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MECHANICAL AGITATION IN THE CARBON-IN-PULP PROCESS

John A. Shaw<sup>\*</sup> & Robert J. McDonough<sup>\*\*</sup>

\* J.A. Shaw is Managing Director of Lightnin Mixers Pty. Ltd., Sydney

\*\* R.J. McDonough is Market Development Manager - Minerals Processing  
with Mixing Equipment Co. Inc., Rochester, N.Y., U.S.A.

1. Abstract
2. Introduction
  - a. Background
  - b. Agitator Function
  - c. Energy-Flow-Shear Relationships
  - d. Shear
  - e. Tip Speed
  - f. Shear Stress
  - g. Carbon Attrition
  - h. Experimental Techniques
  - i. Typical Flow Patterns
3. Open Impeller Agitation
  - a. Tank Design
    - (i) Agitator Selection
    - (ii) Mechanical Design
    - (iii) Start-up in Settled Solids
    - (iv) V-Belt Drive
4. Draft Tube Circulators
  - a. Draft Tube Diameter
  - b. Design Approach
    - (i) System Head vs Flow
  - c. Impeller Head vs Flow
    - (i) Impeller Behaviour
  - d. Power Requirements
  - e. Resuspension
5. Summary

## ABSTRACT

Mechanical agitation is the preferred system for Carbon-in-Pulp processes. Agitator design requires an understanding of the relationships between energy, shear and flow and the nature of shear rates and shear stress distribution in an agitated tank. Laser techniques are useful when studying these phenomena.

The alternatives are open impeller agitation or draft tube circulation. Each has advantages and disadvantages but both require careful choice of impeller and system geometry. Overall the draft tube circulator is the more attractive in many cases.

Open impeller agitators are designed in accordance with established procedures for mineral slurries taking due account of shear effects in tanks containing carbon. Preferred tank proportions are 1:1.

Circulator design methods were developed in the alumina industry before being applied to CIP processes. Optimum draft tube diameter is critical to power requirements. The overall system must be designed as an axial flow pump operating in a closed circuit. Such a system exhibits quite characteristic behaviour which requires an impeller of varying pitch angle and aerofoil section for best stability and efficiency. Tank height may be 2 or more times diameter.

Resuspension of settled solids can be assisted by a series of slots in the draft tube. These slots have the secondary effect of stabilizing hydraulic behaviour and reducing installed power requirements.

Overall, draft tube circulator systems will be the higher capital cost but can be 50%-70% lower in power costs relative to open impeller systems. Directionally, draft tube circulators, if well designed, can be expected to reduce carbon attrition.

## MECHANICAL AGITATION IN THE CARBON-IN-PULP PROCESS

### INTRODUCTION

#### BACKGROUND

The principles of the Carbon-in-Pulp (CIP) process have been known for many years but it is only in the last 8 to 10 years that gold prices have justified the commercial application of the process. One of the first installations was in 1973 at Homestake in U.S.A. and in the late 1970's this was followed by a number of applications in South Africa. In recent years there has been a resurgence of interest in U.S.A. and several Australian developments.

The CIP process requires careful design of agitation for firstly dissolution and secondly the carbon adsorption step. The same basic considerations apply to both operations. The traditional air-lift or pachuca system used in early minerals processing systems is not acceptable because of rising energy and capital costs, so mechanical agitation must be used. This is usually either open impeller agitation or draft tube circulation (although there were some early experiments with rakes and down pumping impellers in pachuca tubes).

This paper discusses some aspects of the design and economics of each system with emphasis on draft tube circulation which is the system currently attracting most interest.

So that design considerations for these systems can be understood more readily some agitation concepts and relationships will first be developed, and there will be a brief reference to experimental techniques for quantifying agitator behaviour.

#### AGITATOR FUNCTION

A mechanical agitator is a device for applying hydraulic energy to a system to achieve some desired process result. This process result should be defined in measurable quantitative terms such as degree of suspension, reaction rate, blend time, heat transfer rate etc. Qualitative terms such as "vigorous agitation", or "mild blending" are of little value as they can mean quite different things in different process contexts, e.g. an input of  $5 \text{ W/m}^3$  would be little short of violent in a tank of crude oil but 200 times that would be unusually mild for a penicillin fermenter. Equally meaningless are design procedures which assign to "degree of agitation" some arbitrary number on a scale of 1-10.

The hydraulic energy can appear as flow or as shear. Some processes require shear energy or, more usually, a combination of flow and shear. e.g. gas-liquid mass trans-

fer, emulsification, dispersion of agglomerates. Other processes are flow controlled, e.g. blending of miscible liquids, suspension of solids.

In minerals processing some operations are fully flow controlled as in solids suspension in a filter feed tank or a slurry storage tank. But some reactions also require shear, e.g. in many leach operations shear is required to strip the boundary layer of pregnant liquor from the ore particle. Couche, 1961, (1), develops the application of these concepts to mineral slurries.

