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1.0 INTRODUCTION

This report evaluates the performance of the Defense Waste Processing Facility (DWPF) feed preparation system using Hanford Waste Vitrification Plant (HWVP) process criteria and feed properties. Included is a proposed test plan to verify performance of the equipment identified in the evaluation. The HWVP is being designed to vitrify high-level liquid defense waste currently stored in double shell tanks on the Hanford site. The following sections describe the background and objectives and the approach used in this evaluation.

1.1 BACKGROUND AND OBJECTIVES

The HWVP will be the second vitrification plant for solidification of defense waste in the United States. The first plant to be constructed is the DWPF in South Carolina. The design of the DWPF is being adopted as appropriate for the HWVP. The feed preparation system will receive high-level waste and concentrate the waste by evaporation. Glass formers and chemicals are then added to the waste, and the slurry is fed to the melter. The feed preparation system must maintain a homogeneous feed within chemical component specifications to fulfill the quality control requirements of the final glass product. In addition, process specifications must be attained to meet time cycle requirements and design production rates. Technology from the DWPF is being adapted for the HWVP feed preparation system due to the similarity of the waste to be processed. However, differences in waste composition and process requirements warrant an evaluation of the DWPF feed preparation system.

The primary objective of this study is to evaluate the adaptability of the DWPF feed preparation system to the HWVP and to propose a test plan to verify equipment performance as necessary using suitable HWVP feeds. Transferral of DWPF feed preparation technology to HWVP design requires evaluation of DWPF design and equipment performance data, evaluation of HWVP process development data, engineering analysis and final assessment. Proper application of these elements could improve equipment design and reduce equipment cost. Verification of process equipment performance will be necessary if DWPF equipment performance cannot be assured using HWVP feed properties or process requirements.

The feed preparation equipment that were evaluated are listed below:

- slurry receipt and adjustment tank (SRAT)
- slurry mix evaporator (SME)
- melter feed tank (MFT)
- spent frit hold tank (SFHT)
- frit slurry make-up tank (FSMT).

In addition, associated equipment, such as condensers, pumps, agitators, and sample systems were included in the analysis.

1.2 APPROACH

The procedure for evaluating the DWPF design of the feed preparation system was as follows. Initially, process requirements were established and technical information was collected. Process requirements define the equipment performance necessary for product quality and design production rates. Technical information consists of equipment specifications, fluid properties, and development and equipment test data from HWVP and DWPF. Once this information was assembled, process requirements and feed properties could be compared between HWVP and DWPF. Engineering relationships were then used to relate fluid properties and equipment specifications to predict process performance.

As feasible equipment was evaluated using data from full-scale testing at the Savannah River Laboratory (SRL), bench-scale testing at Pacific Northwest Laboratory (PNL), and engineering relationships and expert advice. Where engineering analysis and judgement could not ensure equipment performance, verification testing of the equipment was recommended.

A test plan is proposed using the equipment identified during the evaluation. The test plan will describe the process performance that requires verification, the operating data that will be collected, and a test strategy. The test system will assess process performance using full-scale equipment. The test equipment will be more flexible than the reference design to allow modifications as necessary. Additional instrumentation will also be used to better characterize the equipment performance.

The body of the report is comprised of the following sections. Section 2 outlines conclusions and recommendations. Section 3 consists of technical information concerning equipment performance requirements and feed properties.

Section 4 evaluates the DWPF feed preparation system as it applies to the HWVP and identifies euqipment that requires testing. Section 5 proposes a test plan for demonstrating process performance of the equipment identified in Section 4. Appendix A provides a technical background on flud mixing, heat transfer, and fluid transport. Appendix B contains supporting calculations.

2.0 CONCLUSIONS AND RECOMMENDATIONS

Full-scale testing of the feed preparation system will be required to verify process performance of the equipment. The differences in feed properties and process requirements between HWVP and DWPF are too uncertain and vary too greatly to assure acceptable process performance. In addition, testing is necessary to evaluate situations that have occurred during full-scale testing at the TNX facility in the DWPF. The DWPF staff have concurred with the need to perform testing to confirm process performance.

The HWVP will need the flexibility to process a variety of feeds: neutralized current acid waste (NCAW), plutonium finishing plant (PFP) waste, or complex concentrate (CC). The DWPF will blend the Savannah River Plant (SRP) waste to produce a uniform feed. The feed components that affect process performance or design requirements are compared below (NCAW is the initial feed to the HWVP).

	HWVP NCAW Feed	DWPF Feed
• Solids	nominally 2 wt%	nominally 13 wt%
• Mercury	none detected	3 wt% max
• Organics	2 wt% TOC	1 to 3 wt% organics
• Elements	more Zr than DWPF feed	

These feed components can affect design requirements and equipment performance. The more dilute HWVP feed requires a higher evaporation rate to meet time cycle design requirements. This means the boiling heat transfer rate must be higher for HWVP feeds. Moreover, dilute feed from the receipt and lag storage tank (RLST) will be added continuously to the SRAT in the HWVP while in the DWPF it is added in batches. Mercury has not been discovered in analysis of NCAW feed to date. Therefore, the mercury removal and purification equipment of the DWPF may not be necessary in the HWVP. The organics in the DWPF feed are suspected of plugging the condenser tubes and causing a low solids decontamination factor (DF) between the slurry and condensate.

The HWVP feed contains similar organic levels to the DWPF feed, but of different types. The affects of these other organics will have to be assessed by testing. Differences in elemental species and particle size, distribution, and shape can affect slurry properties like viscosity and the potential for foaming. These properties, in turn, impact slurry mixing, homogeneity, and condenser solids DFs.

In addition to differences in feeds, a few problems have surfaced during testing at the DWPF that may require equipment testing using HWVP feeds:

- severe impeller, coil, and vessel erosion due to mechanical abrasion between coils and supports, erosion of the lower impeller blade, and slurry abrasion of the vessel bottom.
- a large vessel heel of residual slurry which is increased by higher impeller speeds
- inaccurate volume measurement while the tank is being agitated
- a low solids DF between condensate and process slurry at HWVP evaporation rates of 10 gpm caused by excessive foaming of the slurry
- condenser tube plugging during TNX feed preparation campaign #3 from organics and solids.
-

Listed below are the key factors that HWVP feed preparation system testing must address:

- impeller, coil, and vessel erosion
- slurry homogeneity throughout the feed preparation cycle
- variable feeds
- high evaporation rates
- high vessel heel
- accurate level measurement
- adaptation of the DWPF melter feed system to the lower HWVP feed rates
- low solids DF between the condensate and the process slurry
- slurry foaming

The DWPF SRAT/SME/MFT vessel design was discussed in a meeting with PNL and HWVP Technology and Design personnel. In the meeting it was suggested that the feed preparation evaluation recommend modifications to the DWPF design. The modifications were confined to maintain the external vessel dimensions and the concept of the coil and agitator system.

The recommended modifications are to increase the impeller diameter and change the type of impeller and modify the coils to increase the heat transfer surface area and reduce erosion. These modifications will improve the following key factors:

- Lower impeller, coil, and vessel erosion by reducing impeller cavitation, coil movement, and slurry velocities.
- Expand the system flexibility to process more difficult feeds.
- Increase the evaporation rate by adding heat transfer surface area.
- Reduce the vessel heel by lowering the impeller speed.

The equipment listed below will be necessary to verify feed preparation system performance.

- A full-scale, prototypic vessel representing the SRAT, SME, and MFT -- the vessel will include a variable speed agitator, condenser with solids de-entrainer, heat transfer system, transfer pump, slurry sampler, and melter feed system. The full-scale system will include additional instrumentation to completely monitor process performance.
- An existing 9-ft-diameter vessel will be used as a hold tank and can model the SFHT, FSMT, and slurry mix evaporator condensate tank (SMECT).

Since the HWVP will be processing a variety of feeds, verifying equipment performance for a single feed is inappropriate. Establishing a process envelope for equipment performance would verify process performance for any feed whose properties fit the requirements. An operating envelope will be established based on full-scale testing. The operating envelope will define process variables (e.g., steam pressure, impeller speed, etc.) and fluid properties (e.g., viscosity, particle size, etc.) within which acceptable process performance is expected. The output of the test program will be as follows:

- Verify equipment performance.
- Establish an operating envelope.
- Provide guidance in equipment operation for variations in feed properties or process requirements.
- Provide experimental data for statistical modeling of the melter feed tank, which could be used for waste form qualification.

3.0 TECHNICAL INFORMATION

This section describes the process requirements, physical and rheological properties of the HWVP and DWPF process slurries, and equipment performance requirements. The process description section (Section 3.1) outlines the process flow scheme and how the equipment operates as a system. The physical and rheological properties section (Section 3.2) contains tables of process fluid physical properties with a discussion of slurry rheology and foaming. Equipment performance requirements (Section 3.3) specify process operating conditions for the HWVP.

3.1 PROCESS DESCRIPTION

The overall function of the feed preparation system is to receive the HWVP feed from the RLST, prepare the feed for vitrification, and transfer it to the melter as shown in Figure 3.1. In the current process, the feed is prepared by concentrating it and adding formic acid in the SRAT. Further concentration is performed if necessary, the slurry is sampled, and the formated and concentrated slurry is transferred to the SME. Slurries of glass frit are added in the SME, followed by additional concentration. The slurry in the SME is analyzed to ensure that it meets the specifications for a melter feed and then transferred to the MFT. The MFT serves as a holding tank for the melter feed; the melter feed system delivers a controlled flow of slurry to the melter. The condensate from the concentration steps in the SRAT and SME is sent to the SMECT.

The SRAT, SME, SMECT, and MFT are identically sized tanks with a 12-ft ID and an 11,000-gal maximum capacity. Agitation in the SRAT, SME, and MFT is provided by a top-mounted agitator using two impellers. The SMECT requires minimal agitation and is mixed with an air sparger. The tank contents in the SRAT and SME are heated and cooled through separate immersed helical coils carrying steam or cooling water. The MFT slurry will not require concentration and therefore has only a cooling coil to remove decay heat. The SMECT requires no coils since it only receives condensate. Vapor evolved in the SRAT, SME, and SMECT is condensed by vessel-mounted, water-cooled vertical tube condensers. All condensate is normally routed to the SMECT. The SRAT condenser is also capable of returning condensate directly to the SRAT tank during total reflux

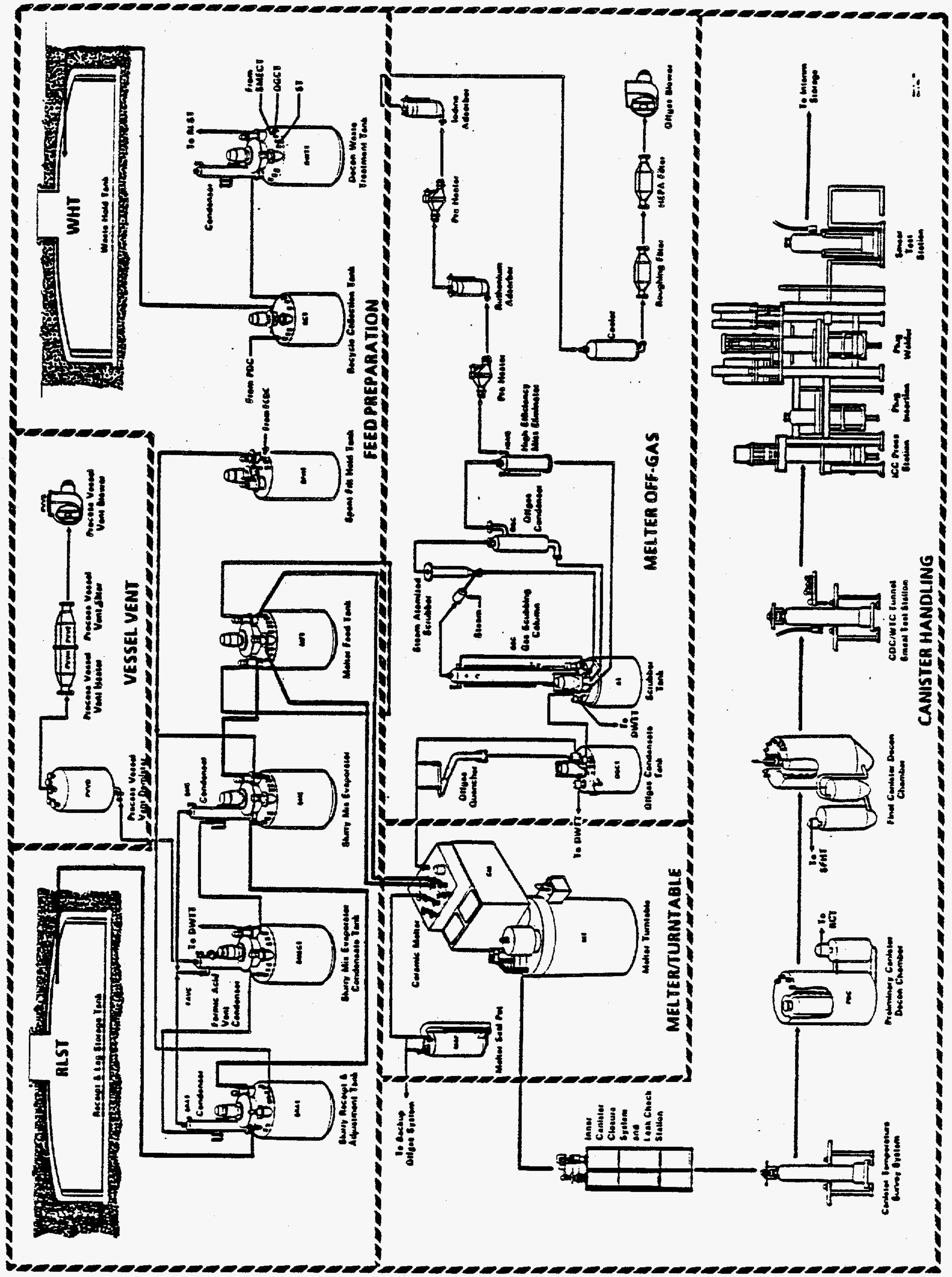


FIGURE 3.1 Process Flow Diagram of the HWVP

operation. Non-condensable gas from the SRAT and SME condensers are vented to the formic acid vent condenser (FAVC) on top of the SMECT. Non-condensable gases from the FAVC are routed to the formic acid vent header (FAVH).

All tanks will have sampling systems composed of a recirculating loop through a sample cell. Intertank transfer will be designed for a flow of 100 gpm.

The SME will be capable of receiving feed directly from the RLST and performing appropriate process operations to deliver design feed to the MFT. The SRAT will also be able to deliver melter feed to the MFT, bypassing the SME. The SME tank contents will also be transferable back to the SRAT. The SRAT and SME contents can be transferred back to the RLST (HWVP TDP 1986).

3.1.1 Receipt and Lag Storage Tank

The RLST stores HWVP feed containing 2 to 15 wt% total solids with a nominal concentration of 2 wt% solids (HWVP TDP 1986). It is an underground double shell tank whose contents are well agitated with a jet pump to create uniform tank composition. Tank contents are sampled for feed analysis. Cooling is provided to maintain contents at less than 122°F (50°C). RLST contents can be delivered to the SRAT or SME at the design transfer flow of 100 gpm or at a controlled low flow of 1 to 10 gpm. Line flushing will follow a slurry transfer.

3.1.2 Slurry Receipt and Adjustment Tank

The HWVP feed is transferred from the RLST to the SRAT. In the SRAT the feed is concentrated up to 150 g waste oxide (WO)/L. The feed is then cooled 10 to 30°F and 90 wt% formic acid is added at a controlled rate. Formic acid has been shown by Savannah River Laboratory (SRL) to improve slurry rheology and also serves as a reductant. Once the formic acid addition is complete, the tank contents are placed on total reflux for 2 to 6 hours with a 1 to 3 gpm boilup. The reflux completes the formic acid reaction. After refluxing, the contents are cooled and analyzed for chemical composition. If the analysis indicates no further adjustment is necessary (eg. more formic acid, feed addition or dilution, or further concentration), then the feed is pumped to the SME. The process line is flushed with water following the transfer.

3.1.3 Slurry Mix Evaporator

The SME normally receives concentrated and formated HWVP feed from the SRAT. After receiving SRAT feed, the SFHT contents are transferred to SME. The SFHT contents are frit and dilute formic acid from the canister decontamination process. The SME contents are then concentrated, followed by another SFHT transfer and concentration step. Frit addition is completed by adding fresh frit from the process frit slurry feed tank (PFSFT). The tank contents are concentrated up to 500 g TO/L to a final volume of 6000 to 8000 gal, cooled to below 122°F and sampled. Once the feed composition requirements are met, tank contents may be transferred to the melter feed tank, followed by a line flush. Provision is also available for the SME to receive HWVP feed directly from the RLST and concentrate and formate it.

3.1.4 Melter Feed Tank

As required, SME tank contents are transferred to the MFT. The tank contents are agitated sufficiently to achieve homogeneity. Samples are then taken and analyzed for accountability and to determine the melter feed composition. The tank contents are agitated continuously. Tank contents are maintained below 122°F, which will require cooling coils to remove decay heat.

The current melter feed system consists of two identical recirculating feed loops. Each loop recirculates melter feed at 100 gpm with a side stream draw through a cross flow strainer. The nominal feed rate in each loop is 0.25 gpm with a range of 0.10 to 0.75 gpm. The flow rate is controlled by varying the pump speed.

The feed line is periodically flushed to prevent plugging and solids accumulation. If the feed pumps are shut down, a full system water flush is performed.

3.1.5 Slurry Mix Evaporator Condensate Tank

Condensate is received from the SME and SRAT condensers, and the FAVC. The condensate temperature is less than 122°F and no cooling is required. Tank contents are agitated with an air sparger before and during sampling and while transferring. The condensate is held for lag storage, sampled and routed to the appropriate process waste system. Prior to transfer, a sample is analyzed to determine whether the tank contents should be routinely sent to the decontamination waste treatment tank (DWTT) for actinides concentration, or routed to the recycle collection tank (RCT). The transfer line normally does not require flushing following a transfer.

3.2 PROCESS SLURRY PHYSICAL AND RHEOLOGICAL PROPERTIES

Slurry rheology and physical properties are important since they affect fluid mixing, heat transfer, and fluid transfer. Chemical composition is important for materials selection and in its affect on physical properties and rheology. All of the physical property data for HWVP slurries are for simulants based on the expected chemical composition of HWVP feed. Tables 3.1 through 3.4 list the process fluid physical and rheological properties and the source of the data. The tables include information on the simulated HWVP feed (Table 3.1), the HWVP simulated slurry during the SRAT cycle (Table 3.2), the HWVP simulated melter feed (Table 3.3), and simulated DWPF slurries (Table 3.4).

The physical and rheological properties of the DWPF process slurries (i.e., SRAT product and melter feed) are not as comprehensive as for the HWVP process slurries. The rheological properties in the tables include: the slurry apparent viscosity, yield stress, consistency index and flow behavior index, and the settled solids shear strength. The physical properties in the tables include slurry density, total solids, total suspended solids, heat capacity, boiling point, pH, maximum particle diameter, mean particle diameter, interface settling rate ("inter. settl. rate" in the table), and ultimate settled solids height. The settling rate of the interface (instead of the particles) is measured because the slurries are too turbid to observe individual particles. The ultimate settled solids height is the final height (in % of total height) of settled solids. The interface settling rate indicates how quickly settled solids will develop and the ultimate solids height indicates the height of the settled solids level. The interface settling rate reflects a minimum settling rate since the interface contains the smallest particles. In comparison, a maximum settling rate would be expected with frit in dilute formic acid. The settling rate of frit in dilute formic acid is less than 1 ft/min, which is not considered a fast settling rate for normal processing.

Table 3.1 Physical Properties of Simulated HWVP Feed

Feed Description	Simulated B-plant processing on PNL-prepared NCAW	Simulated B-plant processing on RHO-prepared NCAW	PNL-simulated NCAW batch HWS-6 @ 19 g WO/L
Data Source	(a)	(a)	(b)
Physical Properties			
Apparent viscosity (cP) at various shear rates, s^{-1}			
10 (50 RPM agitator)	16	5.8	--
25 (130 RPM agitator)	7.5	3.0	--
183	2.5	1.4	2.9 @ 50 °C
Yield stress, $dyne/cm^2$	1.4	0.5	1.9 @ 50 °C
Consistency index, cP	1.7	1.1	--
Flow behavior index	1.0	1.0	--
Slurry density, g/cc	1.02	1.02	1.04
Total solids, wt%	4.1	2.3	1.7
Total susp. solids, wt%	2.8	1.2	--
Settled solids shear strength, $dyne/cm^2$	--	--	20
Inter. settl. rate, cm/h	2.3	4.8	12
Ultimate settled solids height, % total height	42	23	21
Heat capacity, cal/gm °C	--	--	0.88
Boiling point, °C	--	--	101
pH	12.4	13.0	11.3

(a) Fow, et al. 1986.
(b) Larson, et al. 1986.

Table 3.2 Physical Properties of Simulated SRAT Slurry

Feed Description	100 g WO/L without formic acid addition	100 g WO/L with 0.092 g formic/ g WO	109 g WO/L without formic acid addition
Data Source	(a)	(a) and (b)	(b)
Physical Properties			
Apparent viscosity (cP) at various shear rates s^{-1}			
10 (50 RPM agitator)	--	--	94
25 (130 RPM agitator)	--	--	43
183	12	7.4	13
Yield stress, dyne/cm^2	10	5	11
Consistency index, cP	--	--	--
Flow behavior index	--	--	--
Slurry density, g/cc	1.08	1.07	1.09
Total solids, wt%	12	12	12
Total susp. solids, wt%	--	--	--
Settled solids shear strength, dyne/cm^2	23	27	--
Inter. settl. rate, cm/h	--	0.2	--
Ultimate settled solids height, % total height	--	66	--
Heat capacity, cal/gm °C	0.83	0.92	--
Boiling point, °C	101	101	--
pH	11.3	6.7	11.3

(a) Larson, et al. 1986.
(b) Blair, Pulsipher, and Farnsworth 1986.

Table 3.2 (cont.)

Feed Description	145 g WO/L w/o formic acid (DWPF SRAT slurry)	151 g WO/L with 0.092 g formic acid per g WO & (DWPF SRAT slurry)	198 g WO/L w/o formic & (216 g WO/L w/ 0.092 g formic per g WO)
Data Source	(a) and (c)	(a), (b) and (c)	(a) and (b)
Physical Properties			
Apparent viscosity (cP) at various shear rates sec^{-1}			
10 (50 RPM agitator)	390	420	2,100 (2,100)
25 (130 RPM agitator)	170	180	1,100 (940)
183	40	34	198 (176)
Yield stress, dyne/cm^2	39 (<10)	43 (<10)	290 (239)
Consistency index, cP	--	--	--
Flow behavior index	--	--	--
Slurry density, g/cc	1.13 (1.12)	1.13 (1.10)	1.17 (1.19)
Total solids, wt%	16.1 (20.3)	17.3 (25.8)	21.0 (23.5)
Total susp. solids, wt%	--	--	--
Settled solids shear strength, dyne/cm^2	--	85.7	(481)
Inter. settl. rate, cm/h	--	0	--
Ultimate settled solids height, % total height	--	97	--
Heat capacity, cal/gm °C	--	0.80	--
Boiling point, °C	--	102	--
pH	11.3 (5.5)	6.4 (5.4)	--

(a) Larson, D. E., et al. 1986.
(b) Blair, Pulsipher, and Farnsworth 1986.
(c) House, C. M. 1986.

Table 3.3 Physical Properties of Simulated HWVP and DWPF Melter Feed. (HWVP waste oxide:glass oxide ratio of 1:3)

Feed Description	HWVP melter slurry at 390-430 g TO/L	HWVP melter slurry at 480-550 g TO/L	DWPF melter slurry
Data Source	(a) and (b)	(a) and (b)	(c)
Physical Properties			
Apparent viscosity (cP) at various shear rates s^{-1}			
10 (50 RPM agitator)	80 - 100	530 - 850	1,000 (est.)
25 (130 RPM agitator)	30 - 60	250 - 370	440 (est.)
183	9 - 14	22 - 36	95 (est.)
Yield stress, dyne/cm^2	7 - 9	17 - 58	100
Consistency index, cP	3 - 9	--	--
Flow behavior index	--	--	--
Slurry density, g/cc	1.27 - 1.30	1.31 - 1.34	1.33
Total solids, wt%	33 - 36	40 - 41	48
Total susp. solids, wt%	--	--	--
Settled solids shear strength, dyne/cm^2	220	--	--
Inter. settl. rate, cm/h	1.6	0.2	--
Ultimate settled solids height, % total height	75	85	--
Heat capacity, cal/gm °C	0.76	--	--
Boiling point, °C	101	101	--
pH	6.5 - 8.3	6.5 - 8.4	5.6

(a) Larson, D. E., et al. 1986.
(b) Blair, Pulsipher, and Farnsworth 1986.
(c) House, C. M., 1986. HWVP/DWPF Technology Exchange Meeting

Table 3.4 Physical Properties of DWPF Feed

Feed Description	Formated sludge	Melter feed	Frit and water
Data Source	(a) and (b)	(a) and (b)	(a)
Physical Properties			
Consistency index, cP	5 - 12	10 - 40	1
Yield stress, dyne/cm^2	15 - 50	25 - 150	None
Density, g/cc	1.12 - 1.21	1.29 - 1.46	1.1 - 1.2
Total solids, wt%	18 - 25	40 - 50	20 - 30
Transfer velocity, ft/sec	3 - 10	3 - 10	4 - 10
Heat capacity, cal/gm °C	0.73 - 0.90	0.65 - 0.79	--

Feed Description	Formated sludge	Melter feed
Data Source	(c)	(c)
Physical Properties		
Particle size, μm	approx. 10	-80 to +325 mesh
Density, g/cc	1.6	2.43

(a) DWPF Basic Data Report 1985.
(b) Martin 1983.
(c) Bechtel specification for DWPF, job no. 13239, spec. M-13, App. B, Rev 1.

