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Modeling and Analysis of ORNL Horizontal Storage Tank Mobilization and Mixing

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MASTER

SUMMARY

This report presents and analyzes the results of computer modeling of the mixing and mobilization of non-Newtonian sludge by submerged liquid jets in horizontal cylindrical storage tanks. The TEMPEST computational fluid dynamics computer code was used for all of the modeling. The storage tank configuration was designed to represent the Melton Valley Storage Tanks (MVST) at Oak Ridge National Laboratory (ORNL) and the planned 378 m³ (100,000 gallon) tanks of the MVST Capacity Increase Project (MVST-CIP). Sludge mobilization and mixing times have been analyzed for a 378 m³ (100,000 gallon) tank of this configuration, using fluid with the observed non-Newtonian rheology.

During the prior fiscal years' work, which was reported in Terrones and Eyler (1993), it was demonstrated that computer modeling of the non-Newtonian turbulent jet mixing of tank contents is feasible. The rheology of a typical MVST sludge, from the MVST W-28 tank, was fit to a two-parameter power law model, which provided a good description for strain rates above 1 sec⁻¹. With a jet located 1/4 of the way along the tank's long axis, rather than at the center, a minimum recirculating jet velocity of 6.1 m/s (20 ft/s) was estimated to be necessary to mobilize W-28 sludge in a half-filled 1/6 scale tank. Mixing times were calculated for salt tracers in supernatant in the 1/6 and 2/3 scale tanks, of 0.87 m³ (230 gal) and 90.8 m³ (24,000 gallon) capacity respectively, and were in reasonable agreement with measured mixing times. The exact definition of the mixing time, in terms of concentration time history at a finite number of locations, was found to have a strong effect on the predicted mixing time. A consistent criterion for mixing time definition was suggested for future work. Finally, a simulant development effort was completed and a sludge simulant (a kaolin/bentonite suspension in water) was suggested.

This work was continued into the present fiscal year with modeling of mixing and sludge mobilization in the 378 m³ (100,000 gallon) tank. In this tank, the double jets are centrally mounted several centimeters above the tank bottom. The sludge rheology had been determined in previous work (Terrones and Eyler 1993) for sludge in MVST W-28. The effect of the tank internals was not addressed. Two jet diameters, 7.6 and 10 cm (3 and 4 in), were selected as adequate for sludge mobilization and were used in later modeling.

With the jet diameters chosen, the tank mixing times were determined for a tank completely filled with supernatant, without sludge. A neutrally buoyant tracer was used to track mixing. Constant jet velocities of 0.61 to 4.6 m/s (2 to 15 ft/s) were modeled. These corresponded to 0.0126 to 0.0492 m³/s (200 to 780 gpm). Two suction configurations were considered: in the first, the recirculation suction was centrally located and under the jets; in the second, the suction was at the far end of the tank and well above the tank floor. Both sets of mixing times were correlated as functions of the jet diameter and the tank circulation time constant. The mixing time was longer for center suction than for end suction, shorter for the smaller jet diameter (given the same flow rate), and barely affected by the height of the end-suction above the tank floor. Supernatant mixing times of 10 minutes or less were estimated for the highest jet velocities modeled. The mixing times and the correlations that fitted them are shown in Figure S-1.

The mobilization of MVST W-28 sludge was also modeled. Again, center-suction and end-suction configurations were modeled. The tank, which was 5.1 m (16.7 ft) high, was filled with supernatant containing 0.61 m (2 ft) of sludge (center depth). The objective was to find the minimum jet velocity that was needed to achieve sludge mobilization. (In previous work [op.cit.],

it had been found that jet velocities below a minimum "threshold velocity" left "sludge banks" of unremovable material on the floor of the tank.)

For the 378 m³ (100,000 gallon) tank, there were no sludge banks even at very low velocities, and so no "threshold velocities." Thus, the minimum velocity for sludge mobilization could not be defined in terms of what was needed to avoid sludge banks. Mobilization was instead tested by whether the rise velocity of the sludge-supernatant interface at the far end of the tank exceeded the sludge particle settling velocity of 2.9 cm/min. By this criterion, the minimum jet velocity for mobilization was between 3.0 and 4.6 m/s (10 and 15 ft/s) for a 10-cm (4-in) jet diameter and between 4.6 and 6.1 (15 and 20 ft/s) for a 7.6-cm (3-in) jet diameter. Figure S-2 shows the relationship between the sludge-supernatant interface's rise velocity (the slopes of the lines) and the settling velocity for cases in which mobilization is or is not expected. The product of jet diameter and jet velocity was required to be about 0.46 m²/s (5 ft²/s) to cause mobilization. The mixing time was about 80 minutes in both the 4.6 m/s 10-cm case and the 6.1 m/s 7.6-cm case. Apparently, the high non-Newtonian viscosity and the suppression of mixing by the density gradient caused the mixing times to be 6 to 8 times longer than for clear supernatant mixing.

The center-suction configuration was also used to find the effects of smaller sludge depth (0.46 m, or 1.5 ft) and doubled sludge density. The double-density case required about 1.5 m/s (5 ft/s) greater jet velocity for mobilization than did the base case. Decreasing the sludge depth from 0.61 m to 0.46 m had little effect on the minimum velocity required for mobilization. As expected, the smaller sludge depth decreased the jet velocity required for mobilization by decreasing the kinetic energy needed to impart velocity to the sludge.

Sludge mobilization was also modeled for an end-suction tank configuration containing 0.61 m (2 ft) of sludge. Mobilization was judged by whether the upward fluid velocity at the end of the tank was large enough to prevent settling. By this criterion, the minimum jet velocity for mobilization was between 4.6 and 6.1 m/s (15 to 20 ft/s) for a 10-cm (4-in) diameter jet and between 7.6 and 9.1 m/s (25 to 30 ft/s) for a 7.6-cm (3-in) diameter jet. For this case, the requisite minimum diameter-velocity product was about 0.65 m²/s (7 ft²/s). The end-suction configuration required a higher velocity for sludge mobilization than the center-suction configuration because the end-located suction tends to decrease the velocity at the end-wall.

As part of another ongoing waste-tank remediation project (the Flammable Gas Program, which is focused on Hanford Tank 241-SY-101), the MVST W-28 sludge rheological data were analyzed and compared to several rheological models. Biviscous models such as those of Carreau and Cross models (Ellwood et al. 1990, Cross 1965) were found to provide a more complete model of W-28 rheology than the power-law model. The results also suggest that the W-28 sludge can be described better by models that do not include a yield stress than by models that do.

Figure S-1. Correlated and modeled supernatant mixing times as a function of flow rate.

Figure S-2. The heights of the final-concentration contour at the tank end for the center-suction configuration.

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1.0 INTRODUCTION

The retrieval and treatment of radioactive sludges that are stored in tanks constitute a prevalent problem at several U.S. Department of Energy sites. The tanks typically contain a settled sludge layer with non-Newtonian rheological characteristics covered by a layer of supernatant. The first step in retrieval is the mobilization and mixing of the supernatant and sludge in the storage tanks. Submerged jets have been proposed to achieve sludge mobilization in tanks, including the 189 m³ (50,000 gallon) Melton Valley Storage Tanks (MVST) at Oak Ridge National Laboratory (ORNL) and the planned 378 m³ (100,000 gallon) tanks being designed as part of the MVST Capacity Increase Project (MVST-CIP).

This report focuses on the modeling of mixing and mobilization in horizontal cylindrical tanks like those of the MVST design using submerged, recirculating liquid jets. The computer modeling of the mobilization and mixing processes uses the TEMPEST computational fluid dynamics program (Trent and Eyler 1992). The goals of the simulations are to determine under what conditions sludge mobilization using submerged liquid jets is feasible in tanks of this configuration, and to estimate mixing times required to approach homogeneity of the contents.

During earlier phases of this work (Terrones and Eyler 1993), it was demonstrated that computer modeling of the non-Newtonian turbulent jet mixing of tank contents is feasible. The rheological behavior of a typical MVST sludge, from the MVST W-28 tank, was fit to a two-parameter power law model which provided a good description for strain rates above 1 sec⁻¹. A minimum recirculating jet velocity of 6.1 m/s (20 ft/s) was estimated to be necessary to mobilize W-28 sludge in a half-filled 1/6 scale tank where the suction was located above the jets and above the sludge level. At lower velocities, some "sludge banks" remained on the bottom of the 1/6 scale tank. Mixing times were estimated for salt tracers in supernatant in the 1/6 and 2/3 scale tanks, and proved to be in reasonable agreement with the measured mixing times. The exact definition of the mixing time, in terms of concentration time history at a finite number of locations, was found to have a strong effect on the predicted mixing time, and a consistent criterion for mixing time definition was suggested for future work. Finally, a simulant development effort was completed and a sludge simulant (a kaolin/bentonite suspension in water) was suggested.

This work is a continuation of the modeling of mixing and sludge mobilization, this time using the 100,000 gallon tank design; here the double jets (pointing in opposing directions) are centrally mounted several centimeters above the tank bottom along its centerline. The sludge rheology was that which had been determined in previous work (Terrones and Eyler 1993) for ORNL W-28 sludge. The effect of the tank internals was not addressed. In Task 1 of this work, a literature review and preliminary modeling of 7.6-cm-diameter jets were employed to determine jet diameters that would be likely to ensure sludge mobilization. Two diameters, 7.6 and 10 cm (3 and 4 in), were selected and were used in later modeling.

In Tasks 2 and 3, the mixing times were determined for a tracer (neutrally buoyant, non-reacting chemical addition) for two different suction configurations in a tank completely filled with supernatant without sludge. In the first configuration, the recirculation suction was centrally located, just under the jets. In the second configuration, the suction was located at the far end of the tank some distance from the center of the tank floor. Suction heights of 0.7 m (2.3 ft), 1.2 m (4 ft), and 5.1 m above the tank bottom were modeled. Both sets of supernatant mixing times could be accurately expressed as functions of the jet diameter and the tank circulation time constant (tank fluid volume divided by recirculation flow rate). It was also possible to include the "extent of mixing" as part of the correlation. Here "extent of mixing" means the maximum percent deviation

of a locally monitored concentration from the final fully-mixed value. The effect of prolonged injection of the tracer was also explored in the modeling done for Tasks 2 and 3.

The mobilization of W-28 sludge was modeled for Task 4. Again, two suction configurations were considered, both with 0.61 m (2 ft) of sludge (center depth) in the tank. One was a configuration with the suction located at the tank center below the jets and the sludge surface, and the other placed the suction at a different height above the sludge at each end. Jet velocities from 1.5 to 9.1 m/s (5 to 30 ft/s) were modeled.

For both the center-suction and end-suction configurations, there was no clear "threshold velocity" below which undisturbed sludge banks remained on the floor of the tank. (The 1/6 scale mobilization modeling, in which the recirculation suction was located in clear supernatant above the jets, had exhibited definite threshold velocities.) Because of the lack of a threshold velocity, mobilization was tested by whether the rise velocity of the sludge-supernatant interface at the end of the tank was greater than the sludge particle settling velocity. Jet velocities of 4.6 to 9.1 m/s (15 to 30 ft/s) were necessary for sludge mobilization. A higher jet velocity was required for the end-suction configuration than for the center-suction configuration. The center-suction configuration was also used to investigate the effects of smaller sludge depth (0.46 m, or 1.5 ft) and doubled sludge density.

As part of another ongoing DOE waste-tank remediation project (the Flammable Gas Program, which is focused on Hanford Tank 241-SY-101), the MVST W-28 sludge rheological data were analyzed and compared to several rheological models other than the power law.

1.1 Objectives/Statement of Work

The purpose of this work was to assess the requirements for mobilizing non-Newtonian sludges using submerged liquid jets and to estimate the time required to reach homogeneous concentration distributions in horizontal tanks. Simulations in tank configurations similar to the ORNL storage tanks, known as MVSTs, were carried out with the three-dimensional time-dependent fluid-flow simulator program TEMPEST (Trent and Eyler 1992). Simulations were intended to complement experimental work at ORNL and to support investigations in the retrieval of wastes.