#### ENERGY-FLOW-SHEAR RELATIONSHIPS

The split between flow energy and shear energy is a function of impeller type, diameter and speed, as well as the geometrical relationship between the impeller and the tank. Typical relationships are:

|                     |     |                                  |
|---------------------|-----|----------------------------------|
| $P \propto Q.H$     | (1) | $Q = \text{flow}$                |
| $P \propto N^3 D^5$ | (2) | $P = \text{power}$               |
| $Q \propto N D^3$   | (3) | $H = \text{head} = \text{shear}$ |
| $H \propto N^2 D^2$ | (4) | $D = \text{diameter}$            |

In flow controlled applications the shear component of input energy does no useful work and is dissipated as heat.

Directionally, a large impeller running slowly will convert more of the input energy into flow, or, produce the same flow with less input energy. But large high flow agitators are high capital cost. So in flow controlled operations one may be faced with a choice between an expensive agitator which does the same job with less power or a less expensive unit which needs more power because more of the input energy is wasted as shear. The decision should be made on economic grounds unless some other process consideration such as particle degradation points to the low shear option. However carefully designed high flow impellers can reduce agitator cost by operating at somewhat higher speeds without introducing a shear penalty. Note that power per dollar is not a valid criterion for choosing between agitators.

These relationships are illustrated in the early work done by Carpenter and Painter, 1955, (2), on optimising agitation in a purification process for zinc leach liquor.

#### SHEAR

Shear rate is the change in velocity with distance,  $\frac{dv}{dy}$  - the slope the velocity profile or "velocity gradient" as on Fig 1. It has the dimensions m/sec per m =  $\text{sec}^{-1}$ .

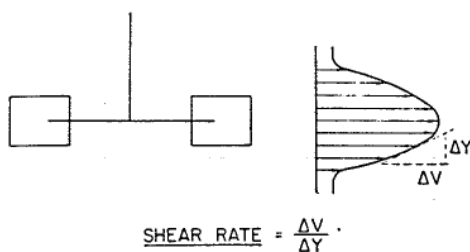


Fig. 1 - Velocity gradient - radial flow impellers.

Oldshue, 1970, (3), has shown that several different shear rates must be considered. These are maximum and average shear rate in the impeller zone and maximum and average shear rate in the bulk of the tank. Each of these may have to be considered in the design sequence particularly if the fluid or slurry is non-Newtonian. One must also consider the frequency of circulation through the high shear zone.

Impeller shear rate absolute values and the maximum to average ratio is related to impeller diameter, rotational speed and type. This is illustrated in Figs. 2 and 3. Tank shear rates are generally an order of magnitude less than those in the impeller zone but can be important in determining the bulk flow behaviour of non-Newtonian fluids.

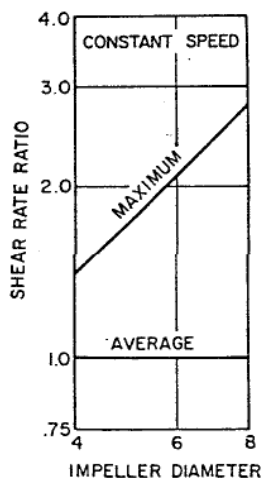


Fig 2. - Max & average shear rates.

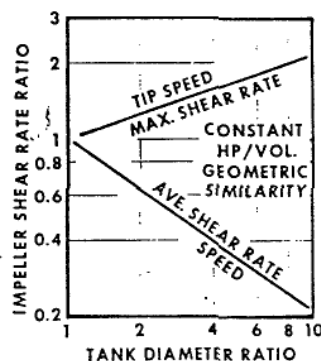


Fig. 3 - Shear rate ratio.

#### TIP SPEED

Impeller tip speed is often used as an indication of shear rate. This is an unsatisfactory parameter as the maximum impeller zone shear rate also depends on impeller

hydraulics and may vary by a factor of 3 or more for the same tip speed with different impellers.

Maximum shear does not occur at the interface between the impeller blade and the liquid. It results from velocity gradients in the liquid generated by the impeller and the point of maximum shear rate is usually some distance away from the impeller.

#### SHEAR STRESS

Particle damage is the result of shear stress which is the product of shear rate and viscosity. This can be a complex relationship as the distribution of velocity gradients and shear rates may be modified by the viscosity itself. Conversely the apparent viscosity of many materials is affected by the instantaneous shear rate and, sometimes, by recent shear history.

#### CARBON ATTRITION

Many elements in the carbon circuit contribute to the rate of attrition as do carbon grade and the characteristics of the particular ore.

It is not yet possible to quantify, say in g/tonne, the contribution to carbon loss due to the agitator system. But there is sufficient experimental and field evidence to permit some qualitative assertions.

