The low values of N_I listed in Table 43 are due to the very slow settling rates exhibited by simulated HW and WV melter feed slurries. The measured

TABLE 42. Full-Scale Impeller Diameters (R = 7.2)

Configuration		Full-Scale Impeller	
Description	Number	Type	Diameter (in.)
DWPF (SR)	1	A310	27.0
CTS (WV)	2	A310	32.4
Industrial A	6	PBT	54.0
Industrial B	6	PBT	32.4

TABLE 43. Calculated Scale of Agitation Values Associated with ± 1 , ± 3 and $\pm 5\%$ Solids Homogeneity for the Large-Scale Mixer

Configuration		Decision Criteria ($\pm\%$)	Scale of Agitation, N_I							
			$m = 0.5$		$m = 0.67$		$m = 0.75$		$m = 1.0$	
			Slurry		Slurry		Slurry		Slurry	
Description	Number		HW	WV	HW	WV	HW	WV	HW	WV
DWPF (SR)	1	1	2.6	1.7	1.9	1.3	1.6	1.1	1.0	0.7
		3	2.2	0.9	1.6	0.6	1.3	0.5	0.8	0.3
		5	0.9	0.6	0.6	0.4	0.5	0.4	0.3	0.2
CTS (WV)	2	1	1.5	2.4	1.1	1.7	0.9	1.5	0.6	0.9
		3	1.1	1.5	0.8	1.1	0.7	0.9	0.4	0.6
		5	1.1	1.1	0.8	0.8	0.7	0.7	0.4	0.4
Industrial A (7.5)	6	1	2.8	2.8	2.0	2.0	1.7	1.7	1.1	1.1
		3	2.1	1.4	1.5	1.0	1.3	0.9	0.8	0.5
		5	1.4	1.0	1.0	0.7	0.9	0.6	0.5	0.4
Industrial B (4.5)	6	1	3.0	2.9	2.2	2.0	1.9	1.8	1.1	1.1
		3	1.5	1.5	1.1	1.1	0.9	0.9	0.6	0.6
		5	1.1	1.1	0.8	0.8	0.7	0.7	0.4	0.4

settling rates (see Section 5.3) at 20°C of HW and WV slurries were 0.45 cm/hr and 0.6 cm/hr, respectively. If the suspended solids in these slurries settled as discrete particles, and had a diameter of 40 microns, these settling velocities would be about 0.3 ft/min or 550 cm/hr (Gates et al. 1976).

Apparently, the yield stress exhibited by each slurry is substantially reducing the settling velocity of the solids in the slurry. A low energy input (N_I of 1 to 3) will keep these particles in suspension.

The large-scale mixer must be able to resuspend solids if they settle out as a result of a power outage or other disturbance. The scale of agitation necessary to resuspend solids is higher than the value needed to maintain solids in suspension. Historically, solids suspension applications where a uniform suspension of solids is required are designed to provide a scale of agitation of 8 to 10 (see Table 5). Values of N_I of 9 to 10 will create surging surface waves. These waves will create wet-dry interfaces and foster the accumulation of solids. This level of mixing should be avoided. An N_I value of 8 is a prudent upper bound for solids suspension applications. An N_I value of 8 implies a bulk fluid velocity of 48 ft/min. Equations (18) through (20) can be used to establish the full-scale rotational speeds needed to create this bulk velocity. These rotational speeds are listed on Table 44.

What shaft power will be consumed by each impeller listed in Table 44? The power consumed by a mixer installed in a given mixing configuration depends on the Reynolds number, Froude number, and the geometry of the mixing system (see Equation 4). The power number, N_p , relates the power consumed by a mixer to its impeller diameter and rotational speed. The power number was defined (see Section 3.3.1) as:

$$N_p = \frac{P}{\rho N^3 D^5} \quad (21)$$

The power number-Reynolds number relationship was established for each bench-scale mixing configuration (see Section 6.3.3).

The first step in estimating the power that will be consumed by a mixer is to identify the Reynolds number at which it will operate. The Reynolds number was defined as:

TABLE 44. Calculated Large-Scale Mixer Shaft Power Needed to Achieve a Scale of Agitation of 8 for Selected Mixing Configuration

Configuration		Rotational Speed		Shear Rate (sec ⁻¹)	Apparent Viscosity (cp)		Reynold's Number		Power Number		Power (hp)	
Description	Number	Calculated	Commercially Available		HW	WV	HW	WV	HW	WV	HW	WV
DWPF (SR)	1	164	190	36.4	141	13	13,600	144,000	3.3	3.1	5.5	5.1
CTS (WV)	2	98	100	19.2	253	19	5,750	74,800	4.5	5.0	7.6	8.2
Industrial A (7.5)	6	21	20	3.8	1,183	57	680	13,900	4.3	4.3	8.1	7.9
Industrial B (4.5)	6	98	100	19.2	253	19	5,750	74,800	5.6	5.7	9.4	9.3

$$N_{Re} = \frac{\rho ND^2}{\mu_a} \quad (22)$$

The viscosity of simulated HW and WV slurries is dependent on the shear rate applied to the slurry. The shear rate applied to each slurry is equal to approximately 11.5 times the rotational speed of the impeller (Section 6.4.2).