The general physical properties are as follows: the process slurry specific gravity ranges from 1.0 with low solids to 1.4 with high solids. Before formic acid addition the slurry pH is from 11.5 to 13.5 and after formic acid addition the pH is from 4 to 6. The total solids varies from 2 to 50 wt%. The settling velocity for the slurry particles is very low while the frit particles have a settling velocity up to 0.5 ft/min in 1% formic acid.

The available DWPF slurries' physical properties do not demonstrate any significant differences from simulated HWVP slurries' physical properties. However, rheological properties of the DWPF and HWVP simulated feeds are dissimilar. The following section introduces slurry rheology by describing how slurry rheology is interpreted and measured and outlines differences between DWPF and HWVP slurries.

3.2.1 Slurry Rheology

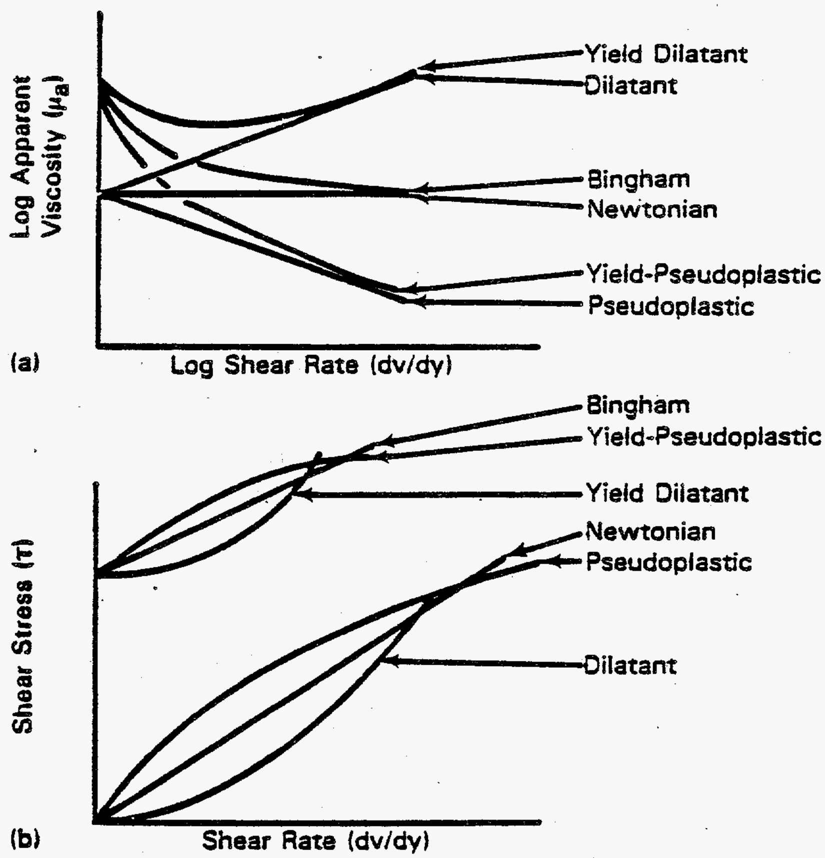
Mixtures of solids and liquids are characterized as either homogeneous or heterogeneous. Homogeneous slurries are composed of small particles (generally less than 40 μm) that can be treated as a uniform fluid. The HWVP waste slurry particles are less than 30 μm and the waste slurries are homogeneous. The small particles' settling velocity is too low to create stratification during normal processing. However, large agglomerates of particles will settle rapidly in the waste slurries and collect in low velocity regions of a vessel. Heterogeneous slurries contain particles whose settling velocities are high enough to cause stratification. The frit particles range up to 180 μm in size and have a settling velocity high enough to cause stratification. Frit in 1% formic acid has the highest settling velocity and the process frit slurries are heterogeneous. The melter feed (composed of frit and waste slurries) is homogeneous during normal processing based on laboratory settling times. This is due to the high viscosity and small particle size of the melter feed slurries.

The rheology of slurries is important for analyzing mixing in the feed preparation vessels, particularly at the low shear rates produced in an agitated tank (1 to 30 s^{-1}). Slurry viscosity can be affected by the solution temperature, % solids, particle size and shape, and solution chemistry. In addition, the HWVP feed preparation process is a series of batch operations; consequently, the properties of the slurries are constantly changing. These changing fluid properties can have different effects on processing. For example, dilute slurries with low viscosities (less than 10 cP) may form a deep vortex during agitation which could affect pump performance and level detection. Later during the processing, concentrated slurries with a high viscosity (greater than 400 cP at 15 s^{-1}) may reduce homogeneity and the heat transfer rate. Concentrated slurries exhibit non-Newtonian fluid behavior and the models and analysis used to describe rheology are explained below.

Viscous fluids are classified according to their shear stress response to an applied shear rate as shown in Figure 3.2. Various models are used to describe the shear stress/shear rate response (rheograms). Fluids whose shear stress responds linearly to shear rate are Newtonian. The slope of the line is the fluid viscosity. Fluids with shear stresses that respond non-linearly to an applied shear rate are non-Newtonian; various models are used to describe this non-linear response. These models are described in Figure 3.2. The viscosity of Newtonian fluids does not change with the shear rate, while the viscosity of non-Newtonian fluids does change with the shear rate.

Most slurries in the feed preparation system are defined as pseudoplastic, yield-pseudoplastic, or Bingham. For pseudoplastic and yield-pseudoplastic material, the flow behavior index (n) is the power to which the shear rate is raised and indicates the deviation from Newtonian behavior. In addition, yield-pseudoplastic and Bingham fluids require a shear stress to be applied before the material will flow. This is exhibited in the rheogram by an off-set from the origin. Once the required shear stress is applied, Bingham fluids will respond linearly with shear rate and yield-pseudoplastic fluids will respond non-linearly. The apparent viscosity (μ_a) of these fluids increases with decreasing shear rate. This behavior is especially non-linear in the regions of low shear rate from 1 to 50 s^{-1}.

Measuring slurry viscosities with consistency is difficult and depends on the type of instrument and how the data is analyzed. Agitated tanks typically have low shear rates, from 10 to 30 s^{-1} near the impeller down to 1 to 5 s^{-1} in remote areas of the vessel. In contrast, the shear rates in pipe flow are much higher, usually above 60 s^{-1}. Pseudoplastic behavior is important in agitated tanks because non-Newtonian performance is most pronounced at low shear rates. Therefore, measuring and comparing viscosities at low shear rates best represents the fluid behavior in an agitated vessel. In addition to



Rheograms for Various Fluids

Equations for various fluids

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

FIGURE 3.2 Rheograms and Models for Various Fluids

comparing viscosity at appropriate shear rates, it is important to appreciate differences between instruments. An absolute viscometer uses a different concept than a relative viscometer (see Appendix A for a description of various viscometers). It is inappropriate to directly compare slurry viscosities from different types of viscometers. Due to analytic variation, equipment suppliers perform their own analysis on slurry samples. This allows them to control the analysis technique and compare results to their past experience.

The SRL analyzes rheograms differently than PNL, which can affect the interpretation of test results. Figure 3.3 shows a typical rheogram and how it would be interpreted by both SRL and PNL. SRL assumes a Bingham fluid and extrapolates a straight line back to the y-axis (shear stress) from the linear portion of the rheogram. This interpretation creates data that can be employed to predict pipeline pressure drop using the Buckingham-Pi equation. This method is appropriate for predicting apparent viscosity at high shear rates. PNL assumes a pseudoplastic fluid and tries to fit the curve of the data with a power law. A power law prediction produces a more accurate prediction of apparent viscosity in the low shear region. Assuming a Bingham plastic will create a more conservative estimate of the slurry viscosity in the low shear region. However, this can create problems when using SRL test data to evaluate mixer performance with HWVP feeds. Testing at SRL could indicate slurry homogeneity is acceptable with a viscosity of 500 cP at 20 s^{-1} shear rate, if the slurry is interpreted as a Bingham fluid. Yet the same material interpreted as a pseudoplastic fluid may only have a viscosity of 100 cP at 20 s^{-1}. Using SRL test results and rheology would indicate slurry homogeneity with a 500-cP fluid, but in reality the slurry may only be 100 cP at low shear rates and a 500-cP pseudoplastic slurry may not be homogeneously mixed.

Another important characteristic of a slurry is its potential for foaming. Foaming of slurries has occurred during testing at the TNX facility of the DWPF and has caused problems in achieving a high solids DF between the slurry and condensate. The following section discusses foaming.

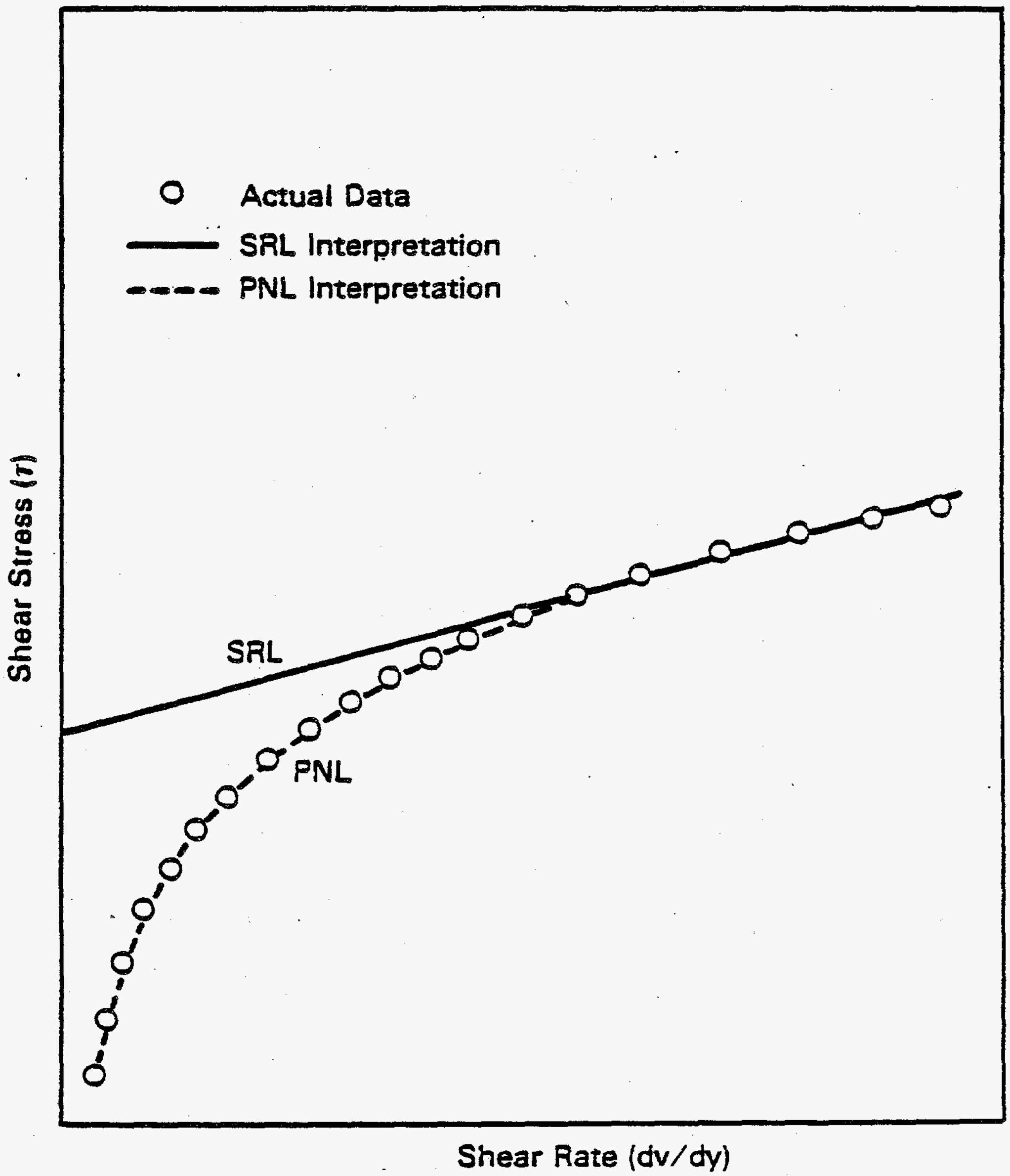


FIGURE 3.3 Comparison of SRL and PNL's interpreptation of rheograms

3.2.2 Foaming

Foaming is a physical property that can be induced or prevented by processing conditions. A fluid can have the potential for foaming, but if the processing does not induce foaming then it will not occur. Foaming is an interfacial phenomenon that can hamper pumping and increase solids entrainment in off-gas equipment. The creation or stability of foam in fluids cannot be accurately predicted without experimental data. It is known that finely divided solids, some surface active organics, and high viscosity fluids can stabilize foam and some organics can increase the creation of foam. In the feed preparation area foam is generated by boiling, agitation, or reaction. The complex chemistry, fine solids, and high viscosity of the feed preparation slurries increases the possibility of a stable foam. This complexity makes it difficult to predict the degree of foaming based on analytic data or bench-scale studies.

3.3 PERFORMANCE REQUIREMENTS

Equipment performance requirements can be expressed in terms of fundamental quantities that are then translated into measurable process results. The essential performance requirements are the evaporation and condensing duty, heating and cooling rate, degree of homogeneity, process line slurry velocity, and equipment design life. Table 3.5 lists HWVP equipment performance requirements. The design life (HWVP FDC 1987) is 20 years for replaceable components and 40 years for non-replaceable components.

Evaporator duty is described in terms of a heat rate in BTU/h and includes the amount necessary to evaporate water at the design rate and compensate for the hot vessel heat losses. The condenser duty is calculated using the design evaporation rate and assuming total condensation and subcooling of the vapor. The overall required heat transfer coefficient (U) for the helical coils or condenser can be calculated using the design heating or condensing rate, heat transfer surface area, and temperature difference (ΔT) for a specific application. The calculations for estimating the duties are in appendix B, pages B-1 to B-5.

The heating and cooling rates are calculated using the heat capacity, density, and volume of the fluid with the required cooling or heating rate in °F/h.

The fundamental measure of homogeneity is to determine the uniformity of the temperature, solids distribution, and chemical composition. For example, if chemical uniformity is required, then samples from various tank locations should be analyzed and compared. A common analytic method developed by Chemineer, Inc. estimates bulk fluid velocities to predict fluid homogeneity in tank designs. This analysis can be used if physical property data is available or a standard tank is being designed.

The following section evaluates the adaptation of the DWPF feed preparation equipment to the HWVP.

Table 3.5 Performance Requirements of the Feed Preparation Equipment

Vessel	SRAT	SME	MFT	SMECT	SFHT	FSMT
Overflow Capacity (gal)	11,000	11,000	11,000	11,000	6,200	2,800

Vessel Coils	Evaporator Design Duty, 10^6 BTU/h	Required U_{boil} (DWPF),[a] BTU/h ft^2 °F	Convective Cooling/Heating Design Rate, °F/h	Required U_{cond} (DWPF), BTU/h ft^2 °F
SRAT/SME	5.5	110 (120)	10	65 (70)
MFT	---	---	--	10

Condenser	Condenser Design Duty, 10^6 BTU/h	Required U_{cond} (DWPF) BTU/h ft^2 °F	Condensing Rate, gpm	Outlet Vapor Temp., °F	Condensate Temp., °F	Solids DF (DWPF)
SRAT/SME	5.33	140 (127-Fouled) (180-Clean)	10	120-150	N/A	(10,000)
FAVC	0.072	10	--	60-80	N/A	--

Pumps	Superficial Velocity ft/s	Pump Rate gpm
Vessel transfer, sample and melter feed recirculation	3 - 10	100
Melter feed loop, each	0.2 - 1.3	0.10 - 0.75

U_{boil} estimated using a 150 psig coil steam pressure, design coil surface area of 340 ft^2, and 10 gpm evaporation rate.

U_{conv} estimated assuming a cooling water (C.W.) temp. of 90°F and a log mean ΔT, and an area of 140 ft^2 (SRAT/SME) or 320 ft^2 (MFT).

U_{cond} estimated using a chilled water temp. of 45°F, C.W. temp. of 90°F, and an area of 350 ft^2 (SRAT/SME) or 190 ft^2 (FAVC).

Melter feed loop superficial velocity calculated using 3/8 in. sch 40 pipe.

4.0 EVALUATION OF THE PROCESS PERFORMANCE OF THE FEED PREPARATION EQUIPMENT

The purpose of the evaluation of the DWPF feed preparation system is to assess the adaptability of DWPF equipment using HWVP feeds and process requiremnts. The objective of this section is to identify the feed preparation equipment that requires testing to demonstrate process performance. A test plan is proposed in Section 5 for equipment requiring process verification.

The feed preparation equipment will be evaluated in groups with similar process performance: the SRAT/SME/MFT vessels, SRAT/SME condensesr, pumps, sampling systems, SFHT, and FSMT. The evaluation will incorporate DWPF development work and equipment testing, engineering relationships, Hanford process experience, and PNL development work. The feed preparation equipment design will be assessed in its ability to meet HWVP design requirements of heat transfer, fluid homogeneity, production requirements, pump transfer rates, and equipment design life.

Section 4.1 discusses the SRAT/SME/MFT vessels. Section 4.2 analyzes the condensers, Section 4.3 evaluates the pumps, Section 4.4 assess the sampling system, and Section 4.5 discusses the SFHT and FSMT. In general, each evaluation will begin with an introduction to the equipment and technical background, followed by an evaluation of full-scale testing at DWPF, an analysis of applicable PNL development work, and an engineering assessment.

4.1 SLURRY RECEIPT AND ADJUSTMENT TANK, SLURRY MIX EVAPORATOR, AND MELTER FEED TANK

The geometry of the DWPF design of the SRAT, SME, and MFT is shown in Figure 4.1. The MFT has only two helical coils that use cooling water for removal of decay heat. The MFT helical coils are the same as the two inner coils shown in the figure. The similarity of the vessels' internals means the analysis for fluid mixing will apply to the SRAT, SME, and MFT. Even though the MFT has only two helical coils (compared to three helical coils in the SRAT and SME), the difference will have minimal impact on overall fluid mixing in the vessel. Accurate level indication is required for precise continuous process control and mass and actinide balances. The following three sections evaluate fluid mixing, heat transfer, and level indication in the SRAT, SME, and MFT.

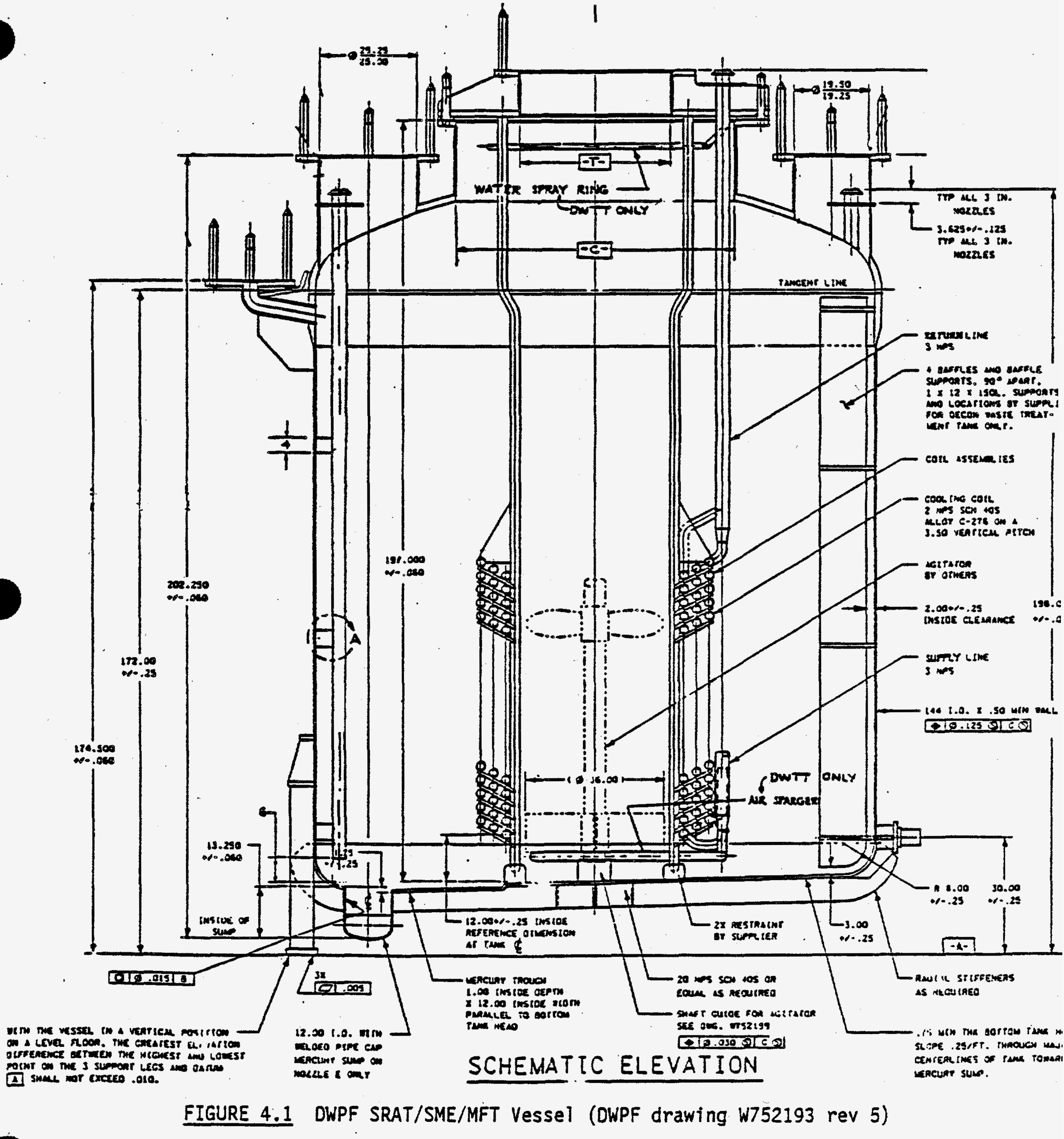


FIGURE 4.1 DWPF SRAT/SME/MFT Vessel (DWPF drawing W752193 rev 5)

4.1.1 Fluid Mixing

The primary purpose of the agitator in the SRAT, SME, and MFT is to provide well-mixed vessel contents; according to the HWVP Functional Design Criteria (Rev. 2) this means homogeneous temperatures, solids, and chemical composition. The homogeneity of the slurry must meet process control requirements for tracking radionuclides and waste form qualification.

The following sections discuss fluid mixing in the SRAT/SME/MFT in terms of full-scale testing (Section 4.1.1.1), bench-scale testing (4.1.1.2), and engineering analysis (Section 4.1.1.3).

4.1.1.1 Full-Scale Testing

The SRL test facility (TNX) has conducted several full-scale tests (campaigns) of a prototypical SRAT/SME vessel using DWPF simulated feeds. During TNX campaign #4 fluid homogeneity was measured by sampling at two different vertical heights in the same radial position using a Hydraguard sampler (Caplan 1987). A third sampler was used as a standard and extracted a sample near the bottom of the tank approximately 180° from the Hydraguard samplers and in the same radial position. The radial position was in the annular region between the vessel wall and outer coil and the slurry level was above the top of the coils.

The samples were analyzed for solids content. With the simulated SRAT slurry, the sampling error was 0.7% while the analytic error was 6.4% for an overall error of 7.1%. With the simulated melter feed the sampling error was 2.4% and the analytic error was 6.4% for an overall error of 8.8%. DWPF expects the solids sampling error to be less than 10%. The test indicated slurry homogeneity at different levels in the SRAT/SME using DWPF feeds in full-scale equipment; however, slurry homogeneity was not tested at different radial positions in the SRAT/SME or for low slurry levels. Data from SRL testing can be adapted for evaluation if the slurries have similar physical properties, the mixing Reynolds number (N_{Re}) is above 1,000, and fluid homogeneity is assessed with an acceptable sampling method.

Viscosity is the most important fluid property in comparing slurry properties for evaluation. However, the viscosities of slurries vary considerably during batch concentration operations and are influenced by the chemical composition and particle distribution and size. The uncertainty in determining viscosity is compounded when comparing slurry viscosities between laboratories with different viscometers and methods. This situation is discussed in more detail in Section 3.2.1. In essence, SRL assumes a conservatively high viscosity and consequently the mixer performance may be incorrectly assumed to produce a homogeneous slurry at high slurry viscosities.

One procedure for analyzing the affect of different slurry viscosities on mixing is to compare the mixing N_{Re}. The flow patterns in the vessel will begin to change as the mixing system approaches the upper end of the transition region, or near an N_{Re} of 1000. The mixing N_{Re} uses the fluid viscosity at the impeller shear rate, which is approximately 25 s^{-1} at 130 rpm. The viscosity at this low shear rate is not directly measured by the DWPF. As described in Section 3.2.1, DWPF defines their slurries as a Bingham plastic and use the data in the linear portion of the rheogram at high shear rates to extrapolate the viscosity at low shear rates.

Figure 4.2 shows the viscosities of different DWPF and HWVP slurries as a function of shear rate. While the slurries have similar viscosities, the previous discussion concerning uncertainties in measurement and interpretation of slurry viscosities should be retained when observing the figure. SRL campaign #4 melter feed had a lower viscosity compared to the HWVP simulated melter feed; the apparent viscosity of the HWVP melter feed was one-third greater than the SRL melter feed. This difference in apparent viscosities lowers the mixing N_{Re} from 8200 using the DWPF feed to 6300 using the HWVP feed. This drop in N_{Re} is not significant, but comparing mixing N_{Re} does not reflect the fluid behavior in the peripheral low-shear regions of the vessel. The HWVP slurries in the feed preparation area are pseudoplastic, therefore they become more viscous at lower shear rates. The HWVP melter feed is more viscous than the DWPF feed that was tested. This means the flow produced by the impeller will not extend as far and the peripheral stagnant region at the vessel wall will expand. The extent of the increase in stagnant slurry cannot be calculated.