Specific objectives of the present work are to

- Simulate mixing and homogenization of Newtonian supernatant in a fully-filled horizontal tank, investigating different suction configurations and jet parameters.
- Determine the required jet velocity to mobilize the non-Newtonian sludge in the tank.

To satisfy these objectives, the statement of work includes four tasks:

- Task 1: Make a preliminary investigation to determine two plausible jet diameters for sludge mobilization. All subsequent simulations for the 100,000 gallon tanks will use these jet diameters.
- Task 2: Determine mixing times for a tank filled with supernatant only, with a centrally-located recirculation line. Two opposed jets are in the middle of the tank at least two

diameters above the floor. Chemical addition (modeled as non-reactive) is near the tank middle (lengthwise) near the top. The design condition is 200 gpm (total from both jets).

- Task 3: The same as Task 2, but with the recirculation suction inlets located at equal heights at the two tank ends.
- Task 4: Determine the minimum velocity to mobilize W-28 sludge for each of two jet diameters. Two recirculation inlet locations are to be considered, one at the tank center under the jets and the other at unequal heights at the tank ends.

Section 2 of this report gives the conclusions and recommendations, based on work performed to fulfill the four tasks. Section 3 describes the general technical approach. The detailed description and results of Task 1 are presented in Section 4, the results of Tasks 2 and 3 in Section 5, and results of Task 4 in Section 6. A summary of an additional analysis of the rheological behavior of W-28 sludge is given in Appendix A.

1.2 Background

The systematic study of jet mixing in tanks started half a century ago and has concentrated on the mixing of emulsions and dilute suspensions, which are principally Newtonian fluids. The preferred geometric configuration in these investigations is the vertical cylindrical tank containing a single submerged jet whose location and orientation within the tank are varied. Horizontal tanks, and highly viscous and non-Newtonian fluids, have not received much attention. Some investigators believe that the tank geometry has little effect on the mixing time, but this has not been conclusively established.

Oak Ridge researchers (Shor and Cummins 1991) have investigated mobilization of sludge similar to that in the MVSTs. Ceo et al. (1990) experimentally determined that sludges and slurries from eight of the MVSTs behave like viscoplastic materials with a small apparent yield stress. Their data were later used by Youngblood et al. (1991) to correlate sludge transport behavior and to calculate the pressure drop in pipes. Graphical comparisons indicate that the power law model renders a better description than the Bingham plastic model, provided the strain rate (shear rate) exceeds 1.0 sec^{-1} .

In much of the literature mixing has been defined in a rather vague or qualitative manner. Some mixing time correlations have been based on concentration information recorded at only two points within the vessel (Maruyama 1986). The particular definition of mixing markedly affects the mixing time. Mixing time estimates computed with different defined criteria, using the same time-history data, can vary by as much as 30 to 40% (Terrones and Eyler 1993). This is significant in comparing computer simulations to experimental data and may, in part, be an explanation for the variation in empirical mixing correlations.

1.3 Motivation

Numerical simulations assist in improving understanding of the phenomena occurring in the mobilization process. They not only complement the ongoing experimental efforts, but also provide information which cannot be obtained from experimentation. Through numerical experiments it is possible to construct accurate correlations for mixing times, because the concentration time history of each computational cell can be computed.

The TEMPEST model has been used extensively in the simulation of three-dimensional time-dependent thermo-fluid problems in complex geometries and its predictions for jet mixing have agreed well with earlier experimental work (Rajaratnam 1976). TEMPEST has also been validated for power-law fluids in pipe and channel flow geometries for which analytic solutions are known. Terrones and Eyler (1993) found that it was feasible to use TEMPEST to model non-Newtonian jet and mixing flow in horizontal cylindrical tanks. Their predictions of mobilization in 1/6 and 2/3 scale tanks have not yet been confirmed experimentally^(a).

^(a) While some of these data have been measured in 1/6 and 2/3 scale MVST experiments, they have not yet been published and made available.

2.0 CONCLUSIONS AND RECOMMENDATIONS

This work investigated the mixing times and minimum jet velocities for sludge mobilization for the 378 m³ (100,000 gallon) tank being designed as part of the MVST-CIP. The tanks are assumed to contain ORNL W-28 sludge. The investigations of supernatant mixing times included the effects of jet diameter and velocity, different suction locations, and chemical additive injection time. The sludge mobilization analysis considered primarily jet diameter and velocity; the sensitivity to sludge depth and sludge density effects was also tested.

2.1 Conclusions

Preliminary modeling and a brief literature review were conducted to determine the appropriate jet diameters for use in more detailed modeling, with the following conclusions:

- The mixing time in the MVST-CIP horizontal cylindrical tank configuration appears to have the same approximate functional dependencies as mixing times in the vertical tanks that had been studied in the literature. In a turbulent jet regime, the mixing time can be expected to be proportional to the recirculation time. When the recirculation time (the flow rate) is held constant, the mixing time is proportional to the jet diameter and inversely proportional to the length of the jet axis (owing to entrainment). The jet axis length is defined as the distance along the jet axis from the nozzle to the nearest blocking surface.
- Jet diameters of 7.6 and 10 cm (3 and 4 in) are small enough to ensure that the jet remains in the low-mixing-time turbulent regime during supernatant mixing.

TEMPEST modeling was conducted for a supernatant-filled tank, with jet diameters of 7.6 and 10 cm (3 and 4 in) and a range of recirculation flow rates from 0.0126 to 0.0492 m³/s (200 to 780 gpm). The conclusions were as follows:

- The location of the suction makes little difference in supernatant mixing. The mixing time for a suction location under the jets (centrally located) is only about 20% greater than for a suction location at the end of the tank. For end-suction, the change in mixing time for suction over a wide range of heights is less than 6%, with the longer mixing times at lower suction heights.
- Adding the tracer over a finite time, rather than instantaneously, has little effect on mixing time. The increase in mixing time that is caused by injecting tracer over a 10-minute period is less than 5 minutes, under the studied range of conditions. For comparison, the mixing times ranged from 15 minutes to more than an hour.
- The mixing time can be correlated with the recirculation time (tank volume divided by flow rate) and the jet diameter, and is nearly linearly proportional to the product of those variables. The correlations are given in Section 5.3.

TEMPEST modeling was also conducted for a tank filled with supernatant and containing 0.61 m (2 ft) of sludge out of a total tank height of 5.1 m (16.7 ft). This 0.61-m depth of settled sludge occupied a bulk volume of 6.1% of the tank. The range of conditions for a jet diameter of 7.6 cm (3 in) included recirculation flow rates from 0.0278 to 0.0694 m³/s (440 to 1100 gpm); these corresponded to jet velocities from 3.0 to 7.6 m/s (10 to 25 ft/s). The range of conditions for a jet diameter of 10 cm (4 in) included flow rates from 0.0271 to 0.108 m³/s (390 to 1570 gpm), corresponding to jet velocities from 1.5 to 6.1 m/s (5 to 20 ft/s). Two different suction locations were considered, one central (under the jets), the other on the end walls, at a different distance above the tank bottom centerline at each end. The following conclusions were drawn:

- The rise velocity of the sludge/supernatant interface was of the same order of magnitude as the settling velocity, showing that settling could have a significant effect in the 378 m³ (100,000 gallon) tank.
- For the configuration in which the suction was located underneath the jet nozzles, at the tank center, the minimum jet velocities for mobilization were found to be between 3.0 and 4.6 m/s (10 and 15 ft/s) for a 10-cm (4-in) diameter jet, and between 4.6 and 6.1 m/s (15 and 20 ft/s) for a 7.6-cm (3-in) diameter jet. For the configuration in which the suction was located at the tank ends, the minimum jet velocities for mobilization were found to be between 4.6 and 6.1 m/s (15 and 20 ft/s) for a 10-cm jet, and between 6.1 and 7.6 m/s (20 and 25 ft/s) for a 7.6-cm jet. Clearly suction location makes a difference to sludge mobilization, though not to pure supernatant mixing.
- Even at the lowest jet velocities, no sludge banks were observed for either the center-suction or end-suction locations. In the 1/6 scale case, sludge banks had formed at one end of the tank because the location of the suction above the jets caused the jet flow to loop back on itself before it reached the tank end. The center-suction and end-suction configurations precluded this "shorted-out" flow pattern.
- Decreasing the sludge depth by 25% (from 0.61 to 0.46 m, or from 6.1% to 4.6% of the tank volume) caused a slight decrease in the minimum velocity required for sludge mobilization.
- Doubling the sludge solids density from 2.0 to 4.4 g/cc required a 30% to 50% increase in the minimum velocity for mobilization.
- At the minimum velocity required for mobilization, the mixing times were about 80 minutes for the central-suction case, 6 to 8 times longer than the supernatant mixing times for the same jet and suction parameters. Mixing times were not determined for the end-suction case, but were expected to exceed 100 minutes.
- The differing heights of the suction points on the end walls (0.7 and 1.2 m, or 2.4 and 4 ft) did not affect the velocity required for mobilization. The height difference did cause the flow rate in the recirculation lines to oscillate by more than 50% during the first 10 minutes. This happened because sludge flowed into one suction line sooner than into the other, and the two suction lines were joined upstream of the nozzles. For a fixed jet flow rate to be maintained, lowered flow in one line had to be offset by higher flow in the other.

- It might be possible to reduce the mixing time by starting the pump at a relatively low flow rate and, once the sludge dilution process is well underway, increasing the flow rate to impel the sludge to the top of the tank.

2.2 Recommendations

Several recommendations are offered as a result of this work.

- If additional mixing analysis is to be conducted, alternate rheological descriptions for sludges (Appendix A) should be investigated more fully. This would establish the extent to which the low-shear-rate inaccuracy of the power-law rheological model (used in the modeling to date) might cause an over-estimation of the required velocity for mobilization and of the mixing time.
- A more extensive set of mixing time calculations is needed to form a complete idea of how sludge mixing time varies with jet and suction parameters and with the extent of filling of the tank.
- A comparison of the TEMPEST model predictions to experimental data should be performed when experimental data become available, especially for sludge mobilization.
- Further sludge modeling could be used to explore whether varying the jet velocity during operation would reduce the required mixing time.

3.0 APPROACH

The present investigations are of supernatant mixing times and sludge layer mobilization in 378 m³ (100,000 gallon), horizontal-cylinder, waste storage tanks at ORNL. These tanks are part of the MVST-CIP. In general, the CIP design is a larger version of the 189 m³ (50,000 gallon) MVST design. Preliminary computer modeling analyses for 1/6 and 2/3 scale experimental configurations of MVST have been reported (Terrones and Eyler 1993). The present investigations continue this work. The two design questions -- the time required to achieve full supernatant mixing and the feasibility of sludge mobilization -- were addressed with somewhat different approaches.

The mixing time in a supernatant-filled tank was determined as in previous work (op. cit.). Tank configuration models were developed that included recirculation suction lines and nozzles for the jets. To simplify computations, the tank was treated as having two symmetry planes. One vertical symmetry plane lay along the length of the tank, giving the modeling domain a semi-circular cross-section. The other symmetry plane was also vertical, located halfway between the ends of the tank. After the jet velocity was initialized and the tracer injected, the tracer concentrations were monitored at several locations throughout the tank. The maximum concentration difference between monitors was used to calculate the extent of mixing and to find the mixing times. The mixing times were correlated as functions of jet diameter and flow rate.

The feasibility of sludge mobilization could not be assessed so quantitatively. In the previous 1/6 scale simulations (op. cit.), the jets had been located 1/4 of the way along the long axis and recirculation suction located in clear supernatant above the jets. These simulations found that low jet velocities allowed the existence of sludge banks at one end of the tank floor. No such sludge banks were evident with the two 378 m³ (100,000 gal) tank suction configurations that were studied. Thus, a different set of criteria for judging feasibility had to be found.

