FLORIDA INTERNATIONAL UNIVERSITY

Miami, Florida

# MECHANICS OF BATCH SLURRY MIXING BY MEANS OF ROTATING IMPELLERS IN A CYLINDRICAL VESSEL

A thesis submitted in partial fulfillment of the requirements for the degree of

MASTER OF SCIENCE

IN

MECHANICAL ENGINEERING

by

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1998

To: Dean Gordon R. Hopkins
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This thesis, written by Victor Manuel Gamez, and entitled Mechanics of Batch Slurry Mixing by Means of Rotating Impellers in a Cylindrical Vessel, having been approved in respect to style an intellectual content, is referred to you for judgement.

We have read this thesis and recommend that it be approved.

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Florida International University, 1998

I dedicate this work to my parents and grandparents, for all their love and constant support.

Thank you. I love you.

## ACKNOWLEDGMENTS

I wish to thank my major professor, Dr. Ebadian, for giving me the opportunity to perform this project. I also want to thank the members of my committee, Dr. Shoephoerster, Dr. Cao, and Dr. Lin for their helpful and valuable comments and guidance in the development of this thesis.

Thanks also to Dr. Mao for his help with the laboratory equipment, Gang Li for his valuable help and recommendations for the experiments, Jon Brady for his unconditional help on many occasions, Nauman Ahmed for his great help while running the experiments, and the rest of the HCET staff that helped me on one form or another to the realization of this work.

To my fellow graduate and undergraduate students at HCET and in the classroom: thanks for your friendship. Special thanks to Alex De La Mata, Osiris Pena and Richard Lopez for being there when I needed it the most.

**ABSTRACT OF THE THESIS**

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**by**

**Victor Manuel Gamez**

**Florida International University, 1998**

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**Professor M. A. Ebadian, Major Professor**

Two slurries with solids concentration of 5 % (wt.) and 20 % (wt.), with approximate chemical and physical properties of melter feeds found in the Defense Waste Processing Facility of the Savannah River Site were prepared, characterized and mixed in a cylindrical vessel with impellers of different diameters pumping the flow in two different axial directions. The electrical power consumed by each impeller was measured and compared for both slurries. A sampling mechanism consisting of tubes positioned parallel to the vertical axis of the tank was also implemented to ascertain the level of homogeneity developed in the mix. The effects of the diameter of the sampling tubes and the sampling speed were evaluated for different impeller rotational speeds acting on the two slurries. Different startup techniques from a fully-settled condition were also assessed. Results indicate significantly lower power consumption when the impeller pumps the fluid downwards, more representative samples when the sampling tube diameter and sampling velocity are increased, and the lowest consumption of power at the startup from fully-settled conditions by the injection of air in the bed of solids.

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## NOMENCLATURE

| | | Units |
|---|---|---|
| $A$ | Area | $m^2$ |
| b | Baffle width | m |
| $C_W$ | Solids concentration | % (weight) |
| $c$ | Volumetric solids concentration | % (volume) |
| D | Tank Diameter | m |
| $D_m$ | Molecular diffusion coefficient | $m^2/s$ |
| $D_r$ | Turbulent diffusion coefficient in radial direction | $m^2/s$ |
| $D_y$ | Turbulent diffusion coefficient in axial direction | $m^2/s$ |
| d, $d$ | Impeller diameter | m |
| $g$ | gravitational constant | $m/s^2$ |
| H | Fluid height in mixing tank | m |
| $K$ | Fluid constant in power-law model for non-Newtonian fluids | |
| $l$ | characteristic length | m |
| $m$ | Mass flow rate | kg/s |
| $m''$ | Mass flux | $kg/m^2$ s |
| $N$ | Impeller rotational speed | rpm |

| | | |
|---|---|---|
| $N_P$ | Power number | |
| $N_{Re}$ | Reynolds number | |
| $n$ | Flow index | |
| n | Function of Reynolds number | |
| $P$ | Power | W |
| Pe | Peclet number | |
| $p$ | Particle diameter | m |
| $r$ | Radial position | m |
| T | Mixing tank height | m |
| $t$ | Time | s |
| $u$ | Convective velocity | m/s |
| $V_{sh}$ | Hindered particle settling velocity | m/s |
| $V_{sl}$ | Relative velocity between particle and the main fluid body | m/s |
| $V_{sD}$ | Settling velocity in a cylindrical container | m/s |
| $V_{st}$ | Terminal settling velocity in the Stoke's regime | m/s |
| $w$ | Weight | N |
| z | Height of impeller from the bottom of the mixing tank | m |
| $\frac{du}{dy}$ | Shear rate in y-direction | $s^{-1}$ |

## Subscripts

| | | |
|---|---|---|
| $l$ | Liquid phase | |
| $m$ | Mixture | |
| $s$ | Solid particles | |
| $y$ | y-direction | |

## Greek Symbols

| | | |
|---|---|---|
| $\alpha$ | Proportion of chemical species | |
| $\rho$ | Density | $kg/m^3$ |
| $\partial$ | Partial differential derivative | |
| $\mu$ | Viscosity | cP |
| $\tau$ | Shear Stress | $N/m^2$ |

# 1. INTRODUCTION

Mechanically agitated vessels by means of rotational impellers are widely used in the chemical, biomedical, agricultural and allied industries when an uniform suspension of solid particles throughout the vessel is desired. In a chemical reactor, an uniform suspension of particles is required in order to ensure that the reactants are efficiently mixed to initiate and maintain the reaction for the creation of the desired products, specially when consistent heat and/or mass transfer are required. Mixing is also essential for the aerobic and anaerobic treatment of farm slurries, such as animal-produced waste. High and low level radioactive waste is found in the form of slurries as well. The vitrification of high-level radioactive waste involve the mixing of the slurry with glass additives in order to produce uniform chemical and physical properties, thus immobilizing the radioactive elements in a durable glass waste form. In operations like this, it is very important to monitor the performance of the process by some means of measurement of the local solids concentration at different positions in the vessel. When the local solids concentrations obtained are very close in value to each other, it can be concluded that there is an homogeneous suspension of particles at any given time.

Different methods can be applied to determine the solids concentration profile in an agitated vessel. These methods include optical techniques, conductivity probes and direct sample withdrawal. Optical techniques, although accurate, are limited to laboratory activities since they require the vessel to be made of a transparent material to permit a beam of light to pass through its walls. They are also only effective for very low

concentrations, in the range of 1% - 5%. Experimentation has also been done with conductivity probes, which allow the determination of solids concentration profiles by measuring the conductivity of the slurry at different points by means of immersible electrodes. The concentration profile can be obtained by means of predetermined slurry resistance versus concentration curves. These curves can be obtained from different relations for non-conducting solids in a conducting liquid. Although this method has proven to be acceptable in many cases, different results can be obtained depending of the relation chosen. Furthermore, output from some probes can be principally a function of the fluid velocity in turbulent flows. This is not desirable since the velocity varies significantly in various positions in the tank and the concentration readings would be erroneous.

Direct sample withdrawal is probably the least accurate method to determine local solid concentrations mixing tanks. It is very difficult to ensure that the concentration obtained by means of a sampling tube or any other draw-off device is an accurate representation of the concentration of solids at a specific point, or a valid measurement of the average, "bulk" concentration of the whole mix. Any of these devices are intrusive on the velocity patterns developed, there are inherent inertial differences between the liquid phase and solids of different sizes and densities, and particle settling in the sampling tubes may occur, making this method unreliable and virtually impossible to obtain accurate results. Despite these shortcomings, direct sampling remains as the most common method in industrial applications due to its simplicity and although inexact, it provides useful information about the degree of homogeneity of a mix, which is a very

good indicator of the performance of the process. It is for these reasons that the activities described on this work focus on sample-withdrawal methods.

There are two main factors that will determine the efficiency of a mixing process for solids suspensions. These two factors are the power consumption of the mixing equipment and the degree of homogeneity in the solids distribution achieved on the slurry. Although it would be ideal to achieve a perfect mix in the majority of the applications, this would require a higher power consumption from the equipment, which would not be very economical in some cases. Ascertaining the relationship between power consumption by the mixing equipment and the solids suspension throughout the tank attributable to the fluid conditions resulting from the power delivered to the fluid and the geometry of the system, will allow the design of mechanisms that will provide acceptable levels of homogeneity at more economical operating costs.

These relationships have been investigated experimentally by the various researchers named in Chapters 2 and 3, but most of them performed their work with a solid phase consisting of a single component, utilizing somewhat "generic" materials such as sand and glass beads. MacTaggart, Nasr-El-Din, and Masliyah (1993) investigated direct sampling techniques, but for suspensions of solids (sand) stirred with a radial impeller, not the most common and efficient impeller for this kind of applications. Their choice of impeller was based on ease of comparison with previous work, mainly by Rushton (1965), Rehakova and Novosad (1971), and Sharma and Das (1980). Although these works have provided very valuable insights on the dynamics and different relationships of slurry agitation systems, there is still a great number of slurries with more than one

component, that is, their solid phase is composed of various types of material, with different densities, settling velocities, and particle size distributions. These differences in the physical properties of the solids may influence in the power consumption of the mixer and the concentrations obtained by sampling. Livestock and radioactive waste slurries are examples of these kind of solid-liquid mixture.

The objective of this work is to investigate the fundamental characteristics of a batch rotational mixing system under different operating conditions, namely rotational speed, impeller and tank geometry and solids concentration for a liquid-solid system with similar characteristics to the slurries kept in the Defense Waste Processing Facility in the Savannah River Site facility of the Department of Energy. These fundamental characteristics are power consumption, turbulence developed in the fluid and the resulting solids concentration profiles. Different sample withdrawal approaches were investigated and compared in order to select the configuration that provides the most accurate results.

# 2. THEORETICAL BACKGROUND

## 2.1 Physical Properties of Slurries

In order to design an efficient slurry handling system, the physical properties of the suspension must be known. These physical properties will dictate the type of equipment to be chosen and the behavior of the slurry under operating conditions. The physical properties of concern in the design of batch rotational mixing systems are the density, the particle diameters of the solids and their size distribution, the settling velocity of the solids, and the viscosity of the slurry. Each slurry presents different physical properties and these properties can change for the same slurry when it is subjected to different conditions such as temperature, time of mixing, and chemical conditions such as pH levels. The following sections describe these physical properties and their relevance in the design and mechanics of mixing systems.

### 2.1.1 Density

The density of the mixture is a function of two other density terms: the density of the particles, and the density of the suspending medium (Wasp, Edward, John P. Kenny, Ramesh L. Gandhi,1977). These two densities, plus the density of the mixture, should be provided when reporting the physical properties of the slurry. The mixture density is also a function of the solids concentration of the suspension.

The density of the mixture is given by:

$$\rho_m = \frac{100}{\frac{C_w}{\rho_s} + \frac{100 - C_w}{\rho_l}} \qquad \textbf{Eq. 2.1}$$

where $\rho_m$ is the mixture density, $\rho_l$ is the liquid density, $\rho_s$ is the particles density, and $C_w$ is the concentration of solids in percent weight (Wasp, Edward, John P. Kenny, Ramesh L. Gandhi,1977). The mixture density can be determined experimentally by weighting a known volume of the well mixed slurry.

### 2.1.2 Particles Diameter and Size Distribution

The size of the particles and their size distribution in the slurry have great influence in its settling behavior. As a general rule, very fine particles (less than 10 μm) are almost always fully suspended whenever the fluid is subjected to some degree of motion; when the particles are bigger than 10 μm, the effect of gravity is enough to cause gradients in the concentration of solids, and the coarser the particles, the more turbulence is needed to maintain suspension (McKay R.L., 1993).

Sieving is the most frequently used method of determining the particle size distribution, due to the simplicity of the equipment and analytical procedure. However, two important items must be taken into account when performing sieve analyses: 1. The apertures of the sieves must be all identical. This can be ensured by utilizing high quality sieves that minimize the dimension deviations. 2. The sieving surfaces are easily damaged during use. (Perry, Robert H., and Cecil H. Chilton, ed.,1973). Extreme care must also be taken when the sieves are washed as to not damage the screens with the impinging water and manipulation with the hands.

### 2.1.3 Settling Velocity

Slurries can be broadly classified as settling and non-settling (Gandhi, R.). In a settling slurry, a minimum velocity must be injected to the fluid to counteract the velocity at which the solid particles settle, and prevent the formation of a bed of solids on the bottom of the container tank. Non-settling slurries contain particles with such a low settling velocity that the suspension is maintained for long intervals of time, even if the fluid is subjected to no motion or to extremely slow motion, i.e. laminar conditions.

The settling velocity of the solids is a function of the particle size and shape, particle density, and the viscosity of the suspending medium. Sherwood, Pigford, and Wilke (1995) propose to determine the terminal settling velocity of a particle first under stagnant or very laminar conditions (Stokes' regime) and then adjust the obtained velocity to account for the effects that the size and density of the particle might have on the flow regime actually developed. The terminal settling velocity under very laminar conditions is derived from the Stokes' law and is given by the expression:

$$V_{st} = \frac{g(\rho_s - \rho_l)p^2}{18\mu} \qquad \textbf{Eq. 2.2}$$

where $g$ the gravitational constant, $\rho_s$ is the density of the particle, $\rho_l$ is the density of the suspending medium, $p$ is the nominal diameter of the particle, $\mu$ is the viscosity of the suspending medium, and $V_{st}$ is the terminal settling velocity under the Stokes' regime. Of course, a relatively coarse particle with a big diameter and high density might not develop laminar conditions while settling. The Reynolds number developed is given by :

$$N_{\mathrm{Re}} = \frac{pV_{sl}\rho_l}{18\mu} \qquad \textbf{Eq. 2.3}$$

where $V_{sl}$ is the relative velocity between the particle and the main fluid body. Since we are assuming that the main body of fluid is stagnant, $V_{sl} = V_{st}$, and the Reynolds number can be expressed as:

$$N_{\mathrm{Re}} = \frac{gd^3\rho_l(\rho_s - \rho_l)}{18\mu} \qquad \textbf{Eq. 2.4}$$

The settling velocity under Stokes conditions is then adjusted to take into account the effect of turbulence conditions that might arise due to the high density and large diameter of the particle. Sherwood, Pigford, and Wilke propose the use of table 2.1 to obtain the $V_s / V_{st}$ ratio, and then obtain $V_s$, the actual settling velocity of the particle.

**Table 2.1. $V_s / V_{st}$ ratios at different Reynolds Numbers.**

| $N_{re}$ | 1 | 10 | 100 | 1000 | 10000 | 100000 |
|---|---|---|---|---|---|---|
| $V_s / V_{st}$ | 0.9 | 0.65 | 0.37 | 0.17 | 0.07 | 0.023 |

Source: Sherwood, Thomas, Robert L. Pigford, and Charles R. Wilke. "Mass Transfer". McGraw-Hill, Inc. 1975

For engineering design purposes, the value obtained with this procedure is conservative, since other external factors such as concentration and wall effects of cylindrical containers tend to hinder the velocity of settling of the particles.

Particle concentration have a hindering effect in the settling velocity of the slurry. The hindered settling velocity due to the effect of the concentration is given by the relation:

$$V_{sh} = V_{st}(1\text{-}C)^n \qquad \textbf{Eq. 2.5}$$

where $C$ is the volume fraction of solid in the suspension, n is a function of the Reynolds number, and $V_{sh}$ is the hindered settling velocity due to the effects of the solids concentration (Perry, Robert H. and Cecil H. Chilton, ed., 1973). Hence, if concentration gradients exists in the container, the settling behavior of the solids will vary from one area of the tank to another.

The walls of the container also have a hindering effect on the settling velocity of the particles. As the ratio between the diameter of the particle and the diameter of the container becomes larger, the hindering effect is stronger. This effect can be calculated utilizing McNown's relationship:

$$\frac{V_{st}}{V_{sD}} = 1 + \frac{9}{4}\frac{p}{D} + \left(\frac{9}{4}\frac{p}{D}\right)^2 \qquad \textbf{Eq. 2.6}$$

where $\frac{V_{st}}{V_{sD}}$ is the ratio between the settling velocity in an infinite medium and the actual settling velocity in the cylindrical container, and $\frac{p}{D}$ is the ratio between the diameter of the particle and the diameter of the cylindrical container (Wasp, Edward, John P. Kenny, Ramesh L. Gandhi,1977).

### 2.1.4 Viscosity

The rheology of slurries can present Newtonian or non-Newtonian characteristics. A slurry can be considered to be Newtonian when its shear stress is directly proportional to the shear rate under the same conditions such as temperature and pressure. This relationship is easily visualized in a rheogram, which is a plot of the shear stress as a

function of the shear rate. The rheogram of a Newtonian fluid will always present a straight line, and the slope of that line is the viscosity of the fluid. The viscosity of a Newtonian fluid is therefore always constant at a specific temperature and pressure. This relation is expressed as:

$$\tau = \mu \frac{du}{dy} \qquad \textbf{Eq. 2.7}$$

where $\tau$ is the shear stress ($ML/T^2$), $\frac{du}{dy}$ is the shear rate in the y direction ($T^{-1}$), and $\mu$ is the viscosity (M/LT). Figure 2.1 is a representation of the rheogram of a Newtonian fluid. An increase in solids concentration will always increase the viscosity of the slurry, and in many occasions induce non-Newtonian characteristics.

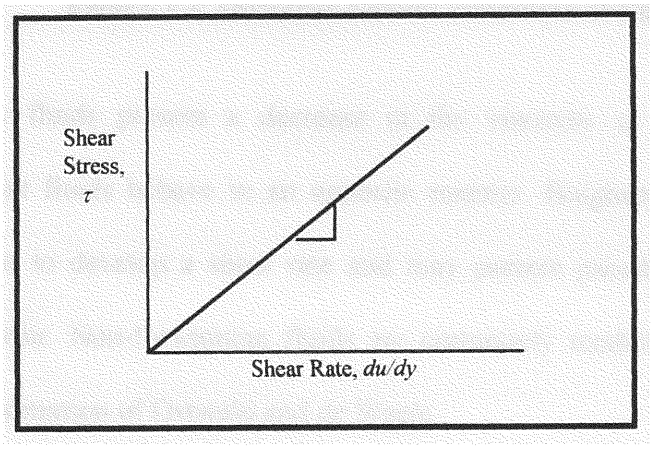


**Figure 2.1. Characteristic behavior of Newtonian fluids.**

Non-Newtonian fluids do not describe a linear characteristic of the relationship between the shear stress and the shear rate, which means that their viscosity is not

constant even at the same temperature and pressure. Figure 2.2 shows a comparison of various viscous behaviors.

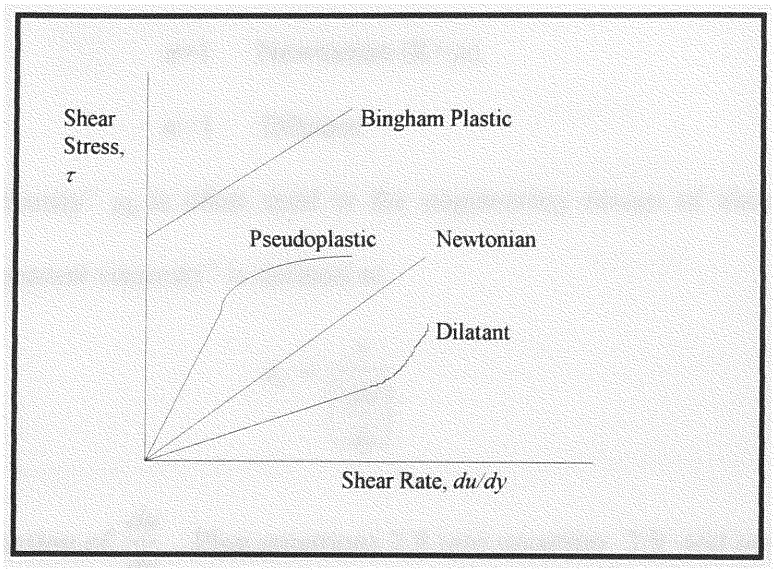


**Figure 2.2. Different viscous behaviors of fluids.**

Pseudoplastic fluids present a decrease in the viscosity as the shear stress is increased. Dilatant fluids behave in an opposite manner. Bingham plastics require an initial yield stress to develop a shear rate and may present pseudoplastic, dilatant, or Newtonian behavior. Non-Newtonian fluids are commonly modeled by means of the power-law approximation of Ostwald and de Waele :

$$\tau = K\left(\frac{du}{dy}\right)^n \qquad \textbf{Eq. 2.8 a}$$

where $K$ and $n$ are constants for a particular fluid at a particular pressure and temperature. $K$ is usually referred as the "consistency" of the fluid while $n$ is know as the "flow index". The higher the value of $K$, the more viscous the fluid. The further $n$

departs from unity, the more pronounced are the non-Newtonian characteristics. The value of $n$ determines the kind of viscous behavior:

$n<1$ Pseudoplastic

$n=1$ Newtonian ($K=\mu$) **Eq. 2.8 b**

$n>1$ Dilatant

An "apparent viscosity" $\mu_a$ is often used in the engineering design of slurry handling systems. The "apparent viscosity" is defined as:

$$\mu_a = \frac{\tau}{\left(\frac{du}{dy}\right)} \qquad \textbf{Eq. 2.9}$$

where $\mu_a$ is a function of $\frac{du}{dy}$. Plug equations 2.8 into equation 2.9, and an expression for $\mu_a$ as a function of $K$ and $n$ can be obtained:

$$\mu_a = K\left(\frac{du}{dy}\right)^{n-1} \qquad \textbf{Eq. 2.10}$$

Since $\mu_a$ is a function of the shear rate, it can only be assumed to be constant for a particular shear rate.

The rheology of the pseudoplastic, dilatant and Bingham plastic fluids described above depend solely on the shear rate. These are Time-Independent Non-Newtonian fluids. In other types of fluids, the shear stress can decrease of increase with time for constant conditions of shear rate. These type of fluids are called Time-Dependent Non-Newtonian, and are basically classified as thixotropic and rheopectic. Figure 2.3 shows the behavior of these types of non-Newtonian fluids, and compares them with Time-Independent fluids.

