AGITATORS IN ALUMINA INDUSTRY: THE DESIGN CRITERIA THAT MAKE YOUR PLANT WORK … OR NOT

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Abstract

In the alumina field as for many other industries, the most obvious duty of an agitator is to prevent solids from settling at the bottom of the tank.

Depending on process requirements, the mixer may also be able to keep the solids concentration homogeneous through the height of the tank, prevent short-circuits in the tank, facilitate the slurry transfer from one tank to another for continuous processes and to achieve sufficient heat transfer performance.

Most of the time, site-specific conditions (electrical power failures, different origins for bauxites, time between cleaning of the mixing tanks) may also influence weather the mixer you have designed will operate properly or not. These conditions are all difficult to predict at the time of design.

In some cases, these “hidden” parameters will have to be the dimensioning factors of the agitators. If not taken into account, this may lead to major failures of the equipment, thus generating bad operability of the whole plant.

This paper outlines the major design parameters to take into account and presents a methodology for defining an optimized mixing system for both precipitators and predesilicators in alumina refineries. This utilises real industrial experiences, in conjunction with numerical simulations of fluid dynamics (CFD) to highlight and explain important points.

Notation and units

Njs	Just off-bottom solids suspension speed	rpm
g	Shear rate	s^{-1}
t	Shear stress	Pa
ty	Yield stress	Pa
µ	Viscosity	Pa.s
n	Flow behaviour index	-
K	Consistency index	$Pa.s^n$
S45PBT4	S coefficient for Pitched 4 blades 45°	
Dm	Impeller diameter	m
C	Impeller clearance from bottom	m
H	Liquid height in the vessel	m
T	Diameter of the vessel	m
Re	Reynolds Number	-

1. Introduction

A great number of papers dealing with agitators in hydrometallurgy and alumina refining in particular focus essentially on solids suspension and homogeneity in the tanks.

Most of the time, they describe many methods to predict just off-bottom solids suspension speed (Njs: Zwietering (1) correlation) including many parameters such as the type of impeller used or the geometric parameters of the system.

In real cases, solids suspension is not the only parameter to focus on.

Other essential parameters, listed below, must also be taken into account:

- ability to prevent dead zones in the tank
- ability to extract properly the slurry without accumulation in the tanks
- ability to restart the agitators in case of electrical shutdowns
- ability to limit scaling in the tanks and preventing the accidental fall of blocks causing failures of the mixers

The aim of this paper is to describe an approach to evaluate these risks when designing agitators in this industry, focusing essentially on desilication and precipitation units.

2. Mixing of medium to high solids contents media

2.1 Rheological characterization of the slurry

The rheological properties of a slurry evolve along with solids concentration. For example, with 20% w/w of phosphate solids in water, dynamic viscosity may be as low as 5.10-3 Pa.S whereas with 68% solids, it may grow up to 5 Pa.S in industrial conditions around the impeller.

In a phosphate storage tank with 68% w/w solids, the agitator has to prevent dead zones from appearing in the tank which automatically imply that solids have to be homogeneously distributed all over the height of the tank. The aim of the agitator in this case is to agitate a viscous media rather than achieving conventional solids suspension.

In red mud storage tanks, yield stress up to 80Pa have been observed, requiring three times more mixing power to move the media compared to conventional mud concentration (450g/L), exhibiting a 5 Pa average yield stress.

Consequently, the rheology of the slurry is the first parameter to take into account and may be the main criteria that will decide the way to design the right agitator.

2.2 Rheology : fundamentals

For fluids undergoing laminar shear, the resistance deformation depends on the dynamic viscosity. For a linear velocity gradient arising due to the movement of one parallel plate over another, the relationship between shear rate g and the shear stress t, is given by:

$$\tau = \mu . \gamma$$

Newtonian fluids

For Newtonian fluids, μ does not depend on shear rate. This behaviour corresponds most of the time to low solids contents slurries (<40% w/w).

