Basic design criteria for agitators in alumina production

Detlef Klatt and Janin Klatt-Eberle
1.Owner and CEO of STC-Engineering GmbH,
2. General Manager of STC-Engineering GmbH
Corresponding author: j.klatt-eberle@stc-engineering.de

Abstract

The cascade of agglomerators, precipitation growth tanks, and other mixed process vessels in a Bayer alumina refinery need a uniform suspension to assure a reliable and efficient process. Basic design rules for agitators for these process steps will be given with respect to solids concentration, solids size distribution and tank geometry. These design rules have been collected, evaluated and summarized from different alumina production installations worldwide. Two principle modes of design end up either with low investment cost or low operational cost. Both possibilities will be compared with their advantages and disadvantages. Local flow velocities, scaling behavior and natural overflow without air support will be examined.

Keywords: Alumina; Bayer; mixing; agitator; precipitation; agglomeration; suspension.

1. Mixing

Mixing is one of the main process tasks in the production of alumina. In the Red Area as well as on the White Area there are many different mixing tasks to be solved. Depending on size of the plant, there are sometimes 100 mixers installed. Mixing has a strong influence on alumina quality and production costs and is involved in the following process steps.

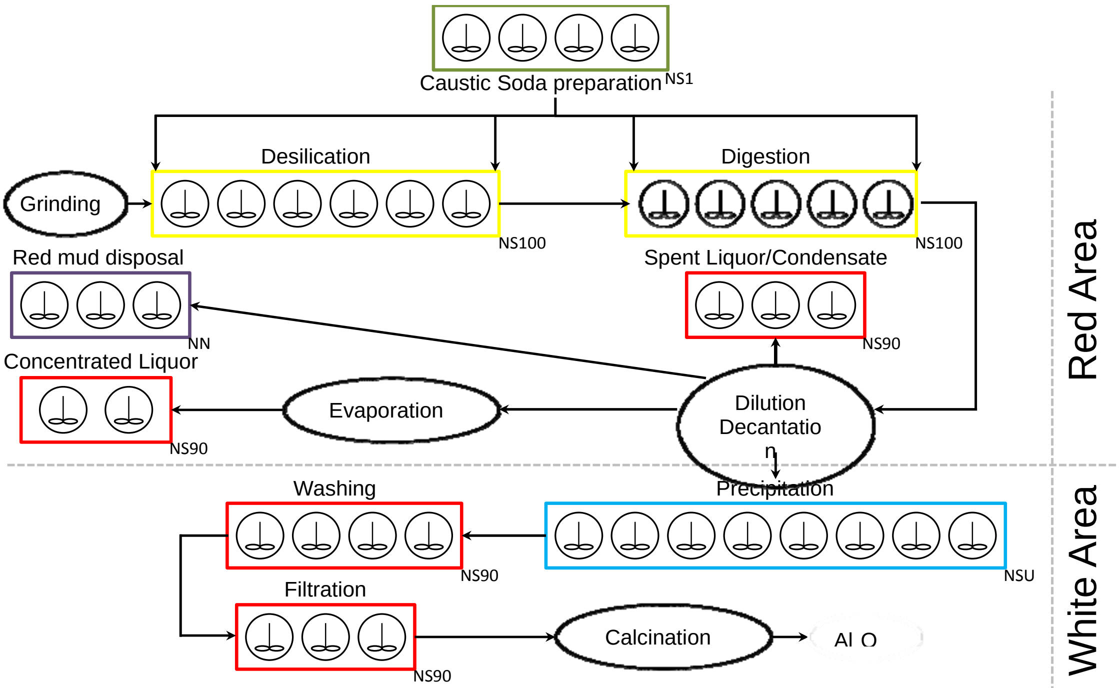
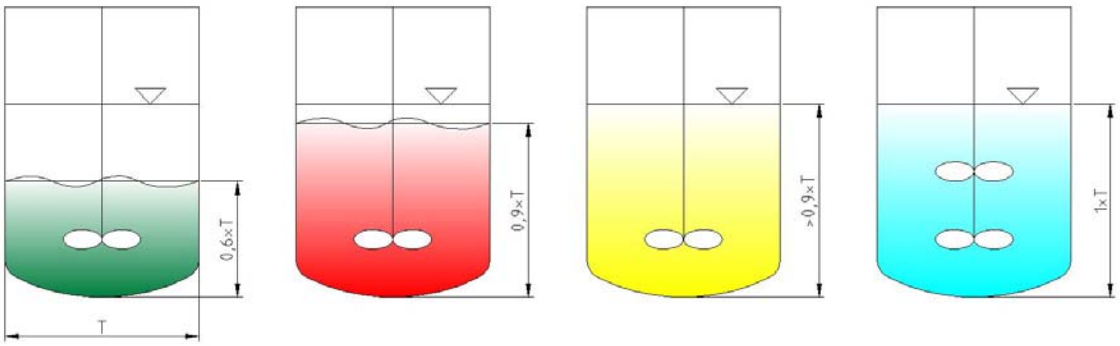


Figure 1. Mixing tasks in an alumina plant.