The projected design life was below specification for the bottom impeller, coils and coil supports, and the vessel bottom due to erosion or abrasion (Good 1987). The bottom radial flow impeller was severely eroded at the top and bottom near the impeller shaft and its projected life was 2.5 years. The suspected cause of the impeller wear was cavitation erosion and turbulent

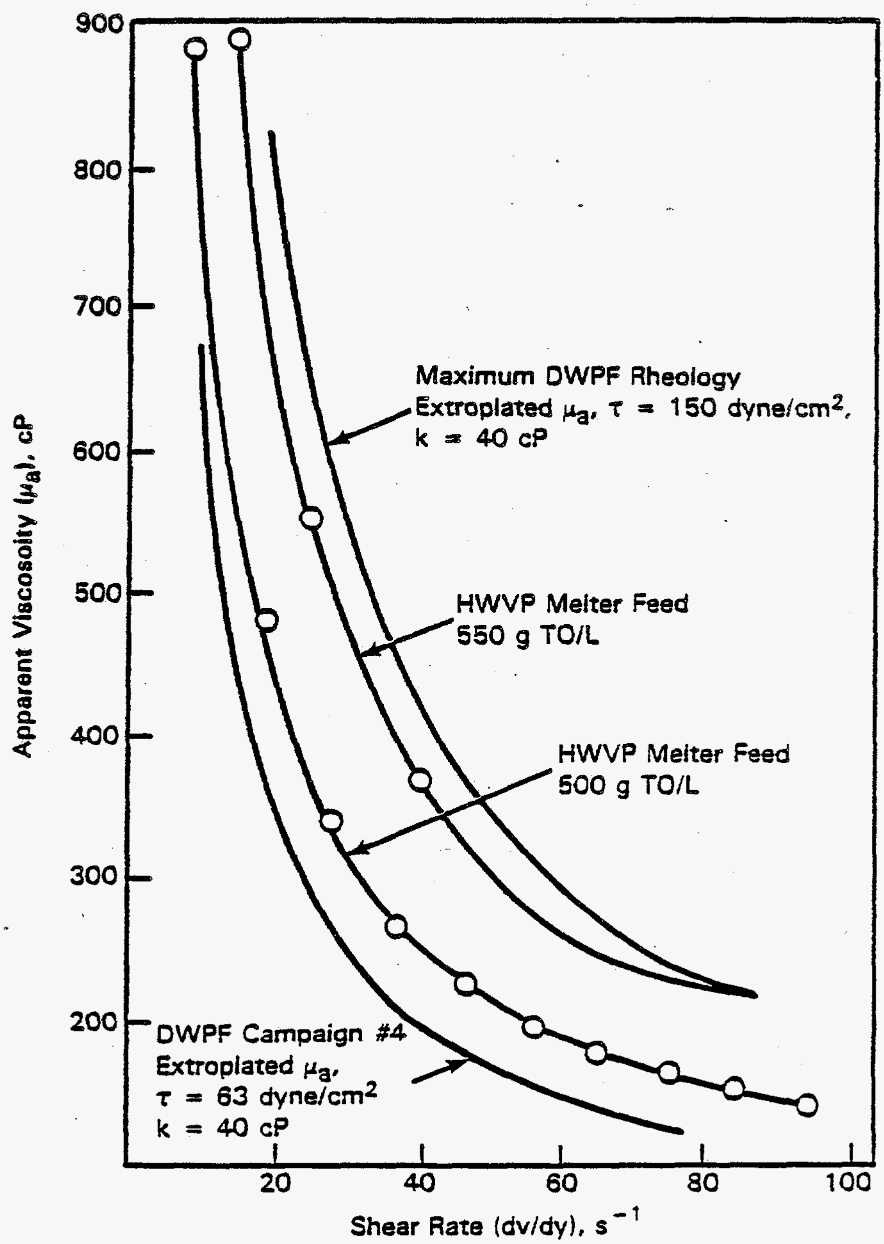


FIGURE 4.2 Apparent Viscosities for HWVP and DWPF Slurries

eddies. The cavitation is created by low pressure regions near the impeller blade as shown in Figure 4.3. In a boiling slurry, these regions of low pressure cause vaporization and bubbles are formed. As the bubbles are collapsed or swept away, the impeller blade is impacted by the slurry and quickly eroded. A pitched blade impeller does not exhibit the turbulence and low pressure regions of a paddle impeller and can be expected to have reduced wear. DWPF will be testing impellers that have a hardened stellite coating. However, the impeller coating will not improve erosion in other areas of the tank.

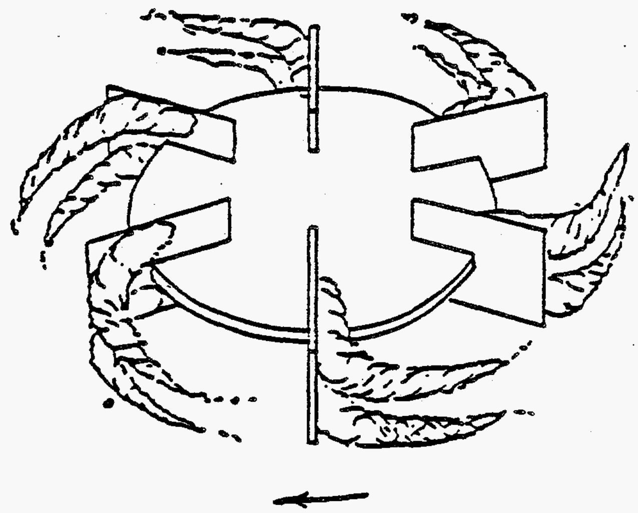


FIGURE 4.3 Gas Vortices from a Radial Flow Impeller (Nienow and Wisdom 1974)

The coils also showed erosion, mostly abrasion of metal-to-metal contact between the coils and supports and the expected design life was 2 to 20 years. The coil abrasion is aggravated by the slurry flow from the bottom flat blade impeller buffeting the coils and supports. The tip of the bottom impeller is moving at 20 ft/s less than 5 in. from the lower coils. DWPF is addressing this situation by testing coils that are welded to the supports. Erosion caused by slurry flow from the bottom impeller will not be reduced by welding the coils to the supports or coating the impeller with stellite .

The vessel bottom had a projected life of 8 years, compared to the design life of 20 years. In addition, a coil support guide was broken during testing. DWPF will be testing a stellite to stellite contact between the coil support and guide.

4.1.1.2 Bench-Scale Testing

Fluid mixing studies were performed (Peterson, McCarthy, and Muhlstein 1986) using representative HWVP melter feed slurries in a 1/10-scale SRAT/SME vessel. The bench-scale vessel included helical coils and dual impellers. Fluid homogeneity was determined by measuring the wt% solids from slurry samples taken at different locations in the vessel. Slurry homogeneity was indicated if the wt% solids were statistically identical between various levels. Several factors were studied to measure their impact on fluid homogeneity. The factors included the type of impeller, impeller speed, and wall baffles. Impeller speed was the most significant factor with lower impeller speeds producing less homogeneity. During the tests, wall baffles improved fluid homogeneity at low impeller speeds. Larger diameter impellers in open tanks (without heat transfer coils) also produced more uniform slurries at lower impeller speeds. The study indicated simulated HWVP slurries can be uniform using the DWPF design vessel if the impeller speed is comparable to DWPF agitator speeds.

4.1.1.3 Engineering Analysis

A fluid mixing analysis estimated a bulk fluid velocity of 18 to 40 ft/min for the SRAT/SME/MFT based on the assumptions below. This level of agitation indicates the SRAT/SME contents will be homogeneous assuming a Newtonian fluid, which is valid for dilute slurries. The analysis is described in Appendix A and the calculations are in Appendix B, pages B-10 to B-12.

The analysis assumed that 1) the bottom impeller supplied all the impeller flow, 2) the flow from the bottom impeller was restricted 50% due to the coils and low bottom clearance, and 3) the fluid was Newtonian. The first and second assumptions were developed through conversations with fluid mixing experts and agitation equipment suppliers; the third assumption is a basis of the calculation method.

The first assumption was based on discussions with personnel at Philadelphia Mixers, the supplier for the DWPF agitation equipment. As part of the DWPF acceptance procedure, Philadelphia Mixers operates the agitator and motor in a basin of water. During these tests additional flow was not created by the top axial flow impeller in the SRAT/SME/MFT agitator design. Furthermore, the coils form a draft tube and the top impeller of a draft tube does not increase total vessel flow (Oldshue 1983), it simply pushes the fluid to the bottom impeller.

The flow from the bottom impeller is restricted to 50% of the normal flow because of obstruction from the coils and the low off-bottom clearance of the impeller. The upper half of the bottom radial flow impeller is obstructed with coils; furthermore, locating a radial flow impeller near the vessel bottom reduces the impeller flow 10 to 15% (Oldshue 1983). Finally, agitator power consumption recorded during TNX SRAT/SME testing was less than half of the calculated consumption (see Appendix B, pages B-12 to B-14). The reduced power consumption indicates a lower flow is being produced by the impellers since agitator power is proportional to impeller flow.

The assumption of a Newtonian fluid is reasonable for a slurry with a low solids content, but less accurate as the solids content is increased. Viscosity affects only the calculation of N_{Re}; as long as the impeller operates in the turbulent region than the flow number (N_P) is fixed. However, the analysis procedure is based on experimental tests using Newtonian fluids, and non-Newtonian fluids will not have the same overall mixing pattern. Non-Newtonian fluids will generally require higher bulk fluid velocities.

4.1.1.4 Agitator Power Requirements

The DWPF design specifies a 100-hp motor for the agitator. This motor size is two to three times larger than necessary during normal operation as indicated by engineering evaluation and DWPF operating experience. The motor sizing is based on start-up in settled solids. The excessive motor size results in additional capital and fabrication cost. Philadelphia Mixers (the DWPF agitator suppliers) commented that the large motor is difficult to fabricate within the space requirements specified by DWPF. A smaller motor would simplify fabrication and reduce installation costs.

Equipment suppliers suggested 20- to 40-hp motors as conservative estimates (Schumacher 1987) of the expected agitator motor power consumption. An engineering evaluation estimated a power requirement of 50 hp for the DWPF agitator motor during normal operation (see calculation in Appendix B, pages B-12 to B-14). Feed preparation system campaigns at TNX confirm the low power requirement for the agitator during normal operation; the power consumption did not exceed 20 hp during TNX feed preparation campaign #4. Testing at TNX has not revealed a settling problem and agitator start-up in settled solids has not been tested.

The DWPF agitator motor was based on start up in settled solids, with a motor power based on a settled solids yield stress of 600 dyne/cm^2. No engineering relationships could be found using yield stress to calculate power consumption. In addition, equipment suppliers do not use the yield stress of settled solids to estimate power consumption. The yield stresses of simulated HWVP settled solids are below 250 dyne/cm^2 (see Table 3.3), therefore the power consumption in HWVP feeds should be lower. However, without knowing how to estimate agitator power consumption using yield stress, the start-up power consumption in HWVP feeds cannot be determined without experimental data.

The problem of settled solids in the fluid mixing industry is normally addressed using specific design elements. Examples of these design additions are an air jet to loosen settled solids or an initial slow rotation of the impeller to fluidize the solids around the impeller.

4.1.1.5 Conclusion

There is no evidence to suggest the HWVP slurry in the SRAT/SME/MFT vessel will be inhomogeneous during normal operation. Therefore verification of slurry homogeneity would not be required. However, SRL has not tested slurry homogeneity at various radial positions and slurry levels. In addition, fluid blending depends strongly on viscosity and the slurry viscosity is non-Newtonian, difficult to measure precisely, and changes an order of magnitude during the feed preparation operation. These factors create uncertainty in determining the degree of homogeneity and using TNX testing results. The HWVP simulated melter feed slurry is more pseudoplastic and viscous than the DWPF slurry that was tested for homogeneity. The increase in viscosity at low shear rates may increase the amount of stagnant areas in the tank and reduce the homogeneity. To certify the degree of homogeneity required for process control and waste form qualification over the range of potential HWVP feeds may compel full-scale testing. The quality assurance requirements of accepting TNX test results for verification of HWVP design must also be addressed.

4.1.1.6 Recommendations

Improvements in DWPF design requiring minor modifications would be to change the type of impellers and use wall baffles. DWPF design uses an axial-flow high-efficiency impeller at the top and a radial-flow flat-blade impeller at the bottom. Literature references and equipment suppliers suggest using two axial-flow pitched-blade impellers instead. Pitched-blade impellers would reduce impeller and coil erosion, improve slurry flow, and reduce agitator power requirements. The study by Peterson, McCarthy, and Muhlstein (1986) indicated using baffles improved homogeneity and standard practice with low viscosity fluids is to employ baffles (Oldshue 1983).

The agitator motor size should be reduced. The DWPF agitator power requirements are larger than necessary and increase fabrication and capital costs. The agitator power consumption will be measured in the feed preparation test facility.

As discussed with WHC HWVP Technology and Engineering staff, the SRAT/SME/MFT vessel design could be improved while maintaining the external vessel dimensions and the concept of an agitator and immersed coils. These modifications address process performance and erosion; plant costs and remote operation would have to be addressed in a more detailed analysis. Using a more conventional coil and agitator design could reduce impeller and vessel erosion, lower agitator power consumption, and improve fluid mixing. Standard coil design uses a large-diameter helical coil located near the vessel wall. Another design that is used occasionally is several immersed small diameter coils placed around the periphery of the tank. The impeller is normally an axial flow impeller with an impeller diameter-to-tank diameter ratio greater than 0.4. The larger impeller diameter means a lower impeller speed can be used to achieve the same degree of mixing. An axial-flow impeller at a lower impeller speed will have less erosion than a flat-blade impeller. This is because an axial-flow impeller does not create a low pressure region like a flat-blade impeller. The vortex at the low pressure region causes turbulence and erodes the impeller blade. In addition, an axial flow impeller produces a more uniform velocity pattern in the vessel. Uniform slurry velocities would reduce vessel and coil erosion by decreasing the magnitude of turbulent eddies. Mechanical abrasion would also be reduced since the coils would not be buffeted by slurry from a close proximity impeller. Axial impellers are more efficient than flat blade impellers and therefore consume less power. Large-diameter impellers would also improve agitation of the slurry because the impeller would be directly mixing a majority of the vessel contents.

4.1.2 Heat Transfer

Two types of heat transfer occur in the SRAT/SME vessels: forced convection and boiling heat transfer. Forced convection consists of heating and cooling while boiling heat transfer involves vaporization. The technical background is described in more detail in Appendix A.

The immersed helical coils in the SRAT/SME are required to heat or cool the slurry at 10 °F/h (convective heat transfer requirement) and evaporate 10 gpm from HWVP slurries (boiling heat transfer requirement). The evaluation of convective heat transfer and boiling heat transfer in the SRAT/SME will be considered separately.

4.1.2.1 Convective Heat Transfer

TNX full-scale testing using DWPF feeds has demonstrated that the required overall convective heat transfer rate is attainable, but some results have been lower than required. Overall convective heat transfer coefficients (U_{conv}) from 54 to 140 BTU/h ft^2 °F have been achieved in testing at the TNX facility using a full-scale SRAT/SME vessel and simulated DWPF feeds (House 1986). These heat transfer coefficients were achieved on clean coils; fouled coils will have lower heat transfer coefficients. During one test, the vessel coils were not cleaned after the vessel contents were drained and an air-dried layer of solids was allowed to form. The boiling heat transfer rate following the application of the air dried solids layer consequently dropped dramatically, but after a few hours the full heat transfer rate was re-established (Weber 1982). The air dried solids were probably removed by abrasion caused by agitation and the effect of boiling. The affect of several consecutive feed preparation cycles on heat transfer has not been evaluated. Testing with several feed preparation cycles is necessary to allow evaluation of the amount of fouling and its affect on heat transfer.

Convective heat transfer is being evaluated at PNL using a 1/10 scale DWPF design SRAT/SME vessel; the HWVP report concerning the testing will be issued in September 1987 by R. K. Nakaoka. Bench-scale studies will identify important factors in heat transfer and will reduce the number of experiments necessary for full-scale testing. The convective heat transfer coefficient was 35 to 40 BTU/h ft^2 °F with simulated WVDP feeds; this heat transfer coefficient is below the required HWVP heat transfer coefficient of 65 BTU/h ft^2 °F. The heat transfer coefficient did not vary significantly with the impeller speed. The lower heat transfer coefficient achieved during small-scale testing could be due to differences in feed properties between the DWPF and the WVDP or a factor of scale down. Small-scale systems cannot match the fluid dynamics in a large-scale tank; the slurry flow past the coils in the small-scale tank is not identical to that in the large-scale tank. Differences in fluid flow could have reduced the small-scale convective heat transfer rate.

As described in Appendix A, the process-side heat transfer coefficient (h_p) can be predicted using a Nusselt relationship (see appendix B, pages B-15 to B-19). However, engineering relationships from the literature cannot be used to verify process performance. This is because the fluid properties of the slurries are unknown at the heat transfer surface and DWPF coil design has not been studied. No literature references could be found using the DWPF coil design; however, the predicted heat transfer coefficient for most coil designs is 50 to 150 BTU/h ft^2 °F. Engineering relationships predict the required HWVP convective heat transfer coefficient (85 BTU/h ft^2 °F) is achievable, though problems could occur with viscous material. Viscosity is the most important fluid property in convective heat transfer. The slurry-side heat transfer coefficient (h_p) is proportional to $(1/\mu)^{1/3}$; i.e., the heat transfer decreases with an increase in slurry viscosity. The effect of viscosity can be observed in Figure 4.4, which displays predicted h_p for a large diameter coil as a function of viscosity. In heat transfer relationships the viscosity at the surface is used. The difficulty in determining the viscosities of slurries is compounded when predicting the viscosity of a pseudoplastic slurry at a hot surface at an unknown shear rate. The uncertainty in estimating viscosity at the heat transfer surface prevents an accurate calculation of the convective heat transfer coefficient.

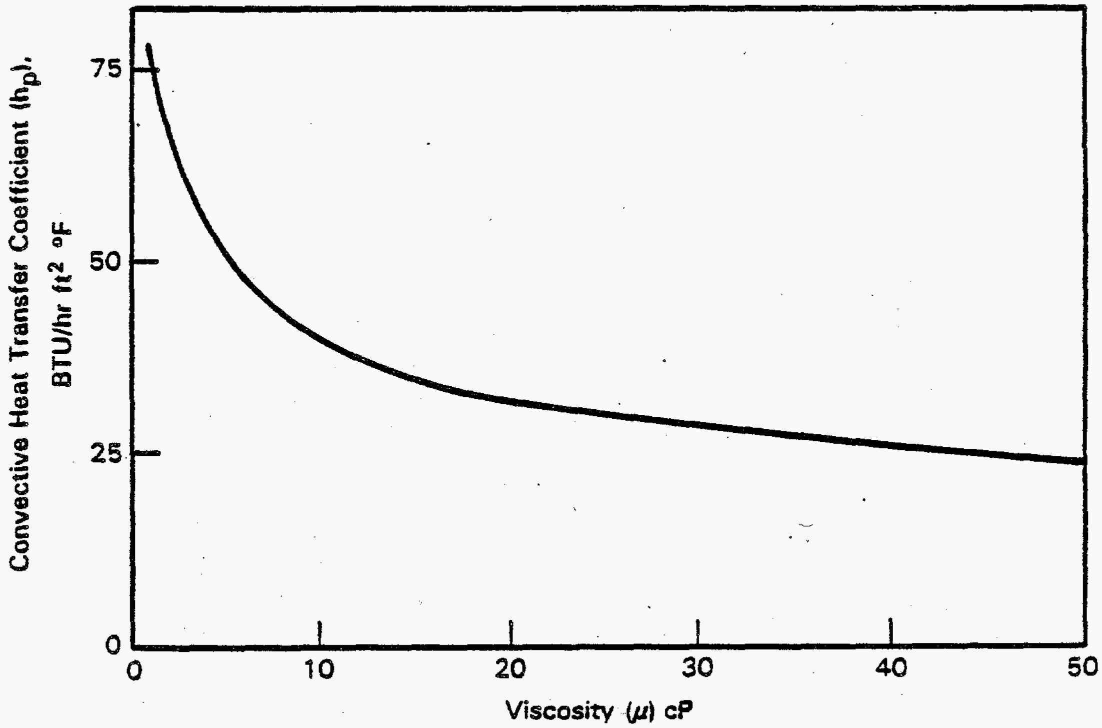


FIGURE 4.4 Affect of Viscosity on Convective Heat Transfer

Engineering relationships do not provide accurate estimates of heat transfer coefficients in complex systems. However, they are valuable in identifying important variables and their affect on equipment performance, and analyzing experimental data. Nusselt relationships are most accurate when confirmed by experimental data from vessel geometries identical to the actual design and using reference feeds. Greater uncertainty is introduced when specific predictive equations are adopted directly from the literature.

4.1.2.2 Boiling Heat Transfer

Boiling heat transfer has been tested at TNX using a full-scale SRAT/SME vessel and coils. The boiling heat transfer coefficient was 120 to 130 BTU/h ft^2 °F during TNX feed preparation campaign #3 (House 1986). This boiling heat transfer coefficient is only 10 to 20% above the required HWVP boiling heat transfer coefficient of 110 BTU/h ft^2 °F. The application of TNX test data to evaluate equipment performance using HWVP feeds is limited by two factors: the testing was performed with clean coils and with simulated DWPF feeds. Boiling of slurries in nested helical coils is a complex phenomenon that depends on numerous factors.

As described in Appendix A, boiling heat transfer is affected by the driving force (ΔT), equipment geometry, heat transfer surface metallurgy and condition, amount of dissolved gases, fluid viscosity and heat capacity, fluid flow across the coils, and fouling. The HWVP and DWPF feeds differ in their rheology and chemical composition; consequently, the expected boiling heat transfer rate on clean coils using HWVP feeds cannot be accurately predicted using full-scale TNX test data. The affect of fouling on boiling heat transfer has not been evaluated during testing at TNX. Any information that is produced at TNX could not be directly used in analyzing the affect of fouling on heat transfer with HWVP feeds. Fouling is a phenomenon that must be assessed using representative feeds because the mechanism for fouling may be different.

In a study currently being conducted at PNL using a 1/10-scale SRAT/SME vessel and coils, the effects on boiling heat transfer of ΔT and fluid properties on boiling heat transfer are being evaluated. The study will be valuable in determining the impact of important variables and estimating the overall boiling heat transfer coefficient. The bench-scale data will make full-scale testing more efficient by identifying optimum operating ranges and important fluid properties. However, boiling heat transfer is sensitive to the size of the equipment; for example, the bubble size is proportionately much larger than the line size in small-scale equipment. Therefore, development data cannot be used directly to verify full-scale equipment performance.

There are no accurate models or relationships for predicting boiling heat transfer in slurries using helical coils. Several relationships have been used to predict boiling heat transfer in pure fluids with simple surface geometries, i.e., single tube, or flat vertical or horizontal surface, etc. (Rohsenow, Hartnett, and Ganic 1985). But no literature reference was applicable to the feed preparation system. Standard industrial practice is to use rules of thumb based on previous experience. A conservative estimate used by Philadelphia Mixers is to assume the boiling heat transfer rate is equal to the convective heat transfer rate. Others (Oldshue 1987) multiply the convective heat transfer coefficient by a factor between 1 and 2; the factor is again based on experience. The experience available for predicting the HWVP performance comes from TNX testing of a feed with a different rheology and chemical composition than HWVP feed. This does not provide an adequate supply of information to predict HWVP process performance.

4.1.2.3 Conclusion

Full-scale testing of boiling and convective heat transfer using reference HWVP feeds is necessary. Testing is required because HWVP feeds are rheologically and chemically different than DWPF feeds, the effect of fouling over several cycles must be determined, and because boiling heat transfer cannot be verified without full-scale testing using simulated HWVP feeds.

4.1.2.4 Recommendations

The heat transfer rate of the DWPF design coils may be increased by using a pitched-blade impeller. A pitched-blade impeller will force more fluid through the coils than high efficiency impeller and may improve the heat transfer rate.

Using a larger-diameter helical coil (as described in Section 4.1.1.6) or several small diameter immersed coils could increase the heat transfer surface area. Additional heat transfer surface area would increase the concentration rate with the most certainty.

4.1.3 Level Measurement

Level measurement is critical for waste form qualification and for process control. The volume, specific gravity, and elemental concentrations of a slurry are used to inventory the chemical components within the feed preparation area. The volume of slurry in a vessel is determined by assuming that the slurry surface is horizontal and measuring the slurry height by using two differential pressure transmitters (DPT) as shown in Figure 4.5. Normally, the agitator is turned off to obtain accurate level indication. However, during continuous operation or with frit slurries this procedure may not be acceptable.

Dip tube bubblers are frequently used to measure liquid levels in radioactive service. Dip tube bubblers bleed a small flow of air out of a set of open tubes into the fluid. The tubes are at different fluid heights (as shown in Figure 4.5) which causes a change in the pressure on the air flowing into each individual tube. The differences in pressures are translated into a level measurement. DWPF design specifies a Holledge level sensor which uses the same principle as a dip tube bubbler, except that a thin metal diaphragm separates the air from the slurry. The air does not enter the slurry and the diaphragm prevents solids from plugging the tube. However, the response from a Holledge level instrument is non-linear and varies with the fluid temperature.

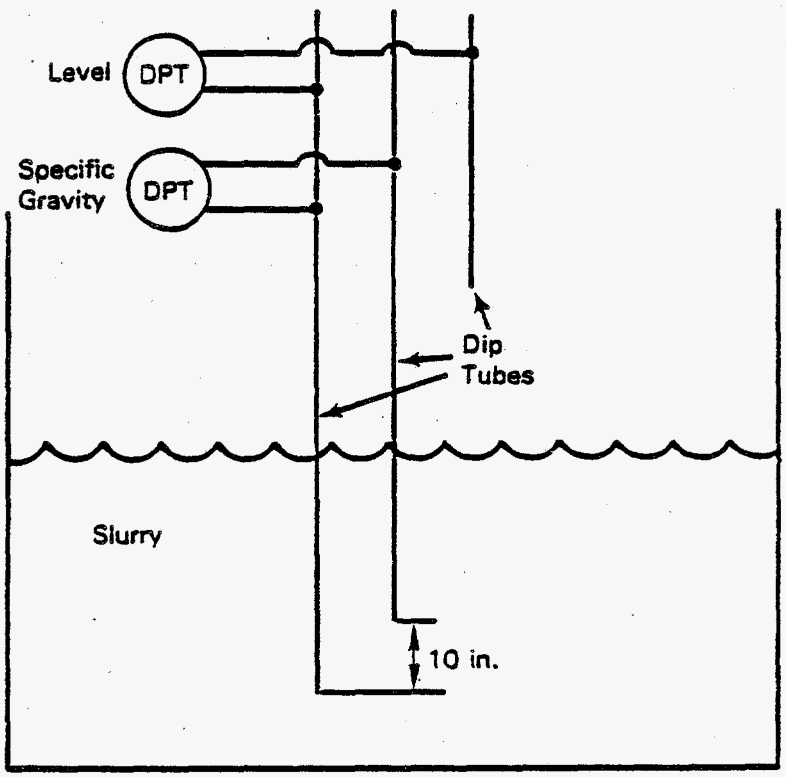


FIGURE 4.5 Level Measurement System

Vigorous agitation of slurries can cause surging surfaces and fluid jets or eddies that cause pressure fluctuations throughout the slurry. These pressure fluctuations in the slurry will affect the DPTs used for determing the slurry level; the level measurement will be erratic. In addition, if a vortex is formed on the slurry surface the estimate of the slurry volume will be inaccurate because the surface will not be horizontal.