Pseudoplastics or shear thinning fluids

$$\mu_a = K . \gamma^{n-1}$$

n = flow behaviour index
K = consistency index

Generally, with common slurries, 0<n<1, which means that the apparent viscosity decreases along with shear rate (eg, $CaCO_3$ slurries up to 80% solids).

Plastic fluids

$$\tau = \tau_y + \mu . \gamma$$

These materials are characterized by a yield stress. In these cases, a minimum stress is required to break down the media structure sufficiently before any movement will occur (eg, red mud at high concentrations).

Plastic fluids with shear thinning

$$\tau = \tau_y + K . \gamma^n \quad \text{Herschel-Buckley model}$$

This case is common (Bauxite from Weipa, phosphate slurries, magnetite, etc.) and is the hardest case to cope with.

In addition, with thixotropic fluids, the apparent viscosity reduces with time as the material is sheared at a constant shear rate.

It is clear that, depending on the rheological behaviour of the slurry, the design method will not be the same.

2.4 Mixer design for solids suspension / low viscosities

The principal method widely used is the Zwietering's correlation, which determines the just-off bottom solids suspension speed.

$$N_{js} = \frac{S . \upsilon^{0.1} . d^{0.2} . \left(g \frac{\Delta_\rho}{\rho_l} \right)^{0.45} . X^{0.13}}{D_m^{0.85}}$$

J.Wu, Y.Zhu and L.Pullum (2) have shown that there was a link between the impeller capacity and the S parameter in this correlation that can be expressed as

$$N_q . S = K_z$$

where Nq is the impeller flow number and Kz is a non-dimensional constant independent of the solid-liquid material property and impeller geometry. See Table 1 for data dealing with different impellers used in solids suspension. Empirical data on S and its variation with impeller/tank geometry has been given by Nienow (3) amongst many others.

From this table, for a given slurry, at constant geometric parameters (Dm,/T, C/T, H/T), the power needed by an HPM 20 3 is far less to achieve solids suspension compared to 45PBT4 which needs +100% more power for the same result.

2.5 Geometry optimization of the mixer

In order to achieve full solids suspension, depending on H/T ratios, the number of impellers on the agitator shaft may increase. This basic design criteria is good for both Newtonian slurries and non-Newtonian slurries, discussed in 2.6.

For instance, when H/T < 1, a single propeller may be enough to cope with uniform solids suspension. When H/T >1, in the case of a single impeller configuration, there is a two circulation loop system that appears in the tank (see Figure 1). In this case, if the velocities in the upper circulation loop are too low to keep solids in suspension, there will be a large solids concentration gradient along with vessel height.

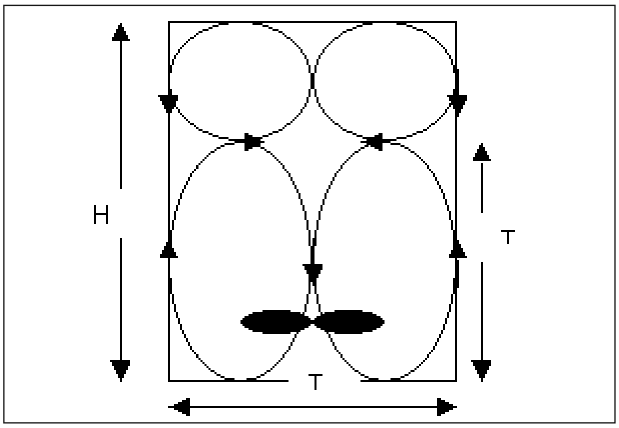

Figure 1.

It is important to note that the loops in the vessel may be different from what is described in Figure 1 when slurry is fed at the top, engendering a modification of velocity fields with a positive effect.

In order to limit the power needed for full solids suspension, it is necessary to have at least one more impeller per cylindrical part of the vessel, where height is equal to T (see Figure 2).