Two criteria were found. First, the rise velocity of the interface between the gradually-diluted sludge and the supernatant was compared to the hindered settling velocity of the sludge particles. This value was derived from the 2.9 cm/min maximum settling velocity observed by Ceo et al. (1990). When the interface rise velocity remained greater than the settling velocity all the way to the top of the tank, the jet velocity was considered adequate to assure mobilization. Second, the upward fluid velocity at the end of the tank floor was checked to confirm that the jet had reached the far wall with enough momentum to maintain mixing.

In modeling sludge mobilization for the end-suction configuration, computer time constraints made it impossible to continue a large number of runs at different velocities to their conclusion. Therefore, a simulation was begun at a low velocity, equal to the velocity required for sludge mobilization for the (less demanding) center-suction configuration. At 50 to 60 minutes into the model run, the jet velocity was suddenly increased by 1.5 or 3.0 m/s (5 or 10 ft/s), and this higher velocity was maintained until the end of the run. If the velocity jump proved to provide an upward fluid velocity that was large enough to prevent settling at the top of the end of the tank, then that velocity was considered to be high enough to ensure sludge mobilization. Though this velocity-change approach saved computer time and allowed more velocities to be tested for mobilization, it made it impossible to determine mixing times.

4.0 DETERMINING JET DIAMETERS

The objective of Task 1 of the present work was to make a preliminary investigation to determine two plausible jet diameters for sludge mobilization. All subsequent simulations for the 100,000 gallon tanks used these diameters. The choice was based on literature correlations for jet mixing in tanks. Preliminary TEMPEST modeling in a simplified geometry was also conducted to check the consistency of the literature correlations for vertical cylindrical geometry with mixing behavior in a horizontal cylindrical tank. This section gives the basis for choosing 3-inch and 4-inch jet diameters for further study.

4.1 Correlations from the Literature

The correlations of mixing time (t_m) found in the literature were all obtained from tank configurations with a vertical axis and nozzles in a variety of locations and orientations, and were based on a qualitative estimate of the degree of mixing. Some of the correlations from past work, which were summarized in Maruyama (1986), Rice (1986), and Tattersson (1991) have included

Fossett and Prosser:

$$t_m = \frac{8D^2}{\sqrt{qu}} \quad (4.1)$$

Van de Vusse:

$$t_m = \frac{8.7D^2 \sin \theta}{du} \quad (4.2)$$

Okita and Oyama:

$$t_m = 5.5 \frac{d}{u} \left(\frac{D}{d} \right)^{1.5} \left(\frac{H}{d} \right)^{0.5} \quad (4.3)$$

Coldrey:

$$t_m = \frac{3.54D^2Hd}{Lq} \quad (4.4)$$

where D = tank diameter (in a vertical cylindrical tank)
 H = liquid depth
 d = jet nozzle diameter
 L = jet axis length (distance along the jet axis from the nozzle to a wall)
 q = recirculation volumetric flow rate (or jet flow rate at the nozzle)
 u = jet velocity at the nozzle
 t_m = mixing time
 θ = jet angle from the horizontal

All of these correlations can be expressed in a single form, as

$$t_m = C t_R \frac{d}{L} \quad (4.5)$$

where C = a constant varying from 2.6 to 7 depending on nozzle angle and location; for a near-floor jet, C is about twice the minimum value, which is seen for a jet at $H/2$ height
 t_R = recirculation time constant, equal to (tank volume / recirculation flow rate)

Equation 4.5 is reliable only for a jet Reynolds number (Re) greater than about 30,000 (Maruyama 1986). In this "circulation" regime, the jet is turbulent and its entrainment rate is independent of Re . Below $Re = 10,000$, where the dimensionless mixing time $[(t_m/t_R) (L/d)]$ reaches a minimum, Maruyama (1986) shows that the mixing time increases strongly with decreasing Re . In this laminar regime the mixing time is proportional to $Re^{-1.1}$, according to Maruyama's Figure 4 (op. cit.). The minimum dimensionless mixing time at $Re = 10,000$ is about half the constant value that is reached above $Re = 30,000$. Only one data set included an estimate of the extent of mixing; in this, the mixing time criterion was defined at the time at which 1% departure from the final mixed concentration was reached, with the measurement locations(s) unstated (ibid.).

The jet diameter should be chosen to assure that, at the lowest design flow rate, the jet Re will remain above the critical value of 30,000, in a range where the dimensionless mixing time is a predictable constant value. Then the Re can be calculated as

$$Re = \frac{du}{\nu} = \frac{4q}{\pi d \nu} \quad (4.6)$$

where ν = kinematic viscosity, = (2.2 cP / 1.285 g/cc) = 1.71×10^{-6} m²/s for W-28 supernatant, based on Ceo et al. (1990)

The design flow rate is 0.0126 m³/s (200 gpm), total for two opposed jets. For this flow rate, the maximum nozzle diameter for which the Re is above the critical value is 15 cm (6 in). Smaller nozzle diameters of 7.6 and 10 cm (3 and 4 in) were selected to increase Re further into the turbulent range and to further decrease mixing time, according to Equation (4.5).

Scaled sludge mobilization experiments have been conducted^(b), using simulants of Hanford waste in a 1/12-scale model of a vertical Hanford tank. The conclusion was that the product of jet velocity and diameter had to be at least 2.7 m²/s (29.4 ft²/s) to mobilize sludge to a distance of 11m (37.5 ft). For a 7.6-cm nozzle, this criterion corresponds to velocities on the order of 37 m/s. Unfortunately, it is very difficult to compare the Hanford simulants, which included very cohesive materials having a yield stress of 1000 Pa, with W-28 sludge, which has no observed yield stress and exhibits a strain rate of about 100 sec⁻¹ at only 10 Pa shear stress.

(b) Fow, C.L., Scott, P.A., Whyatt, G.A., and C.M. Ruecker. 1987. *Pilot-Scale Retrieval Tests Using Simulated NCAW*. TW21-87-15, Pacific Northwest Laboratory, Richland, Washington. Technical Report for Westinghouse Hanford Company.

In other work, Chang and Beaver (1993) conducted scaled experimental work supporting implementation of a mixing pump in Hanford's 241-SY-101 tank. Jet nozzles of 7.1 cm (2.8 in) were ultimately used in 241-SY-101, and indications are that mixing has been sufficient to obviate that tank's problem. Jet velocities of about 15 to 18 m/s (50 to 60 ft/s) were used, giving diameter-velocity product of 1.1 to 1.3 m²/s (12 to 14 ft²/s). Recirculation suction was located directly above the jets, at about 2/3 of the tank depth above the bottom. The initial thickness of the sludge layer in SY-101 was about half the total depth. The SY-101 settled sludge had a yield stress on the order of 1000 Pa, thus was much more cohesive than W-28 sludge.

From these sludge mobilization studies, it was concluded that the 3-inch and 4-inch jet sizes would provide jets whose flow was adequate to mix the tank contents and mobilize the W-28 sludge.

4.2 Preliminary Modeling

After the jet diameter determination was made, a 378 m³ (100,000 gal) tank configuration with estimated nozzle size and location was used as a preliminary model for mixing. The tank was a horizontal cylinder 5.1 m (16.7 ft in) in diameter and 20.2 m (66.2 ft) long, filled with supernatant. Two opposed jets were located in the tank center-plane, halfway between the ends, two nozzle-diameters above the floor. The suction was located under the jets, one diameter above the floor. (Based on symmetries, the model included a half-circle of the tank cross-section, half of its length, and one of the two jets.) The modeled tank volume was about 387 m³ (102,000 gal), a sufficiently close modeling approximation to the 378 m³ (100,000 gal) design volume. The liquid properties were taken as being the same as those measured for the supernatant (pure liquid) in the W-28 tank, a density of 1285 kg/m³ and a dynamic viscosity of 2.2 Pa s (Ceo et al. 1990).

Several simulations were conducted with 7.6 cm (3-in) diameter nozzles at various jet velocities. With all other parameters fixed, the mixing time was found to be inversely proportional to the jet velocity, a result that was consistent with Equation (4.5). Mixing times for 0.1% departure from the final mixed concentration at all monitored tank locations were approximately correlated as

$$t_m = 5.66 / u \quad (4.7)$$

where u is in ft/s and t_m in hr.

Equation (4.5) was evaluated for a jet length L of 10.0 m (half a tank length), a volume of 194 m³ (half a tank volume per jet), and compared to Equation (4.7), using consistent units. The constant C in Equation (4.5) was found to be about 20 for the modeled horizontal tank. This was well outside the range of values for C (2.6 to 7) that had been found for vertical-cylinder tanks. The increase in the constant may have resulted from a more restrictive criterion for the extent of mixing (0.1% rather than 1%) or from the differences in mixing due to the horizontal configuration. Furthermore, it suggests that the horizontal-cylinder experiments in progress at ORNL will be of value in providing a basis for a validation of the computed mixing times^(c).

(c) While some of these data have been measured in 1/6 and 2/3 scaled MVST experiments, they have not yet been published and made available. Personal communication with J. Perona.

5.0 SUPERNATANT MIXING

The objective of Tasks 2 and 3 of the present work was to determine mixing times for 378 m³ (100,000 gallon) tanks filled with supernatant alone. In Task 2, the suction inlet was centrally located directly under the jets. In Task 3 the suction inlet was placed at the ends of the tanks at various distances above the tank bottom centerline. In both tasks, two nozzle diameters, 7.6 and 10 cm (3 and 4 in), were used. Jet velocities were varied corresponding to a range of total flow rates from 0.0126 to 0.0492 m³/s (200 to 780 gpm).

In most calculations, a chemical addition was modeled as instantaneous. The effect of the duration of the addition was modeled for the conditions that predicted the shortest and longest mixing times.

The results of the simulations conducted in these tasks are correlations of mixing time in the supernatant-filled tank. These correlations are presented as a function of jet nozzle diameter and mixedness (percent deviation from the final concentration).

5.1 Modeling

The modeled 378 m³ (100,000 gal) tank was a horizontal cylinder 5.1 m (16.7 ft) in diameter and 20.2 m (66.2 ft) long, filled with supernatant. These dimensions were a modeling approximation of the design dimensions of the tank, 4.9 m (16 ft) OD and 20.8 m (68.4 ft) long. Two symmetry planes were used to simplify modeling. The jets were located in the center of the tank plan-view, with the jet centerline 0.22 m (8.6 in) above the tank bottom. The modeled tank volume was about 387 m³ (102,000 gal). A schematic of the tank configuration used in these computations is presented in Figure 5.1.

In the first suction configuration, the recirculation inlet was located beneath the jets, 8.9 cm (3.5 in) from the bottom of the tank. This central location resembled that used for the Task 1 preliminary modeling (Section 4.2). In the other suction configurations, the recirculation inlet was located at either 0.7 m (2.3 ft) above the tank bottom, 1.2 m (4 ft) above, or the top center of the tank end (5.1 m above the floor). Because the tank was modeled as end-to-end symmetric, one inlet was modeled at the same height at each end of the complete tank.

The liquid properties were those measured for the supernatant (pure liquid) in the W-28 tank, a density of 1285 kg/m³ and a dynamic viscosity of 2.2 mPa s (Ceo et al. 1990). To permit turbulent but not laminar diffusion as modeled by the TEMPEST code, the turbulent Schmidt number was set to 0.71 and the molecular Schmidt number to 10⁹. This choice reflected the relatively low viscosity of the pure supernatant and the expected importance of turbulent mixing.