Recent plant experience suggests that impeller zone shear rate, as measured by tip speed for a given impeller, is not as critical as previously supposed and higher tip speeds are now being used without adverse effects in U.S.A. This is probably because exposure to the high shear zone in the impeller region is quite brief at perhaps 1%-2% of the total residence time for a given particle in draft tube units.

It is quite possible that damage to carbon results, in part, from prolonged exposure to the lower shear stresses in the bulk of the tank particularly in the presence of abrasive ore particles. Open impeller attrition is worse than with draft tube systems because of the broad spectrum of velocity gradients in an open tank. Aerofoil impellers give better results than simple flat plate impellers in both systems because they produce fewer velocity gradients.

It is extremely difficult to make quantitative predictions on carbon attrition from laboratory tests because it is not possible to reproduce on a small scale the same balance of velocities, shear rates, exposure times and energy inputs as would prevail in a full size plant.

## EXPERIMENTAL TECHNIQUES

Historically, the measurement of flow and shear patterns in an agitated tank has been difficult. A recent development described by Weetman & Salzman, 1981, (4), uses a laser velocimeter technique. The tank is scanned by two intersecting laser beams. The associated computer equipment directly plots velocity vectors, integrates them to give flow through any given envelope and differentiates them to give average and maximum shear rates. Simultaneously the equipment can measure and display average and transient torques, bending moments and energy input. This technique is referred to several times in this paper. Fig. 4 is a diagrammatic arrangement of the facility.

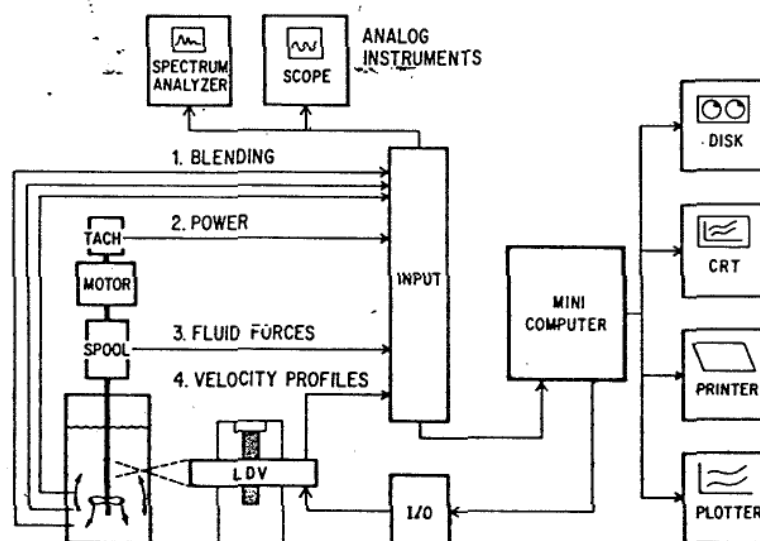


Fig. 4 - Integrated laser test facility.

## TYPICAL FLOW PATTERNS

Fig. 5 shows typical vectors for 2 types of axial flow impeller each mounted 2 diameters off the tank bottom.

'PBT' has 4 flat blades at  $45^\circ$ . It has a strong radial component and inefficient reverse flow patterns at a relatively short distance below the impeller.

'A-310' is a 3 bladed impeller. The blades are cambered and the pitch increases from tip to hub ("twisted"). The flow is strongly axial and continues right to the bottom of the tank. At the same power total flow is 70% above the PBT but speed is higher, reducing torque and cost.

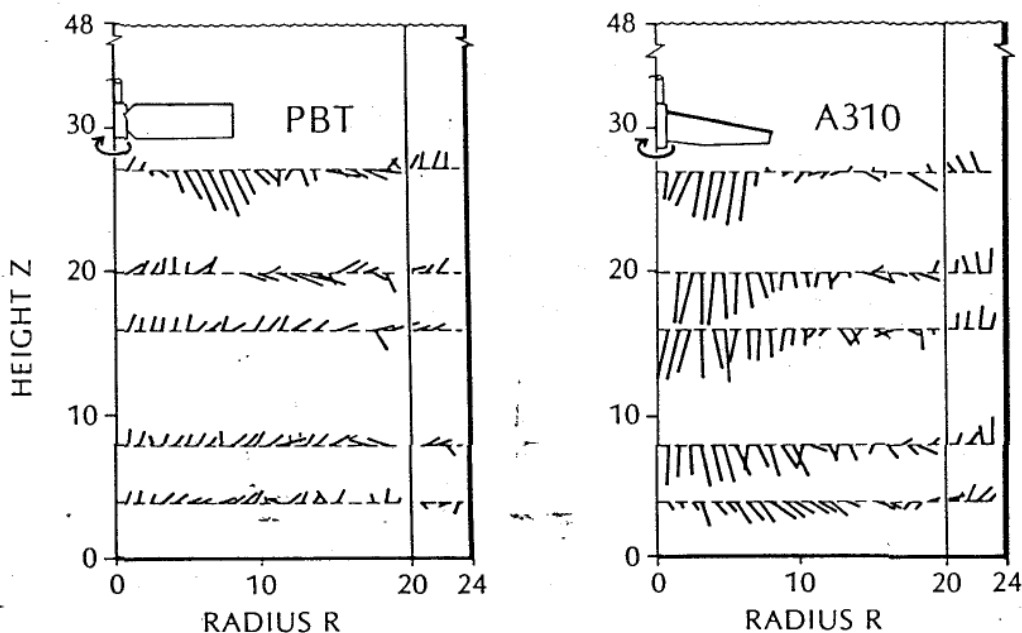


Fig. 5 - Velocity vectors for axial flow impellers.