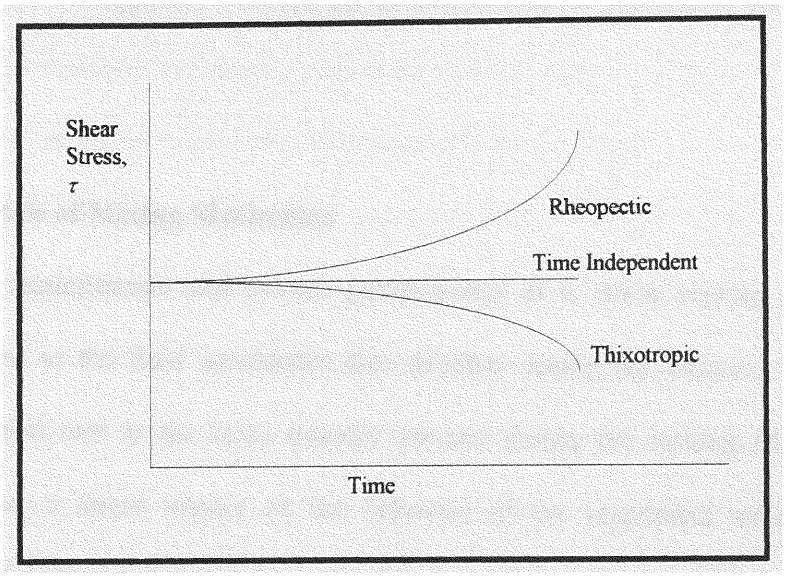


**Figure 2.3. Rheology of Time-Dependent and Time-Independent non-Newtonian fluids.**

Rheopectic fluids present an increase in the shear stress over time even at a constant shear rate. On the other hand, the shear stress in thixotropic fluids decreases at a constant shear rate as time passes.

The particle size, particle shape, and solids concentration are the main factors affecting the viscosity of slurries (Wasp, Edward, John P. Kenny, Ramesh L. Gandhi, 1977). In radioactive waste slurries, such as those of the DWPF, the temperature and pH of the suspension have also been found to have a significant effect on the viscosity. Suspensions of particles symmetrically shaped and of diameters of approximately 50 μm or larger have Newtonian characteristics, while suspensions of smaller particles or of asymmetrically shaped particles present non-Newtonian rheology (Wasp, Edward, John

P. Kenny, Ramesh L. Gandhi, 1977). Most non-Newtonian slurries present pseudoplastic behavior, therefore the apparent viscosity $\mu_a$ decreases with an increase in the shear rate.

## 2.2 Fundamentals of Mixing Mechanics

The power requirements and overall performance of a slurry mixing system are intimately related to the fluid mechanics that develop during the process. Convective effects play a vital role in the mass transfer process during the mixing of two-phase systems and have a direct impact on the behavior of the suspended solid particles. Diffusion mechanisms depend greatly on whether the fluid is laminar of turbulent. The following sections present the theory behind the mechanics of slurry mixing necessary for the design of agitation equipment.

### 2.2.1 Velocity Field

The velocity of the suspending fluid has direct effects on the shear rate, power consumption of the mixer, and on the solids concentration resulting from the process. In many cases, the velocity field for the liquid phase is solved first before attempting to solve the problem including the secondary solid phase.

The velocity fields that develop during the mixing process of a fluid in a stirring tank are very complex, especially with turbulent conditions, and depend on the type of impeller, the geometry of the mixing tank, and the rotational speed of the impeller.

Tatterson (1991) identifies three time-dependent flow systems in an agitated vessel under turbulent conditions:

1. Impeller flows: high-speed discharge flows, impeller blade boundary layers, blade wake regions, boundary layer separation regions on blades and baffles, trailing vortex systems (multiple numbers per blade), and blade bound vortex systems. These type of systems are shown in figure 2. 4 for two different types of impellers.

2. Wall flows: impinging jets that originate directly from the impeller, boundary layers, corner flows, baffle bound vortex systems, and shed vortex systems from the impeller and baffles. Figure 2.5 presents some illustrations of these flows.

3. Bulk tank flows: large recirculation zones, toroidal vortices, and decaying vortex systems.

Tatterson calls these conditions in the tank "a mess", however, he adds, that is the point in order to achieve major solids homogeneity. White (1991), identifies turbulent flow with the following characteristics:

1. Fluctuations in velocity and pressure. If the system involves heat transfer, there are fluctuations in temperature as well.

2. Eddies which intermix and fill the shear layer. The sizes of the eddies vary constantly.

3. The motion is self-sustaining. The turbulent flow is capable of maintaining itself by the production of new eddies, replacing those lost due to viscous dissipation. This phenomenon is more evident in wall-bounded flows, like in a mixing tank.

4. *Mixing* is much more stronger than that due to laminar action, which involves mainly molecular diffusion mechanics. Turbulent eddies move constantly in three dimensions, promoting the rapid diffusion of mass, momentum, energy. The heat transfer is also greatly enhanced during turbulent conditions.

These characteristics make turbulent flow the ideal condition to achieve maximum homogeneity in a solids suspension.

Nevertheless, Tatterson adds, for industrial problems, most of the complexity need not to be studied: only the portion that is relevant to the mixing process and to the level which provides basic understanding of the dominant process mechanisms in the mixing.

Tatterson has also categorized the literature of turbulent flow into three major approaches: gross-scale studies, statistical turbulence studies, and structural turbulence studies. Gross-scale studies deal with the overall patterns of the flow throughout the tank, gross internal flow measurements, and agitator pumping capacity. He points out that although this analysis may appear to be crude and simplistic, it can provide a basis for understanding the complex nature of the velocity field in a gross sense. Statistical turbulence studies treat the instantaneous value of the velocity as the sum of an average component and a fluctuating or random component. Other variables are treated in the same manner. Structural turbulence studies seek a deeper understanding of the flow than what is possible with gross and statistical studies. It involve discriminated averaging, three-dimensional and time-dependent calculations of the flow field. Analytical solutions are not possible, and the numerical computations required are very demanding. (Tatterson, G., 1991)

The literature detailing these approaches is vast. White (1991) offers an excellent introduction to fluctuations and time averaging and two-dimensional turbulent-boundary layer equations. Tatterson also dedicates sections to a complete turbulence statistical analysis description and structural turbulence studies and their applicability in

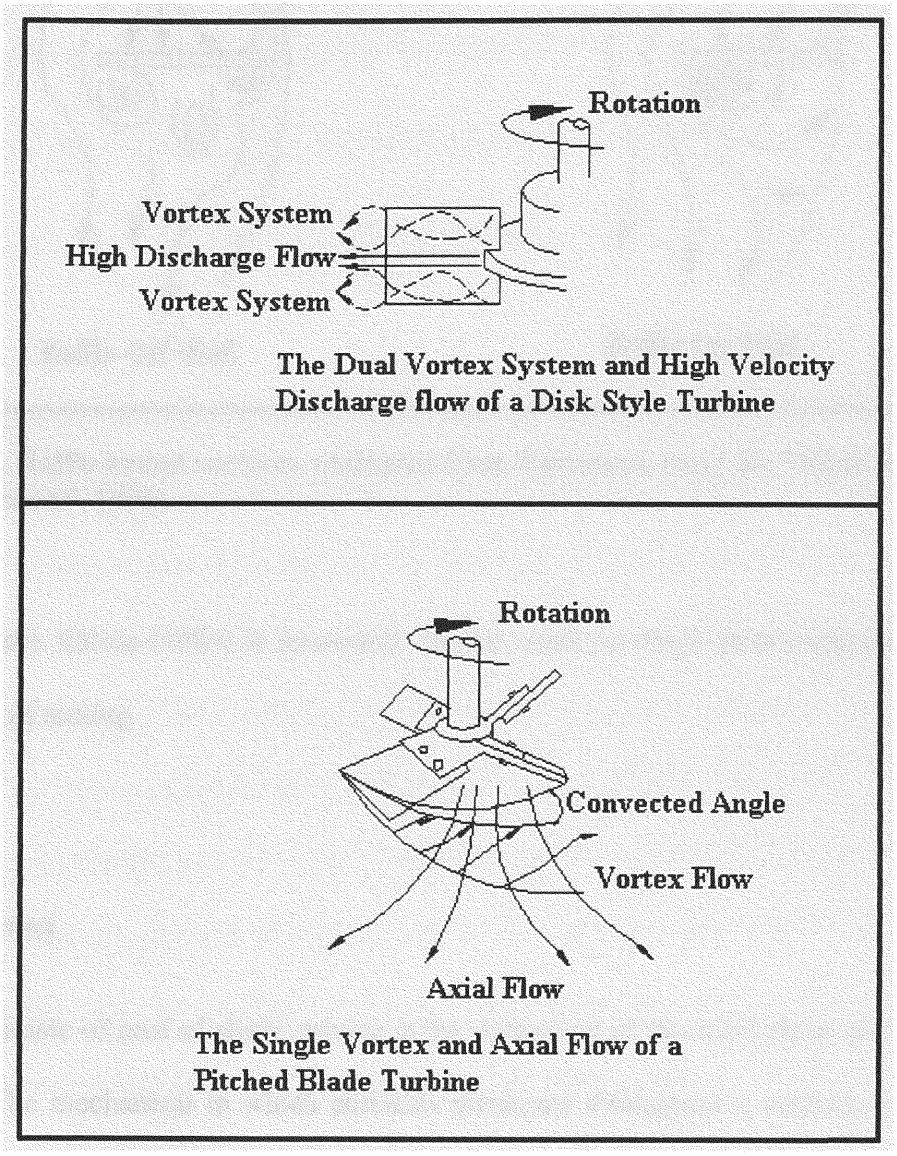


**Figure 2.4. The various flow systems around turbulent agitators. (Adapted from Tatterson, Gary B., "Fluid Mixing and Dispersion", 1991)**

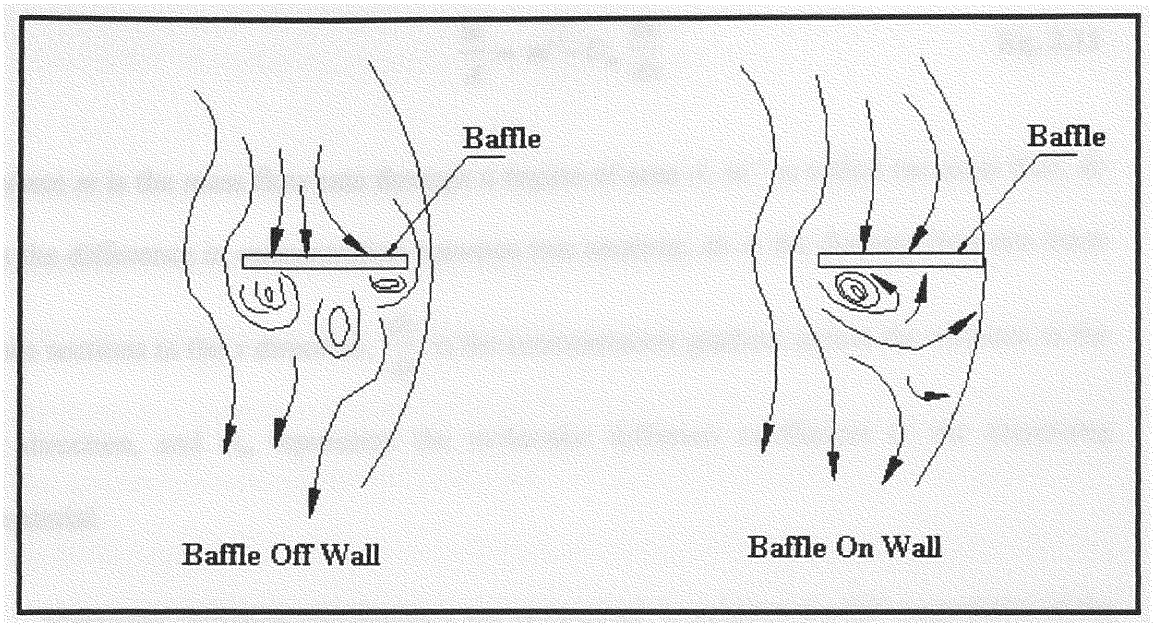


**Figure 2.5. Baffle bound vortices. (Adapted from Tatterson, Gary B., "Fluid Mixing and Dispersion", 1991)**

agitated tanks. Ottino (1989) is renowned for his work on chaos theory applied to the kinematics of mixing.

### 2.2.2 Diffusion

The ultimate of goal of slurry mixing is the dispersion of the solid phase all through the tank. The mechanism in which particles propagate throughout a medium is called diffusion. When particles tend to spread all over a stagnant medium due to concentration gradients that exists in the system, the mechanism is called molecular diffusion. Particles mobilize from high concentration regions to low concentration regions, following Fick's Law, which can be written as:

$$\frac{m}{A} = m'' - D_m \frac{dc}{dx} \qquad \textbf{Eq. 2.11}$$

where $m$ is the mass flow rate through a region of area $A$, $m''$ is called the mass flux, $dc$ is the difference in concentration between two sections, $dx$ is the distance between these two sections in the $x$ direction, $\frac{dc}{dx}$ is the concentration gradient across the medium in the $x$ direction, and $D_m$ represents the molecular diffusion coefficient of the dispersing material.

Molecular diffusion mechanisms are not enough to achieve the full suspension of the particles contained in slurries contained in a tank. The particles must be propagated by means of convective forces generated by the agitation of the system. These convective forces are provided by the action of rotating impellers, which generally induce turbulent conditions in the fluid. The propagation of the particles due to the mixing and transfer of momentum of eddies and convective effects is called turbulent diffusion. Turbulent diffusion is not constant and can vary depending on the location of the tank, and the direction in which the motion takes place, due to the different magnitudes of velocities and eddies that develop in these locations. Since turbulent flow involves random fluctuations of concentration, attention is paid to the temporal average concentration at a point. These can be assumed to be analogous to Fick's Law, when performing statistical turbulence analyses.

The dimensionless number that determines which diffusion mechanism is driving the propagation of species in a mixing process is the Peclet number, which is defined as:

$$Pe = \frac{lu}{D_m} \qquad \textbf{Eq. 2.12}$$

where $Pe$ is the Peclet number, $l$ is the characteristic length of the domain, $u$ is the convective velocity (fluid velocity plus particle settling velocity), and $D_m$ is the molecular diffusion coefficient. A high Peclet number indicates a preponderance of convective effects on the dispersion of mass, while low Peclet numbers indicate slow motion of the flow and therefore dominance of molecular diffusion. In most slurry mixing applications, where the regime is turbulent, the convective effects outweigh the molecular diffusion, which can be safely assumed to be negligible.

### 2.2.3 Settling-Dispersion Model

A simplistic dynamic mass balance in which only changes in two differential directions (axial and radial) are considered, allows the derivation of the settling-dispersion model, which has been widely adopted as the model describing the concentration of particles in a stirred vessel:

$$-\frac{\partial}{\partial y}\left(D_y \frac{\partial c}{\partial y}\right) - \frac{1}{r}\frac{\partial}{\partial r}\left(rD_r \frac{\partial c}{\partial r}\right) + \frac{\partial}{\partial y}\left(u_y c\right) + \frac{\partial}{\partial y}\left(V_{st} c\right) = \frac{\partial c}{\partial t} \qquad \textbf{Eq. 2.13}$$

where $D_y$ and $D_r$ are the turbulent diffusion coefficient in the axial and radial direction respectively, $r$ represents the radial position, $u_y$ is the convective velocity, $V_{st}$ is the settling velocity of the particles, and $c$ is the concentration. This model can be simplified

even further by assuming a steady state for concentration ( $\frac{\partial c}{\partial t} = 0$ ), no variations in the radial direction ( $\frac{\partial}{\partial r} = 0$ ), and constant diffusion coefficient and velocities, at least for broad areas in the tank. The simplified model is then:

$$- D_y\left(\frac{\partial^2 c}{\partial c^2}\right) + u_y\left(\frac{\partial c}{\partial y}\right) + V_{st}\left(\frac{\partial c}{\partial y}\right) = 0 \qquad \textbf{Eq. 2.14}$$

which is the form of the classical one-dimensional convection-diffusion equation. The exact solution for this differential equation over a domain $0 \leq y \leq L$ and with the boundary conditions $c=c_0$ at $y=0$ and $c=c_L$ at $y=L$ is:

$$\frac{c - c_o}{c_L - c_o} = \frac{\exp(\frac{Pe \cdot x}{L}) - 1}{\exp(Pe) - 1} \qquad \textbf{Eq. 2.15}$$

(Patankar, S. 1980). This differential equation can also be discretized following one of the schemes described by the same author.

It must be remembered though, that equation 2.14 is a very simplistic model that although it is a good start in the analysis of solid suspensions, it may not be a true representation of the mixing mechanism due to the actual variations of velocities and diffusion phenomena. It is evident that the more the velocity field developed in each application is understood, the closest is the model to the actual conditions in the tank.

### 2.2.4 Power

According to Oldshue (1983), the impeller power consumption is a function of these seven parameters:

1. Impeller diameter.

2. Impeller design.

3. Rotational speed of the impeller.

4. Physical properties of the fluid medium.

5. Vessel size and geometry.

6. Impeller location relative to the vessel and fluid boundaries and relative to other impellers or obstructions in the vessel.

7. Presence or absence of baffles, their design, and location.

The relationship between the consumed power and the rotational speed can be determined experimentally for each design and size of impeller working on a specific fluid with a known viscosity and density.

As it will be seen, the power drawn by an impeller under specific conditions depends on the laminar or turbulent conditions existing in the tank. The mixing Reynolds number is the dimensionless quantity that indicates the flow regime in an agitation process, and it is expressed as:

$$N_{\mathrm{Re}} = \frac{d^2 N \rho}{\mu} \qquad \textbf{Eq. 2.16}$$

where $N_{re}$ is the mixing Reynolds number, $d$ is the impeller diameter, $N$ is the rotational speed of the impeller, $\rho$ is the density of the fluid, and $\mu$ is the viscosity of the fluid. With British units, equation 2.16 is expressed as:

$$N_{\mathrm{Re}} = \frac{10.754 N d^2 \rho}{\mu} \qquad \textbf{Eq. 2.17}$$

where $N$ is expressed in rpm, $d$ is expressed in inches, $\rho$ is expressed as a specific gravity (which is sometimes taken as unity), and $\mu$ is expressed in cP. Similarly, when SI units are used, equation 2.16 takes the following form:

$$N_{\mathrm{Re}} = \frac{1.667 \times 10^{-5} N d^2 \rho}{\mu} \qquad \textbf{Eq. 2.18}$$

where $N$ is expressed in rpm, $d$ is expressed in millimeters , $\rho$ is expressed as a specific gravity and $\mu$ is expressed in Pa·s.

The turbulent range begins at different Reynolds numbers for different kind of impellers (Oldshue, J., 1983). As a general rule, the turbulent range can be assumed to start between $N_{re}=10^3$ and $N_{Re}=10^5$. An arbitrary value of $N_{re} \geq 10^5$ is used to define turbulent conditions for all kind of impellers. When $N_{re}<10$, the flow is under a complete laminar state.

The relationship between the power consumed, the rotational speed, the impeller diameter, and the density of the fluid is given by the power number. The power number is a measure of the power consumed by a mixing impeller and is expressed as:

$$N_P = \frac{P}{\rho N^3 d^5} \qquad \textbf{Eq. 2.19}$$

where $N_P$ is the power number and $P$ is the power consumed. Equation 2.19 expressed in British units takes the form:

$$N_P = \frac{1.523 \times 10^{13} P}{\rho N^3 d^5} \qquad \textbf{Eq. 2.20}$$

where $P$ is expressed in hp, $N$ is expressed in rpm, $d$ is expressed in inches, and $\rho$ is expressed as the specific gravity of the fluid. The SI equivalent of this expression is:

$$N_P = \frac{2.158 \times 10^{17} P}{\rho N^3 d^5} \qquad \textbf{Eq. 2.21}$$

where $P$ is expressed in W, $d$ is expressed in millimeters, $N$ is expressed in rpm, and $\rho$ is the specific gravity.

The relationship between the power number and the Reynolds number is presented in what is called power curves, which are plots of $N_P$ Vs. $N_{re}$. Figure 2.6 shows a typical power curve. The behavior is the same for all kind of impellers working with different fluids: in the viscous or laminar regime, the power number is inversely proportional to the Reynolds number, which means that if the viscosity is increased, the demanded power also increases. This dependence is less pronounced when the flow enters the

transitional regime, and in fully turbulent conditions, the power consumed is totally independent of the Reynolds number.

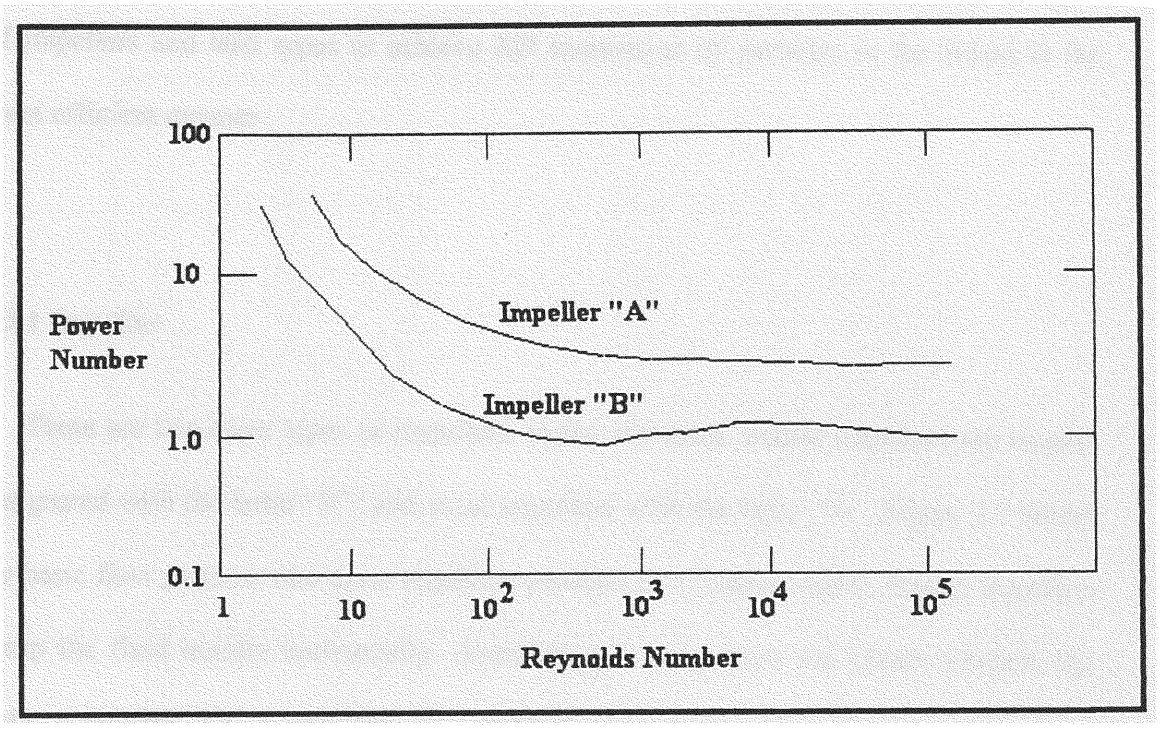


**Figure 2.6. Power curves for two different impellers. When the flow regime becomes more turbulent, the power consumed becomes less dependent of the Reynolds number and becomes practically constant in the fully turbulent regime.**

Suspension applications generally require the fluid to be in the turbulent regime, so the viscous effects do not play an important role in the power consumed by the system.

## 2.3 Agitation Equipment for the Suspension of Solids

The type of vessel, baffles and type of impeller are the most important considerations

when choosing the equipment for the mixing of slurries. The physical characteristics of the slurry that were described previously will determine the kind of equipment, as well as the mixing requirements. The following sections will describe the most appropriate types of impellers and tank types to achieve full suspension of particles in the liquid in the most efficient manner.