Fig 1 provides an overview of the mixing tasks which are involved. It is obvious that the majority deal with suspension. Red mud disposal (NN) is an exceptional case of non-Newtonian mixing for fluidization, but not dealt with in this paper.

Uniform suspension, called here “NSU”, is a special case of high quality suspension, required to manage the natural overflow of the precipitator cascade and for wall scaling reduction or suppression. In most of these cases the mixing task is suspension. The mixing qualities in alumina refineries are defined as follows:



N=NS1	N=NS90	N=NS100	N=NSU
Preparation	Liquor, washing, filtration	Desilication, digestion	Precipitation
Complete	90 % height	Homogenous	Uniform

Figure 2. Basic cases for suspension quality.

N shaft speed (rpm)
NSj required shaft speed (rpm)

- Principally, the quality of suspension is distinguished by three different basic types; Complete suspension is the simplest task where a single particle is not allowed to settle longer than 1 second at the vessel bottom. This case means further that the cloud of particles can be lifted up to 60 % of the filling level. The speed which is necessary to lift particles in this manner can be calculated as a function of the settling velocity of the particles which will be given in detail later on.
- 90 % height suspension means that the particles or the cloud of particles is lifted up to 90 % of the filling level.
- In a homogenous suspension the particles are distributed in 100% of the liquid, up to the liquid surface.
- Further, there is a suspension type which is even better than homogenous – this is the so called “uniform suspension” (NSU) which is applicable in the precipitators. The uniform suspension is evaluated in its quality in detail in the following section.

Quality here means the solids distribution over vessel height, and is defined by statistical variance (Sigma).

The statistical variance of the solids distribution means suspension quality is defined by a statistical measurable standard.

$$\sigma = \sqrt{\frac{1}{n}\sum_{1}^{n}\left(\frac{Cv}{Cvav}-1\right)^{2}} \qquad (1)$$

Cv local concentration
Cvav average concentration
n numbers of measuring locations

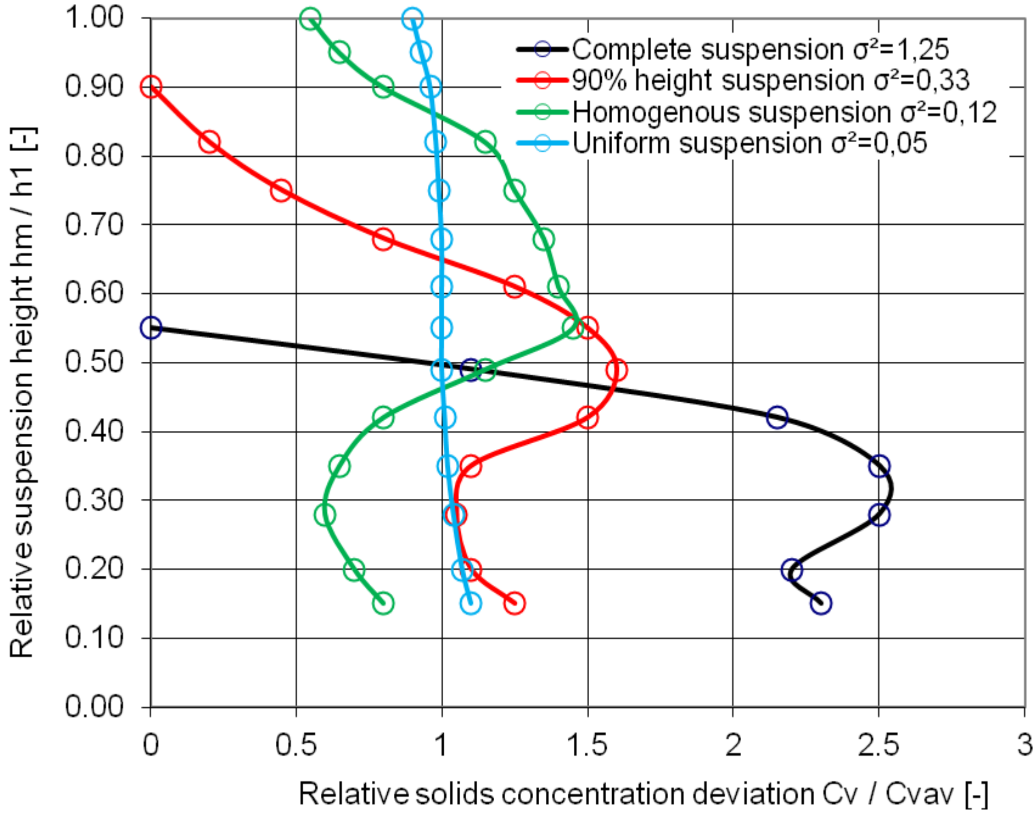
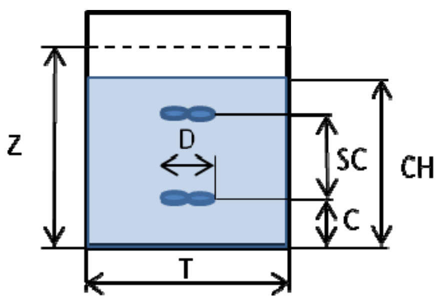