#### OPEN IMPELLER AGITATION

Open impeller agitation systems have been widely used for CIP applications. They are probably the lowest overall capital cost option but these initial cost savings must be balanced against inherently higher power requirements and more severe carbon attrition.

#### TANK DESIGN

The best vessel shape is a cylindrical tank with height equal to or slightly less than its diameter. A simple flat bottom is satisfactory as at the power levels employed there will be negligible deposition of solids in the bottom corner of the tank.

Tank baffles will be required. Typically these will be 4, each with a width of around 5% to 8% of the tank diameter and standing off the tank wall about 1/3rd of their width.

Agitator support beams will be required and these must be of sufficient strength and stiffness to handle static and dynamic loads.

#### Agitator Selection

The agitator must give essentially uniform suspension of the ore and the carbon. This is a flow controlled operation. This requirement should be achieved with minimum power firstly because of the intrinsic energy savings and secondly because this must tend to minimize carbon attrition.

Slurries of relatively low solids content will be free settling and the free settling velocity of the largest particle will determine pumping capacity and power. At the high solids content usually applying in CIP applications settling is hindered and apparent viscosity is high. Selection criteria will be similar to those applying to blending high viscosity fluids.

The best selection is a high efficiency axial flow impeller with minimum radial flow and a uniform velocity profile to minimize velocity gradients. This type of impeller will have an aerofoil or cambered blade section and/or twisted blades to give the right angle of attack at various diameters. Generally simple flat plate impellers will give a less satisfactory flow pattern (see laser plots in Fig. 5) although they may be adequate for small tanks.

Power requirements tend to be higher than for a draft tube circulator so it is essential to minimize the proportion of that power appearing as shear energy. Additional power and/or major additional capital cost will be required if liquid height significantly exceeds diameter.

The design should be reviewed for performance under upset conditions, e.g. a fall in slurry solids content resulting in less hindered settling conditions and higher settling velocities. Or an increase in solids content may increase slurry specific gravity and viscosity to an extent which affects agitator performance.

#### Mechanical Design

Mechanically, the agitator must be designed as a system not as individual elements. A major factor is the magnitude of random fluid forces acting on the impeller. These are much larger than commonly supposed.

These forces cannot be predicted on purely theoretical grounds but must be calculated from actual measurements on models and full size units. Precise data on these forces can be obtained accurately in laser laboratory measurements, but scale-up correlations should be confirmed on instrumented full scale installations.

Shafts may be long and the bending moment on the speed reducer may be considerable. The reducer must be designed for this load as well as axial loads and power transmission. The capacity to handle all these additional loads is often not available in standard commercial gearboxes unless they are grossly oversized from the power transmission point of view.

Critical speed of the shaft-impeller assembly is a key factor and should be at least 25% above the operating speed if the impeller is fitted with stabilizing devices and 2.5 times operating speed if it is not. Shaft deflection can be a limiting factor in some applications including draft tube type equipment.

Note that high efficiency axial flow impellers can be located some distance off-bottom and this helps critical speed design. Of course if the tank must be operated at low liquid level this will dictate the location of the lower impeller and the shaft length.

#### Start-up in Settled Solids

If there is a power failure solids may settle and imbed the impeller. Unless the equipment is designed for this situation an attempt to start an imbedded impeller may result in damage to the impeller, shaft or speed reducer. Sometimes this damage is not immediately apparent but initiates a fatigue failure at some time in the future.

Whenever it is conceivable that the impeller can become imbedded the equipment should be designed accordingly. Methods of protecting against this include one or more of the following.

1. Torque limiting motor coupling.
2. Independant bearing support of the mixer shaft.
3. Heavy duty shaft and shaft bearing design.
4. Limit ring to hold shaft deflection to safe values.
5. Oversize impeller blades and connections.

Reverse rotation or air sparging is effective in some circumstances but the need to use these measures may not become apparent until the damage has been done unless other protection is also designed into the equipment.

#### V-Belt Drive

V-belt drive is sometimes specified for agitators on the grounds that it will "simplify future speed change". Indeed V-belts have become almost traditional in some sectors of the mineral industry. But it is important to recognize both the limitations and the value of this feature.