The rotational speeds listed in Table 44 were used to estimate the shear rate present in the full-scale mixing tank. This shear rate established the apparent viscosity μ_a (Equations 10, 11) of simulated HW and WV in the full-scale mixing configurations. This value of μ_a , the diameter of the full-scale impeller (see Table 42), and the rotational speeds in Table 44 were used to calculate the full-scale Reynolds number for each mixing configuration.

The power number, N_p , associated with each Reynolds number was determined by inspection of the appropriate power number-Reynolds number curve (Section 6.3.3). This power number was used to calculate (Equation 21) the power consumed at each full-scale impeller rotational speed. These power requirements are summarized in Table 44.

The power levels listed in Table 44 are the power levels actually applied to the slurry in the mixing tank. The motor used to operate the mixer will have a power rating higher than those listed. This motor must overcome energy losses that will occur in the motor and the gearbox. It must also possess a

start-up torque that can overcome the yield stress of the slurry in the tank. The yield stress will increase as the length of time the slurry remains motionless in the tank increases. The effect of time on this yield stress value was not evaluated. It is likely that this start-up torque requirement will be the critical factor that will fix the size of the motor needed for melter feed slurry mixing applications.

7.2 DESIGN OF THE FULL-SCALE MIXING TANK SHELL

One goal of this evaluation was to select a reference prototype mixing system for full-scale verification tests. After selection of the reference, fabrication drawings were prepared for a large-scale mixing tank shell that incorporates the geometric features necessary to properly mix simulated HW and WV melter feed slurries. The selection criteria used to specify the large-scale tank shell design are described.

The large-scale mixing tank will be used to prepare simulated HW and WV melter feed slurries. The tank must be located in the vicinity of existing vitrification process equipment. A decision was made to locate the large-scale tank in the space now occupied by an existing tank in the research facility. The largest tank diameter that can fit in this location is 9 feet. This dimension was chosen for the large-scale mixing tank shell. Since mixing operations are most efficient when carried out in vertical, right-circular cylindrical tanks, a tank diameter of 9 feet was selected for the large-scale mixing tank.

The height of the mixing tank must be sufficient to contain the fluid being processed and provide the freeboard necessary to prevent inadvertent spills of the tank contents. Industrial practice is to use an L/T ratio of approximately 1 (See Table 6) for solids suspension applications. The L/T for the DWPF mixing tank is 1.08, where the L/T ratio of the WV tank is 0.80. An L/T of 1 for this large-scale tank provides for a fluid level of 9 feet. The total height of the mixing tank shell was set at 11 feet to allow for 2 feet of freeboard.

The large-scale mixing tank shell was designed to have a flat bottom profile to conform to industrial practice, and to the design installed at WV. The DWPF mixing tank has a slightly sloping bottom to facilitate the removal of

mercury that may settle out in the tank. Since HW and WV slurries contain very little mercury, the need for a sloping bottom on the large-scale tank shell was not considered necessary.

The large-scale tank shell was designed to contain four, equally spaced wall baffles. The baffles were sized according to industrial practice (see Table 4), with a B_w/T ratio of 0.083 and a B_g/T ratio of 0.014. The baffles are removable.

The tank shell and its support structure were designed to withstand a total load of 55,000 pounds. This load includes the weight of the fluid in the tank, as well as the weight of the tank cover, agitator, slurry pump, and other auxillary equipment that the tank shell must support. The tank wall thickness needed to support this load was calculated to be 3/16 inch.

The material of construction specified for all tank components was 300 series stainless steel.

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APPENDIX A

POWER NUMBER--REYNOLDS NUMBER RELATIONSHIP FOR EACH
BENCH-SCALE MIXING CONFIGURATION

FIGURE A.1. Power Number Versus Reynolds Number Relationship with Newtonian Fluids Configurations 5 to 8 Using 3.75 in. Diameter Impellers

FIGURE A.2. Power Number Versus Reynolds Number Relationship with Newtonian Fluid Configurations 1 to 4 Using 3.75 in. Diameter Impeller

FIGURE A.3. Power Number Versus Reynolds Number Relationship with Newtonian Fluids Configurations 1 to 4 Using 4.5 in. Diameter Impellers

FIGURE A.4. Power Number Versus Reynolds Number Relationship with Newtonian Fluids Configurations 6 and 8 Using 7.5 in. Diameter Impeller

FIGURE A.5. Power Number Versus Reynolds Number Relationship with Newtonian Fluids Configurations 2 and 4 Using 7.5 in. Diameter Impeller

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