5.2 Results

The mixing times in hours were estimated by using Criterion 1 (as defined by Terrones and Eyler 1993) with eight concentration monitor cells, whose locations are shown in Figure 5.2. Criterion 1 describes the "extent of mixing" as the maximum concentration difference between any two monitored locations, divided by the final well-mixed concentration. The mixing time results are presented in Figures 5.3 and 5.4 for 7.6-cm (3-in) and 10-cm (4-in) diameter jets respectively. In each figure, two different sets of mixing times are shown, one for a 5% extent of mixing and

one for more complete mixing of 0.1%. Also included in each figure are results for the various recirculation locations as identified in the legends.

The two figures show that the mixing times at 0.1% and 5% mixing vary with flow rate in very similar ways. The effect of varying the extent of mixing is about the same at all flow rates: a mixing time increase of roughly 45% is required to go from 5% mixing to 0.1% mixing.

For the end-suction configuration, the effect of the suction height variation is small. At short mixing times (high flow rates), the mixing-time difference between locating the suction at 1.2-m (4-ft) height and the top of the tank (about 5.1 m) is negligible (Figure 5.3). At long mixing times (low flow rates), varying the suction height from 0.7 m (2.3 ft) to the top of the tank causes less than 6% change in mixing time (Figure 5.4). The mixing time consistently increases as the suction level is lowered.

In Figure 5.3, where the jet diameter was 7.6 cm (3 in), the mixing times are clearly shown to be about 20% longer for center-suction than for end-suction, given the same flow rate. Figure 5.4, for the 10-cm (4-in) jet diameter, shows some irregularities in the relationship of mixing time to flow rate. As can be seen, the modeling results show mixing times for center-suction that can be either longer or shorter than the time required for end-suction at the same flow rate and extent of mixing. The irregular mixing-time behavior may be due to jet interaction with the bottom of the tank. The jet is slightly closer to the tank bottom, in terms of the number of jet diameters, for the 10-cm jet than for the 7.6-cm jet. To provide an upper-bound estimate of mixing time for the center-suction case, only the mixing times that were higher than those for end-suction were used in the correlation (see Section 5.3). This is why more data points are shown in Figures 5.3 and 5.4 than in Figures 5.5 and 5.6, which depict the correlations.

All the data plotted in Figures 5.3 and 5.4 are for instantaneous initial addition of a non-reacting chemical (in effect, a tracer) at the top center of the tank. The effect of adding the tracer at the same location over a 10-minute period was also modeled for two end-suction conditions: a 7.6-cm (3-in) jet with a flow rate of 0.0416 m³/s (660 gpm), and a 10-cm (4-in) jet with a flow rate of 0.0126 m³/s (200 gpm). These were the two conditions leading to shortest and longest mixing times. At 7.6 cm and 0.0416 m³/s, the modeled 5%-extent mixing time increases from 0.253 hr to 0.306 hr when the tracer addition is prolonged. At 10 cm and 0.0126 m³/s, the modeled 5%-extent mixing time increases from 1.217 hr to 1.229 hr when the tracer addition is prolonged. Interestingly, the overall mixing time is almost exactly equal to the root-mean-square of the injection time and the instant-addition mixing time:

$$\begin{aligned} 0.303 \text{ hr} &= \sqrt{(0.253 \text{ hr mixing})^2 + (0.167 \text{ hr injection})^2} \\ 1.228 \text{ hr} &= \sqrt{(1.217 \text{ hr mixing})^2 + (0.167 \text{ hr injection})^2} \end{aligned} \tag{5.1}$$

The values calculated by r-m-s in Equation (5.1) almost equal those estimated by TEMPEST (0.306 and 1.229 hr). It is not clear whether there is a theoretical basis for this correspondence.

5.3 Correlations

The mixing time data presented in Figures 5.3 and 5.4 were correlated in a form similar to that given in Equation 4.5. It proved to be possible to include the extent of mixing as part of the correlation, though there is no theoretical basis for this functional form of the extent of mixing. Equations 5.3 and 5.4 are the correlations that were derived for the data. Figures 5.5 and 5.6 are plots of these correlations and the 5%-extent mixing time data; the mixing times are plotted versus flow rate in Figure 5.5 and versus jet velocity in Figure 5.6.

Correlation for center suction:

$$(t_m/t_R) = d^{0.91} [0.0170 + 0.001197 (\log_{10} p)^2] \quad (5.2)$$

maximum overestimation 5.1%
maximum underestimation 2%

Correlation for end suction:

$$(t_m/t_R) = d^{1.07} [0.01066 + 0.0008494 (\log_{10} p)^2] \quad (5.3)$$

maximum overestimation 11.6%
maximum underestimation 1%

where $t_R = V/q$ (5.4)

- t_m = mixing time
- t_R = residence time
- d = jet diameter, cm
- V = total tank fluid volume, defined as 378 m³ (100,000 gal)
- q = recirculation volumetric flow rate, m³/time
- p = extent of mixing, or the maximum simultaneously monitored concentration difference as a fraction (not a percentage) of the final fully-mixed concentration

These correlations are known to be accurate only for the 378 m³ (100,000 gallon) tank dimensions and geometry, within the parameter ranges of nozzle diameter 7.6 to 10 cm (3 to 4 in), recirculation flow 0.0126 to 0.0492 m³/s (200 to 780 gpm), and extent of mixing 5% to 0.1%.

Figure 5.1. Schematic of tank configuration used in computational models for Task 2 (top) and Task 3 (bottom), supernatant mixing.

Figure 5.2. Schematic showing locations of the monitors used to determine the extent of mixing in the supernatant-filled tank.

Figure 5.3. Supernatant mixing times for two opposed 7.6-cm jets in the 378 m^3 (100,000 gallon) tank, for various suction locations.

Figure 5.4. Supernatant mixing times for two opposed 10-cm jets in the 378 m^3 (100,000 gallon) tank, for various suction locations.

Figure 5.5. Correlated and modeled supernatant mixing times as a function of flow rate.

Figure 5.6. Correlated and modeled supernatant mixing times as a function of jet velocity.

6.0 SLUDGE MOBILIZATION

The objective of Task 4 was to determine the minimum jet velocity to mobilize W-28 sludge for each of two jet diameters, 3-inch and 4-inch. Two opposed jets were centrally located between the tank ends and near the tank floor. The axis of the jets was aligned with the horizontal tank axis. Two recirculation suction configurations were considered, one with the recirculation inlet centrally located just below the jets, and one with a recirculation inlet at each end of the tank. In the latter case, each of the recirculation inlets was at a different height above the tank floor. A limited number of calculations were done to find the effects of sludge solids density (2.0 and 4.4 g/cc) and sludge depth (0.61 and 0.46 m, or 2 and 1.5 ft).

6.1 Modeling

The modeled 378 m³ (100,000 gal) tank was a horizontal cylinder 5.1 m (16.7 ft) in diameter and 20.2 m (66.2 ft) long, filled with supernatant, as in the modeling described in Sections 4 and 5. The jets were located in the center of the tank plan-view with centerline 24 cm (9.4 in) above the tank bottom. The modeled tank volume was about 387 m³ (102,000 gal). These dimensions represent the 378 m³ (100,000 gal) tank within a few percent, adequate for modeling purposes. Figure 6.1 shows the two suction configurations as modeled. The sludge mobilization modeling included a non-viscous tracer species (initially placed at the top center of the tank) as well as the non-Newtonian sludge species.

In the first of the two suction configurations, the recirculation suction was located beneath the jets, 6.1 cm (2.4 in) from the bottom of the tank. This central location resembled that used for the supernatant mixing modeling, and it was possible to simplify the model by using two symmetry planes.

In the other suction configuration, the modeling began by locating the suction at 0.7 m (2.4 ft) above the tank bottom at one end and at 1.2 m (4 ft) above the bottom at the other end. The suction diameter was 0.17 m (6.8 in). Only one symmetry plane (the axial one) could be used to model this configuration, since the two ends were distinguished by their different suction heights. As the modeling proceeded, it became clear that the difference in suction heights affected the distribution of flow between the two suction points only during the first 10 to 20 minutes. At later times, the concentrations and viscosities in the suction lines were close enough to equal to allow the suction flows to equalize. Therefore, the mobilization and mixing at later times were not changed significantly by different suction heights. Accordingly, the end-suction model was changed to an end-symmetric suction model (as in previous work) with the suction located at 4 ft above the bottom. This reduced the computer time required per run.

In the base case runs, the sludge depth was set at 0.61 m (24 in) and the final well-mixed sludge concentration in the tank was 1.0 vol%. A set of sensitivity tests checked the effect of a smaller sludge depth of 0.46 m (18 in).

Sludge and liquid properties were taken from Terrones and Eyler (1993) and Ceo et al. (1990). The liquid properties were those measured for the supernatant (pure liquid) in the W-28 tank, a density of 1.285 g/cc and a dynamic viscosity of 2.2 mPa s. The sludge density was 2.0 g/cc.

Particle settling was not modeled in the central-suction TEMPEST runs, in accord with the insignificant settling effects shown by results of the 1/6 scale modeling (Terrones and Eyler 1993). However, the slow rise velocity of the interface was comparable in magnitude to the sludge settling velocity (2.9 cm/min). Therefore, it was possible that particle settling could offset mixing and prevent mobilization. To make sure that settling was accounted for, the predicted interface rise rates were checked against the settling velocity. Only those cases where the rise velocity exceeded the settling velocity were considered to ensure sludge mobilization.

After the central-suction runs indicated that settling could have significant effects for the 100,000 gallon tank, settling was directly included in the modeling of the end-suction cases. In those cases the following Stokes-regime hindered settling model (Perry and Chilton 1973, p. 5-64) was used to represent the decreased settling velocity at high concentration:

$$V_s = V_{s0}(1 - C_r)^{4.65} \quad (6.1)$$

where V_s = hindered settling velocity
 V_{s0} = low-concentration settling velocity (2.9 cm/min, according to Ceo et al. [1990] sedimentation data)
 C_r = reduced concentration = $C_v / C_{v,max}$
 C_v = volume fraction solids
 $C_{v,max}$ = maximum packed volume fraction (0.154 for W-28 sludge)

The sludge rheology was modeled as that of a concentration-dependent power-law fluid with the following constitutive equation:

$$\tau = m(C_r)\dot{\gamma}^{n(C_r)} \quad (6.2)$$

$$m(C_r) = 0.0022 + 3.0513C_r + 17.631C_r^2 - 10.684C_r^3 \quad (6.3)$$

$$n(C_r) = 12510.6^{-C_r} + (1 - 1.55967^{-C_r}) \quad (6.4)$$

where τ = fluid shear stress (Pa)
 $\dot{\gamma}$ = shear strain rate (sec⁻¹)
 m = consistency factor (Pa s) as a function of reduced concentration
 n = flow behavior index as a function of reduced concentration

The power-law model of W-28 sludge overestimates the shear stress required to produce low strain rates (below 1 sec⁻¹), as shown by Figure 4.1 of Terrones and Eyler (1993). Therefore this rheological model is expected to provide conservative predictions of the jet velocity required for mobilization.

The value of $C_{v,max}$ in the current base-case modeling, 0.154, has been changed from the value of 0.68 that was used in Terrones and Eyler (1993). (The previous higher value included both dissolved and undissolved solids.) However, this previous value of 0.68 was used for the maximum packing in several of the current runs, and the results were compared with those obtained using the base-case value of 0.154. The $C_{v,max} = 0.68$ runs had the same variation of viscosity with reduced concentration as the runs with $C_{v,max} = 0.154$, but gave a higher sludge density for any given viscosity. Because the concentration of the initial sludge layer was always specified at the maximum packing factor (whichever one was used), the results for a $C_{v,max}$ of

0.68 and a sludge solids density of 2.0 g/cc were the same as if a $C_{v,max}$ of 0.154 had been used with a higher solids density of 4.4 g/cc. Thus, the high- $C_{v,max}$ runs served as sensitivity cases to check the effect of increased sludge density on mobilization.