Table 1. Impeller specification, flow number, power number and S parameter, C/T = 1/3, D_m/T = 0.41, T = 0.41m

Impeller	Full name	Flow pattern	Number of blades	Blade width W/D	Flow number Nq	Power number P_0	S / S_{45PBT4}
HPM 20 3	MiltonRoyMixing Propeller	Axial	3	NA	0.66	0.41	1.15
HPM 10 3	MiltonRoyMixing Propeller	Axial	3	NA	0.48	0.221	1.58
31 T	MiltonRoyMixing Propeller	Axial	3	NA	0.37	0.11	2.05
30PBT4	30° pitched 4-bladed Turbine	Axial Radial	4	1/5	0.58	0.56	1.31
45PBT4	45° pitched 4-bladed Turbine	Axial Radial	4	1/5	0.76	1.22	1.00

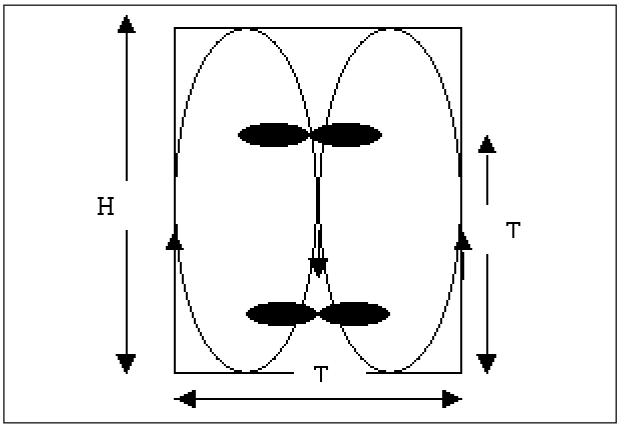

Figure 2.

In addition, the lower propeller position has to be chosen as adequately as possible, so as to limit solid scaling in the vicinity of the bottom of the tank.

Depending on the application, very small clearances can be required (see Figure 3), especially when high scaling conditions are present (eg, first alumina precipitators of the precipitating line, after an attack of bauxite under pressure, exhibiting large oversaturation of alumina in the caustic media).

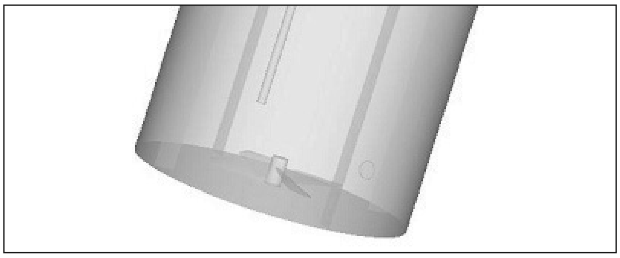

Figure 3.

2.6 Mixer design for solids suspension / shear thinning + plastic fluids

The reduction of Reynolds number below 300 in a stirred vessel partially suppresses the axial component in the discharge of axial and mixed-flow impellers. In this case, the overall pattern of these impellers is then comparable to a radial impeller. The consequence is that the optimum Dm/T in this case increases from 0.35 in water-like viscosity to around 0.5 in this range of Reynolds number. It must be noted that Zwietering's correlation and corresponding S parameter is not valid anymore in this Reynolds range.

M.Zirnsak and D.Stegink (4) highlighted the main problem that may be encountered with shear thinning slurries in predesilication tanks (Eurallumina refinery /Italy): the build-up of thick slurry in the tank up to the whole volume.

K.Wichterle and O.Wein (5) have investigated the origin of such a media behaviour pointing out what is called the "cavern theory", using X-ray flow visualization techniques with xanthan gum solutions (yield stress from 5.8 to 14.2 Pa).

With shear thinning fluids, high apparent viscosities are observed in low-shear areas. Consequently, because of a high resistance to fluid motion, poor mixing may occur.

With a yield stress, the phenomenon has greatest effects: a cavern is formed around the impeller, outside of which the media is stagnant (see Figure 4). In this case, the shear stress engendered by the impeller is lower than the yield stress of the slurry in the outer area of the cavern, usually leading to fast build-up in the tank.