During agitation in campaign #5, the Holledge level sensor on the full-scale SRAT/SME vessel at the TNX facility varied ±12 to 13%. The Holledge level indication was compared to that determined using a weighted string and a measuring tape. The dip tube bubbler was also variable. Both the dip tube bubbler and Holledge were accurate if the vessel was not agitated. In addition, the Holledge level sensor had a hysteresis with a variation in temperature. The Holledge level sensor and a dip tube bubbler level system will be used in the test facility to determine the slurry levels.

4.2 CONDENSERS

The most rigorous condenser performance requirement is for the SRAT/SME condenser; therefore it will be evaluated exclusively. If the SRAT/SME condenser operates successfully, the other condensers are expected to function properly. The geometry of the SRAT/SME condenser is shown in Figure 4.6. The SRAT/SME condenser is designed to condense 10 gpm of condensate generated from the evaporator helical coils. The SRAT/SME condenser, in combination with the SRAT/SME vessel, must achieve a high solids DF between the feeds and condensate. DWPF design expects a DF of 10,000 (10^4); the HWVP Reference Conceptual Design Report process flow diagram uses a DF of 800 (see appendix B page B-20). However, the expected DF is 1000.

Condenser heat transfer coefficients (U_{cond}) of 110 to 140 BTU/h ft^2 °F were obtained during full-scale testing at TNX. The experimental values do not necessarily reflect the maximum U_{cond}. The vapor load to the condenser at the TNX facility produced by the helical heat transfer coils may not generate enough vapor to overload the condenser. Using clean tubes, the SRAT/SME condenser has met process requirements. However, performance below design specifications for the condensing rate and solids DF have occurred during TNX testing. U_{cond} dropped to 66 BTU/h ft^2 °F during the SME cycle of TNX feed preparation campaign #3. Upon examination, the condenser tubes were found to be coated with an organic sludge. The DWPF feed consists partly of a stream (from the salt processing cell) containing a high amount of organics. It was felt the organics from the salt processing cell were causing the plugging. Fouling of the condenser did not recur during TNX campaign #4 even though organics were used in the feed. However, the condensate became muddy with solids at evaporation rates above 7 gpm during campaign #4. The specific cause for the low solids DF has not been positively identified, but excessive foaming of the feed due to organics is suspected.

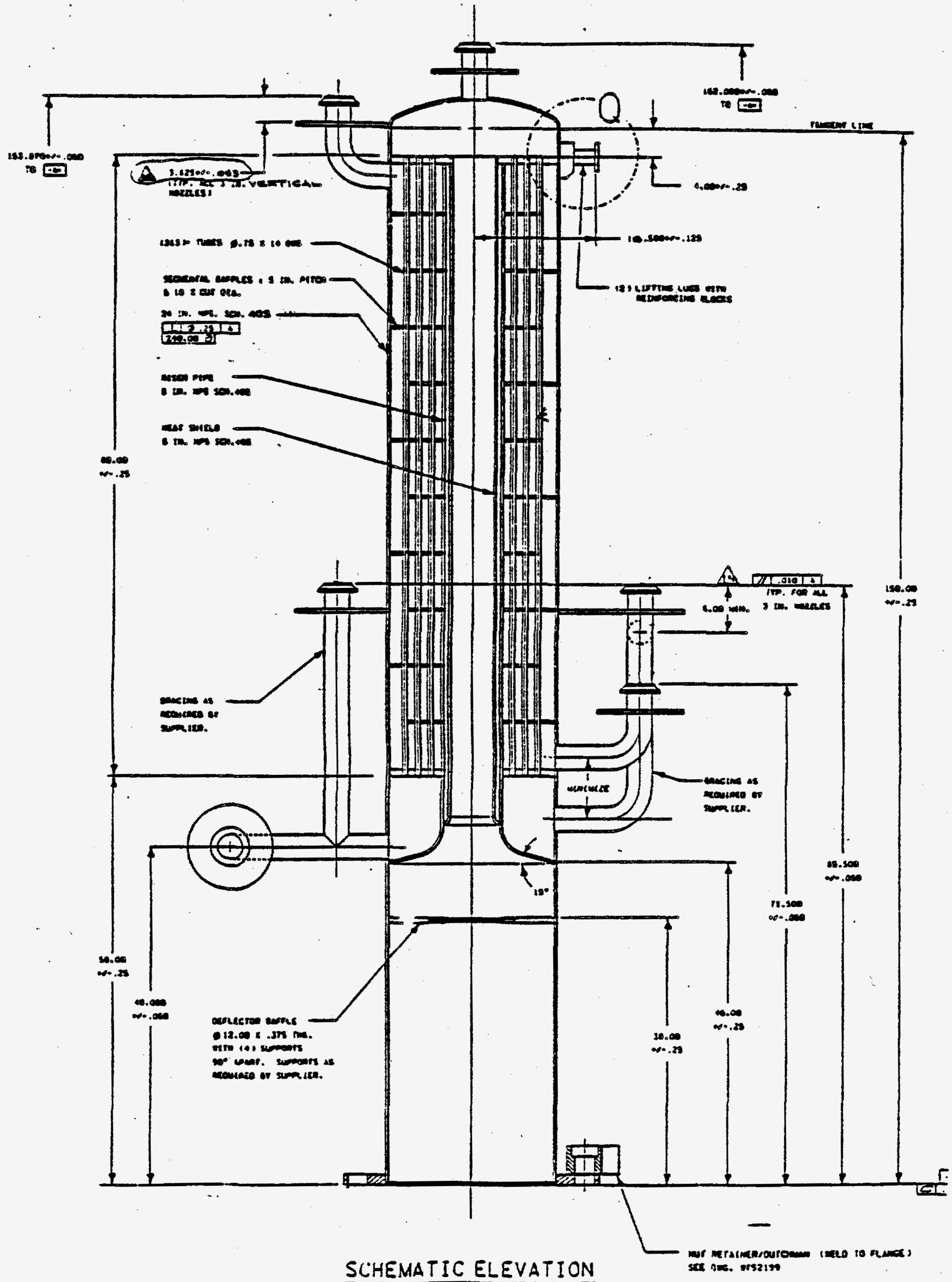


FIGURE 4.6 SRAT/SME Condenser Drawing (DWPF drawing W752132 rev 4)

Current DWPF design does not use a solids de-entrainer in the condenser such as a mesh pad or chevrons. Condensers at the Hanford operations routinely use mesh pad de-misters to produce high solids DFs between the condensate andfeed. Solids DFs up to 10^6 have been observed using two 6-in. mesh pads placed in series (Dunford 1987).

The HWVP feed of pre-treated NCAW contains levels of organics similiar to the DWPF feed, though the types of organics are different. Therefore the TNX problems of condenser tube plugging and solids in the condensate may occur in the HWVP process using the pre-treated NCAW. Several factors influence this assessment. The pre-treated NCAW is not well characterized, the organic material in the HWVP pre-treated NCAW feed will be different from that in the DWPF feed, the effect of other organics on equipment performance is unknown, and feeds with different organic levels (i.e., PFP wastes) may be processed in the HWVP after NCAW.

Organic materials are active, mobile, and comprised of different properties with contrasting process effects. Small amounts of organics can produce a dramatic effect on fluid properties and hence equipment performance. An example is the low amount of organic stabilizers required to reduce foaming. Organics are mobile and can be carried throughout the system, such as the organics in the feed fouling the condenser tubes during TNX campaign #3. Also, organics can produce contrasting effects; for example, while certain organics can reduce foaming, others can increase foaming. While the concentration of organics is low, their presence can strongly influence equipment performance. In addition, the effect of a mixture of organics is difficult to predict. Consequently, condenser performance with representative HWVP feeds containing the expected organics should be verified with full-scale testing.

As described in Appendix A, two mechanisms can be assumed for condensation of a pure vapor: falling film or dropwise. Falling film condensation is the most conservative assumption and is the one normally chosen for design. The predicted U_{cond} is greater than 500 BTU/h ft^2 °F for all the methods using a falling film assumption (see Appendix B, pages B-21 to B-23). The engineering evaluation indicates the condenser will meet design specifications under ideal conditions. However, the effect on process performance of non-condensable gases (e.g., in-leakage air), organics, tube metallurgy, or fouling cannot be quantitatively evaluated. Obviously fouling will reduce condenser performance, but precisely how much or how quickly cannot be determined without testing. The testing at TNX has demonstrated the debilitating effect of organics and fouling on condenser performance. During a single feed preparation cycle the condenser capacity dropped below the design criteria due to fouling by organics. The testing at TNX emphasizes the importance of process verification. Full-scale testing can reveal problems that cannot be foreseen during initial design. In addition, testing with HWVP feeds may create problems not observed at TNX. In other words, processing with HWVP feeds may create a different set of problems than those encountered at TNX.

4.2.1 Conclusion

The SRAT/SME condenser will require full-scale testing due to the possibility of foaming with HWVP feeds and problems encountered during testing at TNX. The condenser will also be used for condensation during equipment testing.

4.2.2 Recommendations

A solids de-entrainer is recommended to improve the solids DF between the feed and condensate. If a clean condensate is produced by the SRAT/SME condenser then further evaporation in the DWTT may be unnecessary. Producing a clean SRAT/SME condensate would eliminate the requirement for further processing to remove radioactive material. Chevron mist eliminators can remove 100% of the particles greater than 10 μm and do not plug as easily as wire mesh pads (Hansen, McNulty, and Monat 1987). Collection efficiency could be improved further by using a wire mesh mist eliminator downstream of the chevron mist eliminator.

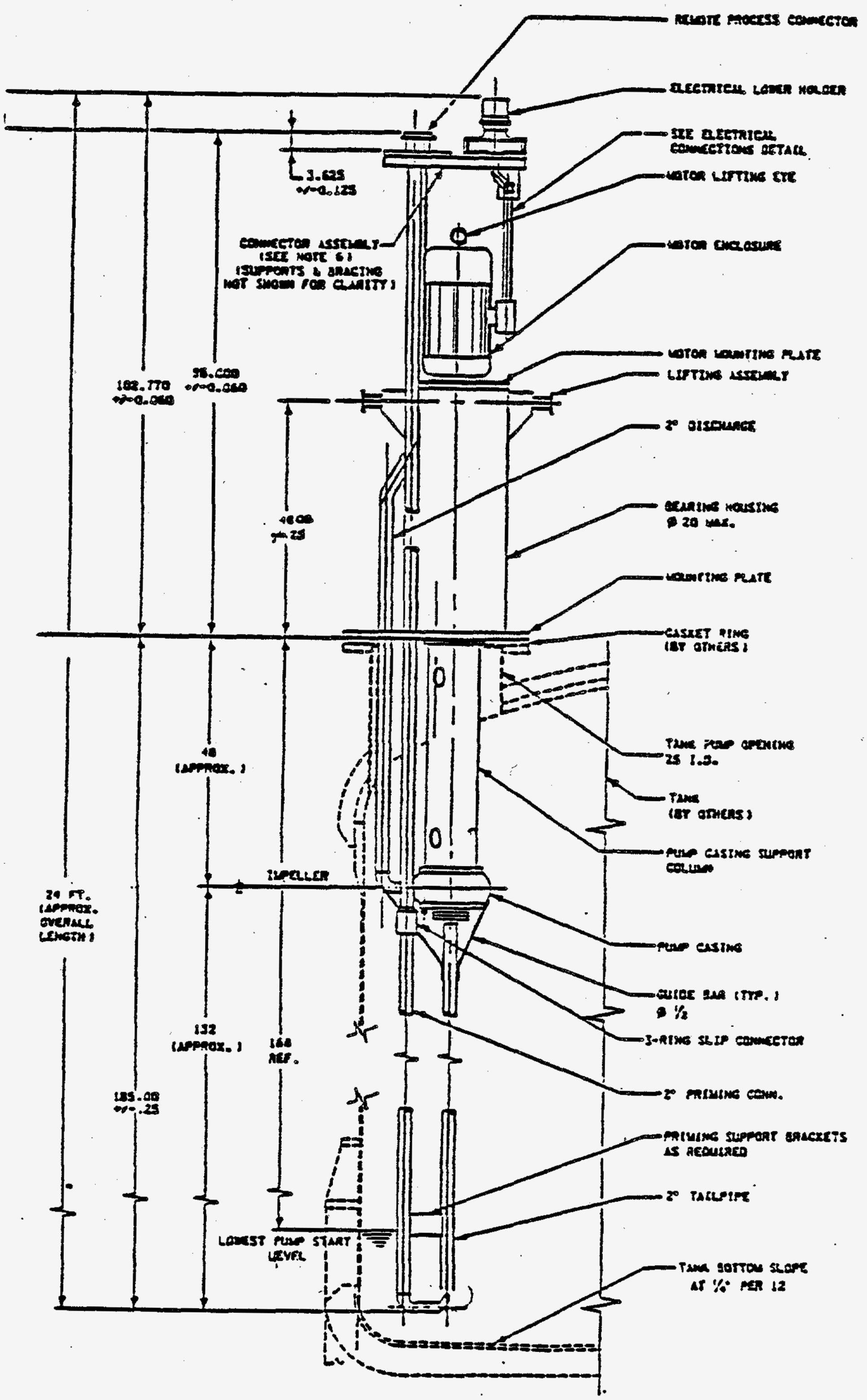
4.3 PUMPS

Pumps are used in the feed preparation area to transfer fluids between tanks and to recirculate slurry through the melter feed and sampling systems. The design rate for tank transfer is 100 gpm. In addition, the transfer pumps are expected to leave a minimum vessel heel, or low amount of residual slurry after a transfer. The HWVP melter feed system is required to deliver melter feed through two loops at a controllable rate of 0.10 to 0.75 gpm per loop (HWVP TDP 1986). The HWVP melter feed rate range is lower than DWPF's; the range for DWPF is 0.4 to 1.2 gpm through each of two loops. DWPF design uses vertical centrifugal cantilevered (VCC) pumps for slurry transfer and sample and melter feed recirculation. VCC pumps are vertical pumps that use an extended shaft between the motor and pump as seen in Figure 4.7; standard centrifugal pumps use a close-coupled shaft between the motor and centrifugal pump.

Testing at TNX demonstrated that the VCC pump can transfer the necessary slurry flow at the required head. However, high vessel heels have occurred after transferring slurries. The vessel heel increased with the agitator impeller speed; this indicates air was entrained due to the formation of a vortex. The vessel heel was reduced by using a shroud around the pump suction inlet; the shroud created a stagnant region to allow air to disengage from the slurry. Vessel heels were also increased when organics were in the feed. The organics probably increased foam stability and formation, causing pump cavitation.

In the feed preparation system, the important fluid properties in pumping are the slurry density, rheology, foaming characteristics, and solids content and distribution. The slurry densities of the HWVP feeds are similar to DWPF feeds; similar enough not to make a substantial difference in equipment performance. The slurry rheologies are reasonably alike at the higher shear rates of 100 to 1000 s^{-1} produced in a centrifugal pump and process line. This similarity in rheology is expected to produce equivalent process performance such as pressure drop and developed head. The foaming characteristics of DWPF or HWVP feeds have not been measured. Therefore, a direct comparison of foam properties is not possible. Information on foam formation and stability is important because a persistent foam can cause the centrifugal pump to cavitate with a large amount of slurry left in the vessel. A large vessel heel will in turn lengthen the feed preparation cycle time.

An net positive suction head (NPSH) calculation indicates the available NPSH is sufficient for the required pump NPSH. When the slurry level is approximately 6 in. above the inlet of the suction pipe, the estimated available NPSH is approximately 8 ft (see appendix B, pages B-24 to B-27). As specified by the manufacturer, the required NPSH for the DWPF pump is 3 to 4 ft of liquid at 100 gpm (see Figure A.6). The pump is theoretically capable of pumping the level in the vessel to the inlet of the suction pipe. Normally the suction



ELEVATION

FIGURE 4.7 DWPF VCC Pump Drawing (DWPF drawing W752260, rev 3) pipe must be submerged 6 in. in the fluid to prevent pump cavitation. Cavitation is caused during low submergence due to air being pulled into the suction pipe and flooding the pump. The calculation assumes ideal operating conditions, therefore it does not include the effect of entrained air.

The melter feed and sample systems are unlike the slurry transfer systems. Both the melter feed and sample systems use a recirculation loop. The sample system recirculates the slurry past the sampler using a VCC sample pump with a total discharge of 58 gpm; 5 gpm circulates in the sample loop while 45 gpm is returned to the tank without entering the sample loop. The melter feed system uses a 100 gpm VCC pump to deliver slurry to two recirculation loops. Each loop contains a cross-flow strainer which uses an orifice to limit flow to the melter feed line. The melter feed rate is controlled by adjusting the recirculation flow past the cross-flow strainer; an increase in recirculation flow (or increase in recirculation line pressure) will increase the feed rate to the melter.

Testing at TNX has demonstrated stable controllable melter feed flow above 0.4 gpm in each loop (HWVP/DWPF Technology Exchange 1980). Below 0.4 gpm, the flow was erratic and uncontrollable (Voogd 1987). This performance is unacceptable for the HWVP because the requirement for the HWVP melter feed system is a controllable feed rate of 0.1 to 0.75 gpm for each loop. In addition, the feed line plugged frequently. Plugging of the feed line was reduced, though not eliminated, by using a suction strainer.

4.3.1 Conclusion

The DWPF melter feed system has not demonstrated stable and controllable melter feed rates below 0.4 gpm per feed loop. Since the nominal HWVP design melter feed rate is 0.25 gpm per feed loop, the DWPF melter feed system requires testing.

4.3.2 Recommendations

To achieve lower HWVP feed rates, it may be necessary to test different orifice sizes in the DWPF melter feed system. As an alternative, the ADS pump has demonstrated a stable controllable flow from 0.1 to 0.8 gpm (Peterson, Perez, and Blair 1986) and will be used as the radioactive melter feed pump at the WVDP. The ADS pump should be included in the test system as an alternative to the VCC system.

Since the DWPF melter feed system uses a standard VCC transfer pump, the pump can be used to evaluate pressure drops in process lines and determine vessel heels with simulated HWVP feeds.

4.4 SLURRY SAMPLING SYSTEMS

Sampling is necessary for process control and waste form qualification. Two issues must be resolved to ensure sampling quality: 1) the vessel sample point must be representative of the vessel contents, and 2) the sampling system must extract a representative sample from the sample point.

As described previously, DWPF design continuously recirculates the slurry past a sample device. The sample device (Hydraguard) diverts a part of the stream to a flush line or sample container as seen in Figure 4.8.

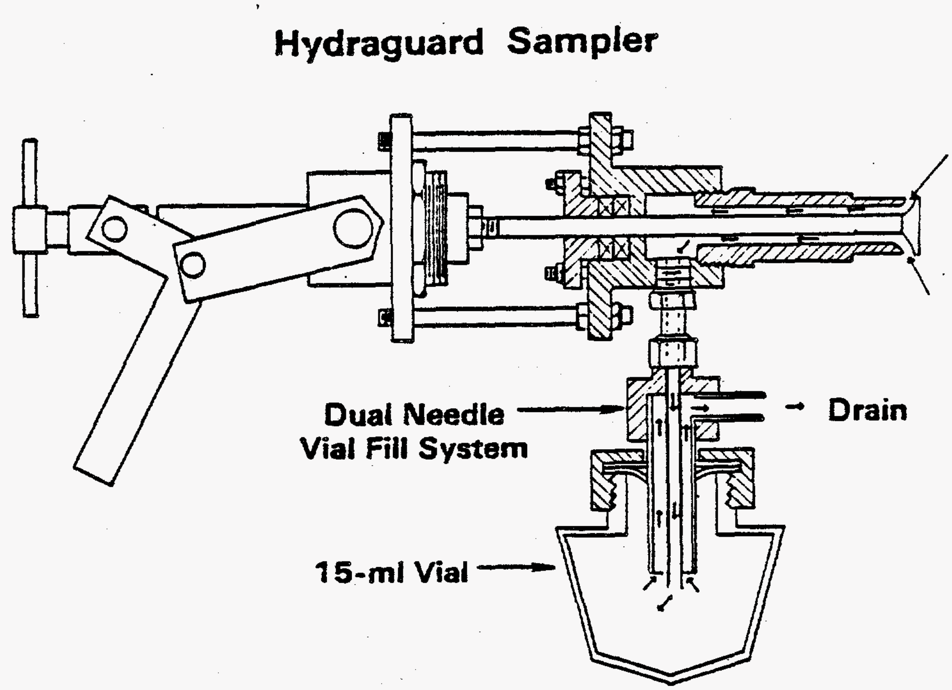


FIGURE 4.8 Hydraguard Sampler

Hydraguard is a registered trademark of Hinds International, Hillsborough, Oregon.

Slurry properties will affect both slurry homogeneity and the sampling system. Above a particular viscosity, the vessel contents will become non-homogeneous. Feeds tested to date have not approached this limit. Slurry rheology will also affect the performance of the sampler. The sampler has many areas with close tolerances and small dimensions; a viscous slurry may not flow like a less viscous material. The presence of solids adds to the problem. Large solid particles may not follow the stream lines, may adhere to walls, and may plug narrow passages; these affects could prevent the sampler from extracting representative samples. Testing at TNX has demonstrated the ability of the Hydraguard sampler to extract representative samples from simulated DWPF slurries.

Sampling systems are not amenable to analysis like an agitated vessel. Consequently, the best method to verify the sampler performance is to test it with representative feeds. The Hydraguard sampler will require testing for two reasons: to verify sampler performance using HWVP feeds and to verify vessel homogeneity.

4.5 THE SPENT FRIT HOLDING TANK AND FRIT SLURRY MAKE-UP TANK

The purpose of the SFHT and FSMT are to maintain suspension of frit in a solution of 1 wt% formic acid in water. The SFHT and FSMT are free of vessel internals and therefore fluid mixing is easier to analyze. A solids suspension criteria is specified for frit/water mixtures. Based on engineering analysis, DWPF design should provide adequate agitation to suspend the frit in water (see Appendix B, pages B-28 to B-32). A settling velocity of 1 ft/min was assumed for the frit/water suspension. It is not possible to estimate how well the system will re-suspend settled solids. The most important factor will be the structure of the settled solids; if the settled solids form a congealed mass, then re-suspension will be very difficult. However, the addition of formic acid is expected to prevent congealing. Re-suspension of loose solids should not be a problem as long as the settled solids do not immobilize the agitator and prevent it from turning. No testing has been performed using the SFHT and FSMT. Testing is not required for the SFHT and FSMT to verify homogeneity during normal operation. However, it might be worthwhile to include the FSMT in a test plan and system to establish lower operating limits and determine the effect of settled solids. In addition, the DWPF design 100 hp agitator motor can be reduced to a smaller size.

5.0 PROPOSED TESTING FOR THE FEED PREPARATION EQUIPMENT

The outline for the test plan is composed of three sections. Section 5.1 covers objectives and background, Section 5.2 details the equipment performance requirements that will be tested, and Section 5.3 consists of the overall strategy of the test plan.

5.1 OBJECTIVES AND BACKGROUND

The objective of the feed preparation test system is to demonstrate process performance of the HWVP feed preparation equipment. This test system will support design of the HWVP by verifying equipment performance. The feed preparation test system will be used to evaluate process performance in the following areas:

- slurry homogeneity
  - -- uniform temperatures, solids, and chemical components
- slurry transfer
  - -- inter-tank transfer
  - -- melter feed system
  - -- sample system
- heat transfer
  - -- convective
  - -- boiling
  - -- condensing
- equipment life
- time cycle.

Testing of the DWPF design for the HWVP is necessary due to differences in feed properties and process requirements, and to evaluate problems that have occurred during full-scale testing at the TNX facility in the DWPF. The HWVP will need the flexibility to process a variety of feeds (NCAW, PFP, or CC), while the DWPF will blend the waste to produce a reasonably uniform feed. In contrasting the feeds, NCAW (the initial feed to the HWVP) is more dilute, has little or no mercury, has more zirconium and less aluminum and iron, and has different elemental and organic species than the DWPF feed. These feed differences affect design requirements and equipment performance. A higher concentration rate is necessary with the more dilute HWVP feed to meet time cycle design requirements. Mercury has not been discovered in analysis of NCAW feed to date. Therefore, the mercury removal and purification equipment of the DWPF should not be necessary in the HWVP. Dilute feed from the RLST will be added continuously to the SRAT in the HWVP while in the DWPF it is added in batches. In addition to differences in feeds, equipment and process problems have surfaced during testing at DWPF that may require similar evaluation using HWVP feeds.

Full-scale equipment testing is being performed in the DWPF slurry feed preparation system located in the TNX facility. The feed preparation system has met DWPF process design requirements with some exceptions. Further equipment testing and evaluation of design modifications are being conducted to improve the performance of the DWPF feed preparation equipment.

The following situations have occurred during testing at the TNX facility; excessive solids have been entrained in the condensate at high evaporation rates of 10 to 12 gpm, severe erosion has been encountered on the bottom flat-blade impeller, coils and coil supports, and vessel bottom, the centrifugal transfer pump has left a vessel heel of 1,000 to 3,000 gal depending on the agitator speed, the condenser has plugged with organic material, and level measurement has been erratic during agitation. The condenser plugging and solids entrainment have not occurred consistently and seem to be driven by the level of organics in the feed. The testing at TNX demonstrates the importance of verifying equipment performance. It is difficult to accurately predict the results of processing a complex feed and the testing at TNX highlights areas that may require special attention.