Power absorbed by an impeller is proportional to the cube of rotational speed and fluid forces also rise sharply as speed increases. So a relatively small speed increase can push a unit beyond its design capacity for power transmission, bending moment and possibly critical speed. Speed increase should be treated with caution. The immediate reaction to an excessive speed increase is motor overload and the typical field response of fitting a larger motor is not recommended!

Sometimes a major increase in specific gravity or a substantial reduction in required agitation level will justify a permanent speed reduction which can be made by changing the V-belt drive, e.g. the benefit of reduced shear rate may, in practice, justify loss of agitation performance. However if shear is not a consideration, trimming the impeller is often a simpler way of reducing agitation level.

In summary, V-belt drives should be specified only when there is a clear potential value for them; not simply because one is not sure what job has to be done.

#### DRAFT TUBE CIRCULATORS

Attempts have been made to provide mechanical circulation by installing a down pumping impeller in a pachuca tube. Although such an arrangement may work in a fashion and even give some power savings it is generally far from optimum because the draft tube is too small. Further, it is extremely difficult to suspend solids from the apex of the pachuca tank cone so either the tank bottom must be modified or supplementary agitation or pump out provided to deal with these solids.

Superficially a draft tube circulator resembles a retrofitted pachuca.

But draft tube, tank and impeller geometry have been optimised to give the required flow with a minimum of shear. Shear stress due to agitation is significantly less than with other systems. Circulator size and power are essentially independent of tank height so in tall tanks power requirements per volume can be reduced by 50-70% if an efficient impeller and draft tube are used. The additional cost of a circulator installation is generally justified by power savings and superior performance.

#### DRAFT TUBE DIAMETER

As Shaw, 1982, (5), notes a key parameter is draft tube diameter. This depends on two independent velocity requirements both of which must be satisfied.

The velocity down the draft tube,  $V_d$ , is set by the requirement to sweep the bottom and prevent solids deposition. This velocity depends on percent solids, solids specific gravity, particle size and slurry rheology.  $V_d$  must be determined experimentally as accurately as possible.

The rise velocity in the annulus,  $V_a$ , must be sufficient to suspend and circulate the pulp. It is a multiple of the settling velocity of the largest particle in a free settling system. Although typical gold ore slurries have hindered settling, the design criterion is usually free settling conditions which can prevail at low solids concentration due to process upset or start-up.

The optimum draft tube diameter,  $D$ , is when both velocities are just satisfied. In Fig. 6 the power required to produce these velocities are plotted against tube/tank diameter ratio. Power requirements are at a minimum where these two curves intersect. This can be shown to occur at

$$\frac{D}{T} = \sqrt{\frac{V_a}{V_a + V_d}} \quad \text{where } T = \text{tank diameter and } D = \text{draft tube diameter.}$$

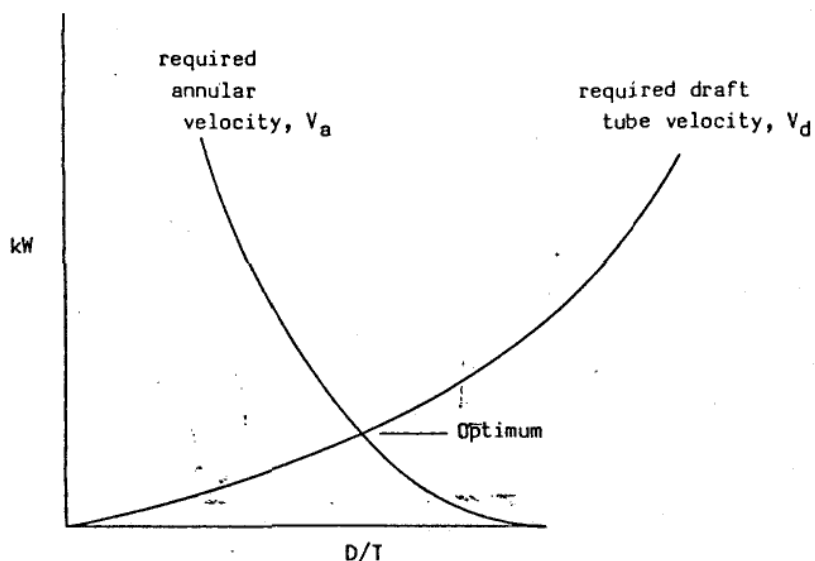


Fig. 6 - Energy optimization in the design of draft tube circulators.

#### DESIGN APPROACH

A circulation impeller in a draft tube behaves like an axial flow pump of high specific speed, and the draft tube circulator system must be designed on the same basis as any other pumped system. The operating point is where the head vs flow curve for the pump intersects the resistance head vs flow curve for the system. Fig. 7. But behaviour must also be examined for start-up and process upset conditions.