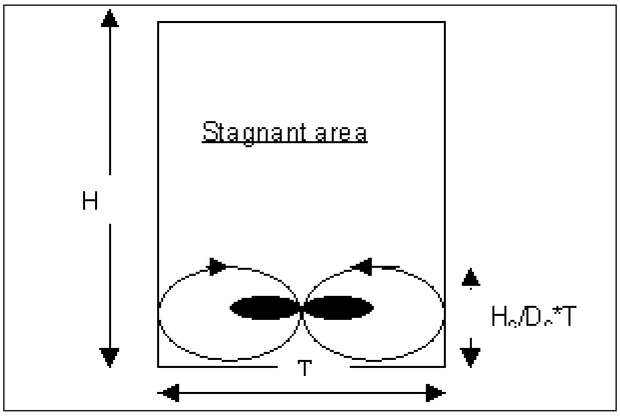

Figure 4.

In the transitional flow regime (between turbulent and laminar), the cavern size may be represented by a cylinder (diameter = Dc), where height (Hc) may be different depending on the type of impeller used (Elson(6)):

- Hc/Dc= 0.40 for disc turbines
- Hc/Dc= 0.45 for two blades paddles
- Hc/Dc= 0.55 for pitched blades turbines/ propellers

A model has been developed using the Yield stress Reynolds number:

$$\mathrm{Re}_y = \frac{\rho . N^2 D_m^2}{\tau_y}$$

which can be written: $\left(\frac{D_c}{D_m}\right)^3 \alpha \; P_0 \, \mathrm{Re}_y$

For a given vessel, with a constant Dm, this means that:

$$P \; \alpha \; \tau_y^{3/2}$$

If the yield stress is multiplied by 2, the power needed to reach the wall of the vessel is multiplied by around 2.8.

Once the cavern has reached the vessel wall, $H_c \; \alpha \; N^{0.87}$.

With pseudoplastic fluids, there is also a highly mixed pseudo-cavern surrounding the impeller which itself is surrounded by relatively slow moving fluid. An analogy can be made between τ_y and the minimum shear stress t_{min} for effective mixing with τ_{min}=K.$\gamma_{min}{}^{n}$ assuming a power law model for the fluid's rheology.

From these results, it is obvious that the location and the number of the impellers all over the height of the vessel will be a key point in the achievement of full mixing of these tanks (see Figure 5).

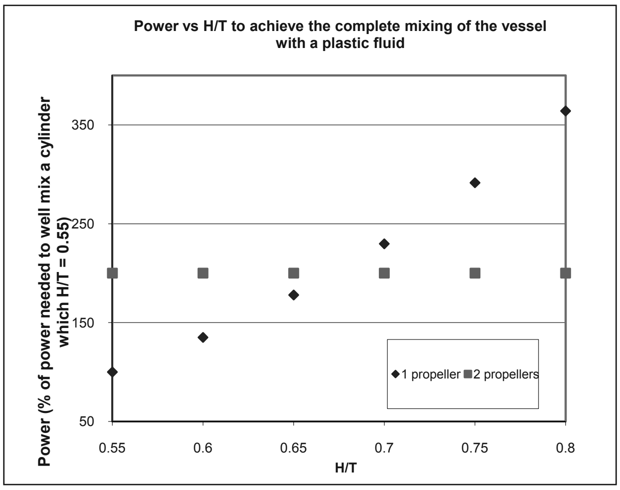

Figure 5.

It is clear that accurate data for the rheology are needed to obtain full confidence in the design of the agitator. The viscometer used for this has to cope eventually with slurries having a particle top size of a few millimetres.

Another way to achieve these measurements is to use an agitator equipped with a precise torque measurement on its shaft and then to plot power vs. rotational speed. With the Power number equal to the f(Re) curve for the impeller used, it is possible to estimate the apparent Reynolds number during the test in slurry and then to calculate the apparent viscosity corresponding to different speed. By knowing the Otto and Metzner constant ks (7) of the impeller (γ = ks.N), it is then possible to have the final curve of $\mu_a=f(\gamma)$.