Figure 3. Solids distribution over precipitation vessel height [1].

A simple technique to visualize suspension functions is the mixing in transparent vessels to show the quality of mixing, but also to quantify parameters to describe their influence. Tests using this method were conducted in vessels with a height/diameter ratio of 1 and 1.75.

The main criteria for influencing the cloud height and quality of a suspension are:
- shaft speed
- impeller diameter
- impeller stages

CH	cloud height
T	tank diameter
Z	filling height
D	impeller diameter
C	impeller bottom distance



SC	stage clearance

Figure 4. Suspension definition by cloud height.

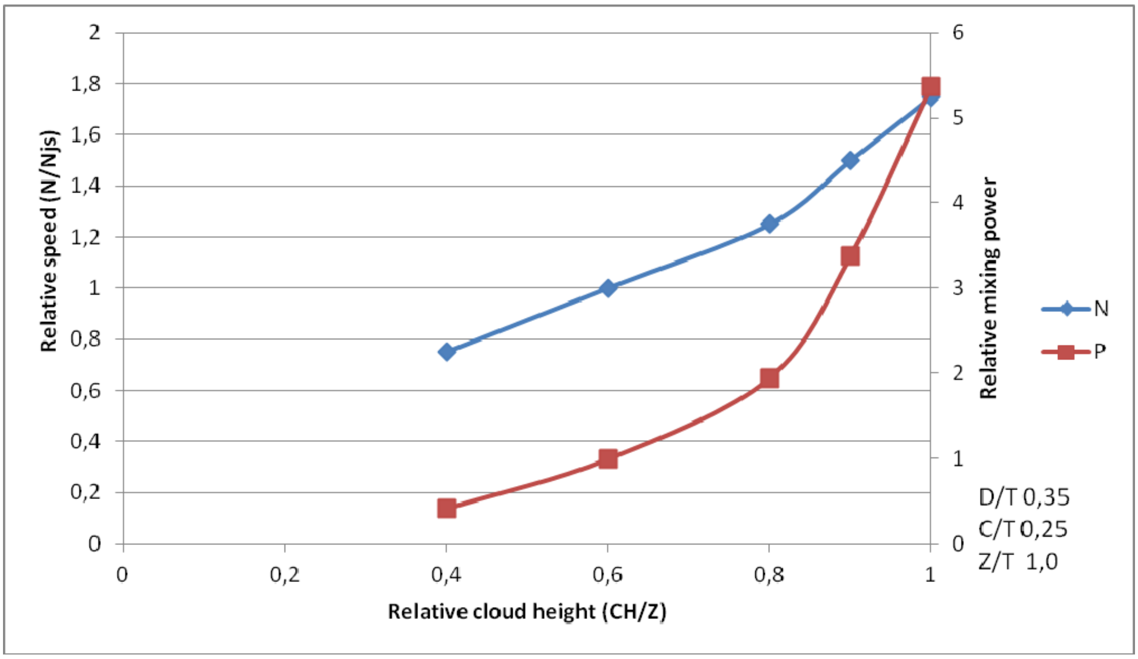


Figure 5. Shaft speed, power relation on cloud height generation for single stage impeller [2].

The diagram shows that the particle cloud can be lifted by increasing speed. As the speed is considered with N^3 in the power equation for agitator impellers, the required mixing power is changing also by the cube. A cloud height necessity of 90% triples the required mixing power compared to 60 % cloud height achieved with a single stage impeller arrangement, operated at NS1 at Z/T=1.

Figure 6 shows the relation between the impeller ratio D/T and the achievable cloud height. The cloud height can be calculated as [3]:

$$CH = \frac{N}{N_{js}}\left[0.84 - 1.05\frac{C}{T} + 0.7\frac{\left(\frac{D}{T}\right)^2}{1-\left(\frac{D}{T}\right)^2}\right] \tag{2}$$

This model, also called the wall jet model by Bittdorf and Kresta, is good for 0.154 < D/T < 0.52. It shows that a bigger impeller ratio generates a higher cloud height with far less mixing power.