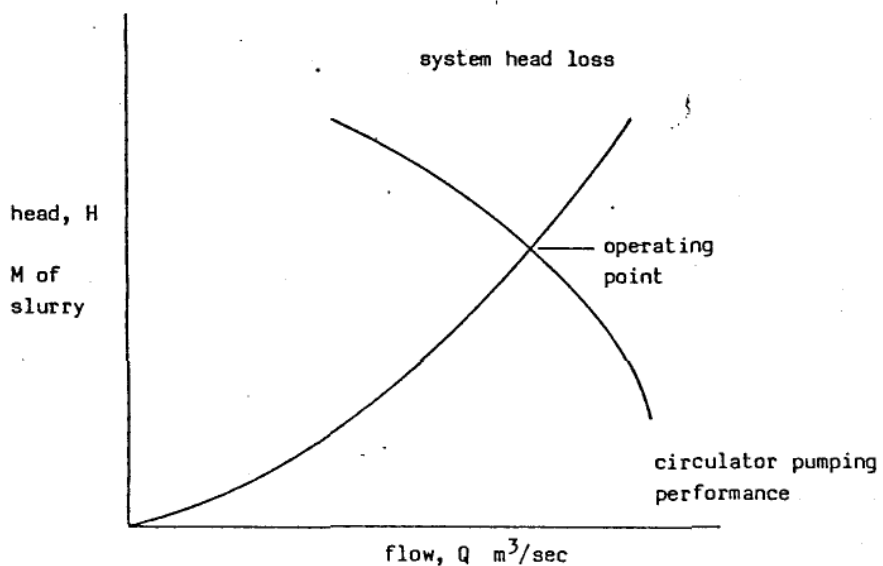


Fig. 7 - Pumping performance in a draft tube circulator.

### System Head vs Flow

The system head - resistance to flow - is made up of several components of head loss due to

1. draft tube entrance
2. draft tube internals
3. draft tube exit
4. reversal at tank bottom
5. friction loss in the draft tube and the annulus.

Friction losses in the annulus are infinitesimal and friction losses in the draft tube are negligible at viscosities below about 1000 cp. Careful design can minimize loss due to draft tube entrance and internals. Reversal losses are the most significant component although the placement of launders and in tank piping must be done with care to avoid introducing additional head loss at the top of the tank.

Determination of system head loss is very important and cannot be predicted accurately enough from hydraulic theory. It must be measured in fully instrumented test tanks of sufficient size to avoid serious Reynolds number discrepancies on scale-up.

It is convenient to express system head loss as a K-factor. i.e. a multiple of velocity heads in the draft tube. A draft tube of sophisticated design operating at a high Reynolds number can have a K of as low as 1.25, although 1.4 or 1.5 is typically assumed for higher solids slurries. A primitive draft tube may have a K-factor of 2.0 or more.

During a process upset the system head vs flow curve may be modified considerably. A partially obstructed draft tube exit will give a much steeper curve - K increases sharply. Low level can cause head loss across the draft tube entrance and superimpose a static head differential. Quite small differences in specific gravity between the draft tube and the annulus can also generate significant static head differentials.

Note that the head loss in the system (and the head generated by the impeller) is typically less than 100 mm so small static head effects can be very significant.

### IMPELLER HEAD VS FLOW

The impeller, or more accurately the impeller, draft tube and related draft tube internals, is an axial flow pump with specific characteristics. These characteristics are very important as they determine not only the hydraulic performance and power requirements of the system but also the tendency of that particular system to contribute to carbon attrition.

Of particular importance is the behaviour of the impeller under upset conditions when it is displaced significantly from its normal operating point.

These factors can best be explored by comparing the behaviour of 2 impellers:

Type A is a 4 bladed flat plate impeller, with blades angled at  $30^\circ$  to  $35^\circ$ , running in a simple draft tube with perhaps a rudimentary conical inlet.

Type B is an impeller with aerofoil section or cambered blades and a progressive increase in pitch angle from tip to hub. The draft tube has a flared entrance, inlet baffles, and stator vanes below the impeller. Tip recirculation is suppressed by a circumferential notch opposite the impeller, devised by Watson, (6).

Typical head vs flow curves for these two impellers are shown on Fig. 8. There are some significant differences.

#### Impeller Behaviour

With many respects the impeller blade can be regarded as analogous to an aeroplane wing. An important consideration is the "angle of attack". This is the angle between the centreline of the blade section and what it perceives to be the direction of flow of liquid across it.

As flow decreases the vertical velocity vector relative to the impeller decreases but the horizontal vector due to rotation remains constant. Fig. 9 shows how as flow decreases the flow vector relative to the impeller becomes more horizontal and the angle of attack of the blade increases. Different response to changing angle of attack explains the differences in impeller behaviour.

In the operating region the type B impeller has a much steeper curve than type A this is because at higher angles of attack an aerofoil generates more lift (i.e. head) than a flat plate.