### 2.3.1 Impeller

There are two basic types of impellers: radial and axial. Radial impellers are usually designated with the letter "R", and axial impellers with the letter "A". Figure 2.7 shows the basic flow patterns that these impellers produce in a baffled vessel. Radial impellers pump the fluid mainly horizontally, discharging it away from the blades towards the walls of the tank. These impellers are suitable for applications that demand a high power per unit volume and/or more shear (head) (McDonough R., 1992). Typical uses are chemical reactors in which the mixing of gases and liquids (gas dispersion) and heat transfer is important. Although radial impellers can be used for the suspension of solids, these are less effective than axial impellers since they require more power to maintain the suspension (Oldshue, J., 1983). Axial impellers discharge the fluid in a vertical fashion, parallel to the vertical axis of the tank. This type of impellers are used when a high pumping efficiency is required (flow-controlled). These applications include low-viscosity blending, and the suspension of solids (McDonough R., 1992). A popular type of axial impeller is the marine propeller (A-1) which is used in portable mixers. For larger tanks, the A-1 impeller may be unsuitable due its high weight, and a pitched

turbine, which is lighter and easier to fabricate due its constant angle, and flat plate design, is used. This type of impeller is designated as A-2.

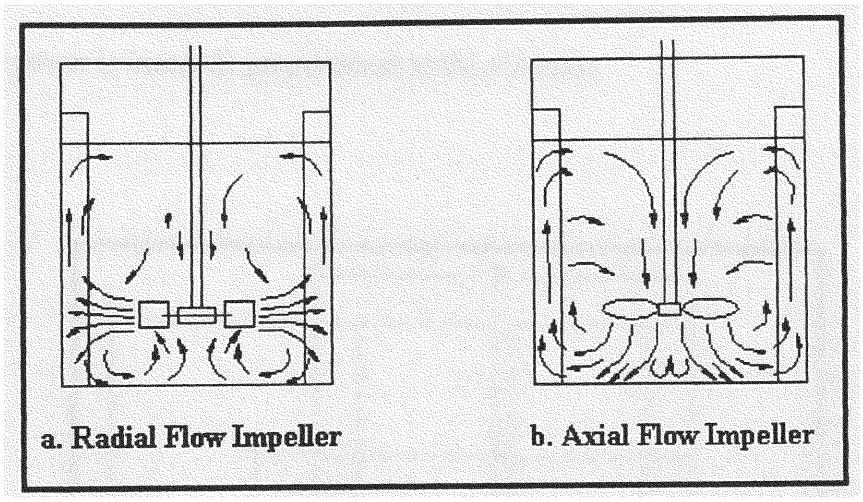


**Figure 2.7. Flow patterns developed with a) radial impellers and b) axial impellers.**

If any of these impellers operated in an unbaffled vessel and were located along the vertical axis of the tank, the resulting flow would be rotary and circular laminar, creating a vortex, and the vertical flow currents would be almost non-existent. These conditions are unsuitable for mixing operations.

Figure 2.8 shows the dimension that must be considered when placing an axial-flow impeller for the suspension of solids. There is are no definitive guidelines as to where to position an A-1 axial flow impeller in the mixing tank, but a distance z/H of about 0.3 has proven to provide good results with any kind of single impeller, of appropriate diameter, used for solids suspensions. McDonough recommends placing the impeller at

a distance of 1/3 impeller diameter off the bottom of the tank for optimal power consumption. The ratio d/D is determined by the power available to operate the mixer. As a rule, the larger the impeller the lower the speed and horsepower required to achieve the same levels of turbulence and hence of particle suspension. In other words, the required power is inversely proportional to the *d/D* ratio.

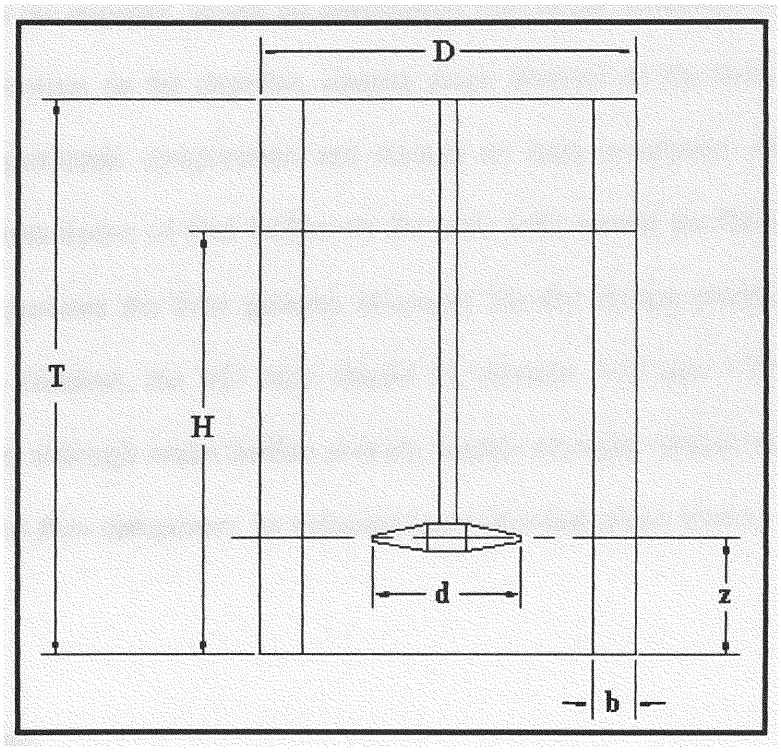


**Figure 2.8. General dimensions for a baffled tank-single impeller batch mixing arrangement.**

### 2.3.2. Tank and Baffles

As the D/H ratio increases, the solids must be suspended further up and therefore the demands of power are increased and it might be necessary to add one or more impellers

to help in the suspension. As a general rule, D/H should be kept as close to unity as possible. When D/H exceeds 1.25, it is recommended to add a second impeller (McDonough, R., 1992). Spherical bottoms or deep cones should be avoided since in severe swirling could be generated in those areas (Oldshue, J., 1983).

Swirls and vortices are generated with the absence of baffles in the mixing tank, regardless of the type of impeller. This condition is often undesirable because if the vortex reaches the impeller, severe air entrainment may occur, unbalanced unidirectional fluid force develops on the impeller, causing stress reversal on the shaft. As the shaft rotates, it experiments compression and tension on each revolution, subjecting it to fatigue. The installation of four baffles on the tank walls avoids the formation of these vortices and promote the flow patterns necessary for the mixing process (figure 2.7). According to Oldshue, the b/D ratio should be between 1/12 and 1/10. McDonough points out that although wider baffles provide slightly stronger vertical mixing currents, they may act as flow dampeners by reducing mass flow and rotary motion.

### 2.3.3 Sampling

As stated before, direct withdrawal of samples during operation may not provide an accurate representation of the solids concentration in a specific location of the tank, or be representative of the average or "bulk" concentration. As it has been seen, the flows resulting from a mixing operation are complex and often chaotic. An accurate sampling procedure would involve the withdrawal of the slurry without disturbing the flow patterns at the location where the sampling takes place. This means that the sampling tube should

be positioned parallel to the velocity vectors at a specific point and that the suction should be done at the same velocity as the local velocity of the fluid (isokinetic velocity). Velocity vectors can now be predicted, even for complicated configurations by means of numerical models, but the implementation of the sampling system to comply with these patterns would be difficult in most applications. The problem becomes even more difficult when treating slurries with particles with high settling velocities. Even if the withdrawal inside the tank could be performed perfectly without inducing any disturbance in the flow field, a certain level of turbulence would still be needed to transport the fluid to the collecting device without stratification. This would involve suction velocities different to those of the actual velocity field of the fluid. Oldshue recommends that when defining a sampling technique, the main considerations should be that the technique has to be defined so that it can be repeated and that reliance be placed mainly on the relationship between sampling and other mixing variables rather than their absolute values.

# 3. PREVIOUS RESEARCH REVIEW

Each type of slurry has its own physical characteristics like viscosity, particles size, particle density, particle falling velocity, and fluid density. When we add the fact that some of these characteristics can change considerably with time, with the intensity of the agitation, or with the time the solids in the slurry have been allowed to settle, not to mention that in some cases the only way to achieve full solids suspension is by means of high turbulence, which is chaotic by nature; it is not surprising that until the last thirty years the mixing of slurries has been addressed on a case by case basis, with the design of agitation systems being the result of trial and error more than anything else.

Rushton was one of the pioneers in ascertaining the power characteristics of mixing impellers and sampling techniques in the 1950's and 1960's. His seminal two-part paper "Power Characteristics of Mixing Impellers" (1950), co-written by Costich and Everett, offered correlations of the most important variables in the form of dimensionless groups, characteristic plots and general equations. Their work involved studies on the impeller, tank, and fluid, over wide ranges of power, size, and physical properties for Newtonian fluids. Years later Holland and Chapman (1966) would publish additional power data for other impeller types such as anchors, curved-bladed turbines, and paddles in baffled and unbaffled vessels.

Rushton was also the first in trying to determine a reliable sampling technique that would indicate the bulk concentration of solids under conditions of full solids suspension (1965). Rushton was able to demonstrate that if the withdrawal of the sample were performed at the same velocity of the fluid (isokinetic velocity), at the impeller plane of

a radial-flow turbine, the sample obtained would have the same concentration as in the body of the tank (bulk). The magnitude of the velocity in a radial-flow impeller is easily determined by a simple expression and is a function of the rotational speed, the impeller diameter and the radial position at which the velocity is of interest. The direction of the velocity in the plane of a radial-type turbine can be safely assumed to be completely horizontal, parallel to the bottom of the tank. Rushton also suggested that if the sampling was executed at the isokinetic velocity for any specific point in the tank, the sample obtained would have the same local concentration of the point where it was withdrawn, and an accurate concentration profile could be obtained. Unfortunately, this technique was very difficult to implement due to the high complexity of the flow in all other sections of the tank other than the plane of a radial impeller.

Among the books in which the general principles and procedures for the design of mixing systems in stirred vessels have been consolidated, are the works by McDonough (1992), Tatterson (1991) (which also provides excellent theoretical backgrounds), Nagata (1975) (also and outstanding source for heat and mass transfer theory applied to mixing systems) and Oldshue (1983), which has become the de-facto handbook for the design engineer.

During the last twenty years, the advent of faster and more powerful computers and numerical methods for modeling, the development of advanced chaos theory, and more advanced experimental equipment, have provided researchers with new and better tools to investigate in detail the mechanics of solid-liquid systems in agitated vessels and ways to improve the determination of the condition of the solids suspension in the fluid. Cumby and Slater (1989) provided a simplified theoretical description for the distribution

of solids in which only the solids settling velocity was used, and compared it to samples withdrawn from their experimental tank. The results presented moderate success, since their models were accurate only at low rotational speeds, probably because at high turbulence the settling velocity of the particles is overthrown by the convective actions. Rasteiro, Figueiredo, and Freire (1993) obtained results that led them to propose the settling-dispersion model as a sufficient basis for the assessment of suspension quality. Other authors that have utilized this model for comparison with experimental results are Magelli, Fajner, Nocentini, and Pasquali (1989), Barresi and Baldi (1987), Shamlow and Koutsakos (1988), Tojo and Miyanami (1982), and Tojo, Yamazaki, and Miyanami (1986). The latter provides models for radial and axial flow impellers. A common factor in these works is the semi-analytical approach when developing the models, in which the Peclet number is adjusted so that the models agree with the experimental profiles. This has been the approach due to the high complexity of the velocity fields developed during turbulent mixing.

Progress in numerical methods and the availability of more powerful computers have permitted a deeper understanding of the fluid mechanics of the mixing process by means of computational fluid dynamics models that are able to calculate the velocity fields developed and the interaction between the solid and liquid phase. Gosman, Lekakou, Politis, Issa, and Looney (1992) have performed work towards the multidimensional modeling of two-phase flows in stirred vessels by extending the $k$-$\epsilon$ turbulence model to two-phase flows. Multiple frames of reference approaches, which generally consist of one frame that rotates at the impeller speed and is used to compute the flow within the impeller in steady-state mode and a second frame which is stationary and is used to

compute the flow away from the impeller, have been developed to model the flow (Luo, Issa, Gossaman, 1992) and have been incorporated in general-purpose commercial codes. Velocity and turbulent energy distribution boundary conditions in different impeller arrangements have also been investigated (Fokema, Kresta, Wood, 1994).

The effect of the sampling velocities has been further studied by MacTaggart, Nasr-El-Din, and Masliyah (1993), corroborating the findings of Rushton and concluding that the sampling errors for local concentration measurements can be minimized by sampling at as high a velocity as possible. The effect of the geometry of the sampling tubes was studied in this work as well. They compared this work later with concentration profiles obtained by means of a conductivity probe designed by the authors (1996).

The work by Ottino (1989) was the first book to present a unified treatment of the mixing of fluids from a kinematical standpoint, utilizing dynamics systems and chaos theory.

# 4. EXPERIMENTAL PROCEDURE

## 4.1 Slurry Characterization

An effort was made to create a slurry that would approximate the actual chemical composition of vitrification melter feeds found in the Defense Waste Processing Facility of the Department of Energy's Savannah River Site. Chemical compounds have inherent physical characteristics at specific states (solid, powder in this case). The use of the same chemical composition resulted in physical properties close to those of the actual slurries. Vitrification melter feeds are composed of waste slurry, supernate and glass-based

**Table 4.1. Typical chemical composition of the Savannah River Site melter feeds.**

| COMPONENTS | CONCENTRATION (wt %) (DRY BASIS) | ANALYTICAL ERROR (+/- wt %) |
|---|---|---|
| Al | 6.67 | 0.40 |
| Ca | 2.67 | 0.15 |
| Cr | 0.18 | 0.02 |
| Fe | 26.8 | 1.55 |
| Mg | 1.31 | 0.08 |
| K | 0.14 | 0.003 |
| Mn | 2.80 | 0.16 |
| Na | 10.2 | 0.60 |
| Ni | 0.24 | 0.02 |
| Pb | 0.10 | 0.01 |
| Si | 0.81 | 0.05 |
| Zn | 0.17 | 0.01 |
| $SO_4$ | 0.33 | 0.02 |
| $PO_4$ | 0.84 | 0.03 |
| $C_2O_4$ | 0.20 | 0.01 |
| $CO_3$ | 1.94 | 0.10 |
| $NO_3$ | 1.96 | 0.05 |
| $NO_2$ | 5.86 | 0.03 |
| OH | 2.95 | 0.05 |
| Total Organic Carbon | < 50 ppm | |
| Total Solids | 26.29 | |

Source: Ebadian, M. A. and F. Mao. "Rheological Properties of Defense Waste Processing Facility Melter Feeds, Fiscal year 1996 year-end report". Hemispheric Center for Environmental Technology, 1997. Table 5.

additives, most commonly silica. Table 4.1 presents the composition of an actual slurry, which include insoluble, soluble, and calcined solids.

The surrogate was prepared with the three major components present in solid-form found in table 4.1, namely aluminum, iron, and sodium. These elements are commonly found in this type of slurries in the form of oxides. It is for this reason that the compounds used for this work were aluminum oxide ($Al_2O_3$), ferric oxide ($Fe_2O_3$) and sodium hydroxide (NaOH). The latter is the responsible of giving the slurry its characteristic pH value of 13 and is the only component of the three that is fully soluble in water, which was used as the supernate. These three different compounds were assumed to constitute the total solids of the slurry with the 26.29% (weight) solids concentration indicated in table 4.1, and the same proportion of components was kept for the slurries prepared. Figure 4.1 shows this distribution.

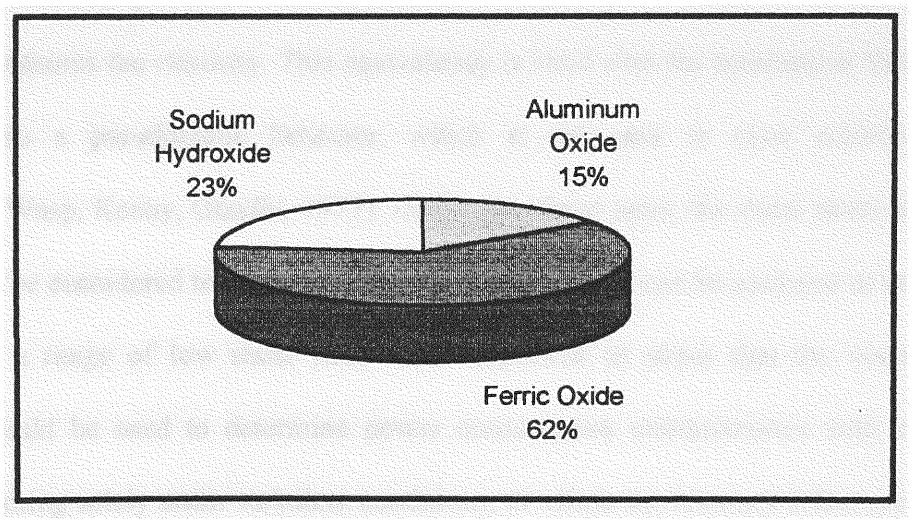


**Figure 4.1. Proportional distribution of the components of the vitrification melter feed surrogates prepared for this investigation.**

The mixing tests of this investigation involved slurries with solids concentrations of 5% and 20% (weight), two values commonly found in DWPF slurries. Following the components distribution of figure 4.1, samples of 300 ml were prepared to measure the apparent viscosity of these two slurries under well-mixed conditions. Table 4.2 presents the exact amount of each compound used in the preparation of the samples. The samples were stirred thoroughly before the viscosity was measured by means of a rotational viscometer. This is important because the instrument can measure the viscosity accurately only when the concentration is uniform. If the solids are allowed to settle, the resulting concentration gradients will produce viscosity readings which would not correspond to the viscosity that an axial impeller would "see" under conditions in which the particles are fully suspended. The power drawn by the impeller is only determined by the shear rate, therefore viscosity, that the impeller "sees" in its discharge zone. An emphasis is made in "axial-impellers", due to the low shear rate developed by this type of impellers in its discharge zone, which is comparable to the low shear rates at which the viscometer measures the viscosity. This equivalency is valid with the assumption that the slurry presents a pseudoplatic behavior, which is the case in most solid-liquid suspensions (Wasp, Kenny, Gandhi, 1977). Under low shear rates, the shear stress-shear rate ratio can be considered to be linear, therefore, the viscosity can be assumed to be the same within a range of low shear rates. It is important to stress that the viscosity measured should be used to determine power consumption characteristics with axial impellers working solely under turbulent conditions, in which the viscosity effect are not too important and errors in viscosity values will not greatly affect power draw (McDonough, 1991). The viscosity values measured were 1 cP for the 5% wt slurry and

9 cP for the 20% wt slurry. The pH was measured by means of a pH electrode meter and the obtained value was 13 for both samples. These tests were executed for a second time with newly prepared samples and the results obtained were the same.

**Table 4.2. Amount of each compound used in the preparation of the samples.**

| Solids Concentration (%wt) | Compound | Amount (g) |
|---|---|---|
| 5 | $H_2O$ | 374.25 |
| | $Al_2O_3$ | 3.01 |
| | $Fe_2O_3$ | 12.09 |
| | NaOH | 4.60 |
| 20 | $H_2O$ | 374.25 |
| | $Al_2O_3$ | 14.29 |
| | $Fe_2O_3$ | 57.42 |
| | NaOH | 21.86 |

Density of water used for calculations: 998 $Kg/m^3$ (White F., "Fluid Mechanics", Table A.1).

The density values of the solid particles were obtained from the information provided by the supplying companies. McMaster-Carr reported a density of 132 $lb/ft^3$ (2114.44 $kg/m^3$) for the aluminum oxide powder, and Fisher Scientific specified a density of 5240 $kg/m^3$ for the ferric oxide supplied. Although no information was provided for the sodium hydroxide, this was not necessary since this compound fully dissolves in water and it was the density of this sodium hydroxide solution, the suspending medium, that was important. The corresponding amount of NaOH for each solids concentration (table 4.2) was diluted in 300 ml of water and then the solution was weighted to determine its density. Table 4.3 presents the densities of each sodium hydroxide solution, their

equivalent concentration in molar form, and in what percentage they increased the density of the unaltered water. As is evident, the density of the water was not changed significantly.

**Table 4.3. Densities of sodium hydroxide solutions for the two solids concentration of the slurry.**

| Total Solids Concentration of Slurry (%wt) | Density of NaOH Solution ($Kg/m^3$) | Molarity (M) | Percentage in which the Density of Water (998Kg/m3) was Increased |
|---|---|---|---|
| 5 | 1012.60 | 0.38 | 1.44 |
| 20 | 1052.77 | 1.82 | 5.20 |

The suspension density for each slurry was calculated analytically by means of equation 2.1. In order to utilize this formula, a correction was made to the concentration of solids ($C_w$) term. Since the amounts of NaOH used for both slurries dissolve completely in water, $C_w$ was modified so it would only take into account the insoluble material, i.e. the suspended solids ($Fe_2O_3$ and $Al_2O_3$). When the sodium hydroxide is disregarded, $C_w$ takes a value of 3.83 % wt. for the 5% wt. slurry, and 15.33 % wt. for the 20 % wt. slurry. We must remember that while the NaOH was not being considered for the solids concentration, its effects on the slurry were accounted in the already measured density of the sodium hydroxide solution. Equation 2.1 assumes a single type of solids in the slurry. Since our slurries contain two different suspended solids with

different densities each, a proportional mean for the density was determined with the following expression:

$$\rho_s = \alpha_1 \rho_1 + \alpha_2 \rho_2 \qquad \textbf{Eq. 4.1}$$

where $\alpha_1$ is the proportion of $Fe_2O_3$ in the slurry disregarding NaOH (80 %), $\alpha_2$ is the proportion of $Al_2O_3$ disregarding NaOH (20 %), $\rho_1$ is the density of $Fe_2O_3$ (5240 Kg/m$^3$), $\rho_2$ is the density of $Al_2O_3$ (2114.44 Kg/m$^3$), and $\rho_s$ is the proportional mean of the densities of the solids. The results for both slurries are presented in table 4.4 and are compared with the densities determined experimentally. The experimental error was considerably low, verifying the validity of the method just described.

The particle size distributions of the insoluble components of the slurry (i.e. ferric oxide and aluminum oxide) were determined by means of an arrangement of four sieves of different sizes. The collected powder for each particle size was collected, weighted and compared with the total weight of the sample. Table 4.5 presents the results for the ferric oxide powder ($Fe_2O_3$) and table 4.6 present its average values. Tables 4.7 and 4.8 present the results for the aluminum oxide powder ($Al_2O_3$).