In addition, a CFD simulation can be very helpful once the rheological parameters are known.

Figure 6 shows a predesilicator of 2000m^3, with bauxite exhibiting a max yield stress of 12 Pa.

Note the low distance between the lower propeller and the bottom of the tank, enabling a full cleaning of the suctions of the extraction pumps.

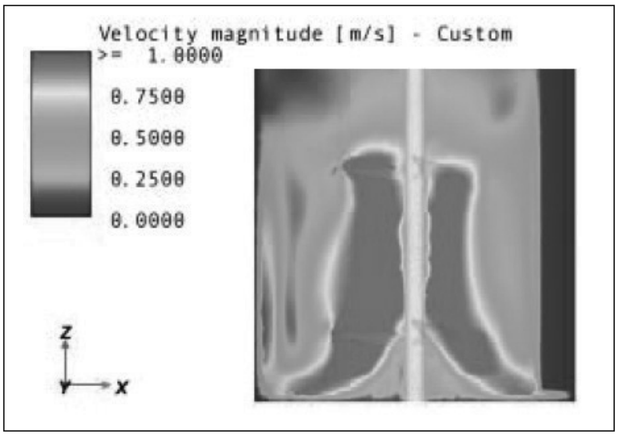

Figure 6.

3. Inlet-outlet design

The way the slurry leaves the vessel will determine how to define the mixer, and this can be achieved in several ways.

3.1 Outlet by pump at the bottom of the vessel

This can involve a single impeller if no homogeneity is required, or multiple impellers depending on the shape of the vessel and the homogeneity needed.

Advantages include:
- less risk of accumulating large solid particles in the vessel.

Drawbacks include
- use of a pump (investment + operating costs)

3.2 Simple overflow

This can involve single or multiple impeller configurations, depending on the vessel height to diameter ratio. Perfect homogeneity is also usually required. For long residence times, the liquid height in the overflow may be very small compared to the other dimensions on the vessel. In this case, there is a great risk for solids to accumulate in the tank as there is always less solids contained near the liquid surface due to a layer effect. To prevent such a situation, the dimensions of the outlet have to be calculated so as to assure a sufficient velocity in this area.

Advantages:
- low cost
- easy to install.

Drawbacks:
- high risk of solids accumulation.

3.3 Dip pipe outlet

Single or multiple impellers may be required depending on the shape of the vessel. For high solids concentration media, homogeneity is required so as to limit specific gravity differences between the top and the bottom of the tank, which could cause direct overflow of the vessel contents into the outlet pipe. The upward velocity in the pipe has to be as high as possible to prevent solids from settling in it (eg, 0.7 to 1 m/s in alumina precipitators).

Advantages:
- low risk of accumulating large solids particles in the vessel (as with the outlet by pumping).

Drawbacks:
- only non-fouling products can be handled by such a method.
- Use of an airlift in the pipe may be needed to control differences between the level of the slurry in the vessel and the level in the outlet pipe (prevent direct overflow). Visual or automatic control is important to check the smooth running of this system.

Figure 7 shows an extraction through a dip pipe in a 4500m^3 hydrate precipitator. A maximum 1.5 to 3% difference in the slurry density between the top and bottom of the tank is required to enable a perfect transfer from one tank to another. If this difference is any greater, or a stoppage of the troughout flow occurs, overflow of the slurry directly into the outlet may happen.

Figure 7.

3.4 Comparison between overflow and dip pipe

To compare a simple overflow outlet to dip pipe outlet, a test was done in a 2 m^3 vessel initially filled with water and a known quantity of solids. The mixer rotational speed has been set visually so as to have complete homogeneity in the vessel. At t = 0h, a known continuous flow rate (1.7 m^3/hr) of water is introduced in the vessel. The liquid + solids flowrate coming out of the vessel is filtered and solids are counted.