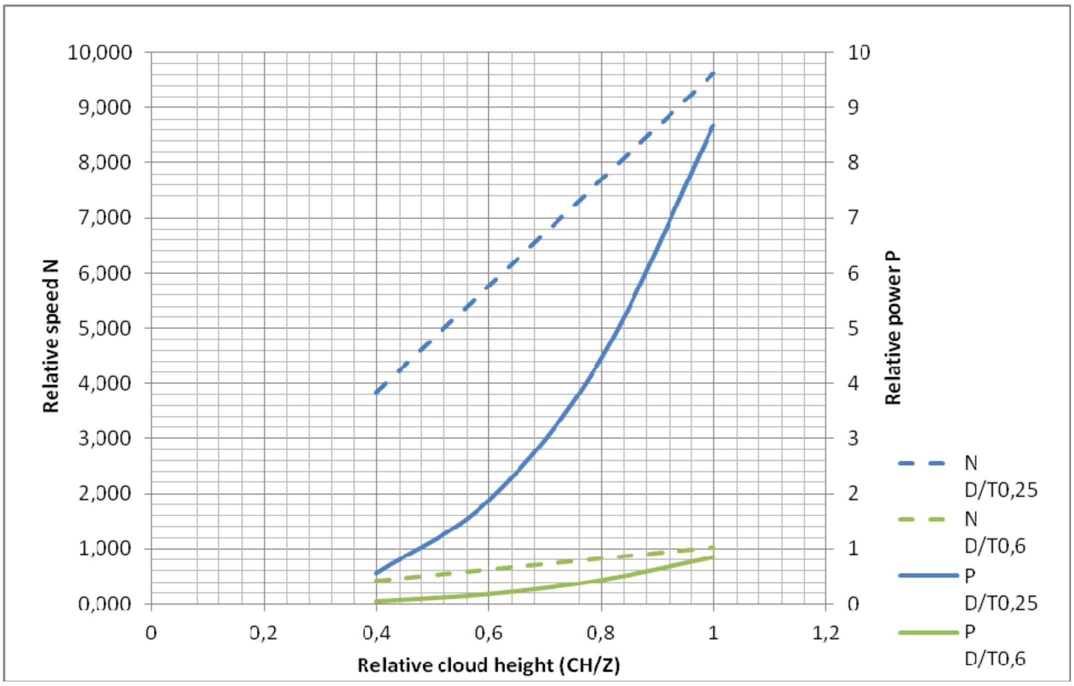


Figure 6. Cloud height generation with different impeller ratios for single stage impeller.

The number of impeller stages is relevant for all suspensions where a better quality than a complete suspension has to be achieved. Especially for suspensions with a very low variance in solids distribution, a dual stage impeller arrangement cuts the required mixing power by 2 or even 3, depending on the stage clearance of the impellers as shown in Figure 7.

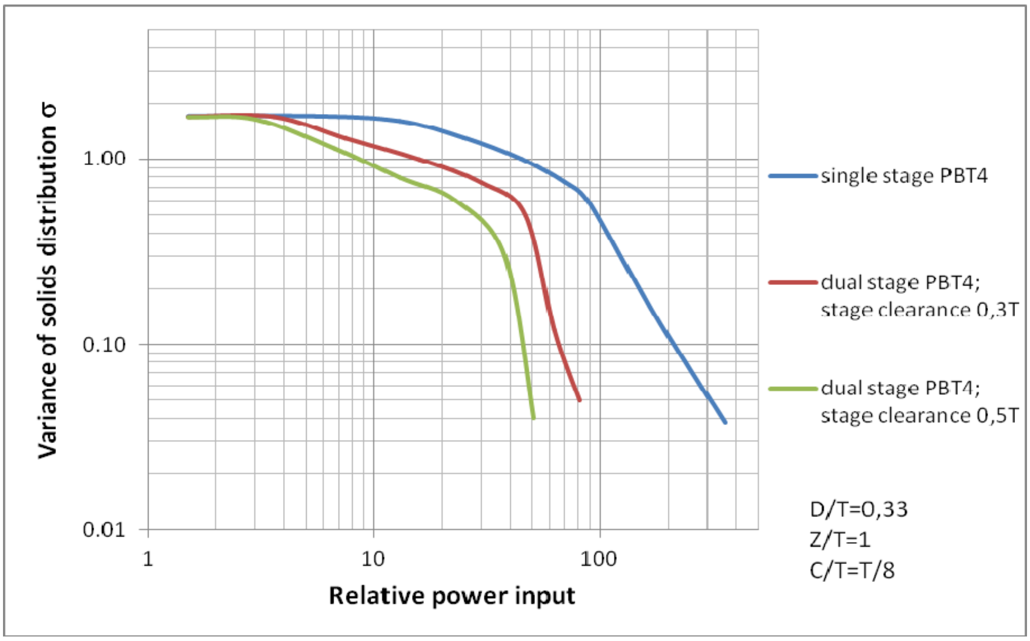


Figure 7. Variance versus power input for dual stage impeller arrangements [4].