**Table 4.4. Comparison between semi-analytically calculated and experimentally determined slurry densities.**

| Total Solids Concentration (% wt.) | Slurry Density | | Error Percentage (%) |
|---|---|---|---|
| | Semi-analytical (Kg/m$^3$) | Experimental (Kg/m$^3$) | |
| 5 | 1043.81 | 994.67 | 4.71 |
| 20 | 1194.06 | 1144.67 | 4.13 |

**Table 4.5. Particle size ranges for the ferric oxide powder obtained after sieving operation.**

| Particle Size Range (μm) | Percent (%) |
|---|---|
| 250-150 | 0 |
| 150-75 | 72.37 |
| 75-45 | 22.09 |
| less than 45 | 5.54 |

**Table 4.6. Average particle size distribution of the ferric oxide powder.**

| Average Particle Size (μm) | Percent (%) |
|---|---|
| 112.5 | 72.37 |
| 60.5 | 22.09 |
| 22.5 | 5.54 |

**Table 4.7. Particle size ranges for the aluminum oxide powder obtained after sieving operation.**

| Particle Size Range (μm) | Percent (%) |
|---|---|
| 250-150 | 0 |
| 150-75 | 0.28 |
| 75-45 | 24.15 |
| less than 45 | 75.57 |

**Table 4.8. Average particle size distribution of the aluminum oxide powder.**

| Average Particle Size (μm) | Percent (%) |
|---|---|
| 112.5 | 0.28 |
| 60.5 | 24.15 |
| 22.5 | 75.57 |

## 4.2 Batch Mixing

The experimental slurry mixing system was built in a way so that we could have a sound starting point from which recommendations could be made to improve the mixing operation for this specific type of slurry, resulting in good homogeneity of the solids in the suspension at the most economical cost of operation and maintenance. The description of the experimental apparatus and the specifications for each of its components can be found in Chapter 5. A scheme of experiments was then planned, in which the performance of the system under different operating conditions was assessed, as well as the approach in which the samples were taken. Different mixer rotational speeds and impeller geometries were tested, as well as different diameters for the sampling probes at different sampling speeds. The following sections provide the descriptions and justifications of all the procedures for this work, from the experimental apparatus conception and implementation, to the execution of the experiments.

### 4.2.1 Mixing Tank and Mixer Design

The configuration of the mixing tank, dimensions and type of impellers used were implemented accordingly to the general guidelines for the design of slurry agitation equipment discussed in Chapter 2. Although these guidelines provide a good starting point, it must be remembered that they are based on suspensions with a single type of solid particles, and as the number of different types of solids in the suspension increases, the slurry may present special characteristics not found in others. Regardless of these general guidelines, experimentation continues in different industries, such as in coal and livestock slurries handling, in order to improve the mixing process.

### 4.2.2 Sampling

MacTaggart, Nasr-El-Din, and Masliyah (1993) experimented with different sizes and geometries of sampling tubes, and different sampling velocities in order to ascertain which configuration provided the most representative sample, and to corroborate Rushton's assertions on sampling procedures. Their findings indicated a strong dependence on the diameter of the sampling tubes and the velocity at which the samples were taken. Their work, however, was performed with radial (Rushton type) impellers; not the type most used in the suspension of solids. It is for this reason that the diameter of the sampling tubes was varied in this work, in order to determine whether the size of the sampling tubes, as well as the sampling rate, have an effect on the concentration of the samples obtained with axial impellers.

The solids concentration profile was determined by withdrawing samples of the agitated slurry at four different positions along the vertical axis of the tank. Table 4.9 shows the different depths at which the samples where removed and their respective dimensionless equivalent . The tubes were located at 0.06 m from the wall of the tank. The sampling was executed three times for each location and the values obtained were then averaged for the presentation of the data in chapter 6. The concentration values obtained did not differ in more than 10 % from each other. Further specifications of the sampling equipment are found on Chapter 5.

Two sizes of sampling tubes where used (0.25 in. and 0.375 in., I.D.) in order to evaluate the effect of the diameter in the concentrations obtained. The withdrawal of samples was achieved by means of a variable-speed, computer-controlled peristaltic pump. In order to observe the effects of the velocity of sampling on the measured concentrations, the sampling was performed at low and high velocities. Before a sample was taken, the fluid was allowed to circulate through the tubing, returning to the tank, in order to obtain a fully developed flow. The slurry was allowed to circulate for 5 minutes for the lowest sampling speeds and 10 seconds for the higher. After several tests withdrawing 325 ml, 100 ml, and 40 ml for each sample, it was decided to utilize 40 ml samples so as not to significantly alter the bulk solids concentration in the mixing tank, and achieve more precision in the measured concentrations. The samples where weighted, placed for two hours in a furnace for drying and the retrieved solids were weighted to establish its percentage relative to the total weight of the slurry. The expression used to determine the concentration of

solids by weight is:

$$C_w = \left(\frac{w_s}{w_m}\right) \times 100 \qquad \textbf{Eq. 4.2}$$

where $C_w$ is the solids concentration by percent weight, $w_s$ is the weight of the solids after drying, and $w_m$ is the weight of the slurry before drying.

**Table 4.9. Different axial positions at which the samples where removed and their respective dimensionless equivalent**

| Sample | Sample Depth (y) (measured from surface) | Dimensionless Depth (y/H) |
|---|---|---|
| 1 | 0.050 m | 0.09 |
| 2 | 0.185 m | 0.34 |
| 3 | 0.320 m | 0.58 |
| 4 | 0.458 m | 0.83 |

H=0.55 m, total depth of fluid.

The samples were returned to the tank before another experiment would be conducted. The liquid level was then brought up to the original level to recover the water lost during the drying of the samples.

### 4.2.3 Power Measurements

As it has been seen, power is one of the parameters that demands close attention when determined the characteristics of an agitation process. The power consumed has a direct impact on the economics of the entire operation.

The electrical power consumption of the mixer was measured under different operating conditions in which the size of the impeller, the rotational speed, start-up conditions, and the solids concentration of the slurry was varied.

A wattmeter was connected to the mixer motor as shown in the diagram in figure 4.2. Voltage and absorbed current could also be measured with the instrument, which was always set to zero before turning the mixer on and doing any measurement.

The power consumption versus rotational speed curves were obtained by starting the mixer at the highest rotational speed possible to ensure maximum solids suspensions, and then bringing the speed down, recording the measured power at selected speeds. Since the mixer motor contained a 4 amps safety fuse, the current was closely monitored in order to prevent the fuse to break the circuit.

### 4.2.4 Air Jet Startup

When the mixer is shut down, the suspended solids will eventually settle. Depending on the nature of the slurry (solids concentration and type of particles), and the time that the material has been allowed to settle, it may be difficult to the impeller to promote the suspension again. The material may be too compacted, and in some cases, the impeller could be buried in the bed of solids. Sometimes the rotating action of the impeller is not enough and external aid is required.

Several tests were made in order to ascertain the effect of air injection on the consumed power during startups from conditions where the solids of the slurry would be fully settled. Air was injected at 0.20 m from the bottom of the tank for 10 seconds before the mixer was turned on . Once the mixer was started and rapidly set to a specific rotational speed, its power consumption was recorded and the air was shut down.

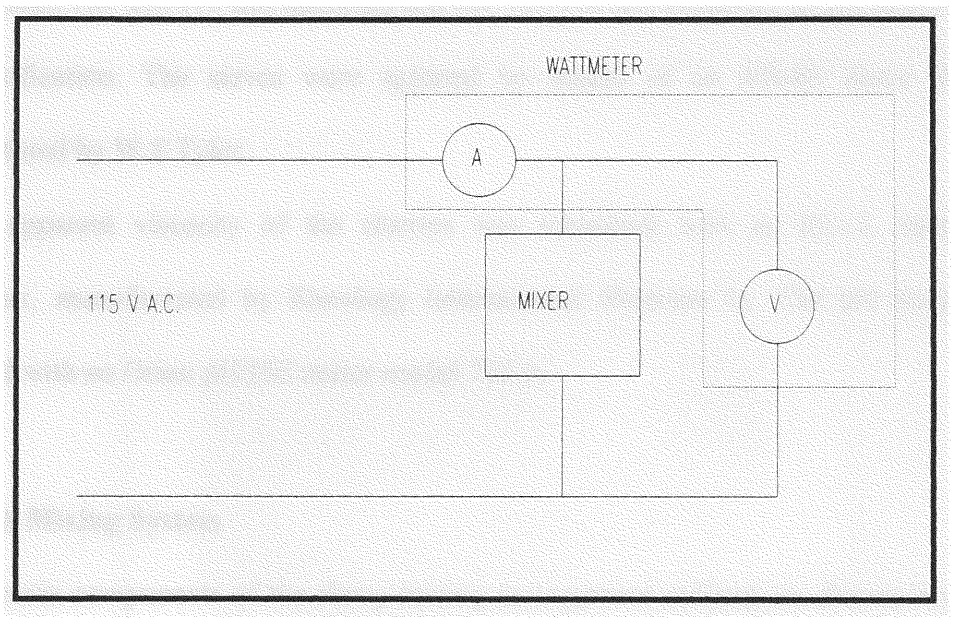


**Figure 4.2. Schematic of the wattmeter connection to the mixer motor.**

# 5. EXPERIMENTAL APPARATUS DESCRIPTION AND EQUIPMENT SPECIFICATIONS.

## 5.1 Slurry Characterization

The particle size distributions of the components of the slurry were determined by means of an arrangement of four U.S.A. Standard Testing sieves of different sizes as shown in figure 5.1. The sieves utilized were No. 60 (250 μm opening), No. 100 (150 μm opening), No. 200 (75 μm opening) and No. 325 (45 μm opening), as per A.S.T.M.E. -11 specification. The sieves were agitated by means of an RX-86 Sieve Shaker manufactured by W.S Tyler.

The apparent viscosity of the slurries was measured with an RI:1:L rotational viscometer, manufactured by Rheology International Shannon Lt. The pH level was measured with an Orion pH/ISE meter model 710 A.

## 5.2 Batch Mixing System

The main components of the slurry mixing arrangement utilized are shown in figure 5.2. The system consisted of the mixing vessel, the mixer driven by an electrical motor, the sampling lines, the computer-controlled sampling pump, a wattmeter for power consumption readings, and a remote optical sensor for rotational speed measurements.

The mixing vessel was a flat-bottomed cylindrical polyethylene tank by Nalgene, with a capacity of 0.110 $m^3$ (110 liters). The inside diameter of the tank is 0.46 m and the total volume that the slurry occupied was 0.090 $m^3$, which corresponds to a liquid height of 0.55 m, measured from the bottom. Four Lexan baffles, 0.006 m thick, were built and

distributed symmetrically in the tank. The top height of the baffles was 0.725 m from the bottom, with a width of 0.043 m and no distance off the wall of the tank. Figure 5.3 shows a diagram of the general dimensions of the tank configuration. Figure 5.4 is a photograph that shows the baffle arrangement in the tank.

The mixer was manufactured by EMI, Inc. and was driven by a variable speed, 115V, 0.25 HP, monophasic electric motor, with a rotational speed range between 87 and 1750 rpm. The impellers utilized were stainless steel, axial, marine-type propellers. Axial pumping directions were up and down and the nominal diameters were 4, 5, 6, 7 and 8 inches, as shown in figure 5.5, for a total of ten impellers. Table 5.1 presents the general dimensions for each of these impellers.

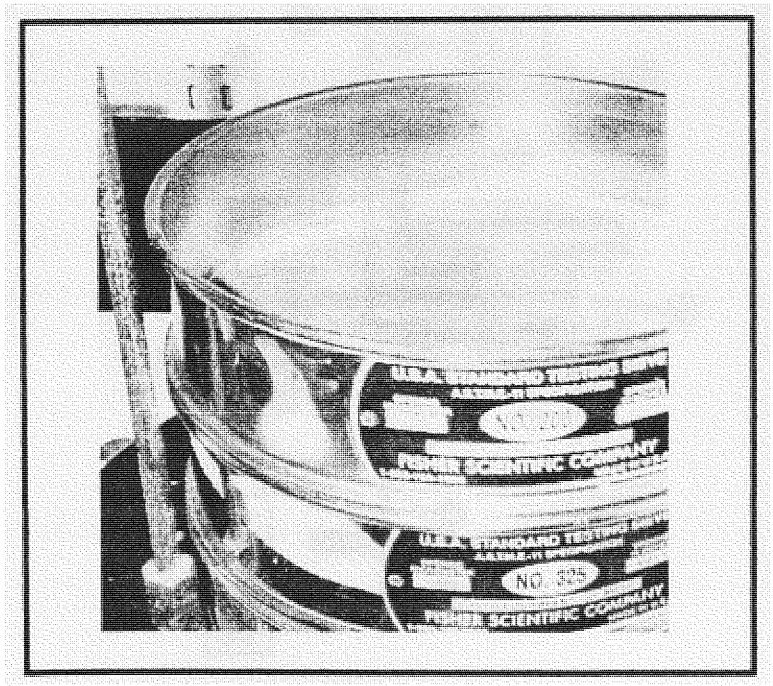

**Figure 5.1. Sieves used to determine the particle size distribution of the solids of the slurry.**

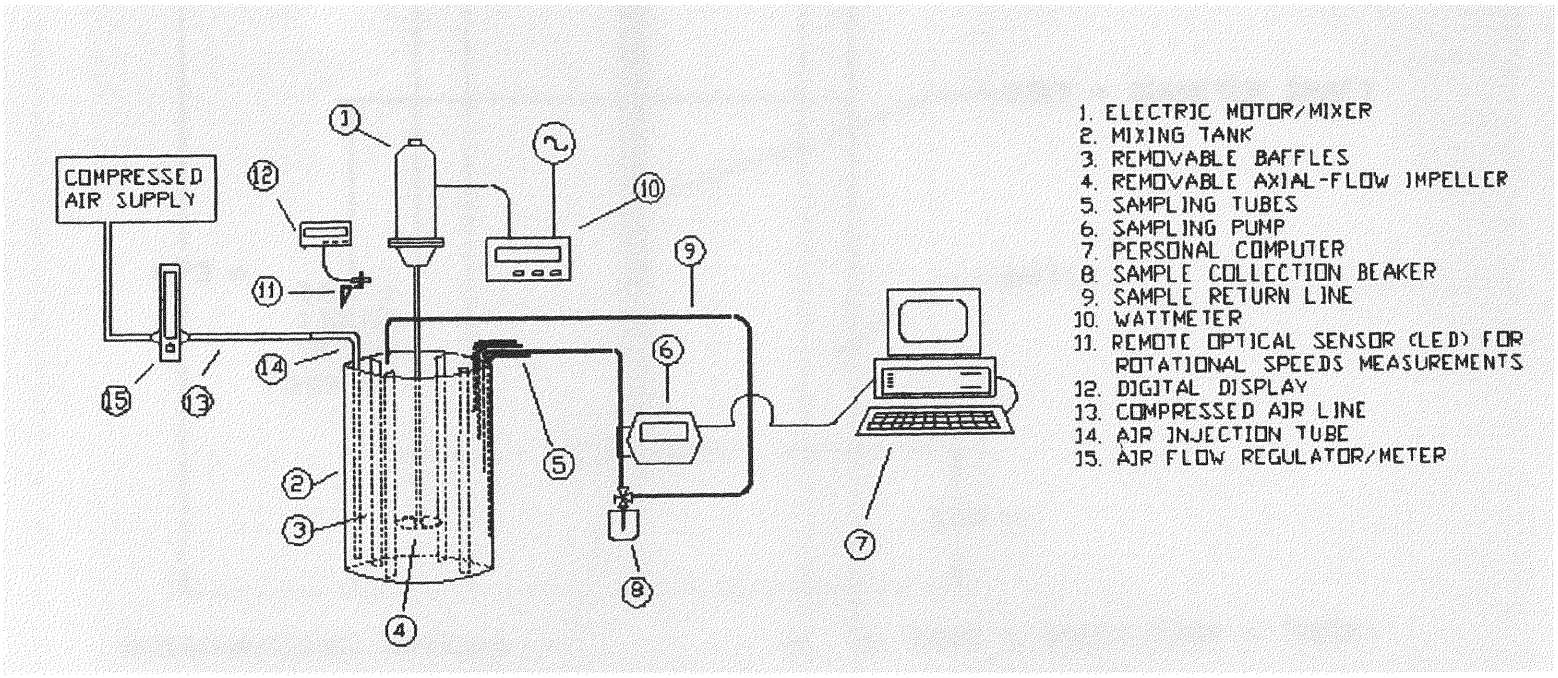


**Figure 5.2. Schematic diagram of the experimental apparatus**

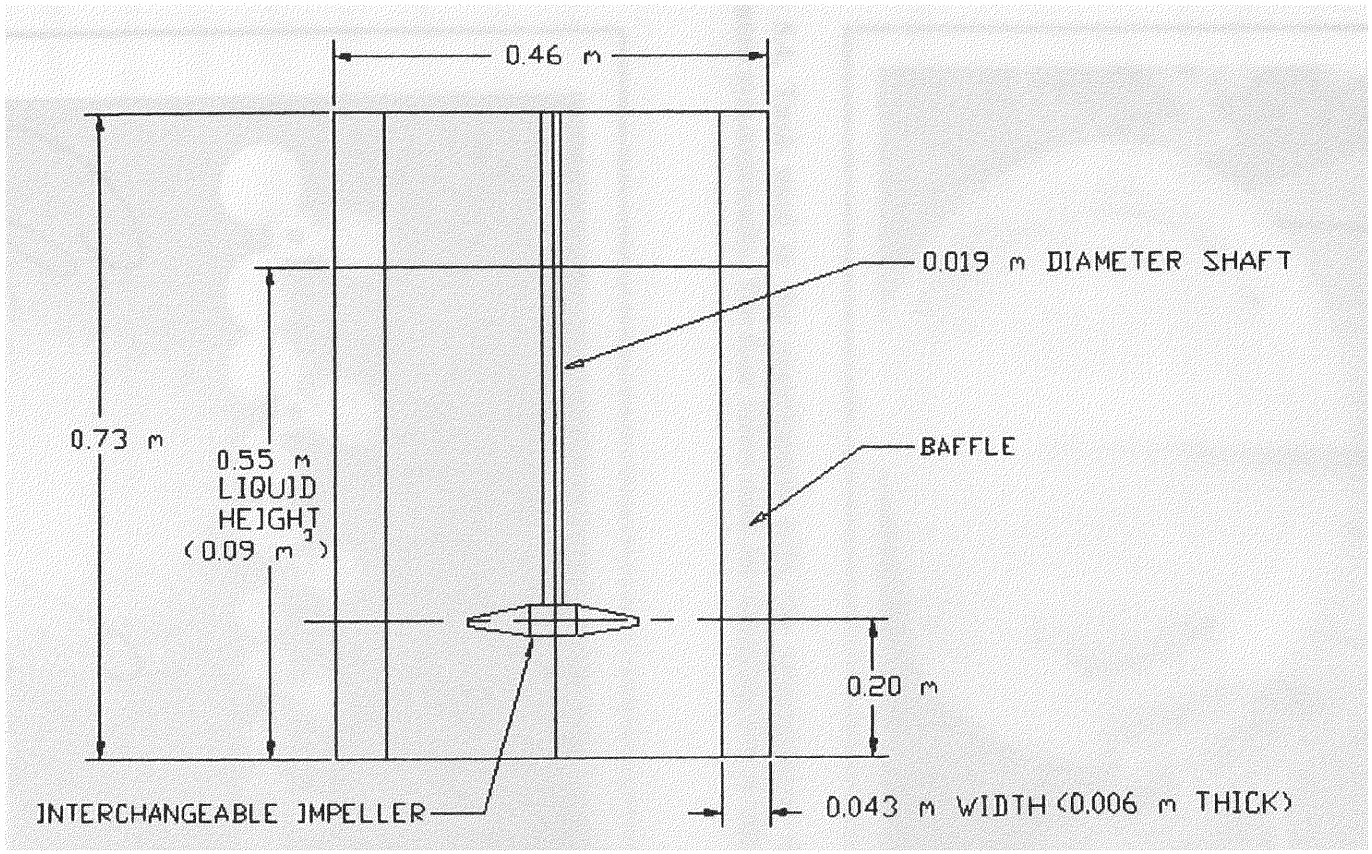


**Figure 5.3. General dimensions of the mixing tank.**

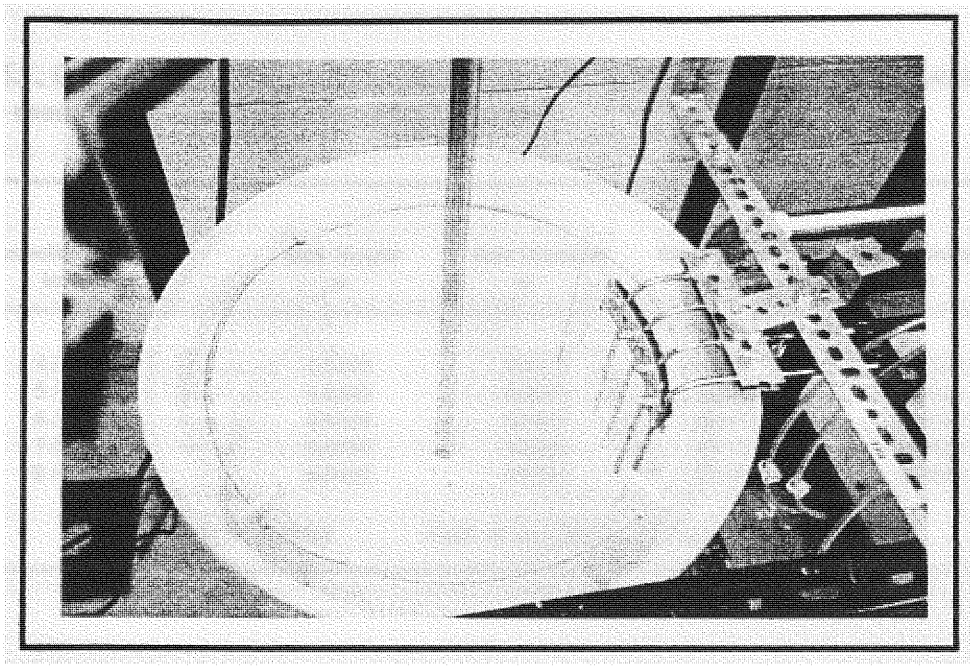

**Figure 5.4. Tank used in the experiments with the arrangement of four baffles distributed symmetrically. The four tubes used for sampling are on the right.**

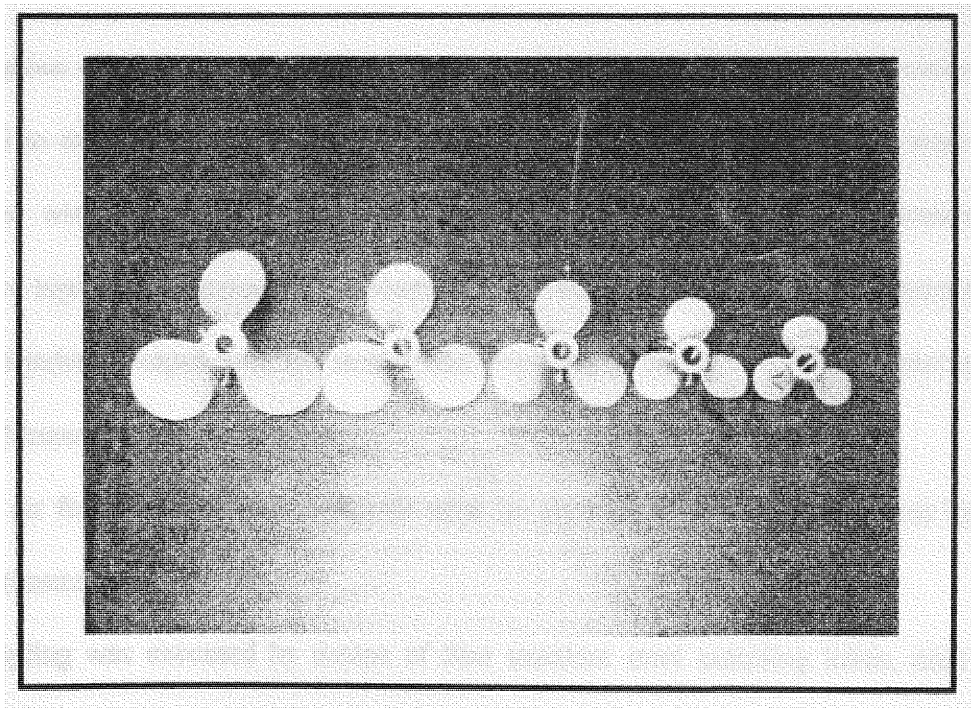

**Figure 5.5. Axial flow impellers (pump-down) utilized. Left to right: 8, 7, 6, 5, and 4 inches of nominal diameter.**

**Table 5.1. General dimensions of the impellers used in this investigation. Dimensions are the same for up and down pumping impellers.**

| Impeller Diameter inch / m | Hub Height m | Hub Diameter m | Blade Width m | Blade Thickness m |
|---|---|---|---|---|
| 4 / 0.1016 | 0.0300 | 0.0310 | 0.042 | 0.0010 |
| 5 / 0.1270 | 0.0300 | 0.0345 | 0.045 | 0.0018 |
| 6 / 0.1524 | 0.0330 | 0.0345 | 0.045 | 0.0020 |
| 7 / 0.1778 | 0.0394 | 0.0345 | 0.072 | 0.0030 |
| 8 / 0.2032 | 0.0465 | 0.0460 | 0.085 | 0.0040 |
| Angle at hub: 45 degrees | | Angle at tip of blade: 15 degrees | | |

The shaft of the mixer was made of stainless steel, with a diameter of 0.019 m and located at a distance of 0.183 m from the bottom of the tank. The electrical power consumption of the mixer was done by means of an Extech DW-6060 Watt meter and attached to the system as shown in figure 4.2. The rotational speed of the shaft was measured with a ROS-5W remote optical sensor by Monarch, which iluminated a reflective tape target located on the shaft with a visible red LED light and detected the reflected pulses. Output from the sensor was displayed in an ACT2A digital panel meter also by Monarch.