As in the earlier modeling (Terrones and Eyler, 1993, Section 5.2.1), it was necessary to specify Schmidt numbers in such a way as to avoid generating unrealistically high molecular and turbulent diffusivities. In TEMPEST, the total diffusivity of a species is calculated as the sum of molecular, laminar, and turbulent contributions:

$$D_{tot} = D_p + \frac{\eta}{\rho Sc_L} + \frac{\eta_{eff}}{\rho Sc_T} \quad (6.5)$$

where	$D_{tot} =$	total species diffusivity
	$D_p =$	particle diffusion coefficient (the only non-zero term under stagnant conditions)
	$\eta =$	dynamic viscosity of the fluid
	$\rho =$	density of the fluid
	$\eta_{eff} =$	turbulent effective viscosity
	$Sc_L =$	laminar Schmidt number (the standard value is 0.71)
	$Sc_T =$	turbulent Schmidt number (the standard value is 0.71)

The high sludge viscosity η leads directly to a large laminar diffusivity (second term) and indirectly to a large turbulent diffusivity (third term). These unrealistically high terms were therefore effectively removed from the sludge model by setting the Schmidt numbers to 10^9 . On the other hand, the resulting very low diffusivities were also unrealistic because they would underestimate the mixing and overestimate the mixing time. Therefore, as a compromise the input representing the particle diffusivity (the only remaining contribution to total diffusivity) was set equal to the value of the laminar diffusivity term for the supernatant:

$$D_p = \frac{0.0022 \text{ Pa s}}{(1285 \text{ kg/m}^3)(0.71)} = 2.4 \times 10^{-6} \text{ m}^2/\text{s} \quad (6.6)$$

For a particle of 26 μm diameter and 2 g/cc density, the particle diffusion coefficient in water would be expected to be about $2 \times 10^{-14} \text{ m}^2/\text{s}$ (CRC, 1976, p. F-270). The particle diffusion coefficient pertains to stagnant conditions and requires some modification to include the additional diffusivity that results from a local laminar velocity gradient (Equation 6.6). The diffusivity used for sludge particles in TEMPEST is therefore greater than the relatively negligible particle diffusion.

6.2 Results

For the configuration in which the suction was located underneath the jet nozzles, at the tank center, the minimum jet velocities for mobilization were found to be between 4.6 and 6.1 m/s (15 and 20 ft/s) for a 7.6-cm (3-in) diameter jet, and between 3.0 and 4.6 m/s (10 and 15 ft/s) for a 10-cm (4-in) diameter jet. For the configuration in which the suction was located at the tank ends, the minimum jet velocities for mobilization were found to be between 7.6 and 9.1 m/s (25 and 30

ft/s) for a 7.6-cm diameter jet, and between 4.6 and 6.1 m/s (15 and 20 ft/s) for a 10-cm diameter jet. These ranges were based on computations which indicate that there is a good probability the lower velocity values might not fully mobilize and mix the sludge. The results for the higher velocity values in the range indicated complete mobilization and mixing.

Figures 6.2 and 6.3 present the modeling results for the center-suction configuration and show the height of the sludge-supernatant interface as a function of time. The interface height was defined as the height above the bottom of the tank at which the final fully-mixed sludge concentration value occurs. This height was measured at the tank end, and was an average across the width of the tank. In performing the analysis leading to these results, it was concluded that the sludge mobilization criterion that was useful in the 1/6 scale modeling, the minimum velocity at which "sludge banks" were absent, was not applicable for the 378 m³ (100,000 gal) tank. No sludge banks (regions of undiluted sludge) were observed in the 378 m³ (100,000 gal) tank modeling at any velocity. Lacking the sludge-bank criterion, the rule applied was that for mobilization to be certain, the rate of rise of the sludge-supernatant interface had to exceed the W-28 sludge particle settling velocity of 2.9 cm/min.

Thus in Figures 6.2 and 6.3 it can be observed that for the 7.6-cm (3-in) jets operating at 6.1 m/s (20 ft/s) and for the 10-cm (4-in) jets operating at 4.6 m/s (15 ft/s), the sludge-supernatant interface has reached the top of the tank after 80 min, indicating complete mobilization and mixing. Furthermore, even at the top of the tank the rate of rise of the interface (the slope of the lines) remained greater than the settling velocity. However, for the 7.6-cm jets operating at 4.6 m/s (15 ft/s) and for the 10-cm jets operating at 3.0 m/s (10 ft/s), the rate of rise of the interface height has dropped below the settling velocity of 2.9 cm/min. In the figures, the settling velocity is indicated by the diagonal straight line with the slope of 2.9 cm/min. The particle settling rate was not explicitly included in the computations of the center-suction scenarios, and therefore is included in this manner, as a criterion.

Sludge-supernatant interface heights are shown in Figures 6.4 and 6.5 for the end-suction configuration. Here again, the interface height was measured at the tank end, half a tank radius out from the middle of the end wall. The rapid increases in interface height at the tank end resulted from the higher-velocity jet reaching and flowing up the end wall. For these calculations, the particle settling velocity was explicitly included in the TEMPEST modeling, and so could not be a mobilization criterion. Hence, the criterion was based on the ability of the convecting fluid to maintain a flow that could carry the particles to the top of the tank.

6.2.1 Details of Central-Suction Results

Even at the lowest jet velocities examined (1.5 m/s, or 5 ft/s), it appeared that there would be no remaining "sludge banks" of the type seen in the 1/6 scale runs. The sludge concentration was decreasing perceptibly with time in all locations in the tank. This suggested that complete mobilization (though perhaps not complete mixing) could occur, given enough time, and that sludge banks could not be used as a definitive criterion of mobilization. The 1/6 scale runs had used a suction location in the tank center above the jets; this location may have allowed more flow to go around the sludge than the bottom suction location, contributing to sludge bank retention.

In all the central-suction runs it was observed that the initial horizontal interface between the increasingly dilute sludge and the essentially clear supernatant remained fairly distinct until mixing was nearly complete. This interface moved up the tank at a rate that decreased with time. The height of the interface could be tracked by plotting it as a concentration contour line set at the final completely-mixed concentration. This interface rise velocity was of the same order of magnitude

as the settling velocity (2.9 cm/min for W-28 sludge), suggesting that settling could significantly counteract sludge dilution and mixing in this large-scale tank.

Therefore, a comparison of the interface upward velocity with the settling velocity was used as the criterion for mobilization in the central-suction configuration. Figures 6.6 and 6.7 plot the movement of the interface heights (at the tank end) for the base-case and sensitivity runs with 7.6 and 10 cm (3 and 4-in) diameter jets. Only the velocities for which the interface remains above the dotted diagonal, which represents the settling velocity, provide enough momentum to make up for the effect of sludge particle settling. Using this criterion, the minimum jet velocities for mobilization were found to be between 3.0 and 4.6 m/s (10 and 15 ft/s) for a 10-cm diameter jet, and between 4.6 and 6.1 m/s (15 and 20 ft/s) for a 7.6-cm diameter jet. The upper-end velocity values consistently show an interface velocity (slope of the lines in the figures) greater than the settling velocity (slope of the dotted line), and so would definitely cause mobilization. The lower-end velocities might eventually effect mobilization but are not sure to do so.

Although the runs performed with an 0.46-m (1.5-ft) sludge depth were short, Figures 6.6 and 6.7 appear to show that after an initial period of accelerated interface rise, mixing of the sludge layer behaves in the same way as for the 0.61-m (2-ft) initial sludge depth. Probably, once the sludge dilution and the interface height reach the same point for both low-depth and base-case-depth cases, the mixing proceeds in the same manner no matter what the initial sludge depth. Thus, much of the mobilization process would be expected to proceed similarly for the two sludge depths. However, the lower final sludge concentration and viscosity would be expected to lead to a lower total mixing time for the smaller sludge depth.

The sludge density exerted a continuing effect on mixing. As discussed in Section 6.1, the $C_{v,max} = 0.68$ cases plotted in Figures 6.6 and 6.7 show the effects equivalent to those of a doubling of sludge solids density. Judging by the figures, an increase in jet velocity of roughly 1.5 m/s (5 ft/s), or an increase of 30 to 50 percent, is enough to make up for this increase in density. This suggests that the governing factor is not momentum but the increase in potential energy (PE) of the rising sludge particles; the doubled PE must be supplied by increasing the jet kinetic energy (KE), the only energy source in the tank, by a factor of two. The velocity must increase by 41 percent to double the KE, and this is approximately consistent with the simulation results.

The mixing times for sludge mixing were considerably longer than those for pure supernatant. Using a 5% sludge concentration difference criterion, the mixing times for the minimum-velocity mobilized cases were found to be 80 min (for 4.6 m/s and 10-cm diameter) and 81 min (for 6.1 m/s and 7.6-cm diameter). Based on the Task 2 results (Section 5.3), the center-suction mixing times predicted for pure supernatant at these jet conditions would have been 13 min and 9.1 min. The high non-Newtonian viscosity and the suppression of turbulence by the sludge density gradient apparently increase the mixing time by a factor of 6 to 8, compared to pure Newtonian supernatant. The final $C_v/C_{v,max}$ is $(0.010 / 0.154) = 0.065$ for all of the base cases. At this reduced concentration, the power-law description of the viscosity is

$$\mu = 0.272\gamma^{0.43} \quad (\text{Terrones and Eyler 1993}) \quad (6.7)$$

where μ = dynamic viscosity (Pa s)
 γ = strain rate (1/s)

Then the final viscosity is as low as 2.2 mPa s (the supernatant viscosity) only at the very high strain rates near the jet nozzle. Possibly these much lower, shear-thinned viscosities near the jet play a large part in determining the mixing time, since the ratio of non-Newtonian to supernatant mixing times is less than a factor of 10.

The presence of sludge also caused a change in the functional dependence of mixing time on jet velocity and diameter. This is shown by the fact that the mixing times are nearly equal for the non-Newtonian simulations, but differ by 50 percent for the estimates based on the Newtonian supernatant. The functional dependence of mixing time for the non-Newtonian sludge was not established, because of computer time limitations. (However, it is interesting to note that the mixing times were the same for the two fully mobilized center-suction cases, for which the product of jet velocity and diameter were also the same. It is possible that the criterion for mobilization could be expressed as the simple requirement that the jet velocity-diameter product equal or exceed 0.46 m²/s, or 5 ft²/s. For comparison, the required value was 1 to 3 m²/s for double-shelled tank waste at Hanford.)

The mixing time would be expected to increase if particle settling were included in the TEMPEST modeling, but no runs were performed to quantify this effect for the center-suction configuration. Figures 6.6 and 6.7 show that the excess interface rise velocity, in the mobilized center-suction cases, is enough greater than the settling velocity to ensure a mixing time (including settling velocity) that is less than three times the mixing time simulated without settling velocity.

The mixing times discussed so far were based on the concentration of the sludge. It was also possible to determine mixing times on the basis of the second species, the neutrally-buoyant tracer that was initially located in a small volume at the top center of the tank. The mixing times based on a 5% difference in tracer concentration were found to be 86 min (for 4.6 m/s and 10-cm diameter) and 87 min (for 6.1 m/s and 7.6-cm diameter). These are only 6 minutes longer than the mixing times that were based on sludge concentrations. Apparently, chemicals located in the stagnant upper supernatant layer will mix slightly more slowly than the sludge, which is directly subject to jet mixing.

6.2.2 Details of End-Suction Results

Initially, the flow rates through the two suction lines were quite different. Flow through the lower suction (0.7 m, or 2.4 ft height) was impeded by the high sludge concentration sucked into it. The flow through the higher suction (1.2 m, or 4 ft) accordingly increased, drew in sludge at a higher rate, and so eventually slowed down until the flow through the two lines equalized. Figures 6.8 and 6.9 show time-traces of the two suction-line velocities and sludge concentrations.