The given values are valid for the stated geometrical configuration. Taller vessels have to be equipped in the same manner with the appropriate number of stages to achieve the same standard deviation. Summarizing all the figures, the central issue of how to design a mixing tank is the question of the suspension quality which is required, as this influences the number of stages, the speed and finally the resulting mixing power requirement.

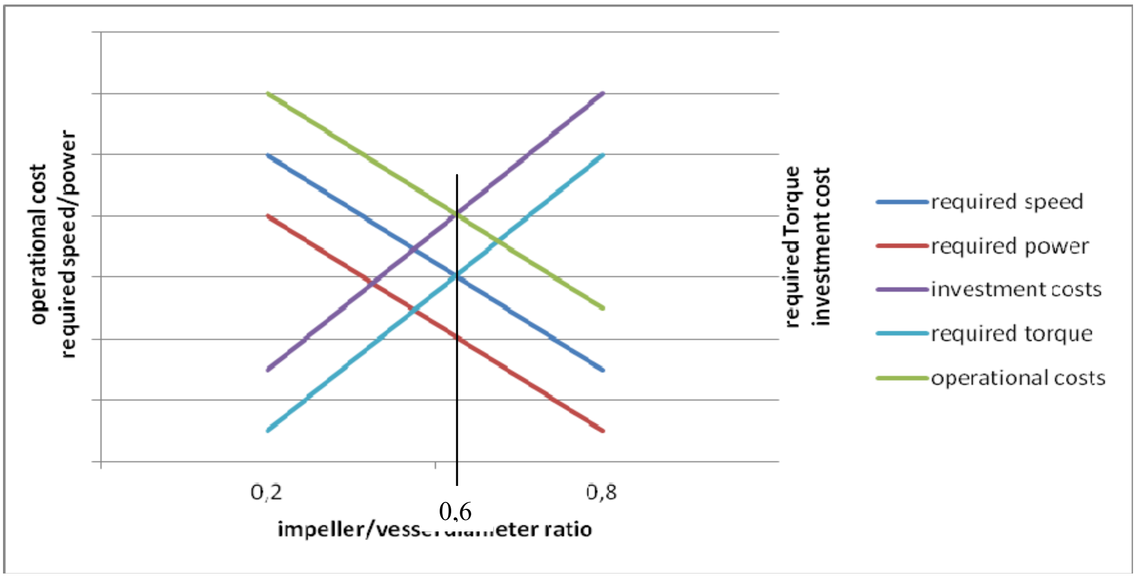


Figure 8. Relationships between design, investment costs and operational costs.

Combining the number of impeller stages, the impeller to vessel diameter ratio and the impeller speed, results in numerous different mixing designs, even for the same suspension quality. Figure 8 gives qualitative correlations between impeller/vessel ratio, required power, investment and operational costs. That means that the design either results in high investment costs and low operational costs or the reverse, low investment but high operational costs. For precipitator agitators optimum results for each aspect are usually achieved with an impeller to vessel diameter ratio of 0,6. This design leads to rather low investment as well as operational costs.

2. Natural overflow

Mixing cascades like the precipitation step in alumina production should be designed for a natural overflow from vessel to vessel without any pump or compressed air supply into the dip tube.

Basically the upward velocity in the overflow dip tube has to be higher than the particle settling velocity of the biggest particle in the stream. Usually a factor of 4 - 5 for the upwards velocity is required to avoid settling of particles in the dip tube. The diameter of the dip tube can easily be calculated by the equation:

$$Q = Av \tag{3}$$

Q Overflow rate
A Cross-sectional area of the dip tube
v Upwards flow velocity in the dip tube