The air flow regulator and indicator used for startups with air injection was by Ki-Key Instruments.

Sampling was executed by means of four stainless steel sampling tubes, each one located at a different positions along the vertical axis of the tank. Tubes with inside diameters of 0.00635 m (0.25 in.) and 0.0095 (0.375 in.) were used. Tygon flexible

plastic tubing, surgical and transparent, was used to transport the samples. The inside diameters of the tubing were 0.00635 m (0.25 in.) and 0.0095 (0.375 in.) as to maintain the same diameter as that of the sampling tubes and not create disturbances in the flow. The samples were withdrawn from the tank with a Masterflex Model 7550-60 variable-speed peristaltic pump, controlled by an IBM PC computer. The pump and the control software were by Cole Palmer. The samples were collected in 50 ml glass beakers manufactured by Pyrex and Kymax. The scales used for weight measurements were a OHAUS Precision Plus and a Denver Instrument Company A-200DS. Samples were placed for drying in a Imperial V Laboratory Oven by Lab-Line.

# 6. RESULTS AND DISCUSSION

The work was divided into several experiments from which the system characteristics, behaviors, and effects previously indicated could be ascertained:

1. Power consumption.
2. Effect of the size of the sampling tubes on the local solids concentration measured.
3. Effect of the speed of sampling on the local solids concentration measured.
4. Effect of the impeller size on the overall performance of the process.
5. Startup from fully-settled solids.

The power consumption under a fully developed flow was measured for solids concentration of 5 and 20 percent (weight). The performance of two different sizes of sampling tubes was evaluated with a concentration of 5 % (weight), suction speed was evaluated with both concentrations, and startup characteristics and techniques from fully-settled conditions were assessed with the slurry with a concentration of 20% (weight).

## 6.1 Power Consumption

The cost of operation depends largely on the power that is consumed by the working impeller. It is therefore important to posses a good understanding of the power characteristic of the system.

The power consumption characteristics under steady state for five different sizes of two different type of axial-flow impellers, namely "up-flow" and "down-flow" impellers were determined first for the slurry of solids concentration of 5% (weight). An impeller installed in an "up-flow" fashion, pumps the flow upwards, towards the surface. With the

same rotational direction, a "down-flow" impeller would pump the flow down, towards the bottom of the tank.

Figure 6.1 shows the electric power consumption of all five different sizes of axial-up-flow impellers used with a concentration of 5% (weight). Figures 6.2 and 6.3 presents the same comparison for the axial-down-flow impellers acting on a slurry with a solids concentration of 5% (weight) and 20% (weight) respectively. As it can be clearly noticed, the bigger the impeller used, more power is drawn by the motor. Although the weight of a bigger impeller might be the driving factor of the power consumed, the higher levels of turbulence produced compared with a smaller diameter impeller may play an important role in the power consumption too, specially in scaled-up systems. The role of the impeller size in the performance of the mixing process under steady state conditions is more evident when the curves of the 4 inch and the 8 inch impellers on any of these three figures are compared. For instance, in figure 6.1, at a rotational speed of 500 rpm, the 8 inch impeller provides much more turbulence (which is highly desirable for mixing purposes) than its 4 inch counterpart. On the other hand, the 8 inch impeller consumes approximately 5 times more electric power than the 4 inch impeller, which makes it much less efficient from an economic standpoint. This is further illustrated by figures 6.4 and 6.5, which are a comparison of the electric power consumed by the different sizes of impellers ("down-flow") at various rotational speeds with slurries of solids concentration of 5% (weight) and 20% (weight) respectively.

The "up-flow" and "down-flow" tests were done in an attempt to determine if the pumping direction of the axial impellers would have an effect in the power consumption of the process. The classic literature states that pumping the flow in an upward fashion

results in power consumption in the neighborhood of 15% above from what would be consumed with the flow being pumped downward. This was corroborated with the results shown in figures 6.6, 6.7, 6.8, and 6.9, which compare the measured power between the "up-flow" and the "down-flow" impellers for the 4", 5", 6", and 7" impellers respectively. As the rotational speed increased, the difference in power consumption becomes more evident. Table 6.1 shows a comparison at three selected rotational speeds.

**Table 6.1. Power consumption comparison between "up-flow" and "down-flow" impellers at three different rotational speeds.**

| 4" (0.1016 m) impeller | | | | 5" (0.1270 m) impeller | | | |
|---|---|---|---|---|---|---|---|
| | Power (W) | | | | Power (W) | | |
| Speed (rpm) | Down | Up | Difference | Speed (rpm) | Down | Up | Difference |
| 200 | 13 | 14 | 7% | 200 | 11 | 14 | 21% |
| 800 | 48 | 61 | 21% | 800 | 55 | 75 | 27% |
| 1300 | 96 | 121 | 20% | 1300 | 140 | 188 | 25% |
| **6" (0.1524 m) impeller** | | | | **7" (0.1778 m) impeller** | | | |
| | Power (W) | | | | Power (W) | | |
| Speed (rpm) | Down | Up | Difference | Speed (rpm) | Down | Up | Difference |
| 200 | 14 | 15 | 7% | 200 | 16 | 18 | 11% |
| 800 | 82 | 114 | 28% | 800 | 136 | 230 | 40% |
| 1300 | 266 | 306 | 13% | 1300 | N/A | N/A | N/A |

As it can be seen, the power consumed with an "up-flow" impeller was in most cases even higher than 15%, making them less cost-effective than the "down-flow" impellers. Although the levels of turbulence generated with "up-flow" impellers are evidently higher, this results in continuous "splashing" that is undesirable in most cases. It is for these reasons that only "down-flow" impellers were used in the rest of the experiments, as the aim of this work is to find the most efficient way of achieve acceptable levels of

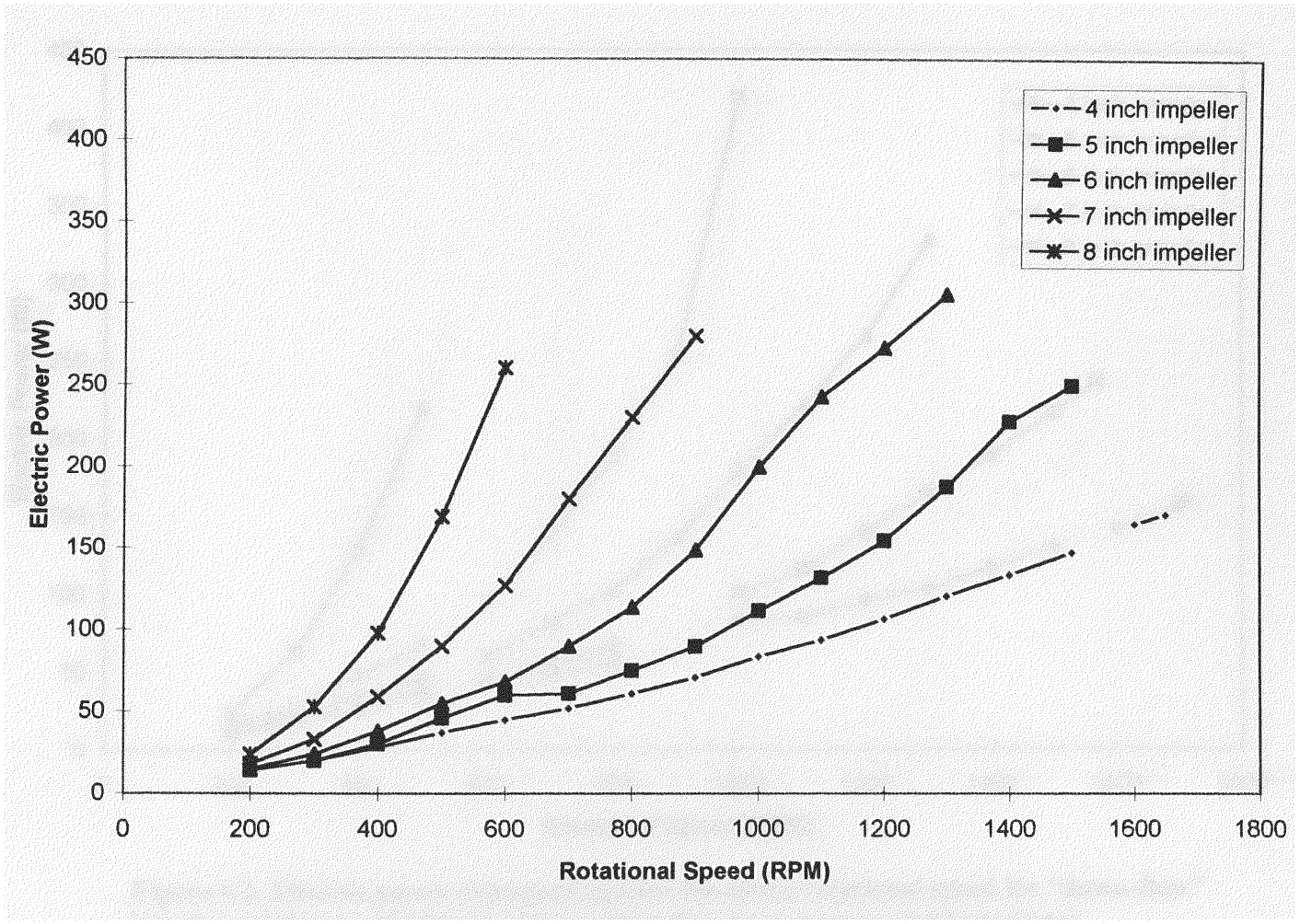


**Figure 6.1. Electric power consumed against the mixer rotational speed for "up-flow" impellers acting on the slurry with a bulk solids concentration of 5 % (weight).**

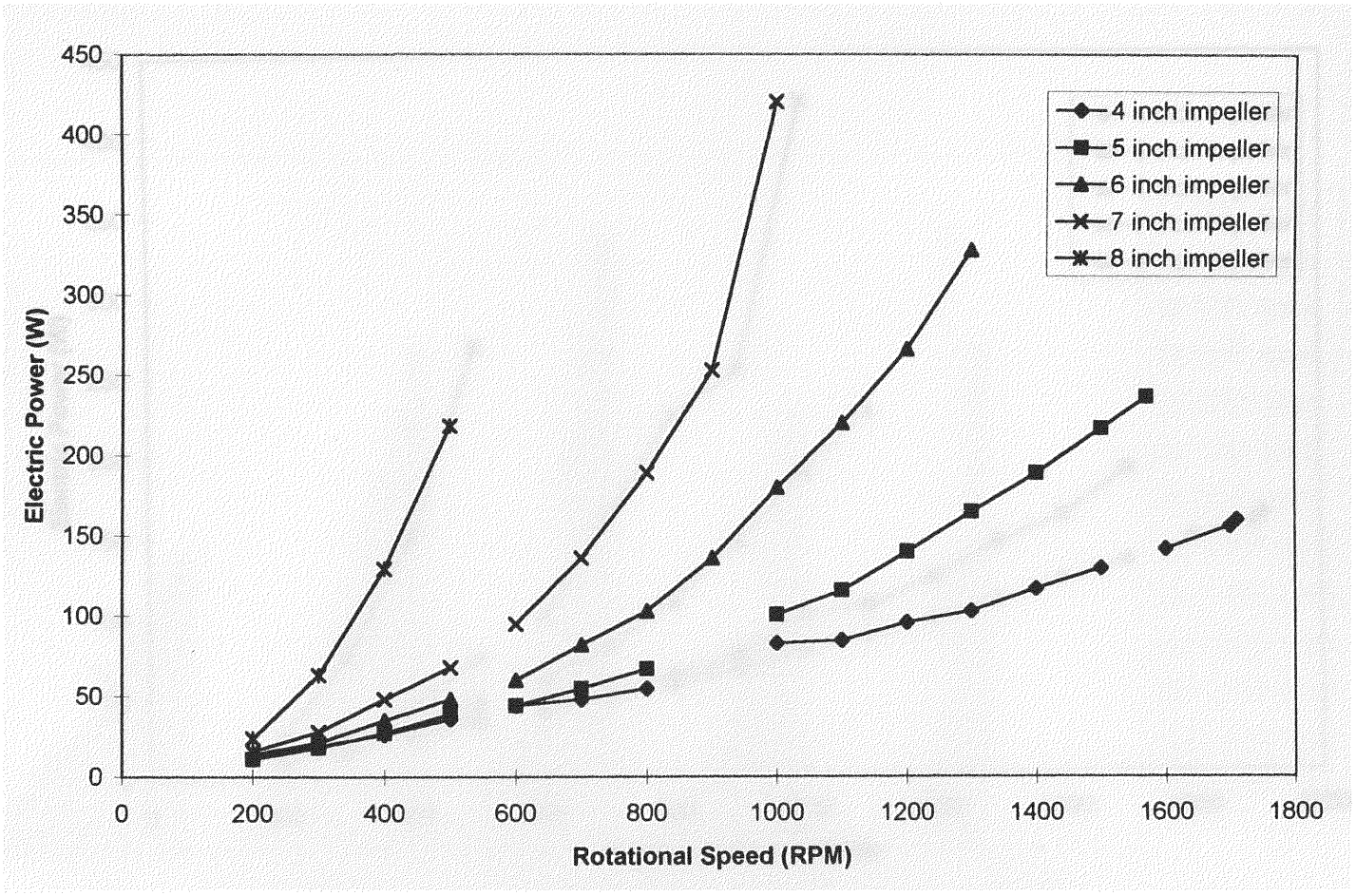


**Figure 6.2. Electric power consumed against the mixer rotational speed for "down-flow" impellers acting on the slurry with a bulk solids concentratoin of 5 % (weight).**

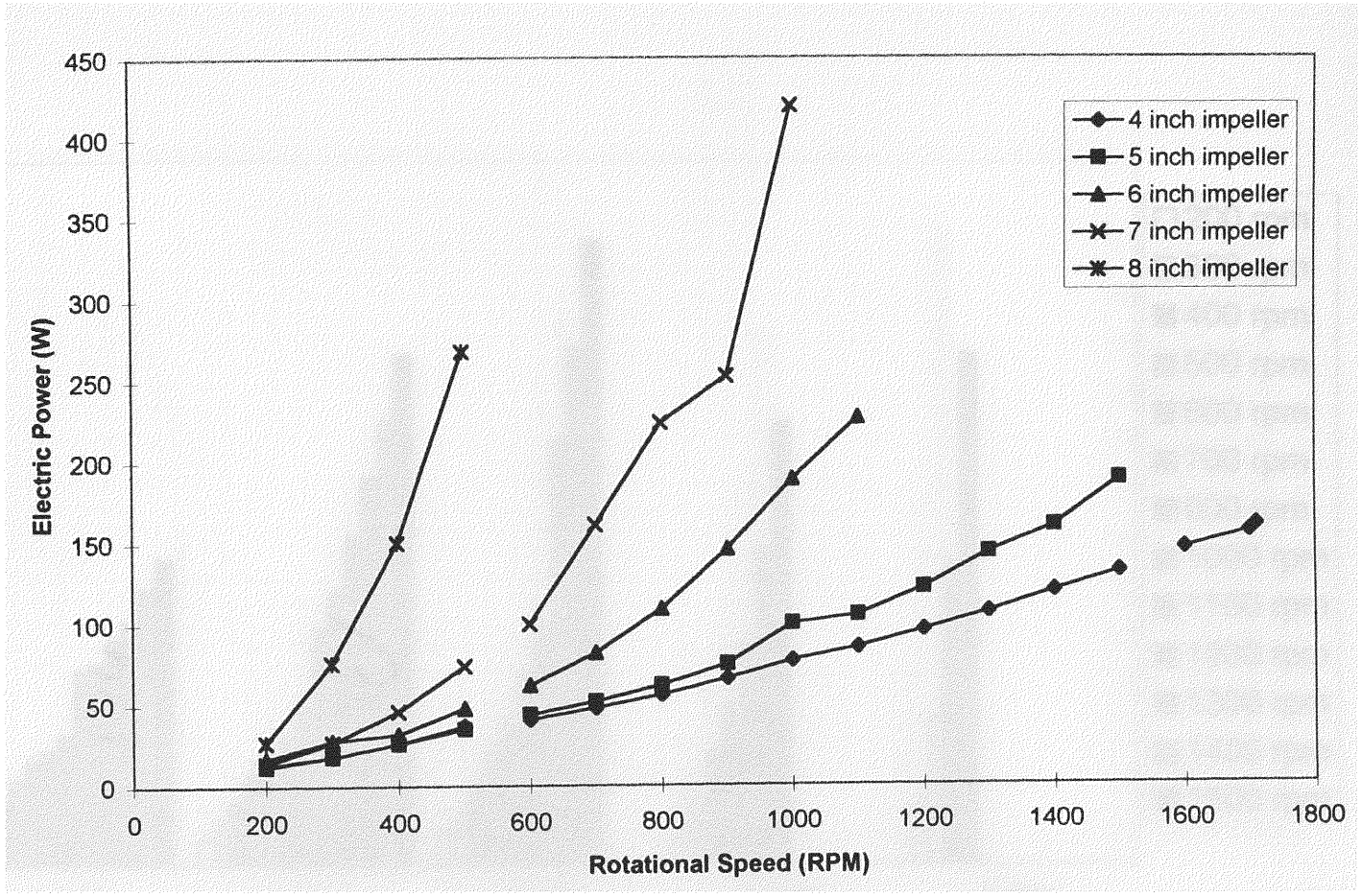


**Figure 6.3. Electric power consumed against the mixer rotational speed for "down-flow" impellers acting on the slurry with a bulk solids concentratoin of 20 % (weight).**

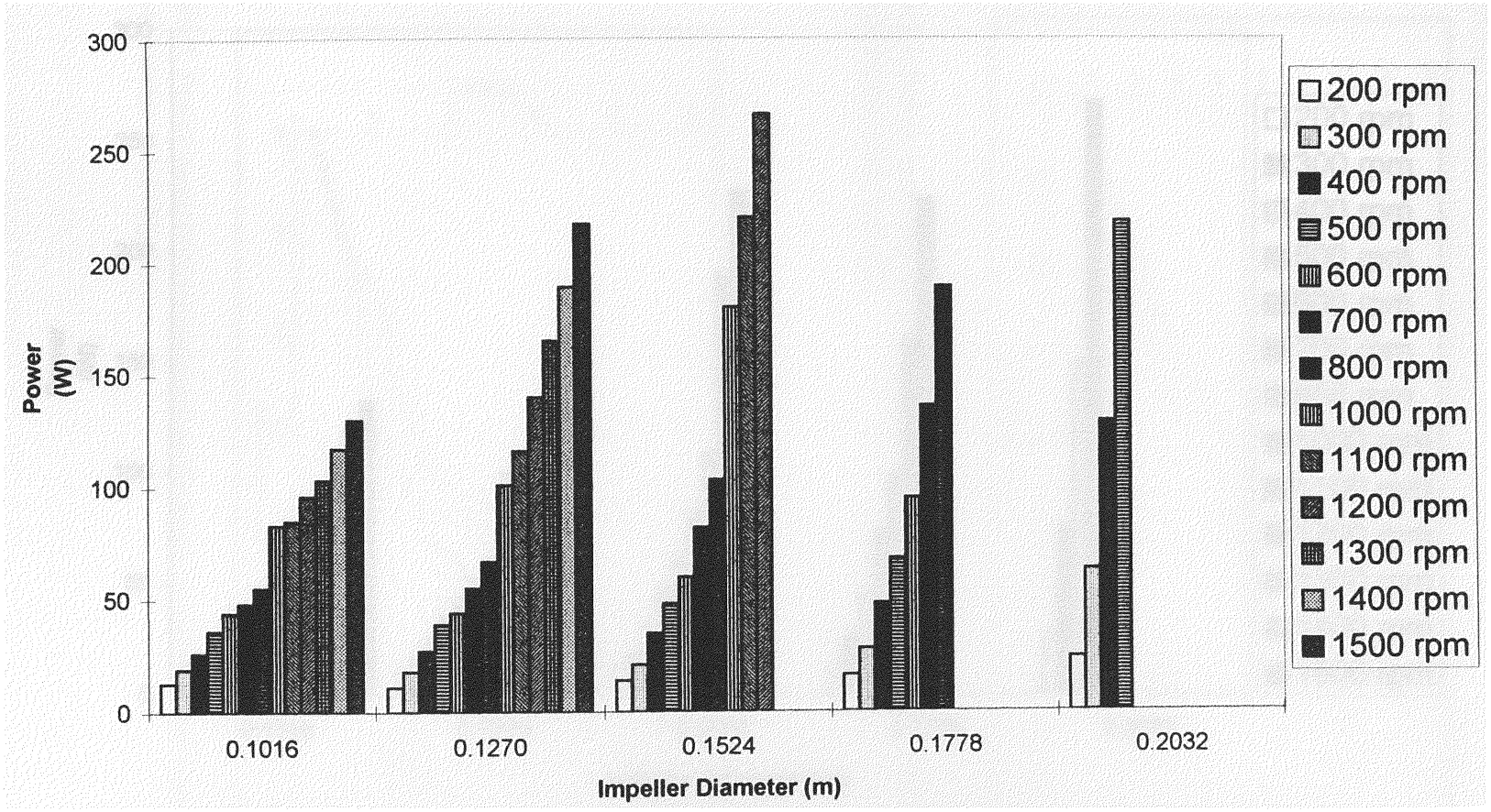


**Figure 6.4. Power consumption against impeller diameter for various rotational speeds. Solids concentration: 5 % (weight)**