Three tests have been done:

- Power = 100 % + outlet by simple overflow
- Power = 100 % + outlet through a dip pipe (upward velocity in the pipe = 1 m/s)
- Power = 300 % (increase of initial speed by 43%) + outlet by simple overflow

The results of the amount of solids filtered along with time for these three tests have been compared to the theoretical values (assuming there is no solids accumulation in the vessel) (see Figure 8).

From the results, it is obvious that the configuration using the dip pipe is far more efficient in terms of efficiency to extract solids from the vessel compared to simple overflow.

Moreover, even by increasing the rotational speed by 43% (300% of initial power), there were still poor performances of the configuration using the simple overflow compared with lower rotational speed using the dip pipe.

4.0 Restarting of the agitator after shut-down

When an agitator is stopped (eg, because of a power failure on the grid), solids settle at the bottom of the tank. When the agitator is restarted, depending on how much solids have settled during the stoppage, it may or may not be able to turn.

Tests have been done on alumina precipitators (4500 m^3, 800g/L) to check the variations of the power needed by the mixer just after it has been restarted in settled solids for different durations of stoppages. The current, power, power factor and voltage have been measured every second and the results for two tests are indicated in Figures 9 and 10.

Several tests have been done with many stoppage times (20s, 1 min, 3min, 7 min, 10 min, 30 min). From Figure 9, it can be seen that there is a noticeable increase in the power used by the agitator during about 30 seconds after restarting.

In fact, after such a duration, the viscosity of the media around the lowest propeller has increased along with stoppage time, changing the flow regime from turbulent during normal running into transitional flow. The increase of power during this period is directly linked to the increase of the Power number P_0 along with Reynolds number.

Theoretically, whatever the case studied, this increase of power can be calculated by knowing the characteristics of the viscosity of the settled slurry. Chemical reactions, thixotropy or other detrimental parameters (depending on process conditions before and during the stoppage) increase considerably the uncertainty of such a calculation.

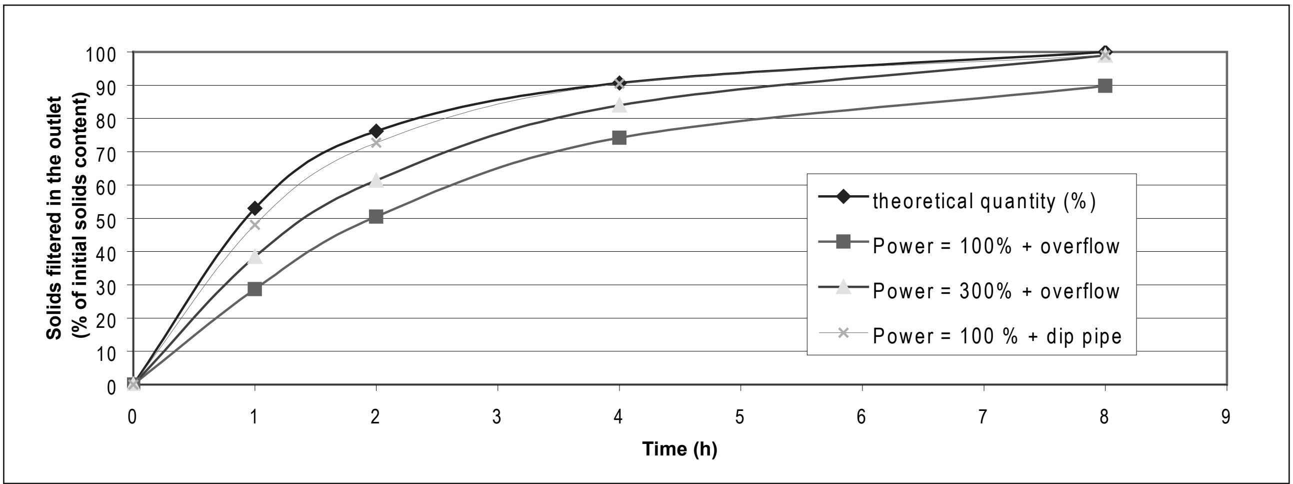

Figure 8.