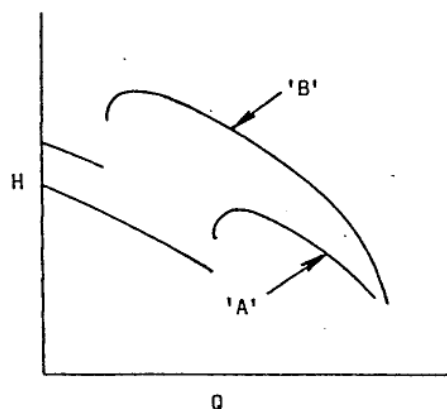


Fig. 8 - Impeller head vs flow characteristics.

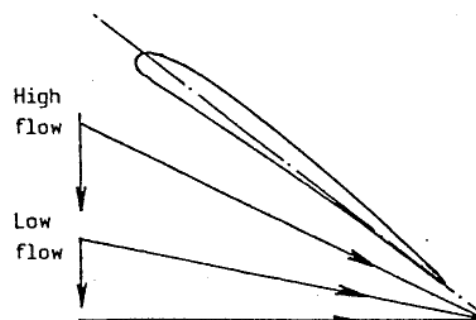


Fig. 9 - Angle of attack.

At lower flows there is a discontinuity in both curves. This is characteristic of all axial flow devices. It corresponds to the stall point of the aerofoil where boundary layer separation and turbulence occurs, lift is lost and the head falls off sharply. This occurs at relatively high flow with the flat plate impeller because (a) it has less tolerance to change in angle of attack and (b) the relative angle of attack varies along the length of the blade. Stall happens at relatively high flow and it is fairly unpredictable. The type B impeller has more tolerance to change in angle of attack because of its aerofoil section and blade twist. Stall occurs at lower flows and is more predictable.

The location of the stall discontinuity governs behaviour during upset conditions. Fig. 10a shows how during a process upset a type A impeller may fall into an H vs Q intersection in the discontinuous region from which it can never recover. This phenomenon has been observed with non-aerofoil type impellers in both alumina and CIP applications.

In contrast Fig. 10b shows how a type B system is stable over a wide operating range and is capable of recovering from even a major process upset.

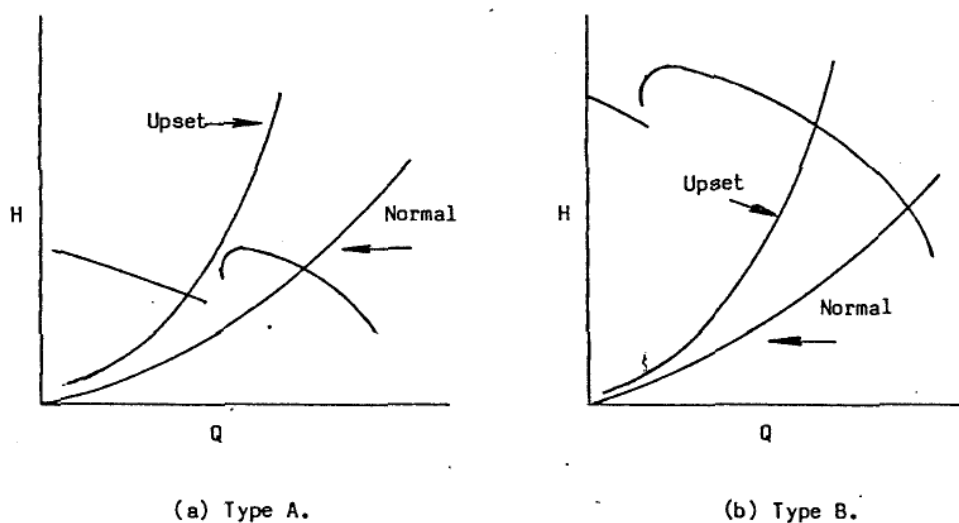


Fig. 10 - Upset conditions.

A typical sequence of events in CIP with a type A impeller is:

1. A temporary fall in solids concentration reduces hindering and increases settling velocity of large particles.
2. Large particles tend to accumulate in the bottom part of the annulus giving a slight increase in annulus specific gravity and a small static head differential.
3. The intersection moves up the H vs Q curve to a point with higher head and lower flow - and lower annulus velocity.

4. Even more solids start to accumulate and/or settle out and the head required goes even higher.
5. The intersection moves further up the curve, falls over the hump into the discontinuity resulting in significant loss of both head and flow. Annulus velocity drops sharply.
6. Solids are dumped on the tank bottom, the draft tube plugs and the system grinds to a halt.

#### POWER REQUIREMENTS

The power required to drive the circulator is given by the following equation.

$$P = \frac{9.788 Q H \rho}{\eta} \quad \text{where } P = \text{kW}$$

$Q = \text{flow in m}^3/\text{sec}$   
 $H = \text{head-m}$   
 $\rho = \text{slurry SG}$   
 $\eta = \text{hydraulic efficiency.}$

From this equation it can be derived that  $P \propto V_D^3$  so accurate experimental determination of  $V_D$  is very important.