Full-scale testing is required because process performance cannot be confirmed exclusively using bench scale equipment. The dynamic slurry behavior during boiling heat transfer and agitation cannot be accurately scaled up from bench scale equipment. In addition, the feed preparation test system could be integrated into the Systems Integration Facility (SIF) to further support the HWVP.

5.2 EQUIPMENT PERFORMANCE REQUIRING TESTING

Since the HWVP will be processing variable feed compositions, verifying equipment performance for a single feed is inappropriate. Establishing a process envelope for equipment performance would verify process performance for any feed whose properties fit the requirements. The process envelope will determine the process operating characteristics (e.g., steam temperature, impeller speed, etc.) and slurry properties (e.g., viscosity, particle size, etc) within which acceptable performance is achieved. This envelope would establish limits on the slurry properties beyond which unacceptable equipment performance may be expected. Generating the performance envelope may also disclose limiting equipment in the process; i.e., equipment that limits further increases in the production rate. For example, during evaporation the limiting equipment may be the immersed helical coils or the condenser or the solids de-entrainer. If increased production is needed, attention can be focused on increasing the performance of the limiting equipment.

The following sections describe the major equipment to be tested, what performance requirements will be evaluated, and the instrumentation that will be used to monitor the process. The sections discuss the SRAT/SME vessel and agitator, heat transfer coils, condenser, and slurry transfer. Following the equipment section, the test strategy is outlined in Section 5.3.

5.2.1 SRAT/SME Vessel and Agitator

The purpose of the SRAT/SME vessel and agitator is to receive and hold the slurries and maintain slurry homogeneity. The vessels are cylindrical with a 12-ft ID.

Process requirements to be evaluated include the following:

- slurry homogeneity
  - -- uniform solids, temperature, and additives
- agitator power requirements
  - -- during normal operation and start-up
- level indication
  - -- accuracy, reliability, and precision
- equipment life
  - -- incorporate corrosion coupons into the test system
  - -- periodically inspect tank internals for wear.

Variables to be measured include the following:

- slurry homogeneity by sampling at different locations using a Hydraguard sampler and sample pump
- slurry temperature at various vessel locations
- agitator power and impeller speed
- level
  - -- Holledge
  - -- dip tube

5.2.2 Heat Transfer System

The heat transfer coils will be capable of carrying steam or cooling water. The purpose of the heat transfer coils are to evaporate the slurries at 10 gpm, heat or cool the slurry at 10 °F/h, and remove radioactive decay heat.

Process requirements to be evaluated include the following:

- heat transfer
  - -- convective
  - -- boiling
  - -- the effect of impeller speed, slurry viscosity, and fouling
- equipment life
  - -- the effect of agitator and coil design, and slurry properties.

Process variables to be measured include the following:

- steam and cooling water flow and ΔT to calculate overall heat transfer coefficients
- coil skin temperatures to detect fouling
- coil wall thickness and coil support dimensions to determine the extent of corrosion/erosion

5.2.3 Condenser

The purpose of the condenser is to condense vapors produced in the SRAT or SME and, together with the vessel, produce a high solids DF between the slurry and condensate. The condenser is vertical and water cooled with a solids de-entrainer. The condensate can be returned to the vessel during total recycle or diverted to an auxiliary tank for disposal. The heat transfer coils may not produce enough vapor to overload the condenser and demonstrate the maximum heat transfer rate. Therefore, a connection will be made upstream of the condenser tubes to introduce additional steam so the condenser performance can be fully evaluated.

Process requirement to be evaluated include the following:

- overall heat transfer coefficient (U_{cond})
  - -- effect of non-condensable gases, solids, fouling, and organics on U_{cond}
- solids DF
  - -- effect of foaming, vapor flowrate, solids de-entrainer, and vessel freeboard on the solids DF
- condenser tube and mist elimator fouling or plugging.

Variables to be measured include the following:

- cooling water flow and ΔT to calculate the overall heat transfer coefficient
- pressure drop across the solids de-entrainer and condenser tubes to monitor fouling or plugging
- samples of the following streams: the vapor upstream and downstream of the solids de-entrainer, condensate, and non-condensable gas
- condensate and non-condensable gas flowrates and temperatures.

5.2.4 Slurry Transport

Pumps are used to transfer vessel contents, recirculate melter feed for the melter feed system, and recirculate slurry through the sample loop. The transfer pump is a 100 gpm, 20 hp, vertical cantilevered centrifugal (VCC) pump. It will serve as a recirculation pump for the melter feed system. The sample pump is a 58-gpm 20-hp VCC pump and recirculates 5 gpm through the sampler; the remaining slurry is diverted to the vessel.

Process requirements to be evaluated include the following:

- melter feed system
  - -- controllable flow of 0.1 to 0.7 gpm
- vessel heel
  - -- effect of impeller speed and slurry properties
  - -- evaluate the suction pipe entrance and priming/flush design
- equipment life.

Variables to be measured include the following:

- pump impeller speed
- power consumption
- slurry flowrates
  - -- melter feed system
  - -- sample system
  - -- inter-tank transfer.

5.3 TEST STRATEGY

The equipment will be tested with a progression of fluids beginning with water, then with sand/water slurries, and finally with simulated feeds. Initially, equipment performance will be tested using non-hazardous fluids for three reasons: it will allow operators to develop familiarity with the equipment and confirm the applicability of test procedures, provide a base level of equipment performance, and allow an evaluation of operating characteristics on process performance. The simulated feed is classified as a hazardous chemical and its disposal will be more involved and expensive than a non-hazardous fluid. If problems occur with a non-hazardous fluid, the material can be sent to the chemical sewer and the equipment easily cleaned; a similar problem with a hazardous material will produce much greater consequences. The simulated HWVP feed runs should be used to produce the maximum of equipment performance data, not to break in the equipment or establish the effect of operating characteristics.

The test procedure will match the actual plant process as closely as possible. Dilute feed will be added continuously during the SRAT cycle. The SME cycle will be a batch cycle. The system will have the flexibility to evaluate different processing steps, equipment design, or operating characteristics as follows: continuous or batch operation, formic acid addition above or below the surface, different types of impellers, various steam pressures, cooling water flows, and impeller speeds, different feed pumps, and various in-leakage air flowrates.

Water and water/sand slurries will be used for functional testing of the test system and to evaluate the effect of operating characteristics on equipment performance. Further testing of the effect of an operating characteristic on equipment performance will be unnecessary if no relationship is established with simple fluids. For example, if the type of impeller does not alter the convective heat transfer rate in water, then the impeller type will not be varied to improve convective heat transfer in simulated slurries.

The first runs will use water as the test fluid. This will allow more accurate estimates of process performance since the physical properties of water are well defined. Convective and boiling water heat transfer coefficients will be determined on the clean heat transfer coils. The effect of impeller speed, type of impeller, and steam temperature on convective and boiling water heat transfer coefficients will be evaluated. In addition, the effect of air in-leakage on boiling heat transfer and condenser performance will be investigated. The heat transfer coils will be used to supply vapor to the condenser and the maximum heat transfer coefficients of the condenser and coils will be determined. If the heat transfer coils cannot supply enough vapor to overload the condenser, then additional steam will be added at the inlet to the condenser.

Following testing with water, runs with sand and water mixtures will be performed. The particle range and concentration of the sand will bracket the range measured in the simulated waste. Testing will be performed with particle ranges above and below the current average particle size and with solids concentrations above the current maximum for the feed preparation system. This will provide information on the effect of particle size and solids concentration on equipment performance. The operating characteristics that had an effect on process performance during the water tests will be evaluated again using sand and water mixtures. The maximum heat transfer coefficients for convection, boiling, and condensing will be determined. The solids DF between the vessel slurry and condensate will be determined and the effectiveness of the solids de-entrainer will be evaluated. A more complete test of the system would also include a foaming agent and assess the impact of foaming on the solids DF.

This test variable should be evaluated based on development at the TNX facility. The sample system could initially be tested using this slurry mixture. The size distribution of the sand would be known and could be compared to the samples extracted by the sample system. Normal operation of continuous dilute feed addition will be performed. All equipment in the feed preparation test system will be operated with a sand and water slurry before testing with a hazardous slurry is begun.

Once testing with non-hazardous material is complete, operations with a simulated feed will begin. The most current feed simulant will be tested using standard processing (e.g., continuous feeding during SRAT cycle). Convective, boiling, and condensing heat transfer coefficients will be determined throughout the concentration process. The feed will be overconcentrated to evaluate the effect of solids concentration on equipment performance. Overconcentrating the feed will be used to establish a performance envelope for the system. As the feed is overconcentrated, the slurry viscosity will increase and the boiling heat transfer coefficient will decrease. When the boiling heat transfer coefficient is below the required amount, then the upper limit of the slurry viscosity is known. Organics have not been conclusively measured in current NCAW analysis; however, other wastes (CC or PFP waste) do contain organics and additional testing with organics should be performed.

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APPENDIX A

TECHNICAL BACKGROUND

TECHNICAL BACKGROUND

The following sections describe viscometers and applicable relationships or models and important factors in agitation, heat transfer, and fluid pumping. This technical background will aid in determining what equipment must be tested and how to efficiently collect and analyze the data. Section A.1 describes various viscometers. Section A.2 discusses fluid mixing, Section A.3 describes heat transfer, and Section A.4 concerns fluid transfer. Each section contains a brief introduction to the subject followed by a review of the technology relevant to equipment and processing in the feed preparation system.

A.1 VISCOMETERS

There are three types of apparatus for determining fluid rheology: absolute, relative, and capillary tube viscometers. The Haake Rotovisco viscometer (a type of absolute viscometer) requires the sample to be placed into a gap between two coaxial cylinders. A motor drives the inner cylinder. A viscosity-related torque, caused by the resistance of the sample to shearing, acts on the inner cylinder. This torque is measured and then translated into a shear stress. Normally, the shear rate is calculated by assuming the fluid at the surface has the velocity of the surface (i.e., no slip) and that the velocity profile is linear through the narrow gap.

Relative viscometers measure a torque necessary to rotate a paddle submerged in the unknown fluid and compare the torque to that obtained with a fluid of known viscosity. Relative viscometers (e.g., a Brookfield viscometer) use the fact that, at low impeller speeds, power consumption is directly proportional to fluid viscosity. The apparent viscosity is then a ratio of the torque of the unknown fluid to that of a fluid with a known viscosity and torque. Relative viscometers cannot create a rheogram of shear stress vs. shear rate as shown in Figure 3.2, but they are commonly used in the fluid mixing industry for measuring the viscosity of slurries. Relative viscometers have been used at PNL, but have not produced consistent results.

The Rotovisco is a registered trademark of HBI Haake-Buchler Instruments, Saddle Brook, New Jersey.
Brookfield is a registered trademark of Engineering Laboratories Inc., Stoughton, Massachusettes.

Capillary tube viscometers pump the fluid at different flowrates through a capillary tube and relate the pressure drop to the fluid viscosity. Capillary tube viscometers can generate a shear stress-shear rate plot or rheogram.

Absolute viscometers operate with a well-defined system, can provide information across a wide range of shear rates, and can be used to analyze fluid behavior in various situations such as pumping, pipe flow, or agitation. Absolute viscometers sometimes have problems with slurries because particle interference may occur across the narrow gap, the assumption of no slip at the wall may be invalid, and slurry particles may settle causing the slurry to be non-uniform. However, two of these problems can be addressed by increasing the gap distance to reduce the interference and using a correction for slip at the wall.

Relative viscometers more closely model fluid behavior of an agitated vessel, but the actual shear rates are unknown. Therefore, relative viscometers require a large data base of fluid rheology and operating experience for effective analysis.

A.2 FLUID MIXING

Evaluation of agitation systems requires a description of the degree of mixing. The size and difficulty of the mixing problem must also be known. A quantitative system for characterizing agitation was detailed by Chemineer, Inc. The system and analysis was published in a series of articles written in 1976 for Chemical Engineering. The articles defined controlling variables and dynamic response for the most common agitation situations of motion and blending of fluids (Hicks, Morton, and Fenic 1976), solids suspension (Gates, Morton, and Fondy 1976), and gas dispersion.

The controlling variable, dynamic response, and system size define the difficulty of the problem. The controlling variable in mixing applications is the key fluid property. The dynamic response relates to the required homogeneity. The controlling variable is fluid viscosity for blending and motion and particle settling rate for solids suspension. The required dynamic response for blending and motion is bulk fluid velocity while for solids suspension it is the level of solids suspension. The degree of homogeneity corresponds directly to the scale of agitation or dynamic response. The dynamic response is scaled from 1 to 10 where 1 is minimal agitation and 10 is vigorous agitation. The system size is based on the mass of fluid and is a product of the volume and specific gravity of the fluid. Increasing the fluid viscosity or particle settling rate will intensify the mixing problem, as will increasing the degree of homogeneity or vessel size. Table A.1 lists the levels of dynamic response and a description of the process result.

Once the problem has been defined, fundamental fluid and transport relationships are used to estimate the process result. Useful dimensionless groups in agitation are the Reynold's number (N_{Re}), the pumping or flow number (N_q), and the power number (N_P);

$$N_{Re} = D^2N\rho/\mu \quad (A.1)$$
$$N_q = Q/ND^3 \quad (A.2)$$
$$N_P = Pg_c/\rho N^3D^5 \quad (A.3)$$

where

D = impeller diameter
N = impeller speed
ρ = fluid density
μ = fluid viscosity
P = power
Q = impeller flow
g_c = gravitational constant

Figure A.1 shows the general relationship between N_{Re}, N_q, and N_P (Oldshue 1983). Three regions are clearly defined. The viscous or laminar region is generally from N_{Re} of 10 to 50, the turbulent region above 1,000 to 50,000, and the transition region between these approximate ranges. The ranges depend on the type of turbine and the system geometry. Most processes operate in the turbulent region where the power number and the flow number are constant for a specific impeller and tank geometry. N_P and N_q have been experimentally determined for many impellers in standard tanks with Newtonian fluids. The power number ranges from 0.5 to 10 and the flow number from 0.3 to 1 (Oldshue 1983). The flow number is an estimate of the flow directly from the impeller and normally does not include entrained flow.

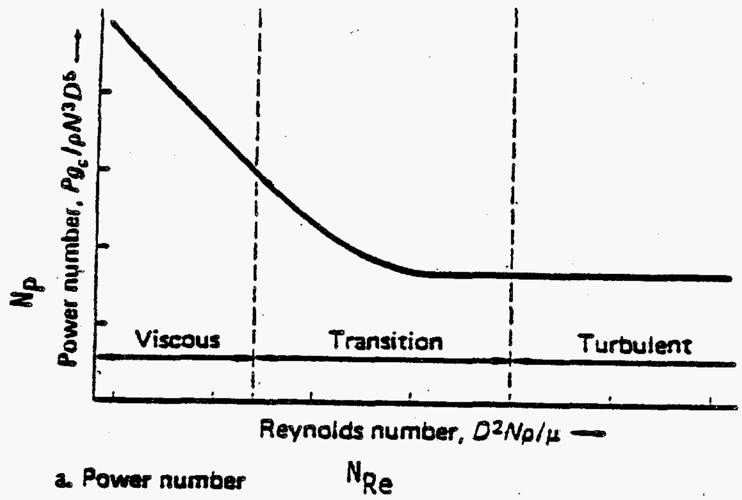
Table A.1

Blending

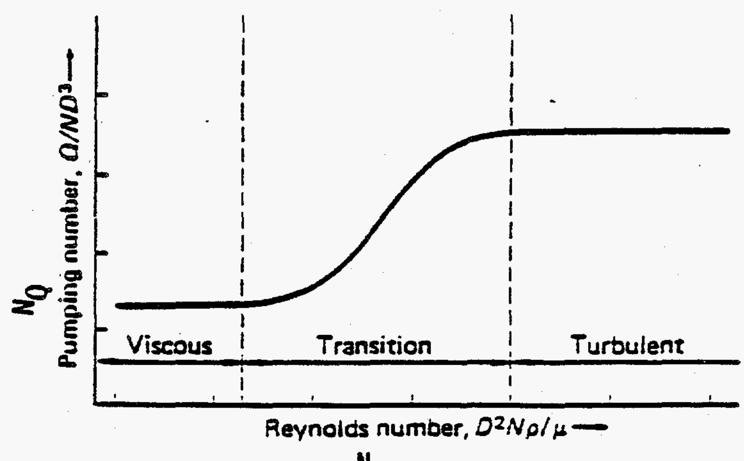
Scale of Agitation	Bulk Fluid Velocity, ft/min	Description
		Agitation levels 1 and 2 are characteristic of applications requiring minimum fluid velocities to achieve the process result.
1	6	Agitators capable of level 2 will:
		· Blend miscible fluids to uniformity if specific-gravity
2	12	differences are less than 0.1.
		· Blend miscible fluids to uniformity if the viscosity of the most viscous is less than 100 times that of the other.
		· Establish complete fluid-batch control.
		· Produce a flat, but moving, fluid-batch surface.
3	18	Agitation levels 3 to 6 are characteristic of fluid velocities in most chemical process industries agitated batches.
		Agitators capable of level 6 will:
4	24	· Blend miscible fluids to uniformity if specific-gravity differences are less than 0.6
5	30	· Blend miscible fluids to uniformity if the viscosity of the most viscous is less than 10,000 times that of the other.
		· Suspend trace solids (<2%) with settling rates of 2 to 4 ft/min.
6	36	· Produce surface rippling at lower viscosities.
7	42	Agitation levels 7 to 10 are characteristic of applications requiring high fluid velocity for the process result, such as in critical reactors.
		Agitators capable of level 10 will:
8	48	· Blend miscible fluids to uniformity if specific-gravity differences are less than 1.0
9	54	· Blend miscible fluids to uniformity if the viscosity of the most viscous is less than 100,000 times that of the other.
10	60	· Suspend trace solids (<2%) with settling rates of 4 to 6 ft/min.
		· Provide surging surfaces at low viscosities.

Solids Suspension

Agitation	Description
	Agitation levels 1-2 characterize applications requiring minimal solids-suspension levels to achieve the process result.
1-2	Agitators capable of scale levels of 1 will: · Produce motion of all of the solids of the design-settling velocity in the vessel. · Permit moving fillets of solids on the tank bottom, which are periodically suspended.
	Agitation levels 3-5 characterize most chemical-process-industries solids-suspension applications. This scale range is typically used for dissolving solids.
3-5	Agitators capable of scale levels of 3 will: · Suspend all of the solids of design-settling velocity complete off the vessel bottom. · Provide slurry uniformity to at least one-third of fluid-batch height. · Be suitable for slurry drawoff at low exit-nozzle elevations. · Produce surface rippling at lower viscosities.
	Agitation levels 6-8 characterize applications where the solids-suspension level approaches uniformity.
6-8	Agitators capable of scale level 6 will: · Provide concentration uniformity of solids to 95% of the fluid-batch height. · Be suitable for slurry drawoff up to 80% of fluid-batch height.
	Agitation levels 9-10 characterize applications where the solids-suspension uniformity is the maximum practical.
9-10	Agitators capable of scale 9 will: · Provide slurry uniformity of solids to 98% of the fluid-batch height. · Be suitable for slurry drawoff by means of overflow.



a. Power number



b. Pumping number

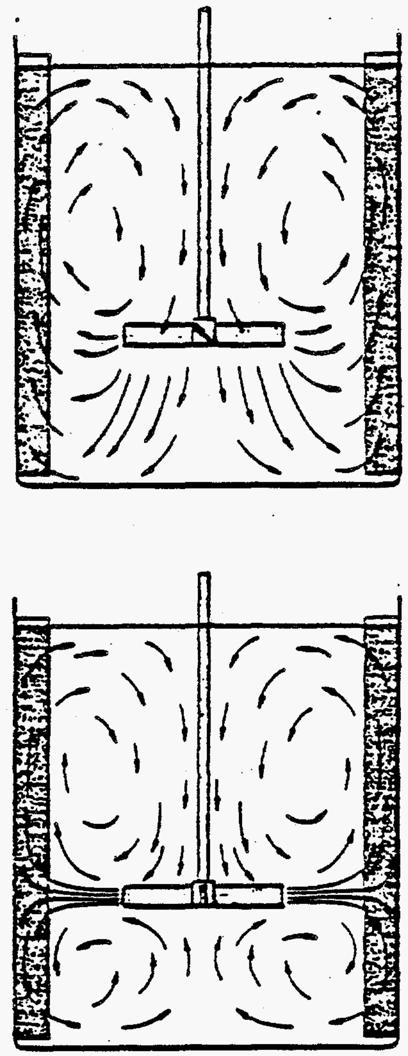
FIGURE A.1 Reynolds Number Correlations for Agitator Power Impeller Flow Number

The dimensionless groups describe relationships between equipment design, process performance, and fluid properties. Agitator power consumption (P) is proportional to N^3 and D^5, while the flow produced by an impeller (Q) isproportional to N and D^3. The power consumption of an agitator can remain constant while the impeller flow is increased by increasing the diameter of the impeller and reducing the impeller speed. Since fluid homogeneity will be increased by increasing impeller flow, larger diameter impellers will enhance slurry homogeneity and not cause an increase in power consumption.

General flow patterns of flow impellers in simple tank geometries have been described and confirmed experimentally. Figure A.2 shows flow patterns in a simple tank from the two types of impellers, radial and axial flow. The flow pattern produced from a radial flow turbine is parallel to the vessel bottom while an axial turbine is parallel to the shaft with little radial velocity. Examples of axial flow impellers are pitched blade or marine impellers; an example of a radial flow impellers is a flat blade impeller. High efficiency impellers are axial flow impellers that produce little radial velocity. Flow patterns in complex geometries with pseudoplastic fluids must be determined experimentally due to the non-Newtonian rheology.

Most slurries are pseudoplastic fluids and cannot be analyzed in the same manner as Newtonian fluids. The viscosity of pseudoplastic fluids increases with decreasing shear rate, unlike Newtonian fluids which have a constant viscosity at all shear rates. In an agitated tank, a pseudoplastic fluid will have a low apparent viscosity in regions of high shear, such as near the impeller, while peripheral areas of low shear will have a much higher apparent viscosity. The net result will be a less effective dispersion of the impeller momentum to the entire tank. The area near the impeller will be well mixed, while distant sections will be stagnant.

Due to rheological behavior and difficulty in transmitting impeller momentum, mixing of highly pseudoplastic fluids is often done in tanks with a high impeller/tank diameter ratio (D/T) of 0.5 or greater. High D/T creates more effective mixing near the tank wall because there is less distance from the impeller tip to the vessel wall.



a. Axial-flow pattern with pitched blade

b. Radial-flow pattern with flat blade

FIGURE A.2 Flow Patterns of Axial and Radial Flow Impellers

Homogeneity of the vessel contents can be determined using the following three methods: 1) perform verification testing with full-scale (prototypic) equipment and similar feeds, 2) perform lab-scale testing, and 3) use engineering relationships. More complex systems will require more rigorous evaluation. The mixing performance of a particular system is most accurately evaluated using prototypic equipment and representative feeds and measuring the desired process result. If the feeds that are used in the prototypic equipment are different from the representative feeds, then the validity of the data will have to be assessed. The evaluation of homogeneity is less precise using bench scale testing and least accurate using engineering relationships. The inaccuracy of the engineering relationships is due to the complex geometry and rheology of the feed preparation system.

In the feed preparation vessels, the fluid temperature, solids distribution, and elemental chemical composition are required to be uniform. If full-scale testing is performed, these properties should be measured to define homogeneity. Two factors must be considered when measuring homogeneity: 1) the sample or measuring point must represent the vessel contents, and 2) the sample or measuring system must obtain accurate and reproducible results. Bench-scale testing is less expensive than full-scale testing, however experimental procedures and results from bench-scale testing must be carefully analyzed. Small-scale equipment dimensions are normally proportional to the dimensions of the large-scale vessel. However, small-scale equipment cannot exactly model the dynamic fluid characteristics of the full-scale system while maintaining proportional dimensions. The smaller vessel produces a proportionally higher flow and lower velocity distribution than a larger vessel. This implies that a small vessel could have good homogeneity while a proportionately sized full-scale vessel may not be homogeneous.

Engineering relationships can be used to evaluate mixing system performance. Engineering analysis only requires knowledge of the vessel geometry, type of impeller and impeller speed, and fluid properties. Engineering relationships are most successful with simple vessel geometries using Newtonian fluids. Complex vessel geometries with immersed structures and non-Newtonian fluids cannot be accurately evaluated with engineering relationships.

In conclusion, explicit relationships can be used to analyze mixing of Newtonian fluids in simple geometries such as the SFHT and FSMT. However, these relationships cannot be used to verify process performance in complex geometries with pseudoplastic fluids such as the SRAT, SME, or MFT. The more complex geometries and fluids will require testing to demonstrate the desired performance. In addition, analyzing slurry viscosities at low shear rates must be done carefully and comparing slurry rheology between laboratories has some limitations.

A.3 HEAT TRANSFER

There are three types of heat transfer: conduction, convection, and radiation. Conduction is heat transmission through a uniform material without significant particle displacement. Convection involves the mixing of one portion of a fluid with another and occurs in two general forms: natural and forced convection. Natural convection is a result of buoyancy forces caused by density differences due to temperature variation. Forced convection is caused by mechanical means and the flow depends on the fluid dynamics of the system, not on the thermal state. Radiation is a transfer of energy by wave motion through space. The wave energy is either transmitted, reflected, or adsorbed as heat by an object.

Controlled heat transfer in the feed preparation area is a surface phenomenon where heat is transmitted from a hot to a cold fluid through a solid surface. The heat rate is then considered in terms of a heat flux, or heat rate per area (i.e., BTU/hr ft^2). The heat rate is commonly represented as a factor of the heat transfer area, the temperature difference between the fluids, and an overall proportionality constant, U.

$$Q = U*A*\Delta T \quad (A.4)$$

where: Q = heat rate, BTU/h
U = overall heat transfer coefficient, BTU/h ft^2 °F
A = heat transfer area, ft^2
ΔT = temperature difference, °F.