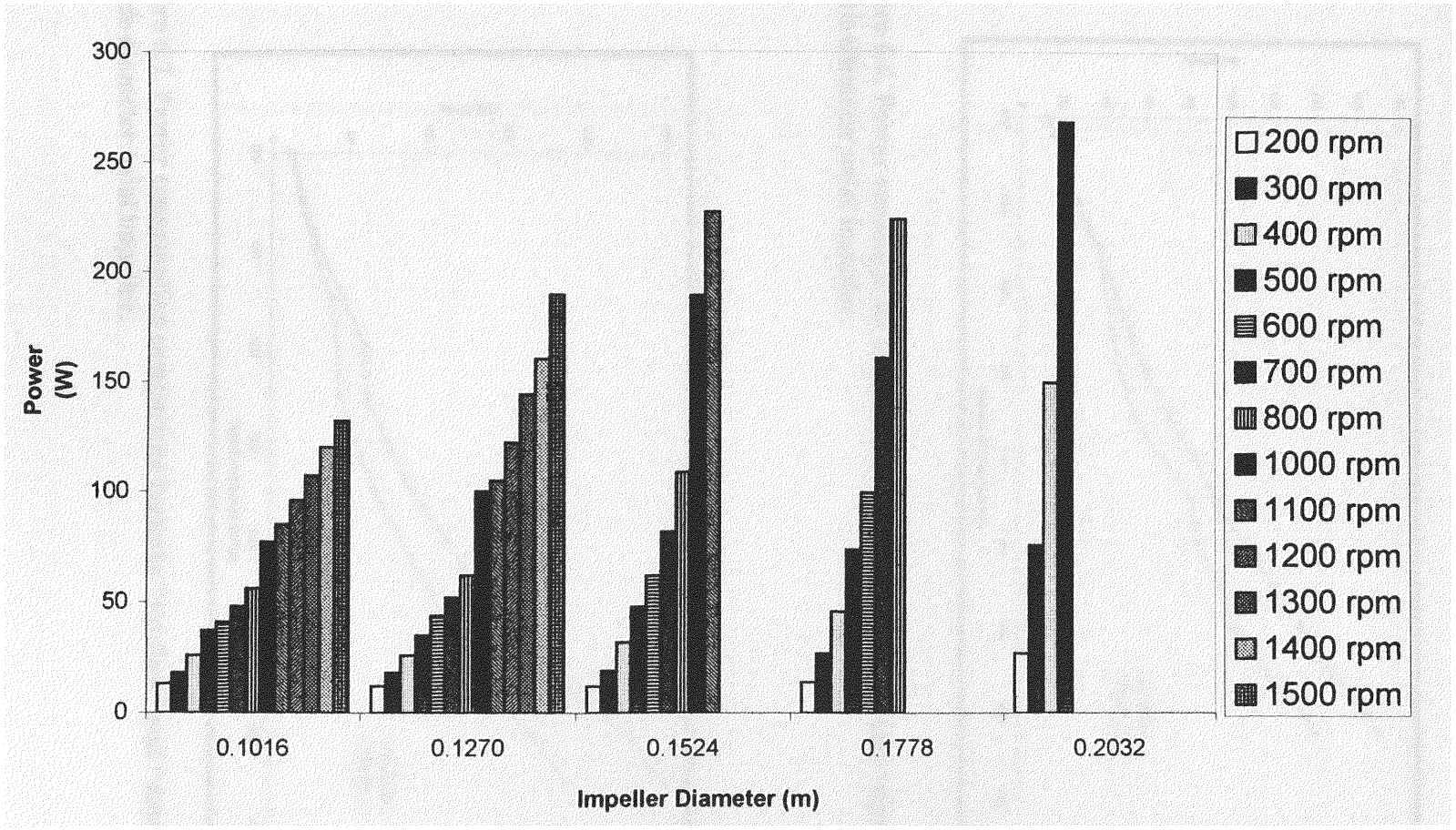


**Figure 6.5. Power consumption against impeller diameter for various rotational speeds. Solids concentration: 20 % (weight)**

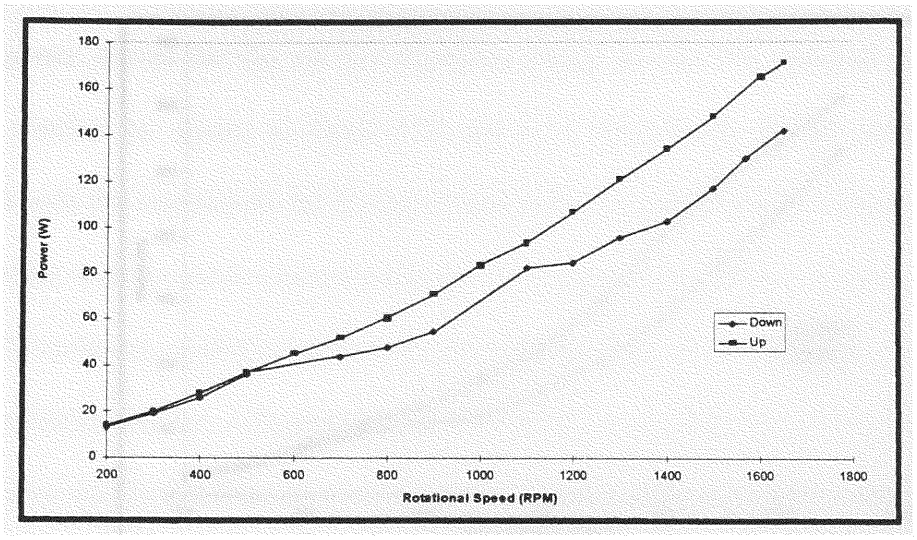


**Figure 6.6. Power consumption comparison between "up-flow" and "down-flow" 4 inch-diameter axial impeller.**

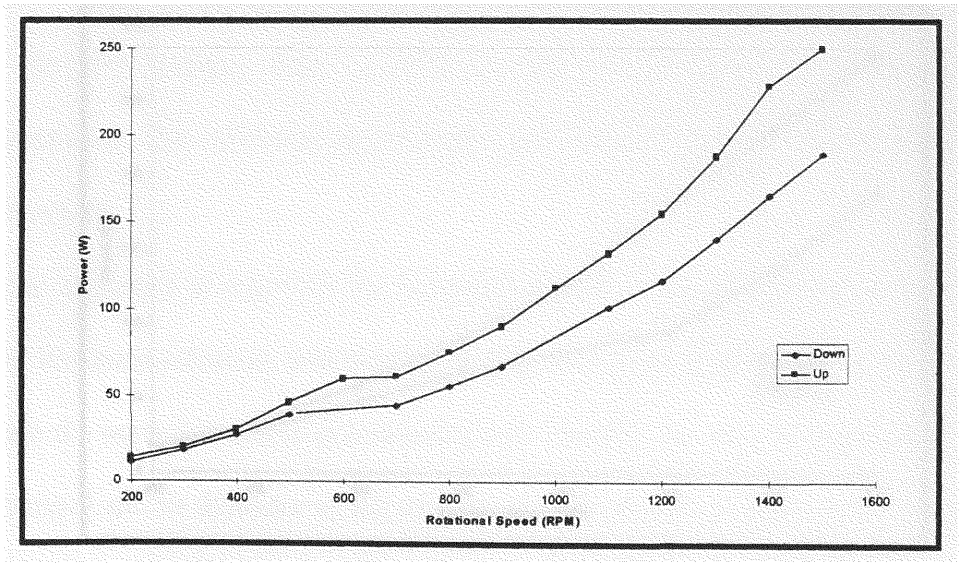


**Figure 6.7. Power consumption comparison between "up-flow" and "down-flow" 5 inch-diameter axial impeller.**

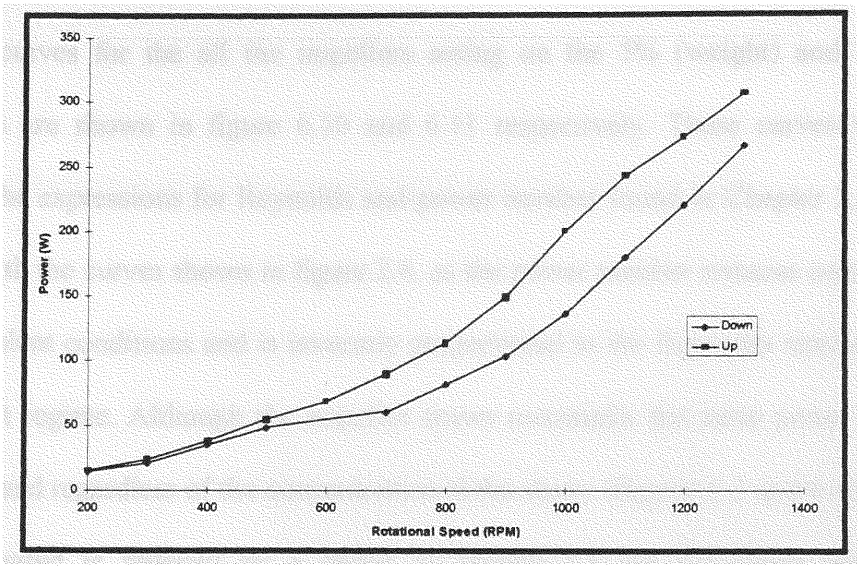


**Figure 6.8. Power consumption comparison between "up-flow" and "down-flow" 6 inch-diameter axial impeller.**

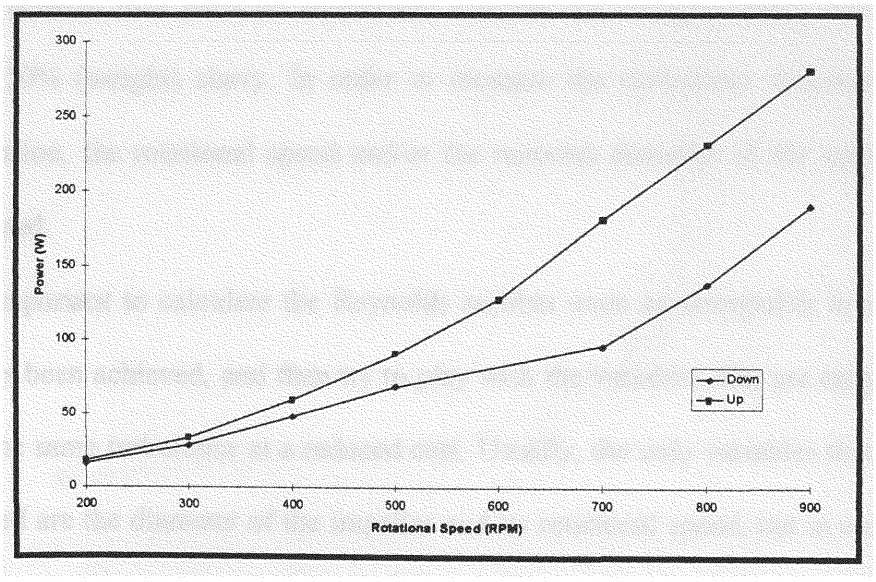


**Figure 6.9. Power consumption comparison between "up-flow" and "down-flow" 7 inch-diameter axial impeller.**

homogeneity on the mix.

The power curves for the all the impellers acting on the 5% (weight) and 20% (weight) slurries are shown in figure 6.10 and 6.11 respectively. These curves were generated with the expressions for Reynolds and power number found in Chapter 2, and are consistent with the curves shown in figure 2.6, as the power number remains constant under fully turbulent conditions and is inversely proportional to the Reynolds number in the non-turbulent regime. Although the impeller draws essentially the same power at a any rotational speed regardless of the concentration of the slurry (figures 6.3 and 6.4), the turbulence generated is reduced by a factor of around 7.8 for the slurry with a concentration of 20% (weight) due to the effect of the increased density and viscosity. The turbulence generated with a certain size of impeller rotating at a certain speed may be sufficient to achieve total particles suspension with the 5% (weight) slurry, but may not be for the 20% (weight) slurry. In order to increase the turbulence necessary to maintain suspension, the rotational speed and/or the nominal diameter of the impeller had to be increased.

It is therefore important to calculate the Reynolds number once an acceptable level of homogeneity has been achieved, and then try to play with the variables that are easier to vary to obtain the same turbulence at a reduced cost. Usually, the only variables that can be easily changed are the diameter of the impeller and its rotational speed, but in certain applications the viscosity could be reduced by the application of heat or by means of chemical agents, in order to increase the Reynolds number. Figures 6.12 and 6.13 are plots of the power number against the impeller diameter-tank diameter ratio for the 5 % (weight) and the 20 % (weight) slurries respectively. These plots can be useful for the

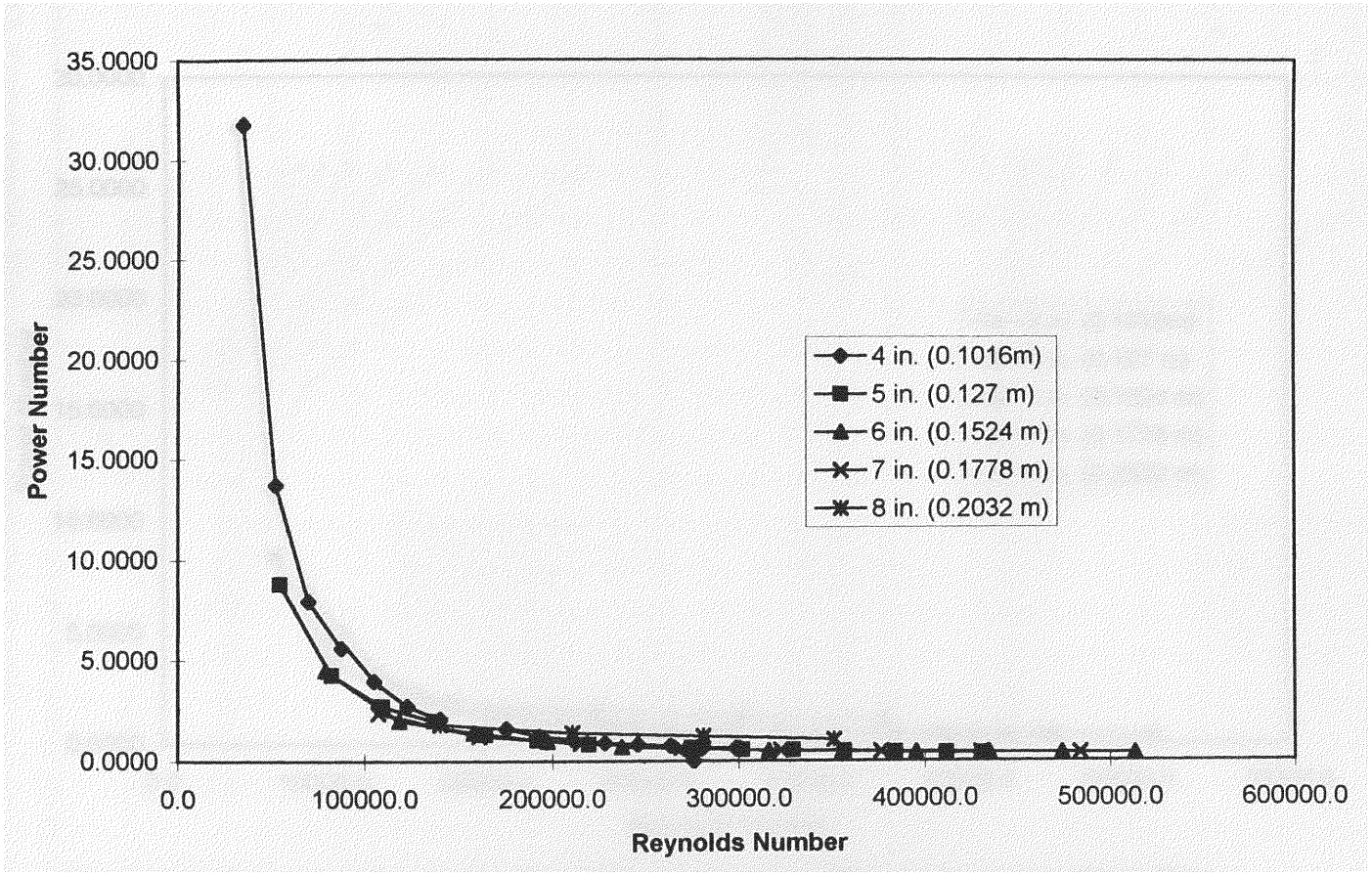


**Figure 6.10. Power number against Reynolds number for different diameters of impellers. Solids concentration: 5 % (weight).**

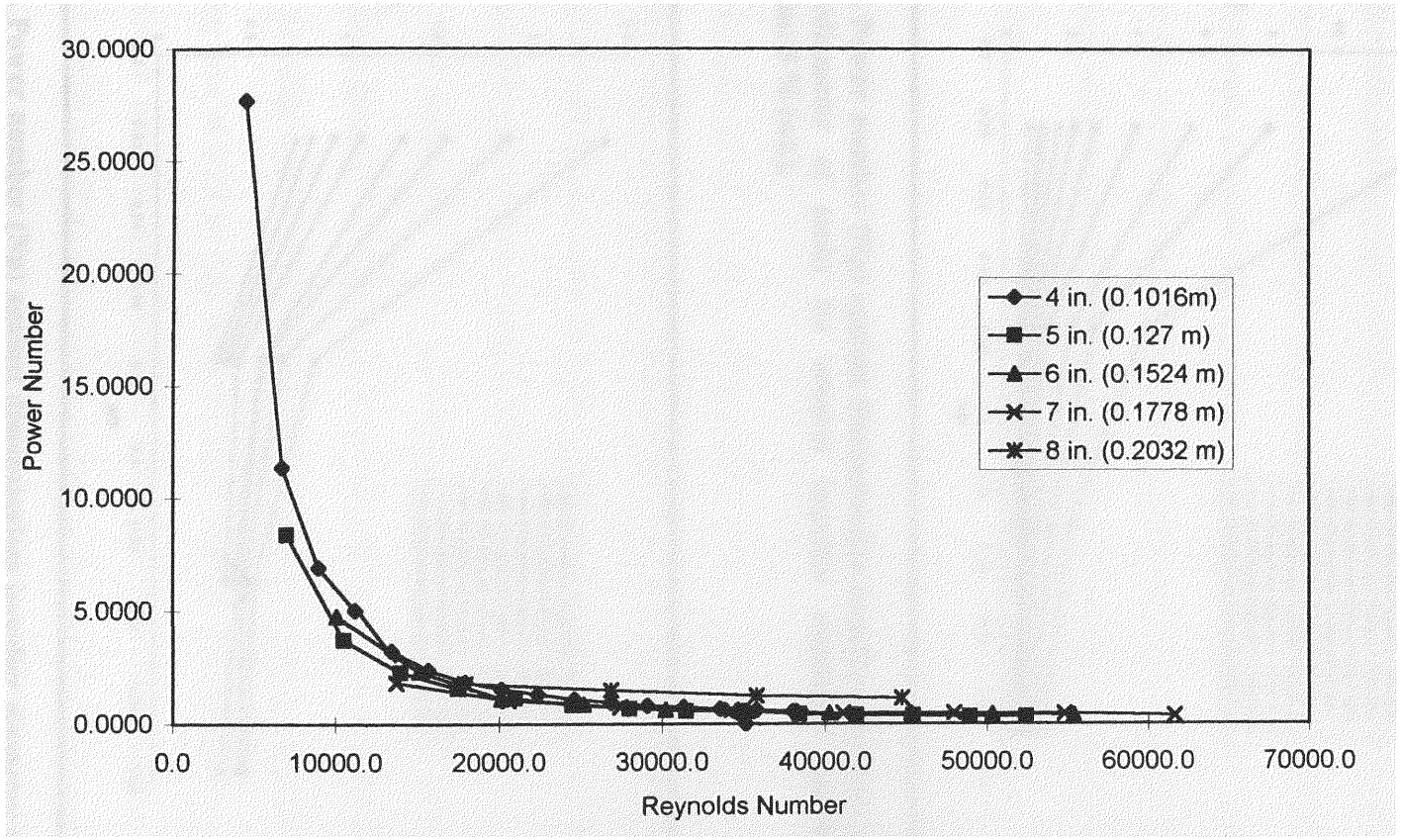


**Figure 6.11. Power number against Reynolds number for different diameters of impellers. Solids concentration: 20 % (weight).**