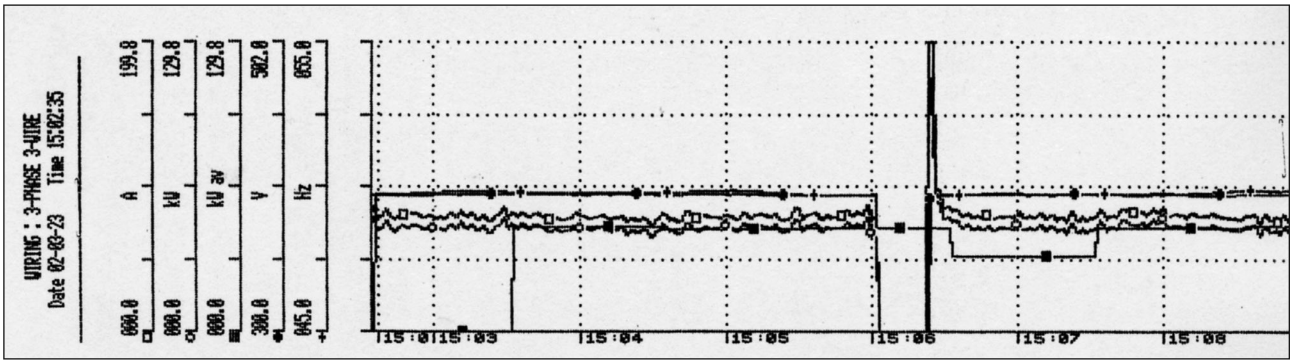

Figure 9. Restart of the agitator after 20 s

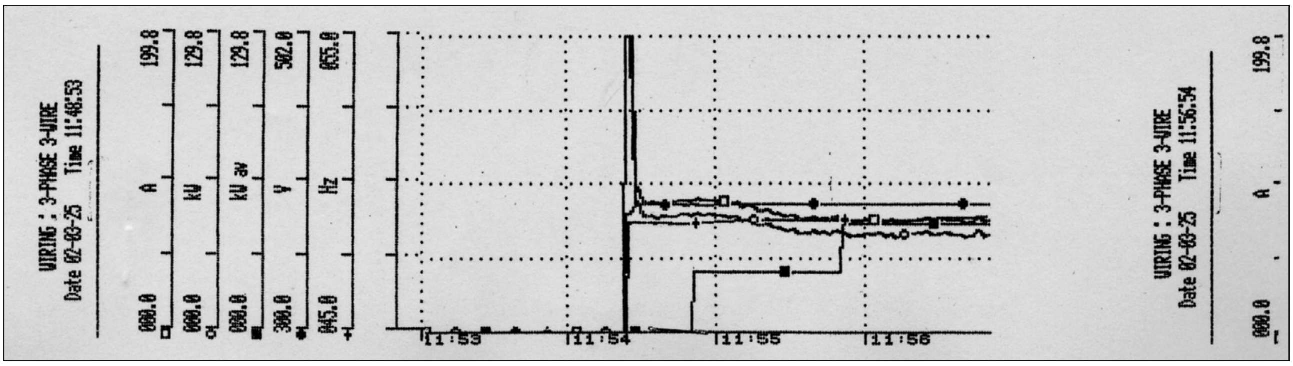

Figure 10. Restart of the agitator after 30 minutes

Depending on the risks of such power failures happening, additional power margins of up to 35% could be needed to face these exceptional conditions.

Moreover, particular attention should be brought to the mechanical strength of the agitator components located in settled media. It can be noted that, because the starting torque of a low voltage motor is generally twice the nominal motor torque, this extra torque can be applied on one impeller only, which is very important when multistage impeller configurations are used.

Conclusions

Many parameters which are not generally included in agitator data sheets at time of design can have catastrophic consequences if they are not taken into account for engineering phases.

In particular, maximum yield stress values for red mud and bauxite slurries, eventual electrical power failures (with average duration) and precise design of inlet-outlet of slurries have to be discussed together with agitator manufacturers, so as to prevent problems appearing during commissioning of these units.

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