The overall heat transfer coefficient (U) is modeled as a series of individual resistances (h) as shown in Figure A.3. Table A.2 lists typical ranges of the individual heat transfer coefficients. As shown in the figure, the most significant individual resistances to heat transfer are the process side heat transfer coefficient and process side fouling factor. All the individual heat transfer coefficients can be estimated except for the fouling factor. The effect of fouling on heat transfer must be determined experimentally.

The feed preparation area has three categories of heat transfer: convective, boiling, and condensing. These three classes of heat transfer are treated in more detail in the following sections.

A.3.1 Convective Heat Transfer

Convective heat transfer in agitated vessels is normally modeled using a Nusselt relationship of dimensionless groups;

$$N_{Nu} = K * N_{Re}^{2/3} * N_{Pr}^{1/3} \quad (A.5)$$

where

$N_{Nu} = h_p D/\kappa$ = Nusselt number

$N_{Re} = \rho D^2 N/\mu$ = Reynolds number

$N_{Pr} = C_p \mu/\kappa$ = Prandtl number

K = constant depending on tank geometry

h_p = process heat transfer coefficient

D = impeller diameter, ft

N = impeller speed, min^{-1}

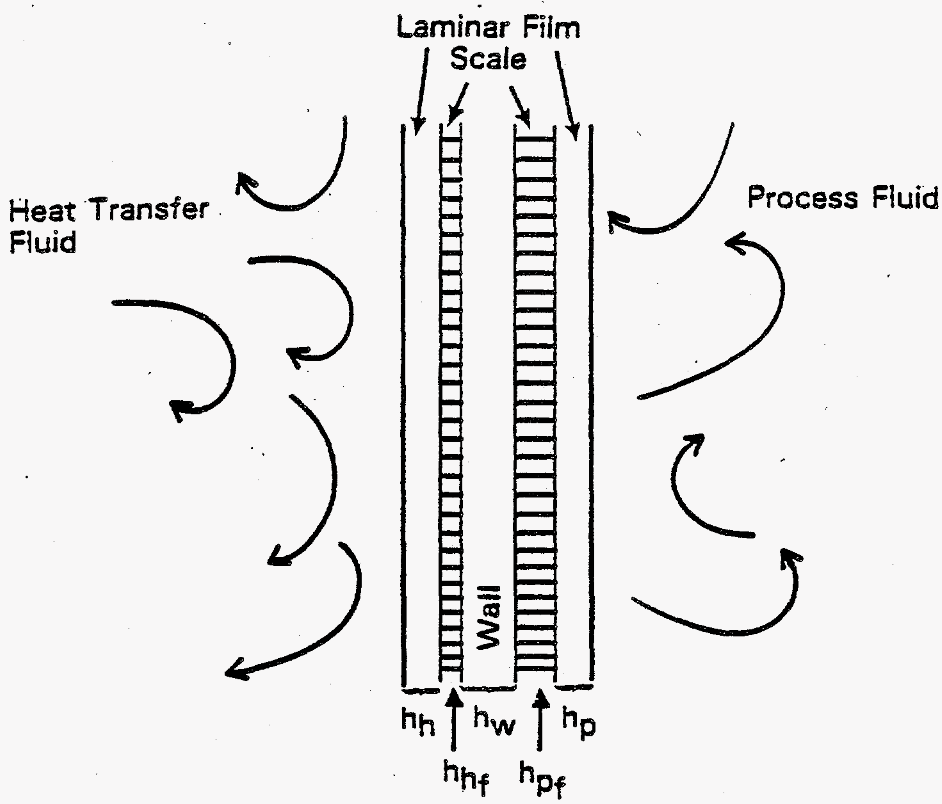
κ = thermal conductivity, BTU/h ft °F

ρ = density, lb/ft^3

μ = viscosity, lb/ft h (Cp*2.42)

C_p = heat capacity, BTU/lb °F.

This relationship (Oldshue 1983) has been successfully used to estimate h_p by several investigators with different vessel geometries and various Newtonian fluids. The most variable component is the constant K, which must be experimentally determined for each specific equipment geometry. No



$$\frac{1}{U} = \frac{1}{h_h} + \frac{1}{h_{hf}} + \frac{1}{h_w} + \frac{1}{h_{pf}} + \frac{1}{h_p}$$

FIGURE A.3 Heat Transfer Model

TABLE A.2 Typical Values for Individual Heat Transfer Coefficients

Heat Transfer Coefficient	Description	Typical value (BTU/hr ft^2 °F)
h_h (steam)	Laminar film	1500 - 5000
h_h (cooling water)	Laminar film	1250 - 2000
h_{hf}	Fouling/Scale	200 - ∞
h_w	Wall	750 - 2500
h_{pf}	Fouling/Scale	50 - ∞
$h_{p,conv}$	Process fluid-convection	30 - 200
$h_{p,cond}$	Process fluid-condensing	100 - 5000
$h_{p,boil}$	Process fluid-boiling	100 - 5000

literature references could be found that used a vessel geometry similar to DWPF design. Consequently, the constant K in the equation would have to be determined from development of full-scale test data. The DWPF has not explored a broad enough range of slurry viscosities and impeller speeds to determine a K. As described in the introduction, the overall heat transfer coefficient is a sum of individual heat transfer resistances. While the most significant individual resistance is often h_p, other resistances cannot be discounted. Over a period of time, fouling can become the controlling heat transfer resistance.

Several fluid properties influence h_p, however fluid viscosity is the most significant since it has the greatest variation. Thermal conductivity and fluid density would not be expected to vary more than 50%, while viscosity could change an order of magnitude during processing and at different locations in the tank. A Philadelphia Mixers, Inc. representative maintained that the relationship was only accurate to ±50% without supporting experimental data (Von Essen 1987). The relationship produces a good empirical data fit, but can be applied only to the experimental system studied. The relationship can be used to predict heat transfer with different fluids in identical equipment. However, the data must be collected in the turbulent regime and the fluid properties identically analyzed (with the same viscometer and identical procedures).

A.3.2 Boiling Heat Transfer

There are several possible mechanisms during pool boiling as exhibited in Figure A.4, which describes the relationship between heat flux and ΔT (temperature difference between the hot surface and boiling fluid). Pool boiling is the boiling of a fluid from a hot surface using only natural convection to move the fluid; there is no forced convection to transport the fluid from the hot surface. The significant regimes are nucleate boiling, transition boiling, and film boiling. The desirable region for heat transfer is nucleate boiling, which generates the highest heat flux (Q/A) at the critical ΔT. The vigorous surface action under nucleate boiling creates good heat transfer and a high heat flux. As can be seen in Figure A.4, increasing the ΔT above the critical ΔT does not increase the heat flux and hence the boiling heat transfer coefficient ($h_b = Q/(A*\Delta T)$) is reduced. This is because a film

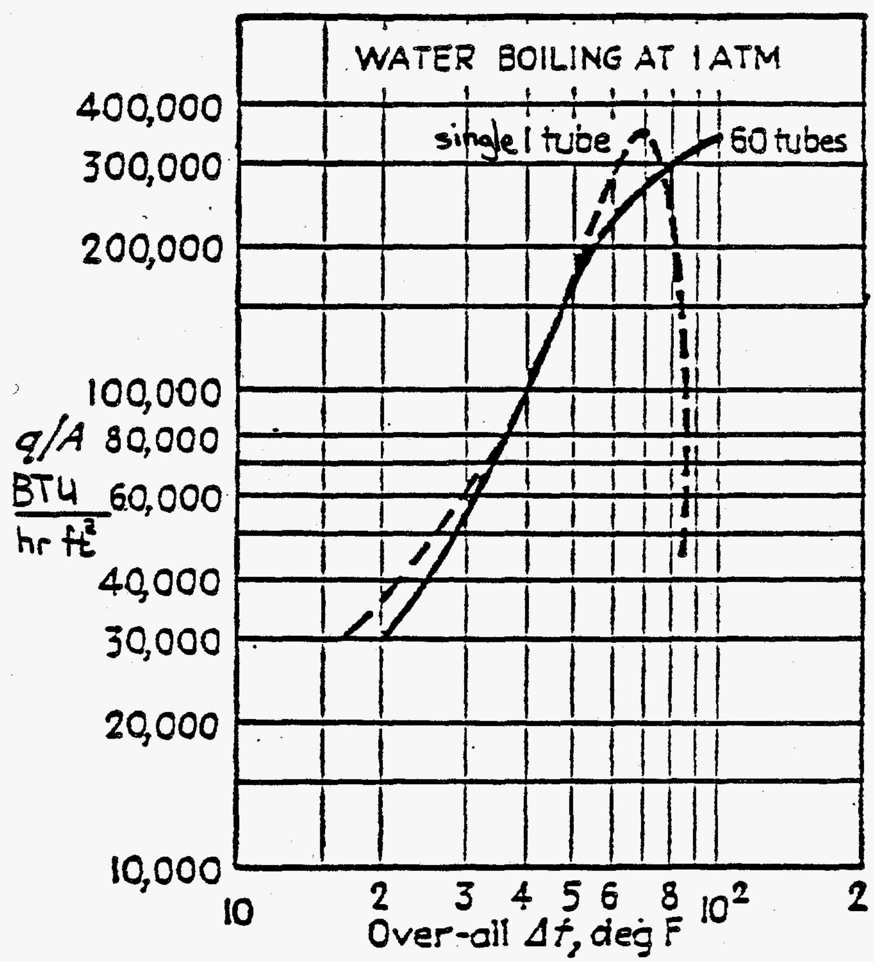


FIGURE A.4 Boiling Curve

of vapor is formed on the surface, increasing the resistance to heat transfer. This phenomenon is significant because it implies that the boiling heat transfer coefficient at higher steam pressures may not be directly extrapolated from data at lower steam pressures. Vigorous fluid flow can greatly improve boiling heat transfer. Thermosiphon reboilers on the Hanford reservation can have 800 to 900 BTU/h ft^2 °F boiling heat transfer coefficients using dilute slurries (Barton 1987). The DWPF design evaporators have a much lower boiling heat transfer coefficient of 100 to 200 BTU/h ft^2 °F. This is largely due to diminished slurry flow across the coils. Increasing slurry flow across the coils would enhance the heat transfer coefficients.

The boiling heat transfer coefficient (h_b) depends on ΔT, fluid properties, material of construction, and surface condition. In contrast, the convective heat transfer coefficient is not strongly affected by material of construction or surface condition. All sources emphasized the difficulty of estimating boiling heat transfer. The imprecision is from the inability to account for surface condition and metallurgy in heat transfer.

Some relationships were found to predict boiling heat transfer (see appendix B, pages B-6 to B-9), but none were accurate with the full-scale DWPF data. The predicted boiling heat transfer coefficients ranged from 400 to 1,000 BTU/h ft^2 °F, four to nine times the actual DWPF boiling heat transfer coefficient of 120 to 130 BTU/hr ft^2 °F. The boiling heat transfer coefficient will have to be determined experimentally; or, a conservative estimate would be to assume the boiling heat transfer coefficient is equal to the convective heat transfer coefficient.

A.3.3 Condenser Heat Transfer

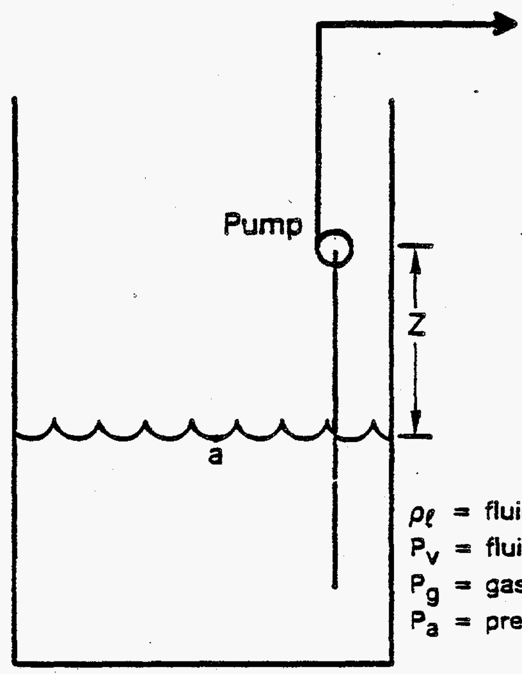
There are two mechanisms for condensation: dropwise and film. Dropwise condensation produces a high heat transfer rate, but conservative predictions should be made assuming film condensation. Film condensation is modeled as a film of uniform thickness, which is calculated using the condensate flowrate and heat transfer surface area. A film N_{Re} is determined that defines the appropriate regime: turbulent, transition, or laminar. The individual heat transfer coefficient (h_{cond}) is then calculated based on fluid properties (fluid viscosity, thermal conductivity, and density) and geometry (vertical or horizontal tubes).

The predicted values for the condenser heat transfer coefficient ranged from 500 to 1000 BTU/h ft^2 °F (see appendix B, pages B-21 to B-23). This is well above the actual overall heat transfer coefficient (U) of 66 to 133 BTU/h ft^2 °F determined during TNX feed preparation campaign #3 (House 1986). The TNX condenser heat transfer coefficients may not reflect the maximum condenser performance since the evaporator may not supply enough vapor to fully test the condenser. The analysis does not include the effect of non-condensable gas, organic vapors, tube metallurgy or fouling. All of these components can affect condenser heat transfer and cannot be estimated without experimental data.

A.4 FLUID TRANSPORT

The purpose of fluid pumping is to deliver a fluid at a prescribed flow and pressure. Standard analysis uses the estimated pipe head loss and required flow to specify a pump and necessary net positive suction head (NPSH). The analysis is well defined for Newtonian fluids and pump performance verification is unnecessary. Remote service introduces special considerations for pump design: low maintenance, simple operation and design, minimum NPSH for low vessel heel, durability, minimum dilution of tank contents using seal or priming flush, vessel entry through the top of the tank, and all-metal construction. The increased requirements for remote service complicate the standard analysis and achieving the required flow and head is only a part of the evaluation.

All pumps require a minimum suction head to operate properly. This necessary suction head varies with the type of pump and its flowrate and speed and can be supplied by the manufacturer. If the minimum NPSH is not supplied the pump will cavitate, causing serious mechanical problems. NPSH is the required amount of head in equivalent feet of fluid between the pump centerline and fluid surface at atmospheric pressure. It is composed of the equivalent head of fluid less the equivalent head of the vapor pressure of the liquid, pressure of any gas in the solution, suction line loss, and entrance loss. Any increase in the head losses will reduce the effective lift of a pump. Figure A.5 shows an example of a typical feed preparation pump system and the equation for calculating NPSH.



ρ_ℓ = fluid density
P_v = fluid vapor pressure
P_g = gas pressure in fluid
P_a = pressure at point a

$$\text{NPSH} = \frac{P_a - P_v - P_g}{\rho_\ell} - h_{f_s} - Z\left(\frac{g}{g_c}\right)$$

where

h_{f_s} = suction line friction loss including entrance losses
Z = distance from liquid level to pump centerline
g/g_c = gravitational conversion factor

FIGURE A.5 Pump Flow System

Each pump has a specific relationship between flow, head, and required NPSH as shown in Figure A.6. In general, as the pump flow is increased, the head is reduced and the required NPSH is increased. Required NPSH is that necessary for the pump to operate and available NPSH is that supplied to the pump. When the available NPSH drops to the required NPSH (as with a drop in liquid level), the pump will no longer deliver the fluid and a heel or residual fluid is left. A large vessel heel results from a low available NPSH. Low available NPSH can be caused by high suction line losses or entrained gases. High suction line loss could be caused by a plugged suction strainer or increased line losses caused by a more viscous fluid. Entrained gases can flood the pump and prevent it from delivering the slurry.

There are two general types of pumps, kinetic and positive displacement pumps. Kinetic pumps transfer momentum to a fluid from another momentum source. Examples of kinetic pumps are centrifugal and steam jet pumps. A centrifugal pump transfers the momentum of a rotating turbine to the fluid while a steam jet pump transfers the momentum of a jet of steam to the fluid. The momentum or velocity of the fluid translates into a head (pressure) and velocity (flow) relationship. Each centrifugal pump produces a characteristic head-flow relationship as shown in a pump curve. As the system head loss is increased, the volumetric flow is reduced. Flow is normally controlled by altering the pressure drop through a control valve, though it can be controlled by varying the speed of the pump. Centrifugal pumps cannot develop enough discharge head with a vapor to displace a liquid-filled line; hence they cavitate and require priming.

Positive displacement pumps either displace the pump chamber volume as with a reciprocating pump or move a fixed volume as in a gear pump. Using incompressible fluids, a fixed volume is transported regardless of the system head loss. Flow is controlled by varying the chamber volume or the speed of displacement. Positive displacement pumps are self-priming because they can develop enough head with a vapor to overcome the head loss of a liquid-filled line.

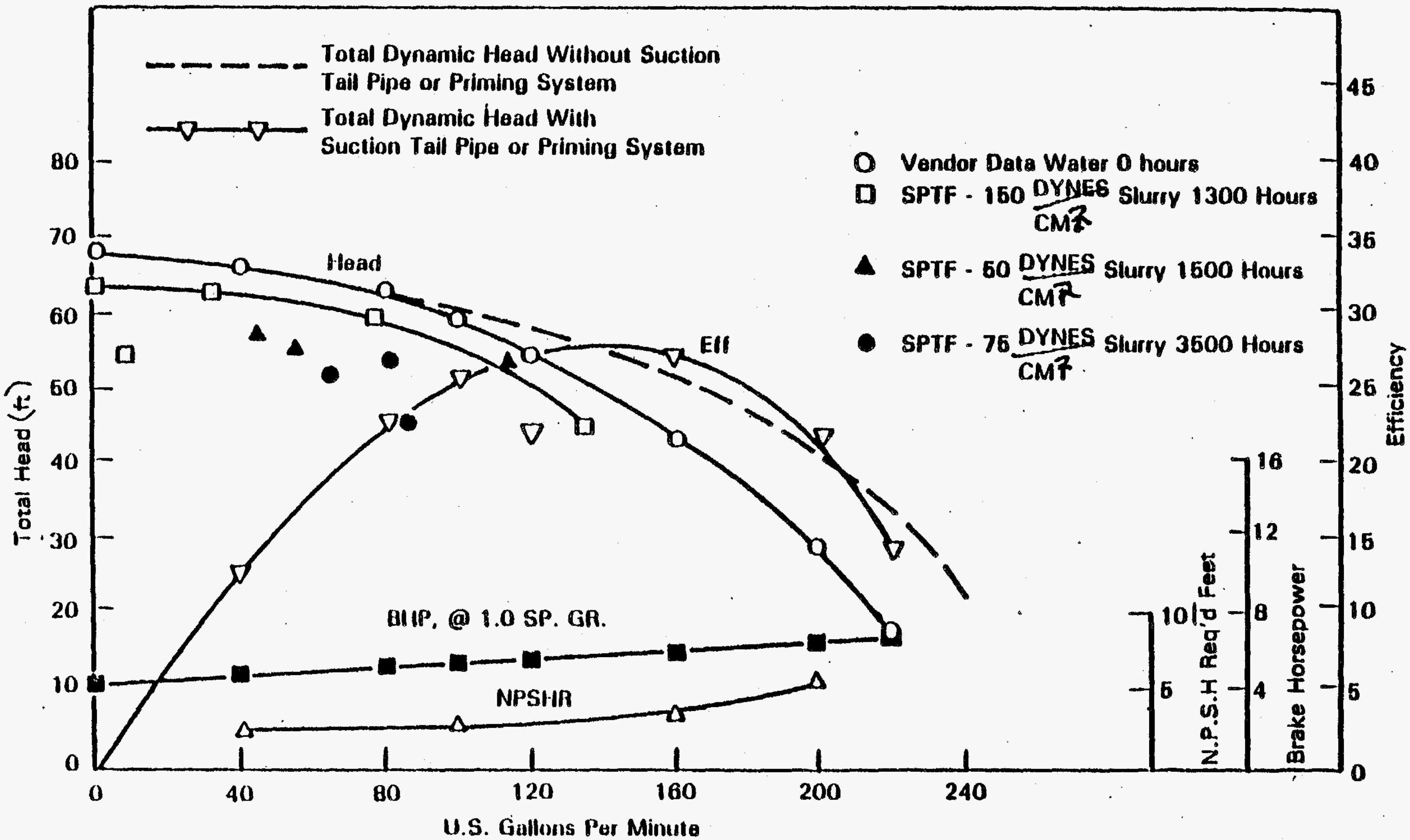


FIGURE A.6 DWPF Lawrence Pump Curve

The advantages of centrifugal or steam jet pumps include the following:

- simplicity - few moving parts and easily controlled
- established performance
- constant flowrate.

The disadvantages of centrifugal or steam jet pumps include the following:

- instability at low flowrates
- necessity of shaft packing glands
- slurry dilution caused by priming or flushing water, or steam.

A volume displacement pump has been developed by PNL for use with slurries at low flow rates. The pump, which uses an air-driven piston to transfer fluid, is called an air displacement slurry (ADS) pump. The ADS pump is being used as the melter feed pump at the West Valley Demonstration Project (WVDP) and the radioactive liquid fed ceramic melter (RLFCM). Reliability has been tested in long-term studies and hot cell operation.

The advantages of the ADS pump include the following:

- small size
- submerged operation creating a low NPSH and minimum vessel heel
- positive flow control
- no slurry dilution.

The disadvantages of the ADS pump include the following:

- pulsating flow
- complex flow control
- does not have an established performance record like centrifugal pumps.

Solids settling in process lines does not usually occur if the slurry flow is turbulent. Turbulent flow generally exists at line velocities above 3 to 5 ft/s in 2-in. pipes (based on N_{Re}= 4,000). High fluid velocities will cause excessive erosion and a maximum line velocity of 10 ft/s is often specified (DWPF Basic Data Report 1985).

APPENDIX B

CALCULATIONS

SRAT/SME coils

Calculation procedure

1) Estimate the heat transfer requirement, $\dot{Q}$, in BTU/hr
2) Estimate the overall heat transfer coefficient, U, BTU/hr ft^2°F

Evaporator duty

Assume a 10 gpm evaporation rate (Larson, 1986, p. 5.3)

$$10 \frac{gal}{min}\left(8.34 \frac{lb\ H_2O}{gal}\right)\left(60 \frac{min}{hr}\right) = 5004\ lb/hr\ H_2O\ \text{evaporated}$$ ✓

$$5004\ lb/hr\ \underbrace{\frac{1000\ BTU}{lb}}_{\Delta H_{vap}} = 5.0 \times 10^6 \frac{BTU}{hr}$$ ✓

assume vessel heat losses add 10% to the total heat load

$$\dot{Q} = 5.0 \times 10^6 \frac{BTU}{hr} * 1.1 = 5.5 \times 10^6 \frac{BTU}{hr}$$ ✓

Required overall heat transfer coefficient, U

$$\dot{Q} = UA\Delta T$$

$$U = \frac{\dot{Q}}{A\Delta T}$$

A = 340 ft^2 Conceptual Design Report (CDR) specification number B595C-3244

Steam pressure = 150 psig or 367°F CDR, p. 3.3, v. 1

Assume a boiling temperature of 214°F

$\therefore \Delta T = 367 - 214 = 153°F$

$$U = \frac{\dot{Q}}{A\Delta T} = \frac{5.5 \times 10^6\ BTU/hr}{(340\ ft^2)(153°F)} = 106 \frac{BTU}{hr\ ft^2\ °F} \approx 110 \frac{BTU}{hr\ ft^2\ °F}$$ ✓

Title	Project:
Process Requirements	HWVP - V11030ZA

Prepared by:	Date	Reviewed by:	Date
Evan O. Jones	3/11/87	C.L. [signature]	6/29/87

SRAT/SME coils (con't)

Cooling heat transfer coefficient

Basis: Cool slurry to 122 °F in 8 hours CDR SK-2-91280, rev A

Batch size: 6000 gal. = V Larson, et al 1986 p. 13.6

$$\text{Cooling rate} = \frac{214°F - 122°F}{8\,hrs} = 11.5\,°F/hr = \dot{T} \quad ✓$$

$\dot{Q}$ = Cooling heat transfer rate = $\dot{T} V \rho C_p$

$\rho = 1.3$ Larson, et al 1986 p 13.7

$\rho = 1.3 \times 8.34 = 10.8\ lb/gal$

$C_p = 1\ \frac{BTU}{lb\,°F}$ ≡ similiar to water

$$\dot{Q} = (11.5°F)(6000\,gal)\left(10.8\,\frac{lb}{gal}\right)\left(1\,\frac{BTU}{lb\,°F}\right) = 7.48\times10^5\ \frac{BTU}{hr} \quad ✓$$

$A = 140\ ft^2$ CDR, specification number B595 C15244

Use a log mean ΔT (ΔT_{LM}) for the decrease in the slurry temperature and use an average cooling water (CW) temperature of 95°F

$$\Delta T_{LM} = \frac{(T_{slurry,hot} - T_{cw}) - (T_{slurry,cold} - T_{cw})}{\ln\left[\frac{T_{slurry,hot} - T_{cw}}{T_{slurry,cold} - T_{cw}}\right]} = \frac{(214-95)-(122-95)}{\ln\left[\frac{214-95}{122-95}\right]} \quad ✓$$

$$\Delta T_{LM} = 62°F \quad ✓$$

$$\underline{\underline{U_{cool}}} = \frac{\dot{Q}}{A\,\Delta T_{LM}} = \frac{7.48\times10^5\ BTU/hr}{(140\,ft^2)(62°F)} = \underline{\underline{86\ \frac{BTU}{hr\,ft^2\,°F}}} \quad ✓$$

Melter Feed Tank cooling U

Basis: cooling load = decay heat cooling = 96,300 $\frac{BTU}{hr}$ Larson, et al, 1986. p 8.2