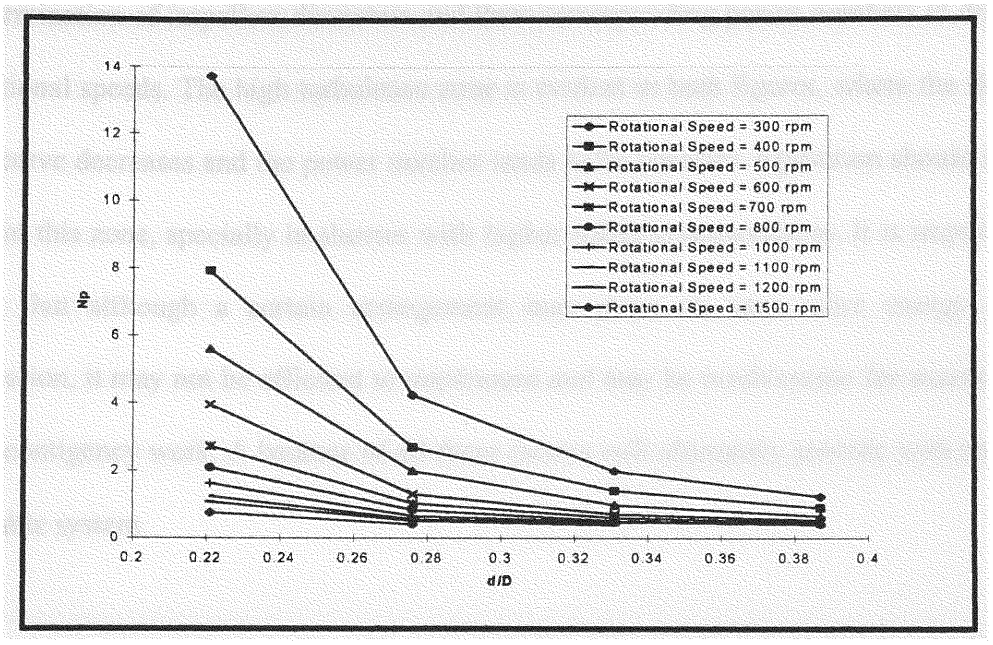


**Figure 6.12. Power number (Np) against dimensionless impeller diameter (diameter of impeller/diameter of tank) for several impeller rotational speeds. Solids concentration: 5 % (wt.).**

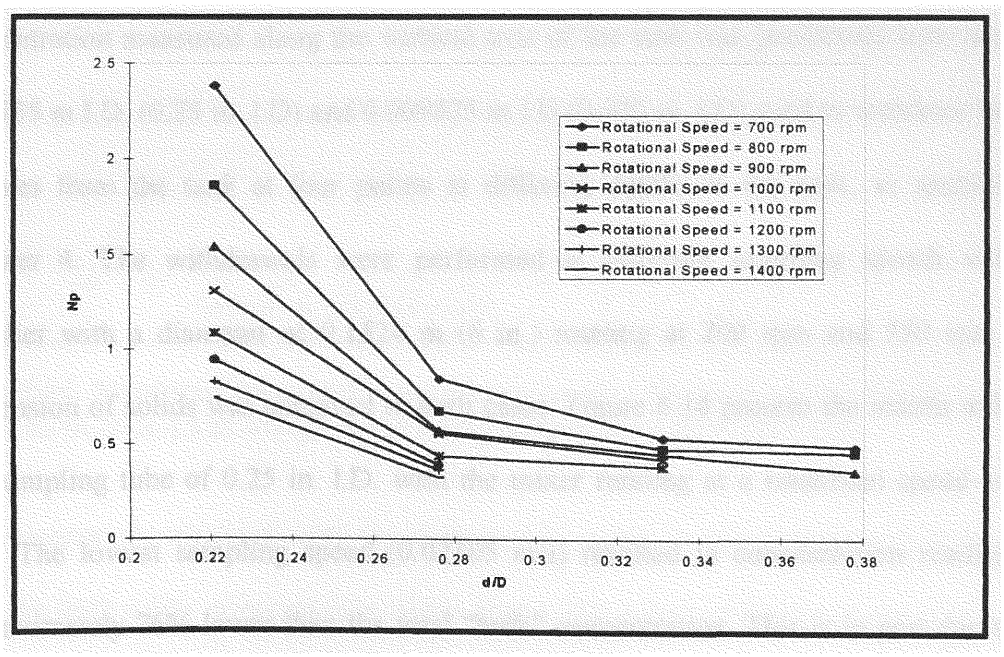


**Figure 6.13. Power number (Np) against dimensionless impeller diameter (diameter of impeller/diameter of tank) for several impeller rotational speeds. Solids concentration: 20 % (wt.).**

determination of impellers diameters and their corresponding power numbers at different rotational speeds. The high turbulence zone is evident in both figures, where the slope of the curve decreases and the power number tends to be constant. Operation should ideally fall in this zone, specially in slurries with higher solids concentrations. It is important to note that although a certain arrangement may prove to save more energy during operation, it may not be efficient to implement and may be troublesome for maintenance and contigency work. A balance of all these factors will ultimately provide with the most suitable system.

## 6.2. Mixing of the 5 % (weight) solids concentration slurry and effect of sampling tubes diameter and sampling velocity in the concentration measurements

The evaluation of the effect of the geometry of the sampling tubes on the solids concentration measured along the vertical axis of the tank was performed with tubes of 0.00635 m I.D. (0.25 in. I.D) and 0.009525 m I.D (0.375 in. I.D) used to withdraw the samples from the tank at four points at different depths in the tank, as specified in Chapter 4. The withdrawals were performed at different sampling speeds with an impeller with a diameter of 0.1524 m (6 in.) rotating at 200 rpm and 550 rpm. Full suspension of solids was achieved in both cases. Figure 6.14 present the results utilizing the sampling tube of 0.25 in. I.D. with the mixer running at a rotational speed of 200 rpm. The lowest sampling speed (0.02105 m/s) resulted in concentration readings of approximately 76% lower than the total "bulk" concentration. This is in part due to the considerable settling of solids settled along the tubing, resulting in samples with a

concentration much lower than those from samples taken at higher speeds. This indicates that it is very important to ensure the highest level of turbulence in the tube so that a greater amount of solids reach the collecting container. The highest velocity (0.63153 m/s) presents higher concentration values but when compared with the results of the sampling done with the 0.375 in. I.D. tube (figure 6.15), it can be concluded that those values are not representative of the concentration of the mixture. Figures 6.16 and 6.17 present similar results for the impeller running at a rotational speed of 550 rpm, although

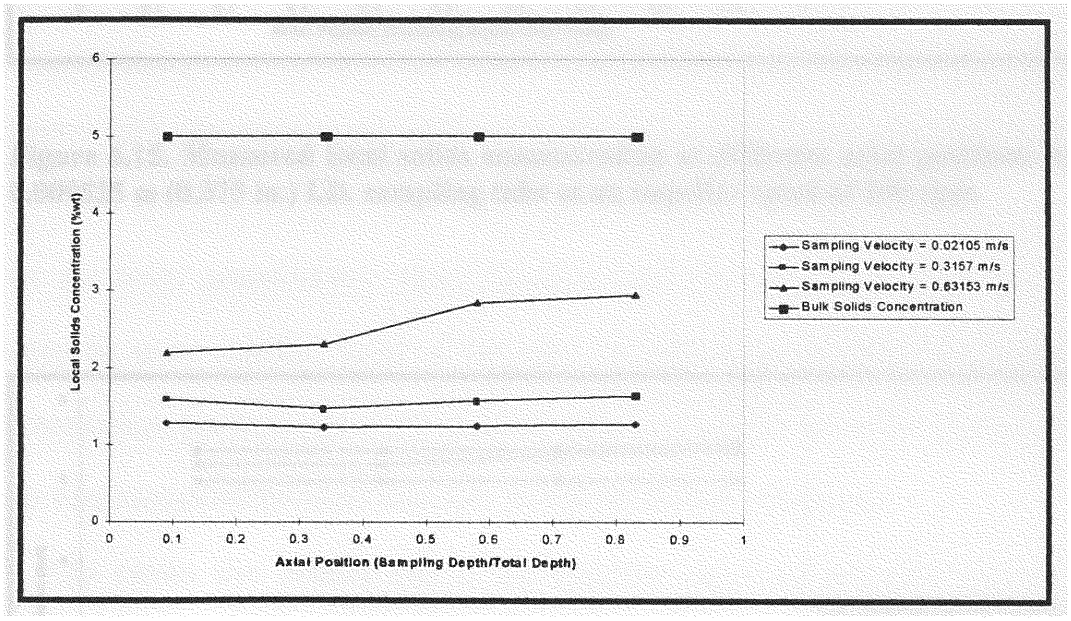


**Figure 6.14. Measured local solids concentration at different axial positions with a 0.00635 m (0.25 in.) I.D. sampling tube at an impeller speed of 200 rpm.**

the concentrations obtained with the 0.25 in. I.D. tubes at the higher sampling rates match the bulk solids concentration of the slurry. The nature of the slurry plays an important role on the accuracy of the samples taken. Once the ferric oxide enters in contact with water, it sticks to the walls, forming a thin paint-like film. This may account

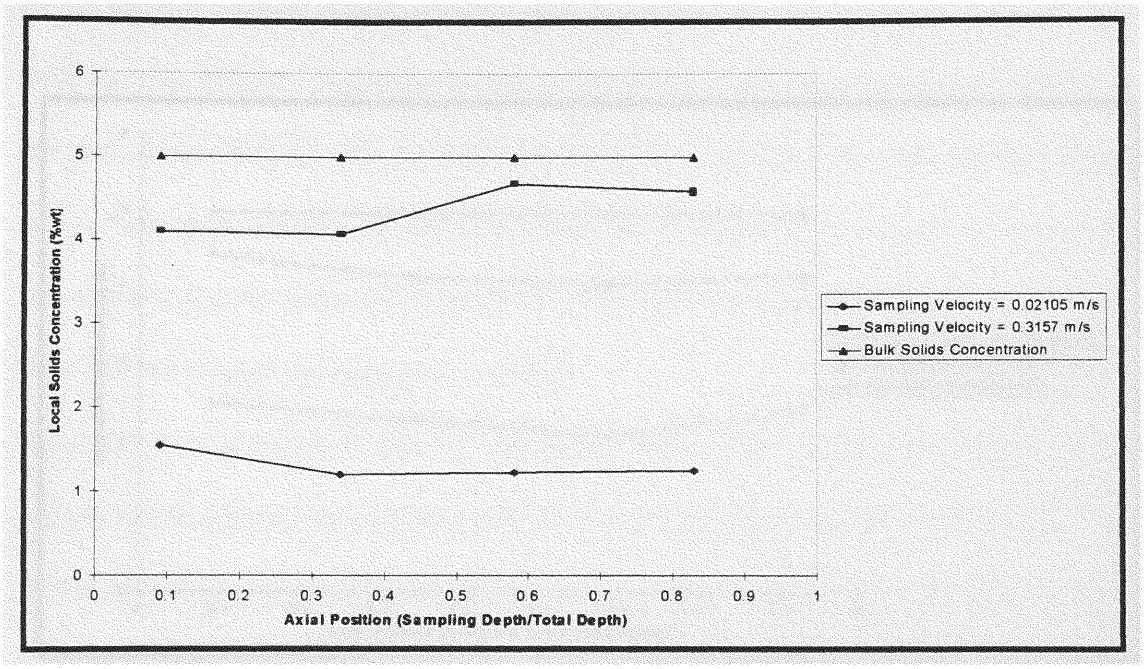


**Figure 6.15. Measured local solids concentration at different axial positions with a 0.009525 m (0.375 in.) I.D. sampling tube at an impeller speed of 200 rpm.**

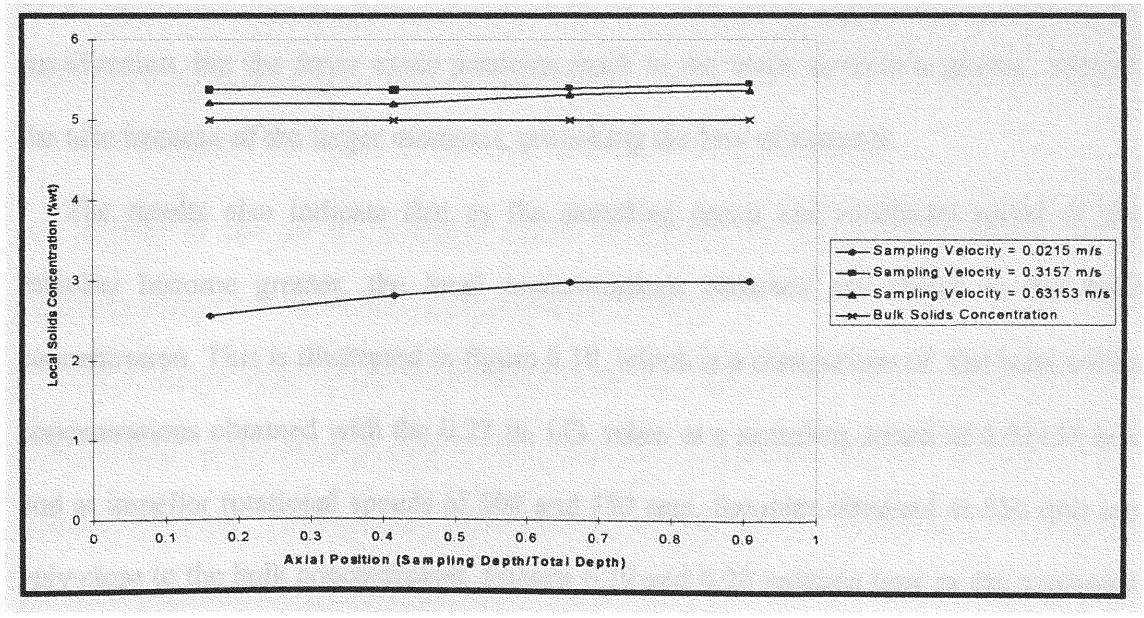


**Figure 6.16. Measured local solids concentration at different axial positions with a 0.00635 m (0.25 in.) I.D. sampling tube at an impeller speed of 550 rpm.**

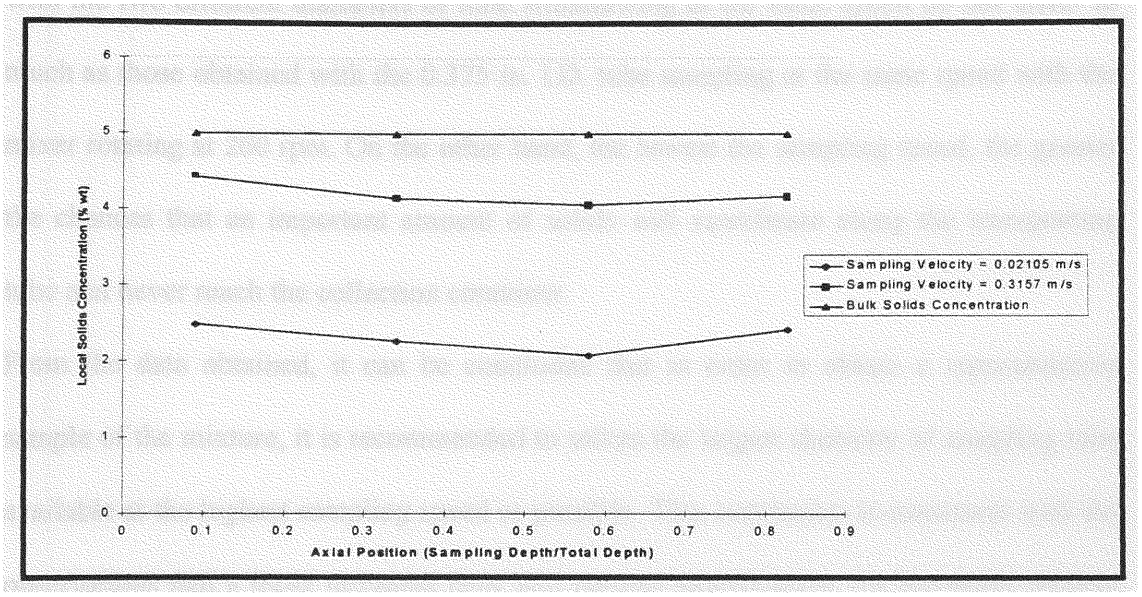


**Figure 6.17. Measured local solids concentration at different axial positions with a 0.009525 m (0.375 in.) I.D. sampling tube at an impeller speed of 550 rpm.**

for the lower concentration values obtained with the bigger sampling tubes when sampling at an impeller speed of 550 rpm: not only some material was lost due to stratification, but the ferric oxide particles stuck to the walls covered a greater area of the tube because of the larger diameter, provoking the loss of material.

The results also indicate that as the sampling speed and rotational speed of the impeller become greater, the local concentrations obtained are closer to the bulk concentration. This is illustrated in figure 6.18, which is a comparison of the local solids concentrations obtained with the 0.25 in. I.D. tubes at a sampling speed of 0.63153 m/s and at impeller rotational speeds of 200 and 550 rpm. Samples obtained at 550 rpm are very close to the bulk concentration. Figures 6.19 and 6.20 indicate how as the rotational speed of the impeller increases, the diameter of the tube becomes less important as long

as the sampling speed is high enough. The concentration profiles obtained at 550 rpm with the two different diameters of tube withdrawing at the same speed do not differ as much as those obtained with the 0.375 in. I.D. tube sampling at the same speed with the mixer rotating at 200 rpm. On the other hand, the lowest the sampling speed, the greater the chances that an important amount of solids will stratificate along the transporting tube and never reach the collection container.

From the data obtained, it can be concluded that in order to obtain a representative sample of the mixture, it is recommended to utilize the largest diameter of sampling tube available at the highest sampling speed as possible. This conclusion is consistent with the observations that a more turbulent flow will prevent stratification: the Reynolds number is proportional to the nominal diameter of the tube and the velocity of the flow. The larger the diameter and greater the fluid velocity, the larger the Reynolds number, more turbulence is produced, and less solids will settle during transportation. The results involving the effect of the speed of sampling are also consistent with the conclusions by MacTaggart, Nasr-El-Din, and Masliyah (1993).

### 6.3. Mixing of the 20 % (weight) solids concentration slurry and the effect of sampling velocity in the concentration measurements

The obtained local solids concentrations of the slurry with a bulk concentration of 20% (weight) were evaluated with the 0.1524 m (6 in.)-diameter impeller rotating at 700 rpm and 1050 rpm. The sampling tubes used had a diameter of 0.009525 m (0.375 in.) as per conclusions on the previous section.

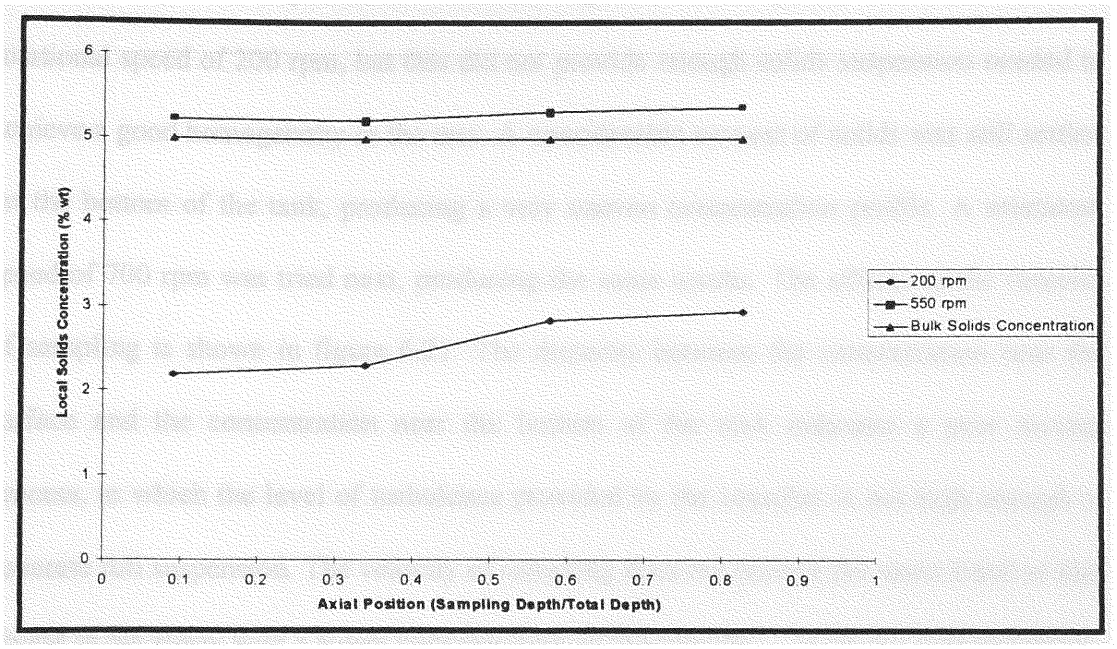


**Figure 6.18. Local concentrations obtained with 0.00635 m (0.25 in.) diameter tubes at a sampling speed of 0.63153 m/s and at impeller rotational speeds of 200 and 550 rpm.**

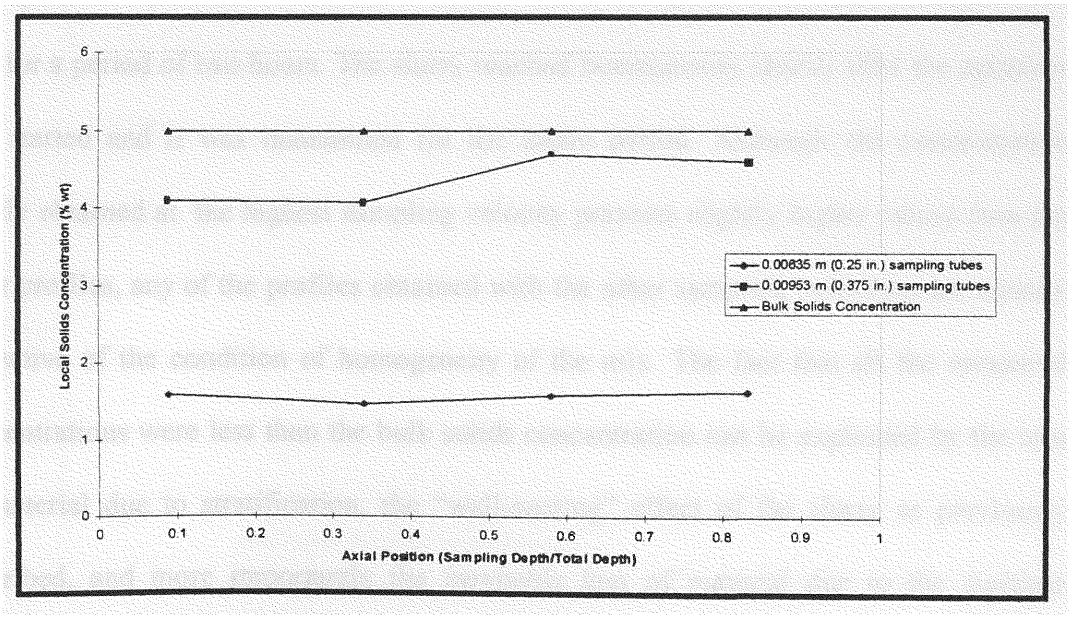


**Figure 6.19. Local concentrations obtained with the 0.00635 m (0.25 in.) I.D. and 0.009525 m (0.375 in.) I.D. tubes, sampling at a speed of 0.3157 m/s and at an impeller rotational speed of 200 rpm.**

The slurry was subjected to a constant agitation for a period of two hours at a rotational speed of 200 rpm, but this did not provide enough solids suspension needed to achieve a good homogeneity in the mix. A considerable amount of solids was still settled on the bottom of the tank, producing a very uneven concentration profile. A rotational speed of 700 rpm was tried next, producing the same results. The effects of the velocity of sampling is shown in figure 6.21. The disparity between the concentration near the surface and the concentration near the bottom of the tank indicates a poor mixing process, in which the level of turbulence provided by the impeller is not high enough to generate full suspension. The velocity of sampling does not present the same trend as that found in the slurry with a solids concentration of 5% (weight), as all the concentrations obtained under different velocities fell under the same range, presenting almost the same concentration profile.