Figure 6.10, which depicts the central vertical plane along the length of the tank, shows that after the initial perturbation the concentration and velocity fields were very nearly identical at the two ends of the tank. This observation was the justification for modeling further runs as end-symmetric with the suction located at a height of 1.2 m (4 ft). The case shown is for a jet diameter of 7.6 cm (3 in) and velocity of 6.1 m/s (20 ft/s), and is typical.

Velocities that produced sludge mobilization for the center-suction case were not expected to be adequate to do so in the end-suction case. The reason for the difference is that the driving force behind sludge motion and mixing is the upward velocity at the end of the tank. In the case of the end-suction configuration, the tank-end upward velocity was reduced by the suction line draw-off, and the residual kinetic energy was therefore less than for the center-suction configuration. Figures 6.11 (for end-suction) and 6.12 (for center-suction) show how the vertical velocity at the

tank end varied with height above the bottom, for a typical case (jet diameter of 7.6 cm and velocity of 6.1 m/s). Clearly there was less kinetic energy in the end-suction case, Figure 6.11. As a result, a higher jet velocity was needed for end-suction to provide the same amount of sludge driving force as for center-suction.

As in the case of the central-suction configuration, no "sludge banks" of undiluted sludge were evident. Although higher particle concentrations were evident on the floor a few feet to the side of the jet axis, the concentrations were decreasing with time and were less than twice as high as the concentrations in the path of the jet.

Once again, the interface velocity at the end of the tank was checked to see whether or not complete mobilization and mixing could be expected to occur. Figures 6.4 and 6.5 show the interface velocities for the 7.6 and 10 cm (3 and 4-in) cases. In each set of runs, the initial run was an asymmetric simulation at the lowest jet velocity studied for the end-suction configuration. The asymmetric run was carried out to nearly an hour, at which time it became clear that the velocity was insufficient to cause mobilization.

At that point, the run was converted to a symmetric model and continued at three different velocities: the original low velocity, and velocities that were higher by 1.5 and 3.0 m/s (5 and 10 ft/s). These are the three "branches" in Figures 6.4 and 6.5. The upper branches, or higher-velocity data, show sudden jumps in interface height. These reflect the increased-velocity jet sweeping the sludge toward (and up) the end wall of the tank; they were not accompanied by an interface height increase over the whole length of the tank. The sludge upsweep can be seen clearly in Figures 6.13 and 6.16. The upper branches can be thought of as what would happen to the interface at the end of the tank if the pumping power were suddenly stepped up partway through a run.

Since the interface height at the tank end is not representative of the interface height throughout the tank for these "velocity step" cases, Figures 6.4 and 6.5 do not give enough information to completely judge mobilization. The apparent stoppage of the higher-velocity interfaces does not necessarily indicate that mobilization failed. However, the figures do show that the change from an asymmetric to a symmetric model has no apparent effect on the behavior of the interface. Given the same jet velocity before and after the symmetry change, the interface continued to rise at the same rate.

An increase in the jet velocity, however, imparts an impulse to the sludge that drives it up the end wall of the tank. Figures 6.13 (10 cm and 6.1 m/s) and 6.14 (10 cm and 7.6 m/s) show the effect over time of the velocity step on the concentration field, and the extent to which the impulse leads to sludge mobilization. Increasing the velocity from 4.6 to 6.1 m/s, for a 10-cm (4-in) jet diameter, did not appear to roll the sludge up to the top of the tank. However, the step to 7.6 m/s at a 10-cm diameter did send the sludge over the top, leading to substantial mixing by 86 min (Figure 6.14). In the same way, as shown in Figures 6.15 and 6.16, a velocity of 7.6 m/s did not appear to send the sludge over the top for a 7.6-cm (3-in) jet diameter, but a velocity of 9.1 m/s (30 ft/s) was adequate.

Data analysis indicated that for end suction the low-end velocities, 4.6 m/s (15 ft/s) for 10-cm (4-in) diameter and 6.1 m/s (20 ft/s) for 7.6-cm (3-in) diameter, might or might not produce sludge mobilization. At an interface height of about 3.1 m, the interface rise velocities for the low-end cases were low, about 1×10^{-4} m/s. At this interface level, about 2/3 of the tank volume contains sludge, and its concentration is about 1.46 vol%. By Equation (6.1), the hindered settling velocity at this concentration is 4.37×10^{-4} m/s, and the hindered settling velocity at the final fully-

mixed concentration -- when the interface reaches the tank top -- is 4.53×10^{-4} m/s. The interface velocity near the tank top would need to be greater than the difference in the settling velocities, or 2×10^{-5} m/s, to ensure full mobilization. Although at 3.1 m height the interface rise velocity was five times the settling velocity difference, it is not clear that the interface would still be moving fast enough at the tank top to resist settling, after "dilution" in the remaining tank volume.

The concentration field plots (Figures 6.13 through 6.16) help to visualize the effect of the velocity step, but the most useful clue to whether a jet velocity would mobilize the sludge came from the velocity step results. Figure 6.17 shows, for the four higher-velocity cases, vertical profiles of the vertical velocities imparted to the fluid at the end wall of the tank. It can be seen that for all except the 7.6-cm, 7.6-m/s (3-inch, 25-ft/s) case, there is an upward velocity all along the tank end-wall, and this upward velocity is much greater than the settling velocity. The failure of the 7.6-cm, 7.6-m/s jet to maintain an upward velocity all along the end wall suggests that its mobilization capacity is doubtful.

These velocity data make it clear that for the end-suction configuration mobilization is possible between 4.6 and 6.1 m/s (15 and 20 ft/s) for the 10-cm (4-inch) jet, and between 7.6 and 9.1 m/s (25 and 30 ft/s) for the 7.6-cm (3-inch) jet. As long as the far-wall velocity is much greater than the settling velocity, all the way to the top of the tank, and there are no sludge banks, mobilization must eventually occur.

The results suggest that a jet velocity-diameter product of $0.58 \text{ m}^2/\text{s}$ ($6.2 \text{ ft}^2/\text{s}$) is not quite adequate for sludge mobilization, but a value of $0.62 \text{ m}^2/\text{s}$ ($6.7 \text{ ft}^2/\text{s}$) is enough. The model results for the "velocity step" cases also suggest that it might be possible to reduce the mixing time by starting the pump at a relatively low flow rate and, once the sludge dilution process is well underway, turning up the flow rate and impelling the sludge to the top of the tank.

6.2.3 Discussion of Results

To summarize, both the central-suction and end-suction configurations of the 378 m^3 (100,000 gal) tank showed mobilization behavior that differed from that of the 1/6 scale tank (0.87 m^3 , or 230 gal volume). Below a certain threshold jet velocity (between 4.0 and 6.1 m/s), sludge banks were retained on the floor of the 1/6 scale tank; the corresponding minimum velocity for mobilization would be expected to be much higher than 6.1 m/s (20 ft/s) in the 378 m^3 (100,000 gal) tank, because of the greater tank volume. Nevertheless, even at low jet velocities no sludge banks were observed in the 378 m^3 (100,000 gal) tank model, and the minimum velocities required for mobilization were only 4.6 to 9.1 m/s (15 to 30 ft/s), depending on jet diameter and suction location.

The 1/6 scale tank model was different from the 378 m^3 (100,000 gal) tank model in several ways besides size: the tank was only half-filled with liquid, the jets were located not at the center of the tank but 1/4 of the way along the length, and the suction was located in clear supernatant several inches above the jets (which were 5.1 cm, or 2 in above the tank floor). These differences led to a significantly different flow field, and in turn to a much different mixing effectiveness.

Figure 6.18 allows a comparison between the well-developed velocity fields for the 1/6 scale model and the two 378 m^3 (100,000 gal) suction configurations. In the 1/6 scale case (the top plot in the figure), the 4.0 m/s (13 ft/s) flow leaves the jet nozzle, meets resistance from the high-concentration high-viscosity sludge at the tank end, and takes the path of least resistance by vertically looping back to the suction point. This "short-circuiting" is aided both by the location of

suction above the jets and by the free surface at the top of the fluid, which provides no frictional resistance to flow along the top. Therefore, a relatively high jet velocity is required to force the jet all the way to the wall before it turns back.

This jet short-circuit is not so readily available in the 378 m³ (100,000 gal) tank configurations, even in cases where the jet velocity is too low for mobilization. The result is more effective mobilization. In the center-suction case, where suction is located near the floor below the jets, the flow (3.0 m/s, or 10 ft/s) cannot reach the suction by looping upwards. Instead, it must loop back down in a nearly horizontal plane; its upward turn where it meets sludge resistance is limited by this requirement. The jet's relatively restricted vertical diversion can be seen in the small vertical component of the back-flow vectors in the middle plot in Figure 6.18. In the case of end-suction in the 100,000 gal tank, the 4.6 m/s (15-ft/s) jet goes directly to the end wall of the tank and is drawn up into the suction; above the suction point, velocities are very low owing to the removal of flow and momentum by the suction line. The end-suction case is shown as the bottom plot in Figure 6.18.

It appears that the location of the suction relative to the jets is one of the more important variables in sludge mobilization performance (though as noted in Section 5, its effect on mixing of a Newtonian fluid was rather small). Of the three configurations modeled, the one with suction located in the tank center beneath the jets provided the most effective mobilization; that is, the minimum jet velocity for mobilization was the lowest for that arrangement. The configuration with the suction located above the jets apparently provided much less effective mobilization -- although other characteristics of the sixth-scale runs, for instance the half-filling of the tank and the low height of the jets, may also have acted to reduce mixing and mobilization.

The literature discussed in Section 4 does not describe the effects of suction location, but sheds a little light on some other tank geometry effects. In a vertical cylindrical tank containing a Newtonian fluid with a free surface, the optimum location for a nozzle is at a height between 25 and 50 percent of the liquid depth. This height allows entrainment from both above and below the jet center and leads to three-dimensional circulations, thus improving mixing. A jet nearer the floor becomes (to some extent) a wall jet and tends to produce two-dimensional circulation. In a horizontal cylindrical tank, moving the jet near to any "wall", whether "side", "top", or "bottom", would tend to convert the three-dimensional jet to a wall jet and reduce its effectiveness. It is not clear how this theory can be applied quantitatively to the MVST-CIP tanks, where the improved mixing effectiveness from raising the jets could be offset by lower solids pickup because of the decreased near-floor velocity.

The literature also showed that, because of entrainment, the mixing time is inversely proportional to the jet axis length (if the tank volume is held constant). In the case of a horizontal cylindrical tank with opposing-direction jets placed at the center of the tank length, the two jets are mixing equal volumes of fluid and have equal jet axis lengths. In a Newtonian fluid, moving the jets along the length should not change the mixing time because the length dependence cancels out; from Equation 4.5, the fluid volume and axis length of each jet change proportionately. In the MVST CIP tanks, jets that are off-center lengthwise would probably increase the chances of sludge-bank retention at the long end, as in the 1/6 scale modeling.

$D = 5.1 \text{ m}$	jet diameter = 7.6 or 10 cm
$L = 20.2 \text{ m}$	jet height = 24 cm (to nozzle centerline)
$V = 387 \text{ m}^3$	suction height = 6.1 cm
sludge depth = 0.61 m or 0.46 m

$D = 5.1 \text{ m}$	jet diameter = 7.6 or 10 cm
$L = 20.2 \text{ m}$	jet height = 24 cm (to nozzle centerline)
$V = 387 \text{ m}^3$	suction heights = 0.7 m and 1.2 m
	suction diameter = 0.17 m
sludge depth = 0.61 m	

Figure 6.1. Schematic of tank configuration used in computational models for Task 4, sludge mobilization and mixing.

Figure 6.2. The heights of the final-concentration contour at the tank end for 7.6-cm jets and center suction (base cases).

Figure 6.3. The heights of the final-concentration contour at the tank end for 10-cm jets and center suction (base cases).

Figure 6.4. The heights of the final-concentration contour at the tank end for 7.6-cm jets and end suction.