#### RESUSPENSION

Landberg, 1970, (7) patented resuspension vents or slots in the lower part of the draft tube. If after a power failure the end of the draft tube is blocked off by settled solids, flow passes through the exposed section of slot at high velocity and progressively erodes and resuspends the solids.

These slots also have an important effect on the behaviour of the system. Because there is always some bypass through the slots a complete "shut-off" zero flow condition is never experienced. Effectively this chops off the unstable low flow area on the head vs flow curve. Indeed with type B impellers the discontinuity may be eliminated giving a stable characteristic over the whole operating range. (Fig. 11a.)

Similarly the power vs flow curve, which rises sharply at low flows is also chopped off. So instead of requiring a drive motor and speed reducer sized for around twice normal operating power, an over capacity of around 33% is sufficient. (Fig. 11b.)

Slurries which are hindered and settle quite slowly behave in a similar fashion to free settling slurries in a slotted system if they contain a proportion of larger particles which still concentrate fairly quickly. This is usually the case with CIP slurries.

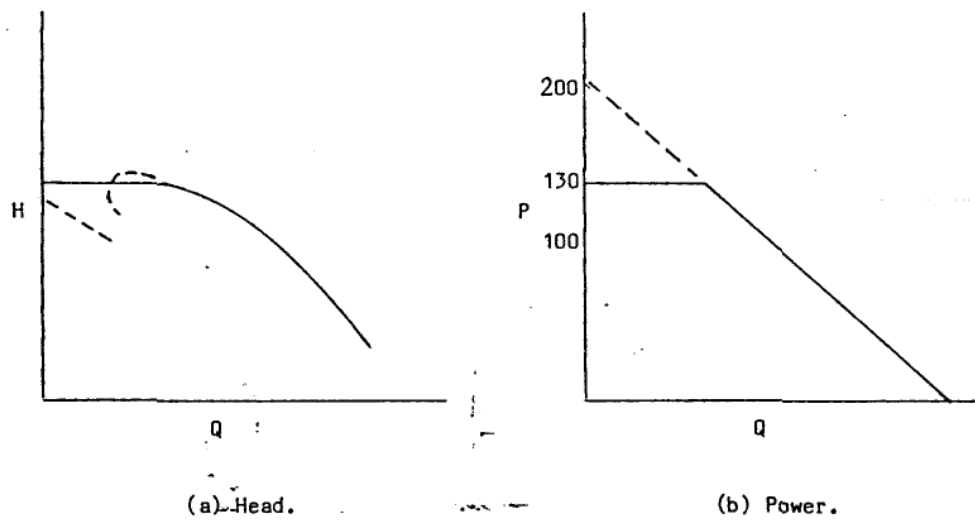


Fig. 11 - Slot bypass effect.

### SUMMARY

Either open impeller agitation or draft tube circulation can be used in CIP systems. Both can be optimised by informed design which maximizes the conversion of input energy into flow. The choice between the two is primarily economic, a trade-off among capital cost, power cost and carbon loss.

Tank geometry is a factor and thus may be set by site conditions, other process requirements or by the need to use existing vessels. Broadly, open impeller agitators are better suited to squat tanks whose height does not exceed the diameter. A draft tube circulator is most effective in a tank with a height/diameter ratio in the range 1.5 to 3.0.

Directionally a draft tube circulator system can be expected to cause less carbon attrition than an open impeller system. But this depends on equipment design and its significance may be masked by the effect of other components in the carbon circuit.

Given that tank geometry is optimized a properly designed draft tube circulator can reduce power requirements by 50-70% of open impeller levels. But this equipment is costly. The impeller can be very sophisticated and costly to make. The draft tube and associated structure may cost almost as much as the circulator. In itself large tank bottom fillets are needed although wall baffles are not.

Draft tube circulators can operate only with a full tank while open impellers can tolerate wide variation in liquid level. However this is usually not a factor in CIP systems.

The open impeller agitator is susceptible to "sanding in" in the event of power failure and must be designed to deal with this situation without damage. The draft tube circulator does not have this problem and with a slotted draft tube readily resuspends solids after a power outage.

Some of the more important differences are tabulated below.

|                                           | Draft Tube<br>Circulator | Open Impeller<br>Agitator |
|-------------------------------------------|--------------------------|---------------------------|
| Best tank shape<br>(T=diameter, Z=height) | $Z/T = 1.5$ to $3.0$     | $Z/T = 0.8$ to $1.0$      |
| Draft tube                                | yes                      | no                        |
| Wall baffles                              | no                       | yes                       |
| Filletted bottom                          | yes                      | no                        |
| Capital cost                              | high                     | lower                     |
| Power cost                                | lower                    | high                      |
| Torque                                    | lower                    | high                      |

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