Figure 6.22 presents the results obtained when the agitation was performed at 1050 rpm for a period of two hours. The slurry reached homogeneity shortly after the agitation was started and it was maintained for the entire period. Although the concentration profile obtained at the highest sampling velocity presents slightly higher values than the other profiles, any of the profiles obtained with the other sampling velocities are equally indicative of the condition of homogeneity of the mix. The fact that all the measured concentrations were less than the bulk solids concentration can be explained by the loss of material due to stratification, the "wall-coating" effect of the slurry as previously described, and more importantly the inevitable loss of material due to the multiple experiments.

As opposed with the slurry with a concentration of 5 % (weight), rotational speeds of 200 rpm and 700 rpm were not sufficient to promote the suspension of particles starting from rest. It takes higher speeds to achieve a level of suspension from which the mix can be safely assumed to be homogenous. This of course translates into a higher consumption of power.

The velocity of sampling was not found to have an important effect in the measured local concentration of the slurry with a bulk concentration of 20 % (weight), although the recommendations made in the previous section about performing the sampling at the highest velocity possible, are still valid in order to prevent plugging in the tubes and collect as much material that was withdrawn as possible.

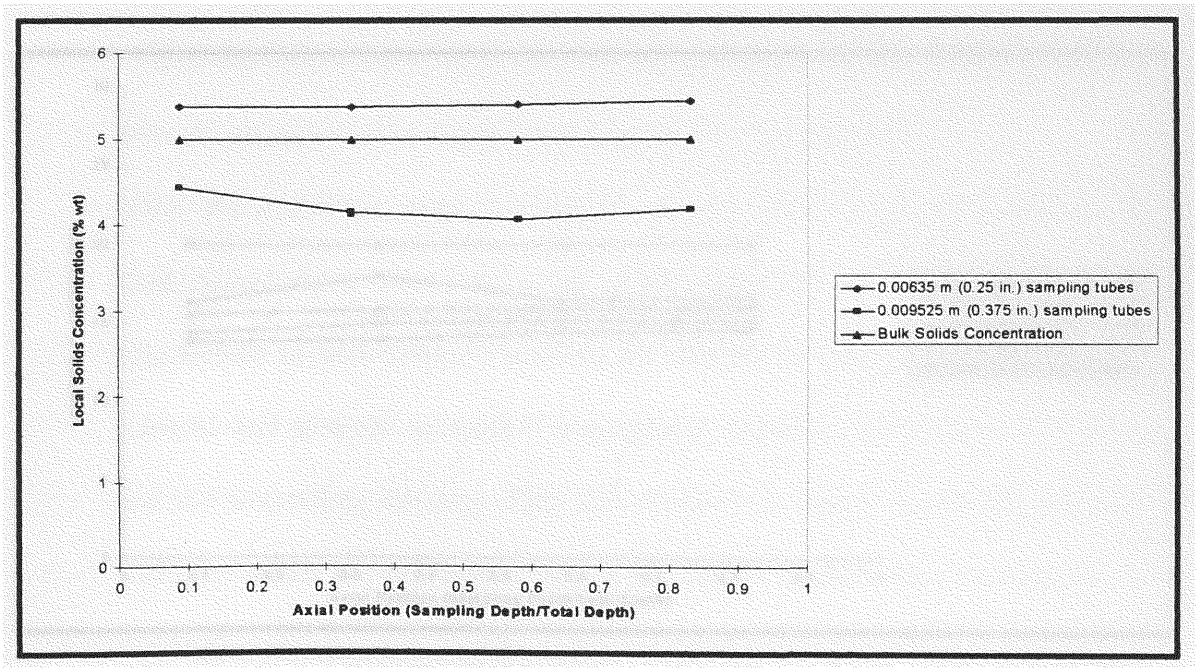


**Figure 6.20. Local concentrations obtained with the 0.00635 m (0.25 in.) I.D. and 0.009525 m (0.375 in.) I.D. tubes, sampling at a speed of 0.3157 m/s and at an impeller rotational speed of 550 rpm.**

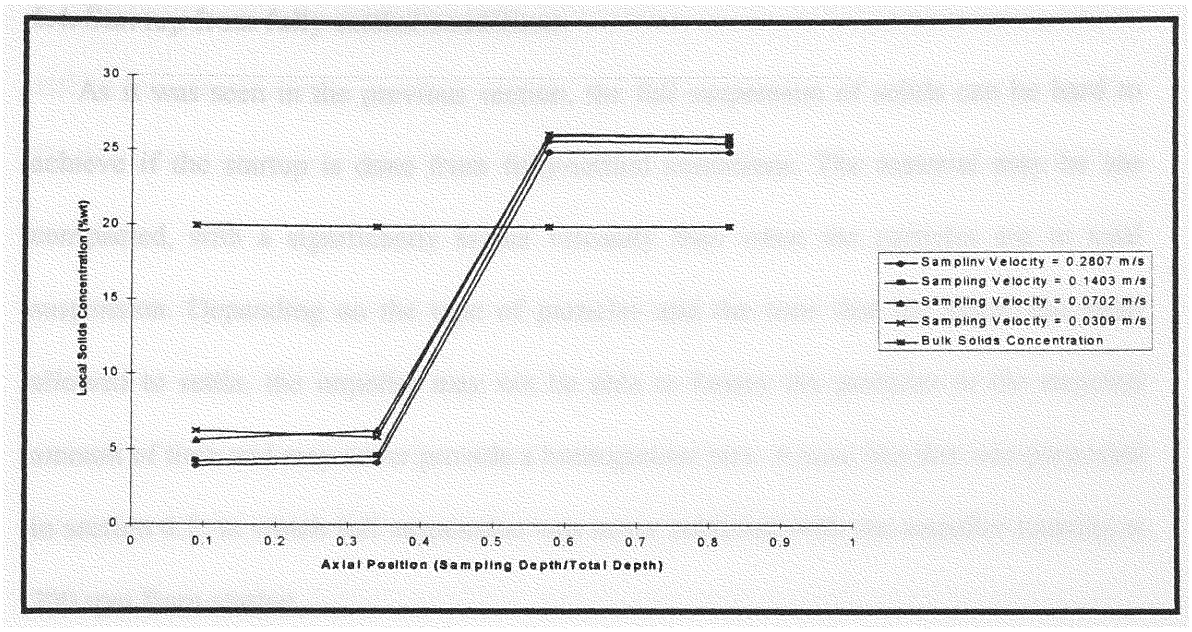


**Figure 6.21. Concentration profiles obtained for the slurry with a solids concentration of 20 % (weight) agitated with a 0.1524 m (6 in.)-diameter impeller rotating at 700 rpm. Agitation was maintained for two hours from a fully-settled condition.**

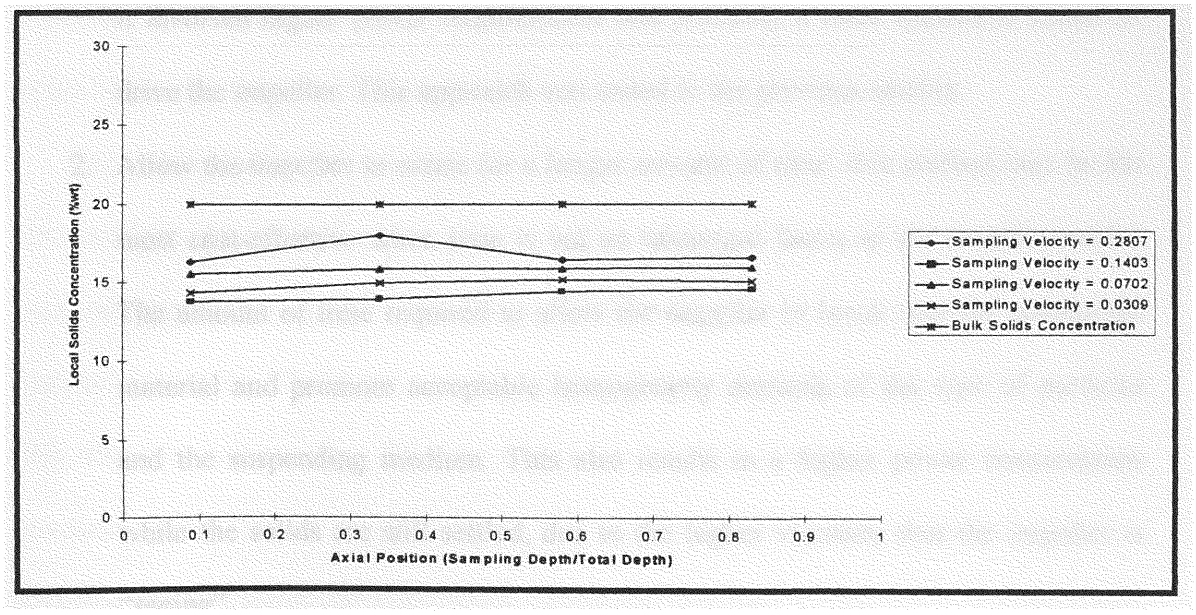


**Figure 6.22. Concentration profiles obtained for the slurry with a solids concentration of 20 % (weight) agitated with a 0.1524 m (6 in.)-diameter impeller rotating at 1050 rpm. Agitation was maintained for two hours from a fully-settled condition**

### 6.4. Startup from fully-settled conditions

As it was seen in the previous section, the full suspension of solids can be hard to achieve if the startup is done from fully-settled conditions. The material may be too compacted, with a significantly higher viscosity than when the particles are in total suspension. Depending on the type of particles and the time that the slurry has been allowed to settle, the impeller may not be able to loosen the particles in the required amount of time and may never provide a homogenous mix. A case like this was presented in section 6.2, in which full suspension was never achieved with the impeller rotating at 700 rpm from startup.

Different solutions for this situation include:

1. Rotate the impeller at a higher speed: although this might be the simplest approach from an operational standpoint, it may not be the most economical since it involves higher power requirements and probably a more expensive motor to drive the impeller. This approach was tested in the previous section.
2. Allow the impeller to rotate for a longer amount of time: this method may be the most cost-effective when time is not an important factor in the overall process. The amount of time required to allow the impeller to break into the compacted material and promote acceptable homogeneity depends of the type of particles and the suspending medium. This also results in a higher power consumption while the solids are still settled, due to the higher viscosity that the impeller is "seeing".
3. Injection of a jet of air in order to loosen the particles before or when starting to rotate the impeller. The advantage of this method is that the impeller will start

conditions very close to those of a fully suspended medium, and the power consumption will be considerable less.

The last two methods were applied to the slurry with a bulk solids concentration of 20 % (weight).

As stated above, when a slurry has been left under repose by a sufficient amount of time so that the solid particles fully settle, it may be difficult to start the mixer and obtain a steady condition in which the particles are fully suspended. Figure 6.21 shows the solids concentration profile obtained after the slurry has been agitated for two hours, starting from a fully settled condition. The mixer was started and left running at a constant rotational speed of 700 rpm. The nominal diameter of the axial flow impeller used was 0.1524 (6 in.). Very poor mixing conditions are evident given the high discrepancy in the solids concentration obtained from samples at different axial positions. Most noticeable is the great difference in concentration values between the upper half of the tank and the lower half, in which the solids were still settled.

The outcome is different after the slurry was agitated gradually from 200 rpm to 700 rpm and left running at that speed for an additional time. The mixer was started at 200 rpm and kept at that speed for 5 minutes before the speed being increased to 300 for another 5 minutes. The speed was increased in the same fashion passing through 400, 500, 600 rpm until reaching 700 rpm, which was kept for three hours. By means of these transient speeds before reaching the steady speed of 700 rpm, and by keeping the agitation for this additional time before taking the samples, a good level of suspension was achieved.

The last test involved the injection of air at 2 SCFM through a 0.00953 m (0.375in.) diameter tube located at .20m from the bottom of the tank for 10 seconds before the mixer was started at 700 rpm. The air was injected vertically towards the bottom of the vessel. This method resulted in the most effective means to achieve the suspension of solids from a fully settled conditions by consuming the least amount of power. Full suspension was noticeable even before the mixer was turned on. The two main advantages of this method are: 1) suspension of solids is achieved immediately, even before starting the mixer, and 2) the power consumption is less and remains constant.

Figure 6.23 presents the local concentrations measured with this two methods and how they compare with the profile resulting from a startup with no aid.

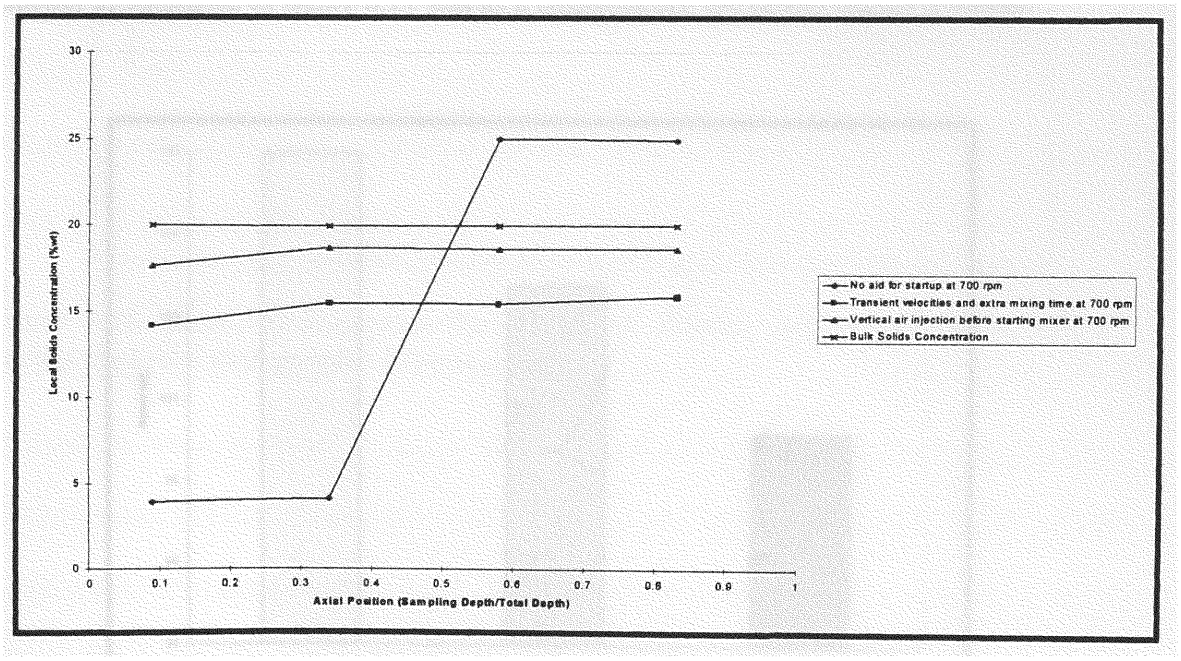


**Figure 6.23. Three local solids concentration obtained with a 0.01524 m (6 in.) diameter impeller agitating the slurry with a bulk concentration of 20 % weight. The two profiles that indicate homogeneity were obtained utilizing two different startup-aid techniques. The non-homogeneous profile was obtained with no start-up help.**

Figure 6.24 presents the power consumption at startup for the three cases. The startup with the injection of air shows a considerable lower consumption of power, approximately 15 % less than the startup with no aid and 10 % less than when the rotational speed reaches 700 after the speed was progressively increased.

Figure 6.25 shows the development of the power consumption of the three cases from startup until the moment when the samples were taken. As expected, the power consumption for the two startup aid methods converged to the same value regardless of the great original difference, indicating the same level of suspension. On the other hand, the case in which no startup procedure was executed, the power consumed at the time of sampling is significantly higher: the concentration of solids is higher in the area of the impeller as indicated in figure 6.21 and as a consequence the power drawn is higher.

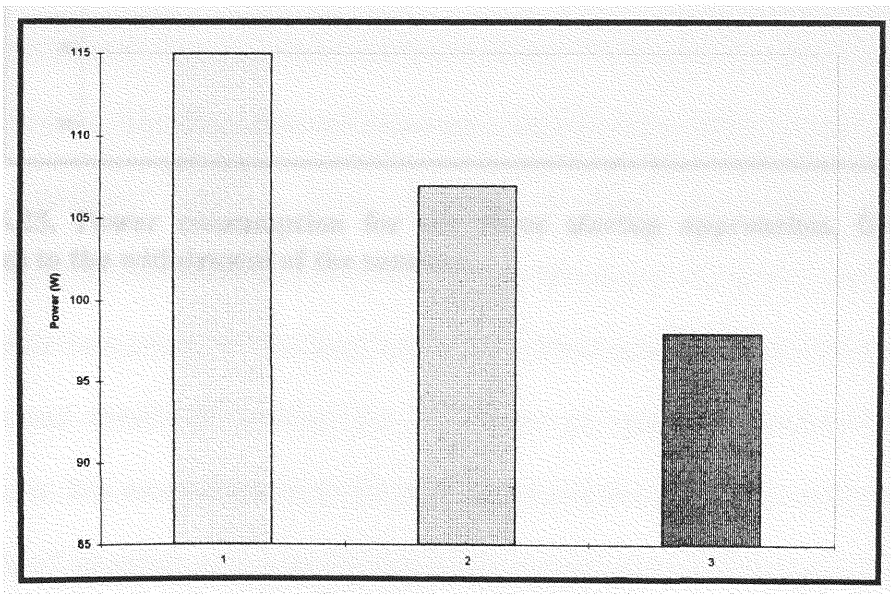


**Figure 6.24. Power consumption comparison between the two startup aid techniques (2. Gradual increase of rotational speed and prolonging the time of mixing, 3. Air-jet injection) and the non-aided startup (1).**

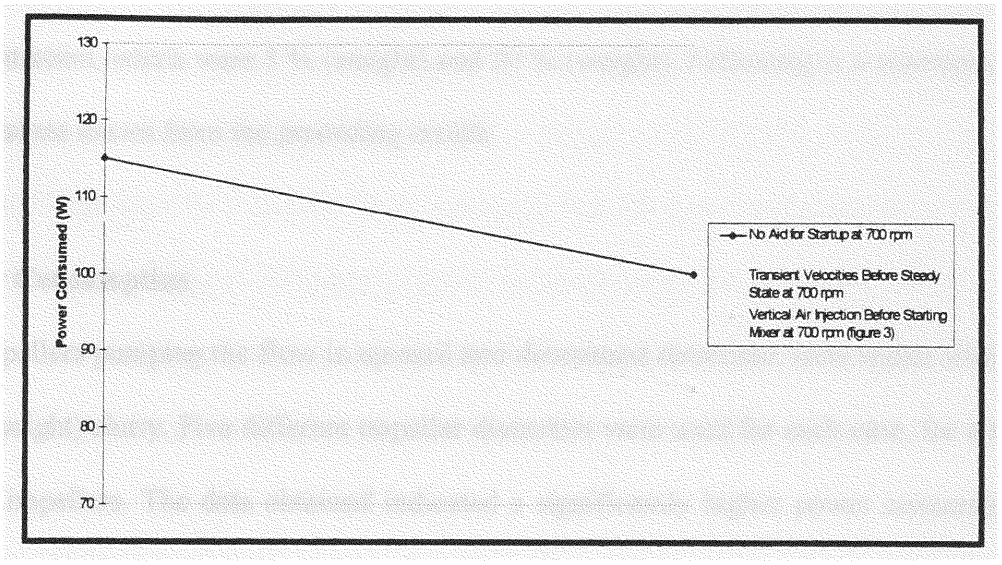


**Figure 6.25. Power consumption for the three startup approaches, from mixer startup up to the withdrawal of the samples.**

# 7. SUMMARY AND CONCLUSIONS

Two surrogate slurries with similar physical and chemical characteristics to those of DWPF melter feeds were prepared and characterized to study their behavior under different mixing conditions, the power consumed by the mixing system, and variables that might affect the accuracy of samples obtained at four different points along the vertical axis of the tank. The only difference between the slurries prepared was the solids concentration, which were 5 % (weight) and 20 % (weight). Following is a summary and conclusions drawn from the preceding results.

## Power Consumption

Impellers pumping the flow in upward and downward directions were tested with the 5 % (weight) slurry. Five different impeller diameters were used for each case, for a total of ten impellers. The data obtained indicated a significantly higher power consumption for the "up-flow" impellers when compared to their "down-flow" counterparts, corroborating the assertions of the literature. Since the quality of the mix was found to be the same for both cases, it was concluded that upward pumping impellers are not cost-effective and should be avoided for the design of agitation equipment for solids suspensions. The measurements of the power consumption of the five different sizes of "down-flow" impellers acting on the slurry with a solids concentration of 20 % (weight) indicated no differences in the power consumption versus rotational speed characteristic of the two slurries. Although the Reynolds number achieved with the 20 % (weight) slurry was decreased by a factor of 7.8 due to its higher viscosity (9cP), it still fell under

what most investigators consider a turbulent regime, suitable for the suspension of solids, and in which viscosity does not play an important role in power consumption. This evidence leads to conclude that the power consumption is not dependent of the concentration of this slurry in the 5 % (weight) to 20 % (weight) range, probably due to the relatively low viscosity that both slurries present under fully mixed conditions.

**Diameter of Sampling Tubes**

The effect of the diameter of the sampling tubes and the velocity of sampling on the concentration of the samples obtained was evaluated with the two slurries. The data obtained with the 5 % (weight) slurry indicated that the larger sampling tube diameter withdrawing the solids at the highest speeds provided samples with a concentration very close to the bulk concentration of the tank. This trend was less pronounced when the impeller rotational speed was increased and with the 20 % (weight) slurry.

**Sampling Velocity**

The velocity of sampling was not found to have any effect in the samples concentration under poor mixing conditions of the 20 % (weight) slurry, in which an important amount of solids were still settled on the bottom of the tank. Nevertheless, it is still recommended to use the largest diameter of sampling tube, collecting the material at the highest flow rate possible, in order to obtain a representative sample of the content of a slurry mixing tank. This conclusion is consistent with the results of previous investigators. The approach helped to produce enough turbulence in the transporting

tubes to prevent stratification. The nature of the slurry will determine the intensity of the flow rate required.

**Startup from Fully-Settled Conditions**

Startups from fully-settled conditions were tested for the 20 % (weight) slurry evaluating different techniques from mix quality and power consumption standpoints. These techniques were: 1) agitation of the slurry at a higher rotational speed, 2) progressive increase of impeller rotational speed and additional mixing time at terminal rotational velocity, and 3) injection of air to loosen the particles before starting to rotate the impeller.

Although all these techniques resulted in acceptable levels of homogeneity, the injection of air resulted in the least consumption of power at the moment of startup, making it a very viable solution in situations when the impeller needs an extra help to promote the suspension of the solids.

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