Figure 6.5. The heights of the final-concentration contour at the tank end for 10-cm jets and end suction.

Figure 6.6. The heights of the final-concentration contour at the tank end for 7.6-cm jets and center suction (all cases).

Figure 6.7. The heights of the final-concentration contour at the tank end for 10-cm jets and center suction (all cases).

Figure 6.8. Comparison of the concentrations in the two end-suction recirculation inlets located at 0.7 and 1.2 m above the tank floor.

Figure 6.9. Comparison of the velocities in the two end-suction recirculation inlets located at 0.7 and 1.2 m above the tank floor.

For all plots, the volume fraction contours are:

— 3 — 4.000E-02
 — 2 — 2.000E-02
 — 1 — 9.500E-03

Time = 20 minutes

Time = 40 minutes

Time = 55.6 minutes

Figure 6.10. Typical velocity and sludge concentration characteristics for a double-ended end-suction configuration (7.6-cm diameter jets, 6.1 m/s).

Figure 6.11. Vertical velocity profile near the tank end for the end-suction configuration (7.6-cm diameter jets, 6.1 m/s).

Figure 6.12. Vertical velocity profile near the tank end for the center-suction configuration (7.6-cm diameter jets, 6.1 m/s).

Each concentration contour shows the location of the final, well-mixed volume fraction at a different time.

Figure 6.13. Velocity and sludge concentration characteristics for end-suction, with 10-cm diameter jets impelled from 4.6 to 6.1 m/s.

Each concentration contour shows the location of the final, well-mixed volume fraction at a different time.
 Figure 6.14. Velocity and sludge concentration characteristics for end-suction, with 10-cm diameter jets impelled from 4.6 to 7.6 m/s.

Each concentration contour shows the location of the final, well-mixed volume fraction at a different time.

Figure 6.15. Velocity and sludge concentration characteristics for end-suction, with 7.6-cm diameter jets impelled from 6.1 to 7.6 m/s.

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Each concentration contour shows the location of the final, well-mixed volume fraction at a different time.

Figure 6.16. Velocity and sludge concentration characteristics for end-suction, with 7.6-cm diameter jets impelled from 6.1 to 9.1 m/s.

Figure 6.17. Vertical velocity profiles near the tank end for four end-suction cases (7.6 cm, 7.6 and 9.1 m/s; 10 cm, 6.1 and 7.6 m/s).

The 1/6-scale tank, 4.0 m/s (13 ft/s).

The 378 m³ (100,000 gal) tank, center-suction case, 10-cm jets, 3.1 m/s.

The 378 m³ (100,000 gal) tank, end-suction case, 10-cm jets, 4.6 m/s.

Each concentration contour shows the location of the final, well-mixed volume fraction.

Figure 6.18. Comparison of the 1/6-scale and full-scale velocity fields for conditions under which sludge is not mobilized.

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APPENDIX A

EXTENDED RHEOLOGICAL CHARACTERIZATION OF MVST W-28 SLUDGE

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EXTENDED RHEOLOGICAL CHARACTERIZATION OF MVST W-28 SLUDGE

As part of another ongoing waste-tank remediation project (the Flammable Gas Program, which is focused on Hanford Tank 241-SY-101), the MVST W-28 sludge rheological data were analyzed and compared to several rheological models other than the power law. This appendix briefly presents the analysis results, which show that biviscous constitutive equations such as the Carreau and Cross models more completely represent W-28 rheology than the power-law model. The results also suggest that the W-28 sludge can be described better by models that do not include a yield stress than by models that do.

A.1 Models in the Literature

The classic rheological description of a viscous fluid is simply

$$\tau = -\eta \mathbf{D} \quad (\text{A.1})$$

In the Newtonian case, the shear stress tensor (τ) is a linear function of the shear rate tensor ($\mathbf{D}$); the viscosity (η) is not a function of $\mathbf{D}$ or of time but only of solids concentration, temperature and possibly other local variables. For a non-Newtonian fluid such as a typical sludge, the viscosity is a function of $\mathbf{D}$, and frequently a complicated one.

Some of the most desirable characteristics of rheological models include a minimal number of adjustable (experimentally obtained) constants, a form that allows easy curve-fitting to data, computational simplicity to hold down computer costs, and physical meaning. The following subsections discuss the models in the literature with respect to these criteria. For convenience, the models are sub-divided into two types, those with and without yield stresses.

A.1.1 No-Yield Models

The first broad set of non-Newtonian rheological models comprises the models without shear yield stresses. In the no-yield models, the presumption is that every stress, no matter how small, leads to a continuing deformation. Fluids in which the viscosity decreases as the shear rate increases are known as "shear-thinning" or "pseudoplastic".

The simplest of the no-yield models is the power law, or Ostwald-DeWaele, model (Bird et al., 1960):

Power law:

$$\tau = -\left\{ K \left| \sqrt{\frac{1}{2} \Pi_D} \right|^{n-1} \right\} \mathbf{D} \quad (\text{A.2})$$

where τ = shear stress tensor
 D = shear rate tensor
 Π_D = second invariant of the shear rate tensor
 K = consistency index
 n = structural (or behavioral) index

For $n < 1$, which is usually the case for suspensions and sludges, the fluid is pseudoplastic. For very concentrated suspensions, n may exceed 1 and the fluid is "dilatant", becoming more viscous with increasing shear rate. The major computational difficulty with the power-law model is that (for $n < 1$) the viscosity becomes infinite at low shear rates.

A more complex non-Newtonian model was proposed by Cross (1965). The original derivation was based on the assumption that pseudoplastic behavior occurred in suspensions because of the formation and breakage of structural linkages between particles. A kinetic approach was used to estimate the average group size as a function of shear, and the viscosity was considered to be proportional to group size (assuming the groups are in the form of kinked chains). Cross-like models are the ones most likely to appear when a model is partly or wholly derived from theory. A generalized Cross model (Evans, 1992) is

$$\tau = \left\{ \eta_{\infty} + \frac{(\eta_0 - \eta_{\infty})}{1 + B \left| \sqrt{\frac{1}{2} \Pi_D} \right|^n} \right\} D \quad (A.3)$$

where η_{∞} = Newtonian viscosity at high shear rate (20-100 sec⁻¹)
 η_0 = Newtonian viscosity at very low shear rate (0.01-0.5 sec⁻¹)
 B = correlating constant specific to the fluid

This four-constant version of the Cross model is practical, given modern curve-fitting software, and can describe a wide variety of non-Newtonian behavior. The Cross model gives a finite viscosity of η_0 as the shear rate approaches zero.

Finite viscosity at very low strain rates is computationally desirable because of its continuity, but poses a new computational inconvenience. The numerical error in any computer model leads to calculation of infinitesimal shear stresses in systems where the true shear stress is zero. For finite viscosities, the shear stress error produces very small but non-zero strain rates, which over a prolonged period overestimate the fluid deformation.

Another no-yield model, the Carreau model, was presented in a handy four-constant version by Ellwood et al. (1990):

$$\tau = \left\{ \eta_{\infty} + (\eta_0 - \eta_{\infty}) \left[1 + (B \left| \sqrt{\frac{1}{2} \Pi_D} \right|)^2 \right]^{(n-1)/2} \right\} D \quad (A.4)$$

Here the parameter B is a characteristic relaxation time. The Carreau model also gives a finite viscosity at low shear rates. This model is potentially useful to describe suspension rheology but seems to be seldom used.

Another type of fluid referred to in passing in the literature (Kalyon et al., 1993) is biviscous -- a fluid that has two different Newtonian viscosities, one at very low and one at high shear rates. A simple three-constant formulation of such a model might be

$$\tau = - \left\{ \eta_0 + \frac{(\eta_0 - \eta_\infty)}{B \sqrt{\frac{1}{2} \Pi_D}} (1 - \exp(-B \sqrt{\frac{1}{2} \Pi_D})) \right\} D \quad (\text{A.5})$$

Here B controls the breadth of the region over which the viscosity changes from its low-shear to its high-shear value.

A.1.2 Yield Stress Models

Several well-known non-Newtonian models include a yield stress, which can be defined as a minimum shear stress below which no continuing deformation occurs. Below the yield stress, such a material behaves like a solid with a finite deformation; above it, behavior is fluid, with a finite shear rate and unbounded deformation. Fluids that display a yield stress are termed "viscoplastic".

By and large, the suspension rheology models in the yield-stress category are correlations of experimental data rather than expressions that are strongly founded in the theory of particle interactions. Four models fit under this heading:

Bingham (Bird et al., 1960):

$$\tau = - \left\{ \eta_\infty + \frac{\tau_0}{\sqrt{\frac{1}{2} \Pi_D}} \right\} D \quad \text{for } \frac{1}{2} \Pi_\tau > \tau_0^2 \quad (\text{A.6})$$

and $D = 0$ for $\frac{1}{2} \Pi_\tau < \tau_0^2$

Casson (Bird et al., 1983):

$$\tau = - \left\{ \sqrt{\frac{\tau_0}{\sqrt{\frac{1}{2} \Pi_D}}} + \sqrt{K} \right\}^2 D \quad \text{for } \frac{1}{2} \Pi_\tau > \tau_0^2 \quad (\text{A.7})$$

and $D = 0$ for $\frac{1}{2} \Pi_\tau < \tau_0^2$

Herschel-Bulkley (Bird et al., 1983):

$$\tau = - \left\{ K \sqrt{\frac{1}{2} \Pi_D}^{n-1} + \frac{\tau_0}{\sqrt{\frac{1}{2} \Pi_D}} \right\} D \quad \text{for } \frac{1}{2} \Pi_\tau > \tau_0^2 \quad (\text{A.8})$$

and $D = 0$ for $\frac{1}{2} \Pi_\tau < \tau_0^2$

Gay (Dabak and Yucel, 1987):

$$\tau = \left\{ \frac{\tau_0}{\sqrt{\frac{1}{2}\Pi_D}} + \eta_\infty + \frac{(\eta_0 - \eta_\infty)}{1 + \frac{(\eta_0 - \eta_\infty)\sqrt{\frac{1}{2}\Pi_D}}{B}} \right\} D \quad \text{for } \frac{1}{2}\Pi_\tau > \tau_0^2 \quad (\text{A.9})$$

and $D = 0$ for $\frac{1}{2}\Pi_\tau < \tau_0^2$

where

- τ = shear stress tensor
- Π_τ = second invariant of the shear stress tensor
- D = shear rate tensor
- Π_D = second invariant of the shear rate tensor
- τ_0 = yield stress
- η_∞ = Newtonian viscosity at high shear rate (20-100 sec⁻¹)
- η_0 = Newtonian viscosity at very low shear rate (0.01-0.5 sec⁻¹)
- K = consistency index
- n = structural (or behavioral) index
- B = correlating constant specific to the fluid

All of these models go to infinite viscosity η as D goes to zero (Bird et al., 1983) because Π_D (which also goes to zero) is in the denominator of the shear stress terms. The yield-stress models also possess a discontinuity at $D = 0$, $\tau = \tau_0$. These computational problems can be avoided by substituting a stiff exponential for τ_0 (Ellwood et al., 1990). For example,

$$\tau_0 = |\tau_0| (1 - e^{-m|\sqrt{\frac{1}{2}\Pi_D}|}) \quad (\text{A.10})$$

where m is a steepness adjustment, usually $m > 50$ (op. cit.).

Usually, m is set for computational convenience rather than determined from experimental data. However, data could also be used, providing a kind of biviscous model. The viscosity of the m -modified Bingham model (for example) approaches a value of $(\eta_{0m} + \eta_\infty)$ as zero is approached, according to L'Hospital's Rule. Thus, $(\eta_{0m} + \eta_\infty)$ can be taken as the value of viscosity at very low shear rates, or $m = (\eta_0 - \eta_\infty)/\eta_0$. The m -modified Bingham model is the simplest (three-constant) form of a biviscous model, but the exponential could equally well be used in the more complicated Herschel-Bulkley, Casson, or Gay forms.