$A = 323\ ft^2$ DWPF drawing, W752133, rev 1

Avg. Cooling water temp. = 95°F

Assume a 120°F slurry temperature with $C_p = 1\ \frac{BTU}{lb\,°F}$

$\Delta T = 120 - 95 = 25°F$

$$U = \frac{\dot{Q}}{A\,\Delta T} = \frac{96{,}300\ \frac{BTU}{hr}}{(323\ ft^2)(25°F)}$$

$$U = 12\ \frac{BTU}{hr\,ft^2\,°F}$$ ✓

SRAT/SME Condenser U

Basis: Condense 10 gpm of vapor (5.0×10^6 BTU/hr, p.1) and cool to 122°F

$$\text{cooling duty, } \dot{q} = 10\ \frac{gal}{min} \Big| \frac{60\ min}{hr} \Big| 212-122°F \Big| \frac{8.34\ lb}{gal} \Big| \frac{1\ BTU}{lb\,°F} =$$

$$\dot{q} = 4.5 \times 10^5\ \frac{BTU}{hr}$$ ✓

$$\dot{Q}_{total} = 5.0 \times 10^6 + 4.5 \times 10^5 = 5.5 \times 10^6\ \frac{BTU}{hr}$$ ✓

Area: use DWPF drawing W752132 rev, 4

3/4" 14 BWG tubing, 7' 4" long, 313 tubes
inside area of tubing = 0.1529 ft²/ft tubing Perry & Chilton 197[illegible]

$$7\tfrac{4}{12}\ \frac{ft}{tube} \Big| 313\ tubes \Big| \frac{0.1529\ ft^2}{ft\ tubing} = 351\ ft^2$$ ✓

ΔT

use 95°F as the surface temperature and 212°F as the saturated vapor temp $\Delta T = 212 - 95 = 117°F$

SRAT/SME condenser

$$U = \frac{\dot{Q}}{A\,\Delta T} = \frac{5.5\times10^{6}\ \frac{BTU}{hr}}{(351\ ft^2)(117°F)} = 134\ \frac{BTU}{hr\ ft^2\,°F} \approx 135\ \frac{BTU}{hr\ ft^2\,°F}$$

Formic acid vent condenser (FAVC)

Basis: $7.2\times10^{4}\ \frac{BTU}{hr}$ CDR, specification number B595C15245

Area: 3/4" 14 BWG, 313 tubes, 4 ft long

$$\frac{4\ ft}{tube}\left|\frac{313\ tubes}{}\right|\frac{0.1529\ ft^2}{ft\ tubing} = 191.4\ ft^2$$ ✓

ΔT:

the FAVC uses a chilled water cooling system, therefore, use a 50°F surface temp.

the FAVC reduces the dewpoint of the vessel vent system off-gas by chilling the off-gas from 150°F to 80°F

$$\Delta T_{LM} = \frac{(150-50)-(80-50)}{\ln\left[\frac{150-50}{80-50}\right]} = 58.1\,°F$$ ✓

$$U = \frac{\dot{Q}}{A\,\Delta T_{LM}} = \frac{7.2\times10^{4}\ \frac{BTU}{hr}}{(191.4\ ft^2)(58.1°F)} = 6.5\ \frac{BTU}{hr\ ft^2\,°F} \approx 10\ \frac{BTU}{hr\ ft^2\,°F}$$ ✓

Melter feed line velocity

Assume a low flowrate of 0.10 gpm and high of 0.75 gpm. (HWVP FDC report)

Use a 3/8" sch 40 pipe (DEL's HWVP Press Chmstry & Tchnlgy, pg 8.26)
& 1" & 2" to bracket possible regions

3/8" sn 40 - 0.00133 ft^2 flow area
1" " - 0.00600 "
2" " - 0.0233 "
} Perry's 5th ed, pg 6-64

$$0.10 \frac{\text{gallon}}{\text{min}} \left| \frac{ft^3}{7.4805\,\text{gal}} \right| \frac{1\,\text{min}}{60\,s} = 2.23 \times 10^{-4} \frac{ft^3}{s}$$

$$0.75\ \text{gpm} \quad '' \quad = 1.671 \times 10^{-3} \frac{ft^3}{s}$$

Linear velocity, ft/s $= \frac{\text{Flow}}{\text{Area}}$

Pipe Flow / Pipe Diameter	3/8"	1"	2"
0.1 gpm	0.17	0.04	0.01
0.75 gpm	1.26	0.28	0.07

Estimate the overall heat transfer coefficient & compare to experimental results with water

Estimate boiling water heat tr coeff using 2 different equations

1) $$h_{predicted} = 0.00658\, P_c^{0.69} \left(\frac{q}{A}\right)^{0.7} \left[1.8\left(\frac{P}{P_c}\right)^{0.17} + 4\left(\frac{P}{P_c}\right)^{1.2} + 10\left(\frac{P}{P_c}\right)^{10}\right]$$ eqn 10-106 Perry's 5th ed

P_c - critical pressure, psia P - system pressure, psia

q/A - heat flux, BTU/hr ft² h - BTU/hr ft² °F

$P = 1$ atm $= 14.7$ psia

$P_c = 218.4$ atm $\times 14.7 = 3210$ psia

$$q/A_{actual} = \underbrace{(15.5°C)}_{\Delta T_{actual}}\underbrace{\left(178\ \frac{BTU}{hr\,ft^2\,°F}\right)}_{U_{actual}}\underbrace{\left(\frac{1.8°F}{1°C}\right)}_{convert\ °C \to °F} = 4970\ \frac{BTU}{hr\,ft^2}$$ from Fig 5-9 Larson et al 1986

$$h_{predicted} = 0.00658(3210\ psia)^{0.69}(4970)^{0.7}\left[1.8\left(\frac{14.7}{3210}\right)^{0.17} + 4\left(\frac{14.7}{3210}\right)^{1.2} + 10\left(\frac{14.7}{3210}\right)^{10}\right]$$

$$h_{predicted} = 486\ \frac{BTU}{hr\,ft^2\,°F}$$

$$h_{actual} = 178\ BTU/hr\,ft^2\,°F$$

2) $$h = 0.225\left(\frac{q\,C_{p_2}}{A\,\lambda}\right)^{0.69}\left(\frac{144\,P\,k_2}{\sigma}\right)^{0.31}\left(\frac{\rho_\ell}{\rho_v} - 1\right)^{0.33}$$ eqn 10-107 Perry's 5th

q/A - heat flux, BTU/ft² hr C_{p_2} - heat cap, BTU/lb °F λ - latent heat, BTU/lb

P - system pressure, psia k_2 - thermal cond, BTU/hr ft °F σ - surface tension, lb_f/ft

Title: Overall heat transfer coefficient

Project: HWVP V110302

Prepared by: Evan O Jones   Date: 1/97

Reviewed by: C L. Taw   Date: 6/30/87

$C_{p_\ell} = 1.01$ BTU/lb°F $\lambda = 970$ BTU/lb = heat of vaporization

$P = 14.7$ psia $k_\ell = 0.393$ BTU/hr ft°F = conductivity of the fluid

surface tension $= \sigma = 58.9 \frac{\text{dynes}}{\text{cm}} = 0.0589 \frac{N}{m} \left| \frac{0.22481 \text{ lb}}{N} \right| \frac{1\text{ m}}{3.2808\text{ ft}} = 4.036\times10^{-3} \frac{\text{lb}}{\text{ft}}$

liquid density $= \rho_\ell = 957.9$ kg/m³ $\rho_s = 0.5956$ kg/m³ = vapor density

$$h = 0.225\left(\frac{4970(1.01)}{970}\right)^{0.69}\left(\frac{144(14.7)(0.393)}{4.036\times10^{-3}}\right)^{0.31}\left(\frac{957.9}{0.5956} - 1\right)^{0.33}$$

$$h_{predicted} = 355 \frac{\text{BTU}}{\text{hr ft}^2\,°\text{F}}$$

Estimate $(q/A)_{max}$

$$(q/A)_{max} = 61.6 \frac{p}{D_o\sqrt{N}}\rho_v \lambda\left[\frac{g\sigma(\rho_\ell - \rho_v)}{\rho_v^2}\right]^{1/4}$$

eqn 10-112 Perry's 3rd ed for tube bundles

p - tube pitch, ft D_o - outside tube diameter N - # tubes

$$(q/A)_{max} = 61.6 \underbrace{\frac{\left(\frac{3.5}{12}\right)\text{ft}}{\left(\frac{2.375}{12}\right)\text{ft}\left(\sqrt{19}\right)}}_{\text{DWPF ref. coil}}\left(0.03718 \frac{\text{lb}}{\text{ft}^3}\right)\left(970 \frac{\text{BTU}}{\text{lb}}\right)\left[\frac{\left(4.18\times10^{8} \frac{\text{lb}_m\text{ ft}}{\text{lb}_f\text{ hr}^2}\right)\left(4.036\times10^{-3}\frac{\text{lb}}{\text{ft}}\right)\left(59.8 - 0.03718\right)}{\left(0.03718\text{ lb/ft}^3\right)^2}\right]$$

$$(q/A)_{max} = (20.83)(18,740)$$

$$(q/A)_{max} = 390,000 \frac{\text{BTU}}{\text{hr ft}^2}$$ ✓

Estimate boiling ht tx coefficient

Use method of Rohsenow, Hartnett, & Ganić Handbook of Heat Transfer Fundamentals p. 13-31

$$\frac{C_{p_l}(T_w - T_{sat})}{\lambda} = C_{sf}\left\{\frac{q''}{\mu_l \lambda}\left[\frac{g_c \sigma}{g(\rho_l - \rho_g)}\right]^{1/2}\right\}^{0.33=r}\left(\frac{C_{p_l}\mu_l}{k_l}\right)^{1.0=s}$$

where

C_{p_l} = heat capacity = 0.8 $\frac{BTU}{lb\ °F}$

λ = heat of vaporization = 970 $\frac{BTU}{lb}$

C_{sf} = constant = 0.013 range 0.003 – 0.015

μ_l = liquid viscosity = 2 cP = 4.838 $\frac{lb}{ft\text{-}hr}$

σ = surface tension = 4.036×10^{-3} $\frac{lb}{ft}$

ρ_l = 68.6 $\frac{lb}{ft^3}$ ρ_g = 0.04 $\frac{lb}{ft^3}$

k_l = 0.42 $\frac{BTU}{hr\ ft\ °F}$

q'' = heat flux, $\frac{BTU}{ft^2\text{-}hr}$

Determine h which equals $\frac{q''}{(T_w - T_{sat})}$

$$\frac{\left(0.8 \frac{BTU}{lb\ °F}\right)(\Delta T)}{970 \frac{BTU}{lb}} = (0.013)\left\{\frac{q''}{4.838\frac{lb}{ft\text{-}hr}\left(970\frac{BTU}{lb}\right)}\left[\frac{4.036\times10^{-3}\frac{lb}{ft}}{(68.6-0.04)\frac{lb}{ft^3}}\right]^{-1/2}\right\}^{0.33}\left(\frac{(0.8)(4.83}{0.42}\right.$$

To isolate $q''/\Delta T$, will need to cube equation to get $(q'')^1$, resulting in $(\Delta T)^3$. Will need to multiply left hand side by $(\Delta T)^2$ & divide q'' by (ΔT)

Assume a ΔT = 100°F, probably a range of 30 – 130°F

$$\frac{q''}{\Delta T} = \left[\frac{(0.8)}{970}\;\frac{(0.42)}{(0.013)(0.8)(4.838)}\right]^3 \frac{(4.838)(970)}{\left[\frac{4.036\times10^{-3}}{68.6}\right]^{1/2}} (100)^2 \; \Delta T$$

$$h_{boil} = \frac{q''}{\Delta T} = 2000 \frac{BTU}{hr\,ft^2\,°F}$$ ✓

U for boiling water (Larson et al. 1986 pag 5-22) was 180 to 220 $\frac{BTU}{hr\,ft^2\,°F}$. McAdams (pg 376-383) range from 280 to 9000 $\frac{BTU}{hr\,ft^2\,°F}$. The low values in McAdams result from dirty surfaces. Rohsenow et al & McAdams make it very clear the importance of surface effects. Both remark that predicting boiling heat transfer coefficients cannot be performed with confidence unless detailed knowledge of the surface is provided. Since this information is not known, particularly after a period of time, heat tx coefficients must be determined experimentally.

ENGINEERING WORKSHEET

Prepared By: Evan O Jones | Date: 12/4/86 | Project: HWVP V110302

Title/Subject: Estimate level of agitation and motor requirements for the SRAT/SME

Reviewed by: C. L. [signature] Date 6/30/87

As an initial assessment, estimate the total flow of the flat blade radial impeller at 130 RPM, estimate the bulk fluid velocity to determine the Chemscale degree of mixing and determine the required hp. Use relationships outlined by Oldshue (1983) Chap 8.

The tank is a cylindrical vessel with a 12 ft ID, 3 ft impellers, nested helical heat trans coils and operating liquid level of 11 ft. See Fig 4.1 in the report

Physical/Chemical data

The melter feed is a slurry of small particles with a low settling velocity. Consequently, the mixture can be treated as a pure fluid. However, as a slurry the melter feed is a non-Newtonian pseudoplastic material.

Oldshue (1983) describes methods for evaluating agitation of non-Newtonian fluids. However, it is concerned with non Newtonian fluids of 5000 to 50,000 Since the melter feed is less than 500 cP, this analysis will not be applied. Alternatively, the system will be analyzed as a Newtonian fluid with a conservative (high) assumption for viscosity. In other words, the fluids analyzed do not exhibit such strongly pseudoplastic behavior to be considered non Newtonian.

Table 3.3 lists viscosities of melter feed at $35\ s^{-1}$. Use 400 cP as a conservative estimate of viscosity.

Calculation

$$\text{Pumping Number} = N_Q = \frac{Q}{ND^3}$$

where
Q = flow, ft^3/min
N = shaft speed, RPM
D = impeller diameter, ft

$$N_{Re} = \text{Reynolds Number} = \frac{10.7\,\rho N D^2}{\mu} \quad \text{eqn 3 Hicks (ref}$$

where
ρ = specific gravity
N = shaft speed, RPM
D = impeller diameter, i.
μ = viscosity, cP

First calculate N_{Re}

$$N_{Re} = \frac{(10.7)(1.4)(130\ RPM)(36\ in)^2}{400}$$

$$N_{Re} = 6,310 \equiv \text{turbulent i.e.} > 1000$$

ENGINEERING WORKSHEET

Prepared By: Evan O Jones | Date: 12/4/86 | Project: HWVP V110302

Title/Subject: SRAT/SME agitation

Reviewed by: C. L. [illegible] Date: 6/30/

Use table 8-2, pg 171, Oldshue 1986 to determine the power number

$$N_Q = 0.9 = \frac{Q}{ND^3}$$

$$Q = (0.9)(130\ \text{min}^{-1})\left(\frac{36}{12}\text{ft}\right)^3$$

$Q = 3160\ \text{ft}^3/\text{min}$ from a "free" 3 ft diameter radial flow paddle impeller

In a radial flow impeller, ~50% of the total flow circulates above the impeller. Locating a radial flow impeller near the vessel bottom reduces impeller total flow, at 1/2 the impeller diameter the flow decreases 10 to 15%. (Oldshue, pg 177-178)

Off bottom clearance

$$\frac{C}{D} = \frac{\text{off bottom clearance (to paddle mid-line)}}{\text{impeller diameter}} = \frac{17\ \text{in}}{36\ \text{in}} = 0.47$$

Therefore, the impeller pumping rate is reduced ~15%

In addition, the upper half of the paddle width is surround by the vessel coils. This is further impedes development of free liquid flow. The presence of the coils adds a resistance or head to the predicted flow. No documentation was found that predicted the head or flow reduction caused by the coils. Assume the impeller flow is reduced ~1/3 by the coils.

Total flow reduction amounts to approximately 50%

$$\text{Net flow} = 3160\frac{\text{ft}^3}{\text{min}} \times 1/2 = 1580\frac{\text{ft}^3}{\text{min}}$$

Estimate the annular area

$$\frac{\pi}{4}\left(\underbrace{(12\ \text{ft})^2}_{\text{tank diameter}} - \underbrace{\left(\frac{60}{12}\text{ft}\right)^2}_{\text{coil diameter}}\right) = 93.5\ \text{ft}^2 = A_{\text{Cross Section}}$$

Bulk annular velocity, u_b

$$u_b = \frac{Q}{A_{cs}} = \frac{1580\ \text{ft}^3/\text{min}}{93.5\ \text{ft}^2} = 17\ \text{ft/min} \equiv \text{Chemscale agitation level of 3}$$

A bulk fluid velocity of 17 ft/min corresponds to a Chemscale level of agitation of 3 for blending & motion. If full impeller flow is assumed, then $u_b = 34$ ft/min and a Chemscale agitation level of 5 to 6 is obtained. The top impeller does not create additional flow in the tank as described in the following section

Prepared By: Evan O Jones | Date: 12/5/86 | Project: HWVP - V110302
Title/Subject: SRAT/SME agitation
Reviewed by: C. L. [illegible] Date: 6/30/8

Axial flow turbine

In the SME, the axial flow turbine is located about 1/4 of the coil length inside the coil. (see Fig 4.1). It has approximately a 1 1/2" gap between the impeller tip and coil supports and 4 in between the tip and coils. These close gaps create a draft tube effect, consequently, the turbine.. will be modeled by a method outlined by Oldshue, Chap 20, section C, and Chapter 8

Fluid properties

$$\left.\begin{array}{l}\rho_{sl} = 1.4 \text{ g/ml} \\ \mu_{sl} = 400 \text{ cP}\end{array}\right\} \text{pg B-10}$$

For the reasoning in previous calculation it will be assumed the slurry can be modeled as a Newtonian fluid with given viscosity.

Impeller flow rate

Use similiar analysis as for a radial flow impeller, only the values of the power number, Np, and pumping number, N_Q, will change.

N_{Re} is independent of the type of impeller, therefore the high effici impeller operates in the turbulent regime as well.

Data for a high efficiency impeller indicate a decrease in power number and increase in pumping number for draft tube service (p. 181, Oldshue, 1983). Conservati estimates would use upper limits of N_Q & lower limits of N_p for high efficiency impellers open tanks.

$N_Q = 0.64$ Table 8-3, pg 173, Oldshue, 1983

$$N_Q = \frac{Q}{ND^3}$$

$$Q = (0.64)(130 \text{ RPM})\left(\frac{36}{12}\text{ ft}\right)^3$$

$$\underline{Q = 2250 \text{ ft}^3/\text{min} =}$$

ENGINEERING WORKSHEET

Prepared By: Evan O Jones | Date: 12/5/86 | Project: HWVP V110302

Title/Subject: SRAT/SME agitation

Reviewed by: C. L. Fow Date: 6/30/87

The open tank flowrate of the lower radial flow impeller is 3160 ft^3/min. The top axial flow impeller produces $\frac{2250\ ft^3/min}{3160\ ft^3/min} = 70\%$ of the "free" lower impeller flowrate. The calculation of the actual flow from the lower impeller assumed the actual, impeded flow would be 50% of the "free" flow. However, the margin of error for this estimate is probably ± 20%. Therefore the flow produced by the top propeller could removed by the bottom impeller. This implies no significant flow occurs through the coil bank and all of the slurry passes down the coil and out the bottom

Annulus superficial velocity and ChemScale agitation level

The top axial flow impeller produces no additional flow to the tank annulus. As described above, all the slurry flow produced by the top impeller exits out the bottom of the coil with the bottom impeller. Therefore, the actual flow is estimated to be flowrate of the top impeller, or 2250 ft^3/min

$$u_b = \frac{Q}{A_{cs}} = \frac{2250\ ft^3/min}{93.5\ ft^2} = 24\ \frac{ft}{min} \equiv \text{Chemscale of } 4 \quad \text{for blending \& motion}$$

Power requirement

Axial Flow Impeller

$N_p = 0.43$ Table 8.3, pg 172, Oldshue

$$P = \frac{N_p N^3 D_a^5}{1.523\times10^{13}} = \frac{(0.43)(130\ min^{-1})^3(36\ in)^5(1.4)}{1.523\times10^{13}}$$ p. B-14

$P = 5$ hp

Agitator Power Requirements (cont)

Radial flow impeller

$$N_p = \frac{1.523 \times 10^{13} P}{N^3 D^5 \rho}$$ eqn 3-12 p. 51 Oldshue 1983

where

P = impeller hp
N = impeller speed, RPM
D = impeller diameter, in.
ρ = fluid specific gravity

N_p = power number = 3.3 Table 8.2, p. 171, Oldshue 1983

Therefore,

$$P = \frac{(3.3)(130\,RPM)^3(36\,in)^5(1.4)}{1.523 \times 10^{13}}$$

P = 40 hp for the radial flow impeller

Impeller	Open Tank Flowrate ft^3/min	Actual Flowrate ft^3/min	Power Consumption hp
top - axial flow	1930	1930 - does not increase total vessel flow	5
bottom - radial flow	3160	1580	40
Total vessel flow and power consumption		1580	45

Assuming a 90% motor efficiency

$$\text{motor hp} = \frac{45\,hp}{0.9} = 50\,hp$$

Estimate h_o for DWPF helical coils

Use relationship by Oldshue to estimate h_o. Relationship is for single coil, nested coils will have a reduced effectiveness of 70-90%. (Oldshue, pg 284)

$$N_{Nu} = 0.17\, N_{Re}^{0.67}\, N_{Pr}^{0.37} \left(\frac{D}{T}\right)^{0.1} \left(\frac{d}{T}\right)^{0.5} \left(\frac{\mu}{\mu_s}\right)^{m}$$ ✓ eqn 14-7, p284, Oldshue 1983

$$\left.\begin{array}{l} D - \text{impeller diameter} = 36'' = 3\,ft. \\ d - \text{tube diameter} = \frac{2.375}{12} = 0.1979\,ft. \\ T - \text{tank diameter} = 12\,ft \end{array}\right\}$$ DWPF drawing W752193 rev 5

Using Larson et al 1986 for μ; use 50 cP for fluid & $\mu/\mu_s = 1.05$. In addition, let $m = 0.25$ (p. 284; Uhl and Gray, 1966)

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

✓

$\rho = 1.1 \times 62.4 = 68.6\ lb/ft^3$ House 1986

$N = 130\ \frac{1}{min}$

$D = 3\,ft$

$\mu = 50\,cP \times 2.419\ \frac{lb}{ft\text{-}hr} \times \frac{hr}{60\,min} = 2.016\ \frac{lb}{ft\ min}$ ✓

$$N_{Re} = \frac{(3ft)^2 (130\ \frac{1}{min})(68.6\ lb/ft^3)}{(2.016\ \frac{lb}{ft\ min})}$$

$N_{Re} = 39{,}820$ turbulent ✓

$$N_{Pr} = \frac{c_p \mu}{k}$$

$c_p = 0.8\ \frac{BTU}{lb\,°F}$ Table D-6, pg D-18, Larson et al 1986

$$\mu = 50\,cP \times 2.419 \frac{lb}{ft\text{-}hr} = 121 \frac{lb}{ft\text{-}hr}$$ ✓

Estimate k assuming a slurry of SiO_2 & water. Use a method described in Perry's 5th ed, pg 3-243

$$k_m = k_c \frac{2k_c + k_d - 2\phi_d(k_c - k_d)}{2k_c + k_d + \phi_d(k_c - k_d)}$$ ✓

where

k_m - mean or slurry conductivity

k_d - discontinuous or solid k

k_c - continuous or fluid k

ϕ_d - volume fraction of solids

Estimate ϕ_d

use 25.8 wt% solids (CM House) & water density = 1 & solid density = 2.5

Basis: 100 gm of slurry

$$74.2\,g\,H_2O \left| \frac{cc}{1\,g} \right| = 74.2\,cc$$

$$25.8\,g\,\text{solid} \left| \frac{cc}{2.5\,g\,\text{solid}} \right| = 10.3\,cc$$

$$\phi_d = \frac{10.3}{74.2 + 10.3} = 12.2\text{ vol \%}$$

$k_c = 0.4$ BTU/hr ft °F (water)

$k_d = 0.6$ BTU/hr ft °F (McAdams, 1954, Table A-5, borosilicate glass)

$$k_m = 0.4\left[\frac{2(0.4) + 0.6 - 2(.122)(0.4 - 0.6)}{2(0.4) + 0.6 + 0.122(0.4 - 0.6)}\right]$$

$k_m = 0.42$ BTU/hr ft °F

Title: helical coil ht. x. coeff.	Project: HWVP V1103==
Prepared by: Evan O Jones, Date: 3/87	Reviewed by: C. L. [illegible], Date: 6/30/87

$$N_{Pr} = \frac{c_p \mu}{k} = \frac{\left(0.8 \frac{BTU}{lb\,°F}\right)\left(121 \frac{lb}{ft\text{-}hr}\right)}{\left(0.42 \frac{BTU}{hr\,ft\,°F}\right)}$$

$$N_{Pr} = 230$$

Use eqn 14-7, p 284, Oldshue 1983

$$N_{Nu} = \frac{h_o D}{k} = (0.17)(39{,}820)^{0.67}(230)^{0.37}\left(\frac{3}{12}\right)^{0.1}\left(\frac{0.1979}{12}\right)^{0.5}(1.05)^{0.25}$$

$$= 173.8 \checkmark$$

$$h_o = \frac{173.8\,(0.42\ BTU/hr\,ft\,°F)}{(3\ ft)}$$

$$h_o = 24 \frac{BTU}{hr\,ft^2\,°F} \quad w/\ \mu = 50\ cP \checkmark$$

Assuming various viscosities and all other properties constant, estimate the heat transfer coefficient

Isolate μ from equation 14-7. It's inside N_{Re} & N_{Pr}

$$N_{Nu} = 0.17\left(D^2 N \rho\right)^{0.67}\left(\frac{1}{\mu}\right)^{0.67}\left(\frac{c_p}{k}\right)^{0.37}(\mu)^{0.37}\left(\frac{D}{T}\right)^{0.1}\left(\frac{d}{T}\right)^{0.5}\left(\frac{\mu}{\mu_s}\right)^{0.25}$$

$$N_{Nu} = \left(\frac{1}{\underset{\frac{lb}{ft\text{-}hr}}{\mu}}\right)^{0.3}\left[0.17\left(D^2 \overset{hr^{-1}}{N} \rho\right)^{0.67}\left(\frac{c_p}{k}\right)^{0.37}\left(\frac{D}{T}\right)^{0.1}\left(\frac{d}{T}\right)^{0.5}\left(\frac{\mu}{\mu_s}\right)^{0.25}\right]$$