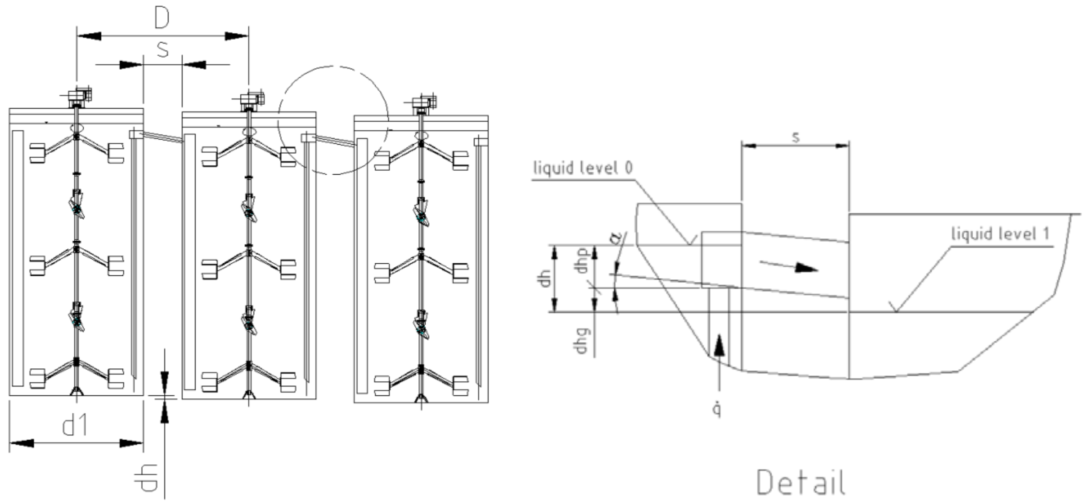


Figure 9. Arrangement of natural overflow in a mixing cascade.

The dip tube diameter should be calculated for a 4 - 5 times settling velocity for a flow velocity of 0.7 - 0.8 m/s. This is to avoid particle settling from the upwards flow stream and to suppress excessive scaling inside the tube.

The pressure difference or potential energy resulting from the different liquid heights of the vessels in the cascade has to compensate for the pressure drop through the upward flow of the dip tube.

$$p_f = (\zeta_A + \zeta_E + \zeta_r)\rho_2 \frac{v^2}{2} \tag{4}$$

ζ pressure loss coefficient for inlet, outlet and pipe
p_f pressure loss due to flow

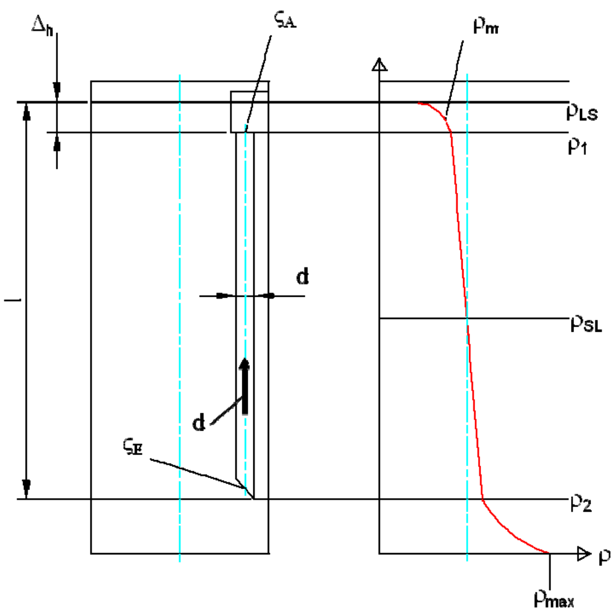


Figure 10. Density profile over height.

In addition the average density in the mixing process with ρ_{SL} is less than the density ρ_2 which has to be lifted up into the next vessel. This lower potential energy on the mixing side has to be compensated by the driving head pressure. The value of this potential energy depends directly on the defined allowable density fluctuation from top to bottom of the suspension quality with its solids concentration deviation inside the vessel. Experience with precipitators shows that a variance $\sigma \leq 0.1$ has to be achieved to operate a trouble free unsupported natural overflow with a sensible driving head Δh. Common practice for such applications is a Δh value of 400 to 500 mm. That requires a maximum density fluctuation of $\Delta\rho$ = 3 % - 3.5 %, equal to $\sigma \approx 0.1$ and a uniform suspension.

The height difference of the vessels for natural overflow can be calculated as:

$$\Delta h = (p_f + p_d)\frac{1}{\rho_m g} \tag{5}$$

p_d	pressure loss due to density difference
g	gravity
Δh	height difference

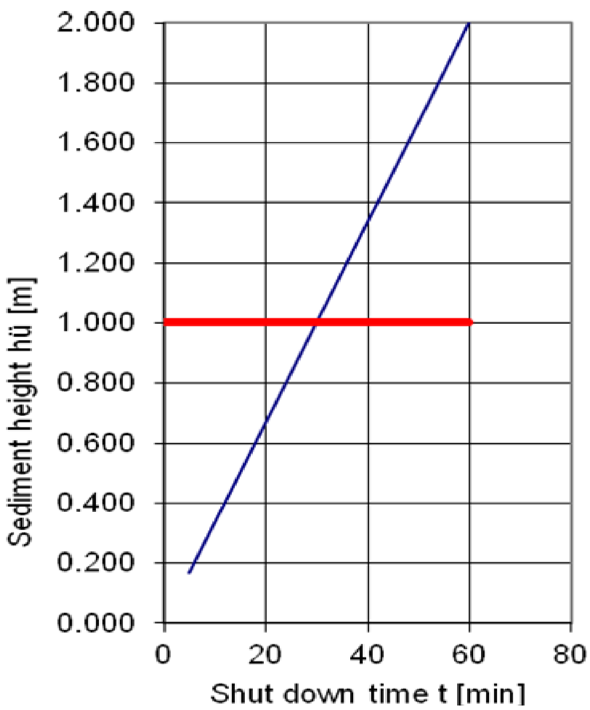
3. Restart in settled solids

Power failure is quite often a problem in alumina refineries. In case of a power failure the suspension in a mixed vessel cannot be maintained. The solids begin settling uniformly on the vessel bottom. The settling velocity and the resulting solids height depends on the grain size, the solids concentration and the densities of the liquor and solids. A higher solids concentration reduces the settling velocity as the particles hinder each other from settling. That means vessels with a higher solids concentration have a longer possible shut down time before re-suspension becomes difficult.

The sediment height over shut down time for a usual standard application of a precipitator shows a sediment height of 1.0 metre after a shut down time of approximately 30 minutes and 2.0 metres after 60 minutes as shown in Figure 11.

From the picture in Figure 12 it looks like the high speed hydrofoil impeller with a big bottom clearance can easier deal with the settled solids than the low speed system with a very narrow bottom impeller. The problem with the high speed system is that the impeller needs the big bottom clearance to generate the vortex which lifts the solids. If the impeller is too close to the bottom the energy is wasted in compacting the material at the bottom, not for raising the particles. This may even cause complete sanding-in of the impeller [5].

The low speed design has to consider the mechanical aspects of the partly or fully submerged impeller stage. Therefore this bottom stage has to be designed to withstand the complete start-up torque of the drive. The required power and torque for the low speed case can be calculated according to [6]:



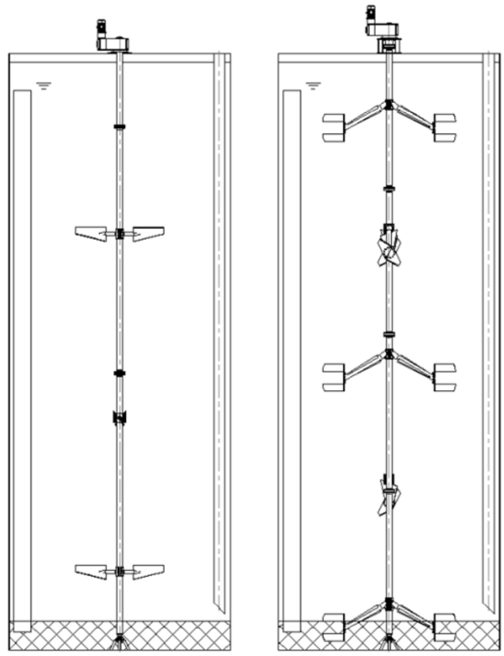


Figure 11. Shut down time. **Figure 12. Restart conditions.**

Grain size dk_{80} = 110 µm, ρ_{Fl} = 1270 kg/m³, ρ_S = 2440 kg/m³, C_V = 33 %

$$\mathrm{Re}^* = \frac{N^2 D^2}{\left(1 - \frac{\rho_L}{\rho_{set}}\right) y_0 g} \quad (6)$$

*Re** Mmodified Reynolds number
ρ_L Liquid density
ρ_{set} Settling density
y_0 Submerged height of the impeller
g Acceleration of gravity.

This modified Reynolds number takes into account the axial pressure on the impeller resulting from the weight of settle solids above. The resulting Reynolds numbers are comparably low and the specific power number curve for each impeller type has to be determined.

The power calculation then follows the standard equation but with Ne* for the solids bed:

$$P_R = Ne^* \rho_{set} N^3 D^5 \quad (7)$$

P_R Required mixing power for restart
*Ne** Power number of the impeller for a bed of solids.

4. Scaling mechanism and erosion

Scaling is an important factor in defining maintenance requirements and plant downtime. Finally the operational costs of an alumina plant and even its competitiveness are significantly determined by the costs for removing scale. Some sources quote 25 % of maintenance costs go into descaling. Hence all measures leading to a reduced scaling are worthy of evaluation. Scaling in the different areas of the alumina production process can be influenced by the local chemistry, but also by the installed agitator. The velocity of the fluid flow plays the most important role herein. The flow velocity generates an erosion effect which suppresses scale growth.