Of these models, the Herschel-Bulkley has the most general applicability; it and its subsets, the Bingham and power-law models, are the most commonly used correlations in the experimental literature. However, two researchers (Turian et al., 1992; Fam et al., 1989) found that the Casson or Bingham models better fit the rheological data for their suspensions than did the Herschel-Bulkley model.

A.2 Experimental Data

In order to choose rheological models for inclusion in TEMPEST and use in tank sludge modeling, it would be useful to know which are most likely to be applicable to tank sludge. Making this judgment requires analysis of sludge rheological data.

Two sources of data on radioactive waste sludge rheology were reviewed. One (Fow et al., 1986) described measurements made on a non-radioactive simulant of different dilutions of neutralized current acid waste (NCAW) from the Hanford double-shell tanks (DSTs). The measurements were made at ambient and elevated temperatures and were intended to support design of a pump-pipe network to transport NCAW slurries. Although the sludge type is very relevant, for current purposes these data are incomplete because the lowest strain rate used in the measurements was 50 sec^{-1} . (This was an appropriate lower end for the design purpose of the reference, though.)

A more immediately useful data set was found in Ceo et al. (1990). Samples were taken from typical radioactive sludges and supernatant liquids in the Melton Valley Storage Tanks (MVST). The sludge samples were taken in cores up to 0.51 m (20 in) deep. All the samples, supernatant, sludge, and sludge diluted with supernatant (from the same tank), were assayed for viscosity using a Brookfield rotational viscometer. The sludge rheological measurements were made on samples that had been handled as little as possible and had not been mixed or homogenized. All measurements were made at 21°C and over a strain rate range of 0.08 to 17 sec^{-1} .

Table A.1 shows the data for the MVST W-28 tank sludge, the most complete data set from Ceo et al. (1990). The supernatant from tank W-28 was Newtonian, and substantial non-Newtonian behavior can be seen in sludge data. There was also some evidence of time-dependent behavior, though without enough data to allow assessment of thixotropy.

The densities of the bulk and interstitial liquid were equal at 1.29 g/cc . The density of the (centrifuged) bulk sludge was 1.40 g/cc and the undissolved solids density was 2.00 g/cc . The bulk sludge contained 51.4 wt% total solids, comprised of 29.4 wt% dissolved solids in the interstitial liquid and 22.0 wt% undissolved solids. Based on the sedimentation rate of 2.9 cm/min , the W-28 sludge particle agglomerates were estimated to have a radius of $26 \mu\text{m}$ (Ceo et al., 1990). The sludges studied were all apparently gas-free.

The W-28 tank at MVST contains predominantly low-level waste (LLW) concentrates. The sludge is mostly calcium, magnesium and aluminum hydroxides and carbonates; the supernatant liquid is a basic aqueous solution of sodium nitrate and fission products (Youngblood et al., 1991).

For comparison, the viscosity of the Hanford Tank 101-SY supernatant is 24 cP , the supernatant density is 1.44 g/cc , the bulk density of the sludge is 1.79 g/cc , and the intrinsic solid density is 2.18 g/cc . The sludge composition (in the absence of gas) is believed to be between 47 and 65 volume percent undissolved solids. The maximum gas content is 19 to 24 volume percent. Both the sludge and the supernatant in Tank SY-101 appear to be more concentrated than the MVST W-28 counterparts, and may therefore exhibit considerably different rheology.

Table A.1. Shear stresses and strain rates of bulk liquid and sludge, ORNL MVST tank W-28.

Shear rate (sec ⁻¹)	Shear stress (dyne/cm ²)
<u>W28-L3 (supernatant)</u>	
36	0.81(constant viscosity of 2.2 mPa s)
<u>W28-S3 (undiluted sludge)^a</u>	
0.08	32
0.17	35
0.42	51
0.84	82
1.7	150
3.4	off scale
<u>W28-S3 sludge, diluted 1:1 with W-28-L3 supernatant</u>	
0.08	5.0 *
0.17	7.4 *
0.42	18
0.84	35
1.7	52
3.4	69
8.4	79
16.8	87
<u>W28-S3 sludge, diluted 1:3 with W-28-L3 supernatant</u>	
0.08	2.2 *
0.17	3.5 *
0.42	12 *
0.84	15 *
1.7	19
3.4	22
8.4	26
16.8	29

^a Data are doubtful because some slip was observed.

* These data were taken at stress levels with low instrument response and may have more than 10% error for that reason.

A.3 Applying Models to Sludge Data

The data from Ceo et al. (1990) were analyzed to find which of the models listed in Section A.1 best described the W-28 sludge rheology. Figures A.1 through A.7 show the data and the curve fits that were obtained for the several models. (Note: The power-law model was already fit to the W-28 data once, by Terrones and Eyler [1993]. That fit was chosen to match only the regime above 1 sec⁻¹; this fit includes the whole data range, accounting for the apparent discrepancies between the results.)

The two-constant models, the power-law and Casson models, do not well represent the "knee" of the stress-strain data (Figures A.1 and A.6). Of the two, the power-law model provides a slightly closer fit. The Herschel-Bulkley model (Figure A.5), a three-constant model, comes much closer to the 1:1 and 1:3 dilution data than does its subset the power-law, but it needs an unphysical negative yield-stress to do so. The other three-constant model, the biviscous model (Figure A.4), is clearly rather well adapted to modeling this type of rheology.

The remaining models, the Cross, Carreau, and Gay models, are all four-constant models that include two different constant viscosities for low and high strain rates. The Gay model (Figure A.7) also includes a yield stress and fits the data slightly less well than the other yield-less models. The Cross and Carreau models (Figures A.2 and A.3) both provide fits which are very close to the data.

A little more insight about model qualifications can be gained by looking at the curve-fit constants for the models. These constants are given below in Table A.2. In all cases the units of shear stress are dyne/cm² and the units of strain rate are sec⁻¹, so viscosities are in poise (g/cm s) and yield stresses in dyne/cm² (where 1 Pa = 10 dyne/cm²).

It appears that the yield-stress models require negative yield stresses τ_0 to provide accurate stress predictions at low strain rates for the whole range of concentrations. The physical meaning of this correlation feature is unclear, and in general the yield-stress models do not seem the best choices to model this sludge.

Of the no-yield models, the Cross, Carreau, and biviscous models are clearly the best. The Cross and biviscous models seem slightly preferable because the viscosity at high strain rates is always greater than the 2.2 cP value found for the supernatant liquid (as noted in Table A.1). The Carreau model, on the other hand, predicts an η_∞ of 0.78 cP for the 1:3 diluted sludge. This incongruity would be an inconvenience in computational modeling.

In addition, the low value of the Carreau infinite-strain viscosity leads to implausibly low shear stresses at high shear rates. For example, at 100 sec⁻¹ the predicted shear stress for W28 diluted 1:3 is 69 dyne/cm² for the Cross model and 40 dyne/cm² for the Carreau model. At 1000 sec⁻¹, for the same dilution, the shear stress is calculated at 530 dyne/cm² for the Cross model and 64 dyne/cm² for the Carreau. The Carreau model plateaus (for these particular sludges) and the Cross model does not.

One danger of using the Cross model is that the equation can, with certain parameters, predict a shear stress that reaches a maximum as strain rate increases and then decreases at higher strain rates. In the case of the W-28 data, no such maximum is found between 0.01 and 1000 sec⁻¹ and this problem is not a concern. This question does not arise with the biviscous model, where the relation between stress and strain is unambiguously monotonic. Furthermore, the biviscous model requires only three constants and the Cross model needs four.

Taking everything into account, the biviscous model (Equation A.5) appears to be best suited to model the MVST Tank W-28 sludge rheology. This may provide a clue as to the best choice of models for the Hanford tank sludges; however, the higher solids content of the Tank 241-SY-101 sludges may require a yield-stress model, while the W-28 sludge does not.

Table A.2. Correlation parameters for rheological models fit to the ORNL MVST W-28 stress/strain data.

Power-law (Equation A.2):

W28 --	$K = 102;$	$n = 0.666;$	$r^2 = 0.97$
W28 1:1 --	$K = 35.4;$	$n = 0.355;$	$r^2 = 0.91$
W28 1:3 --	$K = 13.7;$	$n = 0.295;$	$r^2 = 0.91$

Cross (Equation A.3):

W28 --	$\eta_0 = 3800;$	$\eta_{\infty} = 81.9;$	$B = 293;$	$n = 1.31;$	$r^2 = 1.00$
W28 1:1 --	$\eta_0 = 49.4;$	$\eta_{\infty} = 1.86;$	$B = 0.312;$	$n = 1.33;$	$r^2 = 1.00$
W28 1:3 --	$\eta_0 = 36.3;$	$\eta_{\infty} = 0.515;$	$B = 1.28;$	$n = 1.097;$	$r^2 = 0.99$

Carreau (Equation A.4):

W28 --	$\eta_0 = 110;$	$\eta_{\infty} = 110;$	$B = 0;$	$n = 1.00;$	$r^2 = 1.00$
W28 1:1 --	$\eta_0 = 45.6;$	$\eta_{\infty} = 0.320;$	$B = 0.659;$	$n = 0.0727;$	$r^2 = 1.00$
W28 1:3 --	$\eta_0 = 30.8;$	$\eta_{\infty} = .0078;$	$B = 1.87;$	$n = 0.164;$	$r^2 = 0.99$

Biviscous (Equation A.5):

W28 --	$\eta_0 = 13400;$	$\eta_{\infty} = 73.9;$	$B = 591;$	$r^2 = 1.00$
W28 1:1 --	$\eta_0 = 53.1;$	$\eta_{\infty} = 0.904;$	$B = 0.727;$	$r^2 = 1.00$
W28 1:3 --	$\eta_0 = 33.0;$	$\eta_{\infty} = 0.547;$	$B = 1.60;$	$r^2 = 0.99$

Casson (Equation A.7):

W28 --	$\tau_0 = 8.54;$	$K = 49.4;$	$r^2 = 0.99$
W28 1:1 --	$\tau_0 = 18.4;$	$K = 1.96;$	$r^2 = 0.78$
W28 1:3 --	$\tau_0 = 8.06;$	$K = 0.497;$	$r^2 = 0.76$

Herschel-Bulkley (Equation A.8):

W28 --	$\tau_0 = 27.6;$	$K = 65.8;$	$n = 1.17;$	$r^2 = 1.00$
W28 1:1 --	$\tau_0 = -355;$	$K = 395;$	$n = 0.043;$	$r^2 = 0.97$
W28 1:3 --	$\tau_0 = -129;$	$K = 144;$	$n = 0.0358;$	$r^2 = 0.98$

Gay (Equation A.9):

W28 --	$\tau_0 = 22.5;$	$\eta_0 = 73.9;$	$\eta_{\infty} = 73.9;$	$B = 0;$	$r^2 = 1.00$
W28 1:1 --	$\tau_0 = -2.19;$	$\eta_0 = 69.4;$	$\eta_{\infty} = -0.18;$	$B = 99.3;$	$r^2 = 1.00$
W28 1:3 --	$\tau_0 = -2.73;$	$\eta_0 = 63.3;$	$\eta_{\infty} = 0.346;$	$B = 293;$	$r^2 = 0.99$

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Figure A.1. Tank W-28 data as fitted by a power-law model.

Figure A.2. Tank W-28 data as fitted by the Cross model.

Figure A.3. Tank W-28 data as fitted by the Carreau model.

Figure A.4. Tank W-28 data as fitted by a biviscous model.

Figure A.5. Tank W-28 data as fitted by the Herschel-Bulkley model.

Figure A.6. Tank W-28 data as fitted by the Casson model.

Figure A.7. Tank W-28 data as fitted by the Gay model.

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