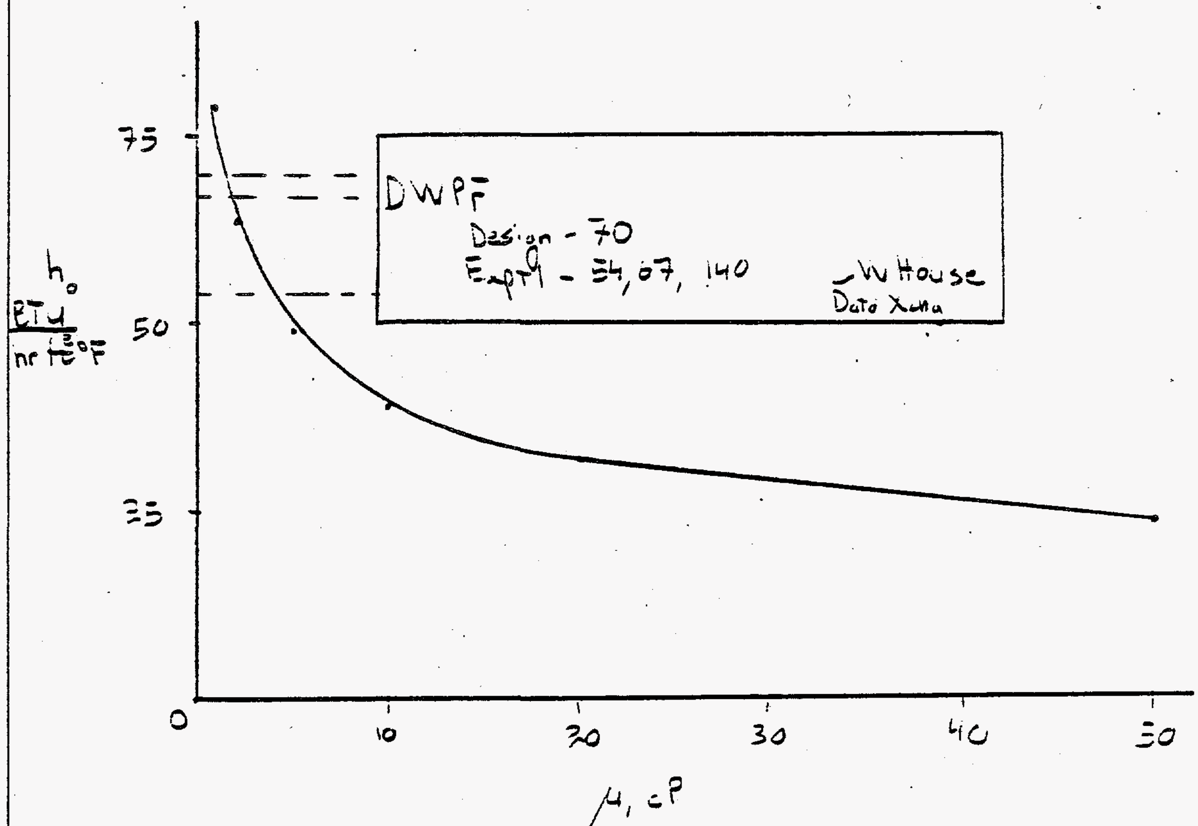
Substituting values above

$$N_{Nu} = \left(\frac{1}{\mu}\right)^{0.3}\left[0.17\left(3^2\cdot(130\times60)\cdot 68.6\right)^{0.67}\left(\frac{0.8}{0.42}\right)^{0.37}\left(\frac{3}{12}\right)^{0.1}\left(\frac{0.1979}{12}\right)^{0.5}(1.05)^{0.25}\right]$$

$$N_{Nu} = \left(\frac{1}{\mu}\right)^{0.3}[733.0] \Rightarrow h_o = \left(\frac{1}{\mu}\right)^{0.3}[733.0]\left(\frac{0.42}{3}\right) = \left(\frac{1}{\mu}\right)^{0.3}(103)$$

μ cP (lb/ft-hr)	N_{Nu}	h_o BTU/hr ft^2 °F
1 (2.419)	562	79
2 (4.84)	457	64
5 (12.10)	347	49
10 (24.19)	282	39
20 (48.38)	229	32
50 (121.0)	174	24
100 (242)	141	20

This calculation is based on a single helical coil. Oldshue (pg 284) states that additional banks of coils will only be 70 to 90% as effective as a single bank. Assuming an effective μ of 1 to 2 cP, h_o is reduced from 45 to 70 BTU/hr ft^2 °F.



Forced Convection - Tube Banks

Use method suggested by McAdams, 1954 pg 275-276

$N_{Re} = 790.2$

From Fig 10-20, McAdams 1954

$$Y = \frac{h' D_o}{k} N_{Pr}^{-1/3} = 20$$

$$\frac{h' D_o}{k} = \frac{20}{(9.22)^{-1/3}} = 41.9 \checkmark$$

$$h' = \frac{41.9 \times 0.42}{0.1979} = 89 \ \frac{BTU}{hr\ ft^2\ °F} \checkmark$$ for bank of 10 tubes

For a bank of 3 tubes, $h_3 = h' N$ & $N = 0.82$, Table 10-6, McAdams 1954

$h = 89\ (0.82)$

$$h_{\substack{3\ tubes \\ forced\ conv}} = 73 \ \frac{BTU}{hr\ ft^2\ °F} \checkmark$$

Solids DF between the condensate and feed

Data source: Fluor drawing SK-2-91287 rev. A

SRAT

	SRAT product	SRAT condensate
Total solids, wt%	12.94	0.01687

$$\text{SRAT solids DF} = \frac{12.94}{0.01687} = 770 \approx 800$$

SME

	SME product	SME condensate
Total solids, wt%	34.45	0.04458

$$\text{SME solids DF} = \frac{34.45}{0.04458} = 770 \approx 800$$

Condenser Heat Transfer Coefficient

Initially assume film condensation of a pure saturated vapor. Assume the geometry in DWPF drawings; namely, downflowing vapor inside 3/4" x 14 BWG 7' 4" long, 313 tubes

Assume:

Condensing 10 gpm steam

Cooling Water temp: 70 & 90°F

Condensing temp: 212°F

Important parameters

Γ - mass flow rate, $\frac{w}{\pi D}$

λ - latent heat of vaporization, $\frac{BTU}{lb}$

μ_f - Viscosity of fluid, $\frac{lb\text{-}ft}{hr}$

ρ_f - density, $\frac{lb}{ft^3}$

ΔT - temperature difference between vapor & wall temp

k_f - conductivity of condensate, $\frac{BTU}{hr\ ft\ °F}$

L - tube length

Use 2 methods outlined by McAdams, 1954 pg 325-333

Method 1

Estimate Γ

$$w = 10 \frac{gal}{min} \left| \frac{8.34 \text{ lb water}}{1 \text{ gal}} \right| \frac{60 \text{ min}}{hr} \left| \frac{}{313 \text{ tubes}} \right| = 16 \frac{lb}{hr} \text{ per tube}$$

$$\Gamma = \frac{16 \text{ lb/hr}}{\pi \left(\frac{0.584}{12}\right)} = 104.6 \frac{lb}{hr\ ft}$$

Perry's 5th, Table 11-2, pg 11-2
3/4" 14 BWG

Title: Condenser Heat Transfer Coefficient

Project: AWVP V110302

Prepared by: Evan O Jones | Date: 3/87 | Reviewed by: C. L. [illegible] | Date: 6/30/87

Estimate $\frac{4\Gamma}{\mu_f} = N_{Re,f}$

$$T_f = T_{sat\ vapor} - 3/4\left(T_{sat\ vapor} - T_{wall}\right)$$

$$T_f = 212 - 3/4(212 - 90) = 120°F \quad (105°F \text{ for } T_{wall} = 70°F)$$

Perry and Chilton 5th ed 1973 $\left\{ \mu_{120°F} = 0.6\ cP \times 2.419 = 1.451 \frac{lb}{hr\,ft} \qquad \mu_{105°F} = 0.69\ cP \times 2.419 = 1.669 \frac{lb}{hr\,ft} \right.$

With $T_f = 120°F$ ($T_{wall} = 90°F$) and $T_f = 105°F$ ($T_{wall} = 70°F$)

$$\frac{4\Gamma}{\mu_f} = \frac{4\left(104.6 \frac{lb}{hr\,ft}\right)}{1.451 \frac{lb}{hr\,ft}} = 288 \text{ @ } T_f = 120°F\ (T_{wall} = 90°F)$$
$$= 251 \text{ @ } T_f = 105°F\ (T_{wall} = 70°F)$$

From Fig 13-6, line $A' - B'$ or $h_m\left(\frac{\mu_f^2}{k_f^3 \rho_f^2 g}\right) = 1.28\left[1.47\left(\frac{4\Gamma}{\mu_f}\right)^{-1/3}\right]$ eqn 13-4 mult by 1.28

$$h_m\left(\frac{\mu_f^2}{k_f^3 \rho_f^2 g}\right)^{1/3} = 1.882\underbrace{(288)^{-1/3}}_{T_f = 120°F} = 0.285 \quad (0.298 \text{ @ } T_f = 105°F)$$

At $T_f = 120°F$
$\mu_f = 1.451 \frac{lb}{hr\,ft}$
$k_f = 0.309 \frac{BTU}{hr\,ft\,°F}$
$\rho_f = 61.8 \frac{lb}{ft^3}$

At $T_f = 105°F$
$\mu_f = 1.669 \frac{lb}{hr\,ft}$
$k_f = 0.363 \frac{BTU}{hr\,ft\,°F}$
$\rho_f = 62.1 \frac{lb}{ft^3}$

McAdams 1954

$$g = 4.17 \times 10^8 \frac{ft}{hr^2}$$

$$h_m = 0.285\left(\frac{\left(1.451 \frac{lb}{hr\,ft}\right)^2}{\left(0.369 \frac{BTU}{hr\,ft\,°F}\right)^3 \left(61.8 \frac{lb}{ft^3}\right)^2 \left(4.17 \times 10^8 \frac{ft}{hr^2}\right)}\right)^{-1/3}$$

$$h_m = 958 \frac{BTU}{hr\,ft^2\,°F} \text{ @ } T_f = 120°F \quad (\text{cooling water temp} = 90°F)$$
$$h_m = 901 \frac{BTU}{hr\,ft^2\,°F} \text{ @ } T_f = 105°F \quad (\text{cooling water temp} = 70°F)$$

Method 2

Using the theoretical value for steam condensing @ 1 atm w/ film condensation

$$h_m = \frac{4000}{L^{1/4}\,\Delta T^{1/3}} \quad \text{eqn 13-5 McAdams 1954}$$

where

L = length, feet = 7.33 ft

ΔT = temp diff, °F = 120 or 105 °F

$T_f = 120°F$
$$h_m = \frac{4000}{(7.333)^{1/4}(120)^{1/3}} = 493\ \frac{BTU}{hr\,ft^2\,°F}$$

$T_f = 105°F$
$$h_m = \frac{4000}{(7.333)^{1/4}(105)^{1/3}} = 515\ \frac{BTU}{hr\,ft^2\,°F}$$

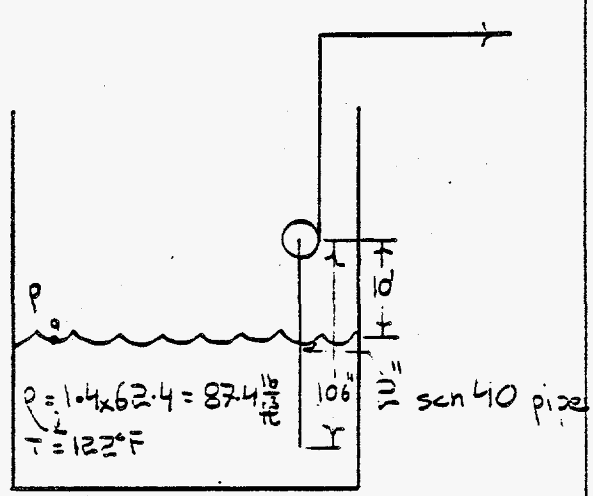
Theoretical values; the higher values calculated before were based on empirical results

This analysis did not include the effect of

- Dropwise condensation – results in higher h_m
- Non-condensible gases – decrease h_m (see Fig 3, pg 11-40, Rohsenow et al)
- Organic vapors – can increase or decrease h_m
- Metallurgy – h_m can vary ≥ 100% depending on tube metallurgy (see Fig 13-12, McAdams or Fig 3
- Fouling – reduce h_m

Example calculation of NPSH available

$$NPSH = \frac{P_a - P_v - P_g}{\rho_l} - Z\frac{g}{g_c} - h_{f_s}$$



$P_a = 14.6$ psig

$P_v = 1.8$ psig - H_2O @ 122°F

$P_g = 0$ = pressure of non-condensible gas in liquid

$Z = 10$ ft

Estimate the suction pipe head loss, h_{f_s}

Use two different methods to estimate h_{f_s}

a) Assume a Newtonian fluid with $\mu = 10$ cP and using a traditional Fanning friction factor

b) Assume a pseudoplastic fluid with $n = 0.8$ and $K = 0.01$ Pa-s (or 10 cP) and a Hanks N_{Re}

a) Newtonian fluid

$$N_{Re} = \frac{D \bar{v} \rho}{\mu}$$

2" sch 40 pipe

$D = \frac{2.067}{12} = 0.1723$ ft Perrys 5th, pg 6-64 & area $= 0.0233$ ft²

$\rho = 87.4 \frac{lb}{ft^3}$ $\mu = \frac{10}{1488} = 6.72 \times 10^{-3} \frac{lb}{ft\text{-}s}$

$\dot{V} = 100 \frac{gallon}{min} \left| \frac{1\ min}{60\ s} \right| \frac{ft^3}{7.4805\ gallon} \Big| \times \frac{1}{0.0233\ ft^2} = 9.562 \frac{ft}{s}$

$$N_{Re} = \frac{(0.1723\ ft)(9.562\ \frac{ft}{s})(87.4\ \frac{lb}{ft^3})}{6.72\times10^{-3}\ \frac{lb}{ft\text{-}s}} = 21,400$$ ✓

Fanning friction factor, $f = 0.0065$ Fig 5-26, Perry's 5^{th} ed, 1973
for hydraulically smooth pipe

$$\Delta P_{pipe} = \rho \underbrace{\left(\frac{4fL}{D}\right)}_{\text{velocity head losses}} \underbrace{\left(\frac{\dot{v}^2}{2g_c}\right)}_{\text{velocity head}}$$ ✓ Perry's 5^{th} ed, pg 5-21

$$\Delta P_{pipe} = (81.4\ \tfrac{lb}{ft^3})\left(\frac{4(0.0065)(10.5\ ft)}{(0.1723\ ft)}\right)\left(\frac{(9.562\ ft/s)^2}{2(32.17\ ft/s^2)}\right)$$

$$\Delta P_{pipe} = 81.4(1.58)(1.42) = 183\ \frac{lb}{ft^2}$$ ✓

$$\Delta P_{pipe} = 183\ \frac{lb}{ft^2} \times \frac{1\ ft^2}{144\ in^2} \times \frac{34\ ft}{14.7\ psia} \times \frac{1}{1.4} = 2.1$$ equivalent feet of head ✓

$$\Delta P_{entrance} = \rho K_c\left(\frac{\dot{v}^2}{2g_c}\right)$$

assume $K_c = 0.5$ Perry's 5^{th} ed, Table 5-18

$$\Delta P_{entrance} = (81.4)(0.5)(1.58) \times \frac{1}{144} \times \frac{34}{14.7} \times \frac{1}{1.4} = 0.7$$ equivalent feet of head

$\underline{\underline{h_{fs}}} = 2.1 + 0.7 = \underline{\underline{2.8\ \text{ft of equivalent head}}}$ ✓

b) Use method outlined in McCarty, Chan, and Lokken 1986 p 5.14 - 5.17 and Figure 25

Calculate a Metzner-Reed N_{Re} $(N_{Re})_{M-R}$

$$(N_{Re})_{M-R} = \frac{D^n V^{2-n} \rho}{\gamma} \quad \text{eqn 20 p 5.15}$$

where

$$\gamma = g_c \left[K \left(\frac{3n+1}{4n} \right)^n \right] 8^{n-1} \quad \text{eqn 17 \& 19}$$

$$K = 0.01 Pa\text{-}s \times 1000 \frac{cP}{Pa\text{-}s} \times 0.00067197 \frac{\frac{lb_s}{ft\text{-}s}}{cP} = 0.000672 \frac{lb}{ft\text{-}s}$$

$$\gamma = (32.2)(0.000672)\left(\frac{3(0.8)+1}{4(0.8)} \right)^{0.8} \left(8^{0.8-1} \right) = 0.15 \checkmark$$

$$(N_{Re})_{M-R} = \frac{(0.1723\ ft)^{0.8} (9.562\ ft/s)^{2-0.8} (87.4\ lb/ft^3)}{(0.15)}$$

$(N_{Re})_{M-R} = 2100$ ✓ ∴ $f = 0.008$ Figure 25 ✓

$$\Delta P\ pipe = \underbrace{(81.4\ lb/ft^3)}_{\text{density}} \underbrace{\left[\frac{4(0.008)(10.5\ ft)}{(0.1723\ ft)} \right]}_{\text{velocity head loss}} \underbrace{(1.42)}_{\text{velocity head}} = (81.4)(1.95)(1.42)$$

✓

$$\Delta P\ pipe = 225\ lb/ft^2 \times \frac{ft^2}{144\ in^2} \times \frac{34\ ft}{14.7\ psia} \times \frac{1}{1.4} = 2.6 \text{ equivalent feet of head}$$

$$h_{fs} = 2.6 + \underbrace{0.7}_{\text{entrance loss}} = 3.3 \text{ ft of equivalent head} \checkmark$$

For assumption a)

$$NPSH = \frac{(14.6 - 1.8 - 0)\frac{34}{14.7}}{1.4} - 10\ \text{ft} - 2.8$$

$$NPSH = 11.1 - 2.8$$

$$\underline{\underline{NPSH = 8.3 \text{ ft of liquid}}}$$

For assumption b)

$$NPSH = 11.1 - 3.3$$

$$\underline{\underline{NPSH = 7.8 \text{ ft of liquid}}}$$

The Lawrence pump requires ~ 6 feet of NPSH, therefore the available NPSH should be adequate to prevent cavitation. This assumes there are no addition suction losses due to increased rheology or entrance losses

ENGINEERING WORKSHEET

Prepared By: Evan O Jones | Date: 12/2/86 | Project: HWVP -V110302

Title/Subject: Agitator Requirements - Frit Slurry Make-up Tank

Reviewed by: C L Fr [illegible] Date: 1/30/87

Use method described by Chem Eng. articles (December 8, 1975 - Dec 6, 1976) written by Chemineer personnel. If solid settling velocity is less than 0.5 ft/min the problem becomes a motion & blending problem

Calculation procedure - for solids suspension, (Gates et al (1976))

1. Calculate the equivalent volume, V_{eq}
2. Calculate the solids settling velocity
3. Choose level of solids suspension and scale of agitation
4. Calculate the required power and shaft speed
5. Determine system geometry
6. Determine impeller design: axial flow

(1)

$V_{eq} = (\rho_{sl}) V_T$ (eqn 1, Gates et al (1976))

where

ρ_{sl} - slurry specific gravity

V_T - volume of agitated slurry

$V_T = 2800$ gal (Fluor PFD, drawing # SK-2-90720)

$\rho_{sl} = 1.50$ (Fluor spec sheet, # B-595-C-18210, Frit Slurry Tank's Agitator)

$V_{eq} = 2800 \times 1.5 = \underline{4200 \text{ gal}}$ ✓

ENGINEERING WORKSHEET

Prepared By: Evan O Jones | Date: 12/2/86 | Project: HWVP \ 110302

Title/Subject: Agitator Power Requirements - FSMT

Reviewed by: C L [signature] Date: 6/30/8

(2) First need to calculate the difference in specific gravity between the solid frit and water

$\rho_s = 156\ lb/ft^3$ (Fluor spec sheet, Frit Weigh Bin w/ Actuator, #B-595-C-1456

$\rho_s = \frac{156}{62.4} = 2.5$

$\rho_l = 1$

$\rho_s - \rho_l = 1.5$

From Fig 2 (Gates et al (1976)) and size range -80 - +200 (Fluor spec #B595-C-014562) determine terminal settling velocity

Use 80 mesh & $\Delta\rho = 1.5$

$U_t = 5\ ft/min$; $U_d = U_t f_w$ where f_w is from Table 1, Gates et al (1976); U_d is settling velocity corrected by slurry density (hindered settling)

with solids conc = 60%, $f_w \cong 2$

$U_d = 2(5) = 10\ ft/min$

(3) Choose an agitation level of 9 which provides slurry uniformity of solids to 98% of the fluid batch height. Slurry drawoff suitable by overflow.

(4) From Table III, Gates et al (1976) the power/speed combinations for 5000 gal, with $U_d = 10\ ft/min$ and scale of agitation of 9 are below

hp	rpm	
40	155	DWPF falls between these brackets, though the 100hp motor is oversized; use the higher speed (155RPM) and
30	100	
25	84	also use 68 RPM for a comparison
20	68	

ENGINEERING WORKSHEET

Prepared By: Evan O Jones | Date: 12/3/86 | Project: HWVP V110302

Title/Subject: Frit Slurry Make Up Tank

Reviewed by: C. L. [illegible] Date: 6/30/

(5) Estimate # impellers, bottom clearance height and impeller diameter, (D_T)

Calculate $\frac{Z}{T} = \frac{\text{liq depth}}{\text{vessel diameter}} = 1$ since vessel height = 7' 6" and T = 8'

Therefore, 1 impeller at a bottom clearance of Z/4 is required

Bottom clearance = $\frac{8 ft}{4} = 2$ ft   Baffles - 4 baffles, equally spaced, width = $\frac{T}{12}$, offset from wall $\frac{T}{72} = 1\frac{1}{2}"$

(6) Estimate D_T

$$D_T = 394 \left(\frac{H_p}{n N^3 (\rho)_{sl}} \right)^{0.2}$$ eqn 3, Gates et al, (1976)

where H_p - power, hp
n - # impellers
N - shaft speed, RPM
ρ_{sl} - slurry specific gravity

choose the 20 hp / 68 RPM

$$D_T = 394 \left(\frac{20}{(1)(68)^3 (1.5)} \right)^{0.2}$$

$D_T = 53$ in $= 4' 5"$ $D_T/T = \frac{53 in}{8 \times 12 in} = 0.55 = \frac{\text{impeller diameter}}{\text{tank diameter}}$

choose 40 hp and 155 RPM

$$D_T = 394 \left(\frac{40}{(1)(155)^3 (1.5)} \right)^{0.2}$$

$D_T = 37" = 3' 1"$

$$\frac{D_T}{T} = \frac{37"}{8 \times 12"} = 0.39$$

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ENGINEERING WORKSHEET

Prepared By: Evan O Jones | Date: 12/3/86 | Project: HWVP V110302
Title/Subject: FSMT

Reviewed by: C L [signature] Date: 6/30/8

The following is typical for blending & motion mixing, as to mix slurrys of fine particles
Estimate the bulk fluid velocity, V_b, using blending and motion analysis and assuming a slurry viscosity of 100 cP

Estimate N_{Re}

$$N_{Re} = \frac{10.7 \rho_{sl} N D^2}{\mu}$$ eqn 3 Hicks, et al (1976)

where ρ_{sl} - fluid specific gravity
N - shaft speed, RPM
D - impeller diameter, in
μ - viscosity, cP

Use 68 RPM & 53" impeller, D/T = 0.55

$$N_{Re} = \frac{10.7(1.5)(68)(53)^2}{100} = 3.07 \times 10^4$$

Use 155 RPM & 37" impeller, D/T = 0.4

$$N_{Re} = \frac{10.7(1.5)(155)(37)^2}{100} = 3.41 \times 10^4$$

} turbulent

Use Fig 2, Hicks et al (1976) to determine the pumping number, N_Q
With D/T = 0.55 (N = 68 RPM)

Pumping number, $N_Q = 0.57 = \frac{Q}{ND^3}$

$$Q = (0.57)(68 \text{ RPM})\left(\frac{53}{12} \text{ ft}\right)^3 = 3.34 \times 10^3 \text{ ft}^3/\text{min}$$

T = 8 ft Tank cs area = $\frac{\pi}{4}(8 \text{ ft})^2 = 50.3 \text{ ft}^2$

$$V_b = \frac{Q}{T} = \frac{3.34 \times 10^3 \text{ ft}^3/\text{min}}{50.3 \text{ ft}^2} = 66 \text{ ft/min}$$ = level 10 agitation for blending and motion mixing

ENGINEERING WORKSHEET

Prepared By: Evan O Jones | Date: 12/3/86 | Project: HWVP V110302
Title/Subject: FSMT

Reviewed by C L [signature] Date: 6/30/87

With D/T = 0.4, 155 RPM & 37" impeller (DWPF)

$$N_Q = 0.67 = \frac{Q}{ND^3}$$

$$Q = (0.67)(155\,RPM)\left(\frac{37}{12}ft\right)^3 = 3.04 \times 10^3\ ft^3/min$$

$$Area = 50.3\ ft^2$$

$$V_b = \frac{3.04 \times 10^3\ ft^3/min}{50.3\ ft^2} = 60\ ft/min$$ = level 10 agitation for blending and motion mixing

Required information

Physical / Chemical data

ρ_{sl} = slurry specific gravity — μ_{sl} - slurry viscosity

ρ_l - pure fluid specific gravity — μ_l - pure fluid viscosity

ρ_s - solid particle specific gravity

particle size

solids concentration

Vessel data

V - largest working volume, gal

T - vessel diameter

Z - fluid height

$\left.\begin{array}{l} n - \text{number of impellers} \\ N - \text{shaft speed, RPM} \\ D - \text{impeller diameter, in} \\ H_p - \text{prime mover power, hp} \end{array}\right\}$ may be estimated

type of impeller

baffles - width, wall gap, length

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CALCULATION WORKSHEET

Calc. No. ______
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Title: References	Project: HWVP V110363
Prepared by: Evan O Jones	Date: 6/87
Reviewed by:	Date:

DOE Richland, WA

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