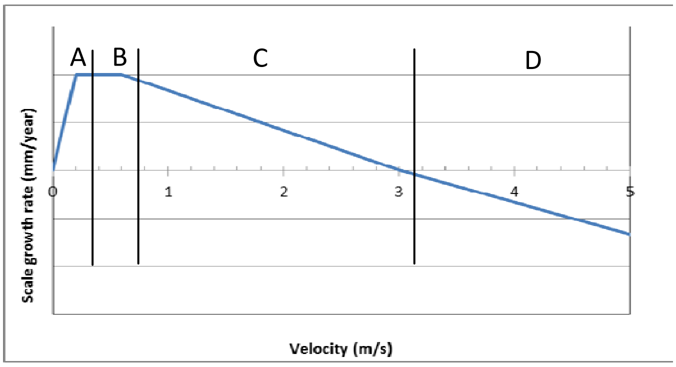


Figure 13. Scaling versus velocity [7].
A – mass transfer control, B – chemical reaction control, C – Scale suppression by erosion D – erosion damage

As shown above, at an average velocity of 3 m/s the velocity of the suspension leads to erosion of the metal parts and has to be avoided under all circumstances.
In a real plant example, a desilication vessel shows scaling at the wall as well as erosion at the outer tips of the impellers.

The reason is a speed above 3 m/s at the impeller tips, but a far too low velocity at the vessel wall. The velocity at the vessel wall is a combined velocity of the upstream flow resulting from the pump rate of the impeller and the tangential velocity resulting from the tip speed of the impeller. The bigger the distance between the impeller and the wall the less influence the tip speed has on improving the wall velocity. That means small impellers show a tendency to have erosion on the one hand and wall scaling on the other hand. Installations with bigger impeller diameters show neither erosion nor excessive scaling. Scale washing cycles of more than 12 month could be easily achieved.

Figure 14. View inside a cleaned desilicator.

Installed precipitators with a small impeller design show exactly the problem described. The flow velocity at the wall is 0.5 m/s, however a comparable installation with slightly bigger impellers has a 20 % higher flow velocity at the wall though a 20 % lower tip speed. The lower investment cost with the smaller impellers has to be paid for by higher operational costs which are growing disproportionately with inflation.

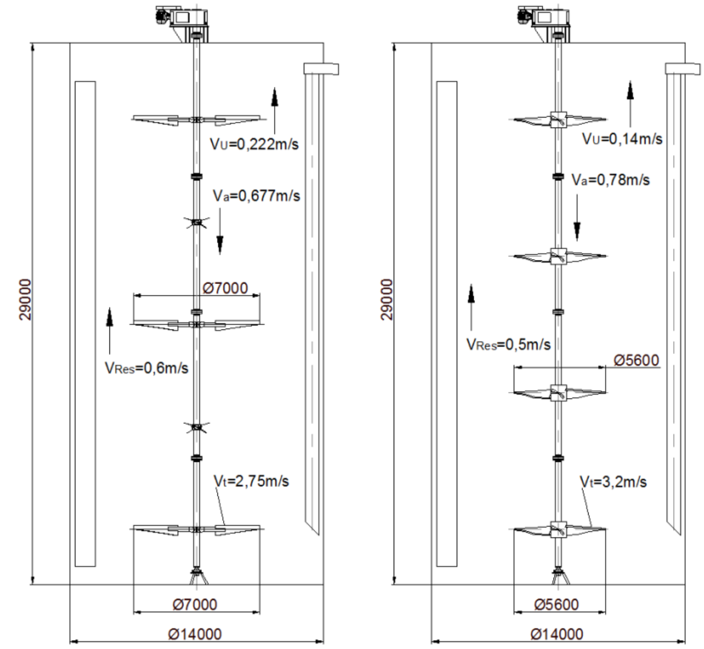


Figure 15. Design comparison for precipitators.

5. References

1. Zlokarnik, M. *Rührtechnik, Theorie und Praxis.* Springer Verlag, 1999.
2. Hicks, M. T., Myers, K. J., & Bakker, A., Cloud height in solids suspension agitation. *Chemical Engineering Comm*, 1997, Vol. 160, pp. 137-155.
3. Paul, E. L., Atiemo-Obeng, V. A., & Kresta, S. M., *Solid-Liquid Mixing.* John Wiley & Sons, Inc., 2003.
4. Mak, A. T.-C., Solid-liquid mixing in mechanically agitated vessels. *Thesis submitted for the degree of Doctor of Philosophy*, June 1992, London, England.
5. Oldshue, J. Y.. *Fluid mixing technology,* Chemical Engineering McGraw-Hill Pub. Co., 1983.
6. Kipke, K. D. (1983, 2). Anfahren in abgesetzten Suspensionen. *Chemie Ingenieur Technik* , 1983, 2, pp. 144-145.
7. Khan, M., & Hassan, N., *Thermofluid Modeling for Energy Efficiency Applications.* Academic Press, 2015.