DIGESTER VESSEL DESIGN

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Abstract

The Ras Al-Khair aluminium complex is a fully integrated aluminium production facility consisting of a 1.8 Million tonne per year alumina refinery. The Ma'aden alumina refinery is adopting a single stream digestion process which couples jacketed pipe unit technology with traditional autoclave digestion technology. The bauxite to be digested has a total available alumina of approximately 50% half of which is boehmitic with a broad particle size distribution, requiring a high digestion temperature and appropriate residence time. A combination of physical modelling and direct steam injection design principles were used to assess residence time distributions of the selected particle sizes of the bauxite. Two types of arrangements of feed slurry geometries were studied. Residence time distribution of solid phase bauxite particles were shown to be sensitive to different slurry inlet configurations.

1. Notation and Units

Symbol	Definition	Units
B	Condensation Potential	Dimensionless
C_D	Drag Coefficient	Dimensionless
Cp	Specific Heat	$kJkg^{-1}C^{-1}$
D	Diameter	m
g	Gravitational Constant	mS^{-2}
h	Enthalpy	$kJkg^{-1}$
t	time	s
T	Temperature	C
T_o	Superficial residence time	s
V	Velocity	ms^{-1}
Greek Symbols		
ρ	Density	kgm^{-3}
μ	Viscosity	Pa*s
Subscripts and Superscripts		
L	Liquid phase	
S	Steam Phase	
P	Particles/Solid Phase	

2. Introduction

The Ras Al-Khair autoclave digesters consist of two high aspect ratio (~8) pressure vessels operating in series. Direct steam injection takes place in the first vessel to raise the digester slurry to its temperature set point. This paper will focus on two distinct areas for the study of digester geometry:

A) Residence time distribution of the digester using physical modelling.
B) Design considerations of the steam injection nozzles.

2.1 Residence time distribution of the digester using physical modelling

There is very little public information on the fluid dynamic modelling of digester vessels for the alumina industry. Woloshyn *et. al*., (2006) conducted a computational fluid dynamic analysis on digester vessels of similar aspect ratio, with three different inlet configurations, specifically a central inlet, fully tangential inlet and normal to the top of the digester vessel with a deflector plate. The study of a massless pulse of fluid distinctly highlighted how the impact of swirl can make on the improvement on the mean residence time and the extent of which semi-plug flow conditions are achieved.

For the purpose of understanding both the residence time distribution and the governing flow structures within the vessel, physical modelling was chosen as the most suitable technique. Physical modelling has been used over a wide range of industries in a wide range of applications.

Understanding physical modelling requires comprehension of similarity theory. For most systems with multiple phases, extreme temperature gradients and chemical reactions, it is impossible to achieve complete similarity. Successful physical modelling requires knowledge of which dimensionless variables to relax and which to pursue, this technique is known as '*partial modelling*' (Spalding, 1963). The current digesters were modelled with live steam, and hence understanding the amount of steam to be injected given the limitations of steam pressure supplied by a small scale electrical boiler is required.

CSIRO was contracted by Hatch-Outotec (HOT) to experimentally investigate the following:

- Liquid phase residence time distribution
- Solid phase residence time distribution

In addition the effect of an increase in angular momentum on both these quantities was to be investigated, by alternating between a 52 mm and 40 mm inlet nozzle simulating the slurry flow into the digester.

In order to achieve these objectives a 1:10 Perspex digester was modified to conform to the preselected geometry of the lead digesters present in the Ma'aden alumina refinery. Water was used as the working fluid simulating slurry flow. Steam at 120 C was supplied with a laboratory scale boiler plant (Figure 1) through six steam nozzles evenly spaced around the circumference of the digester at a defined height from the water (slurry) nozzle. Two different size water nozzles were used during the course of this campaign; namely 52 and 40 mm. In order to assess the liquid phase residence time distribution the water stream was laced with a pulse of a saline solution. Glass beads of known size distribution were used to simulate the behaviour of bauxite in the full scale digester. The water nozzle is located semi tangentially, 0.3 digester diameters from the digester axis (Figure 1).

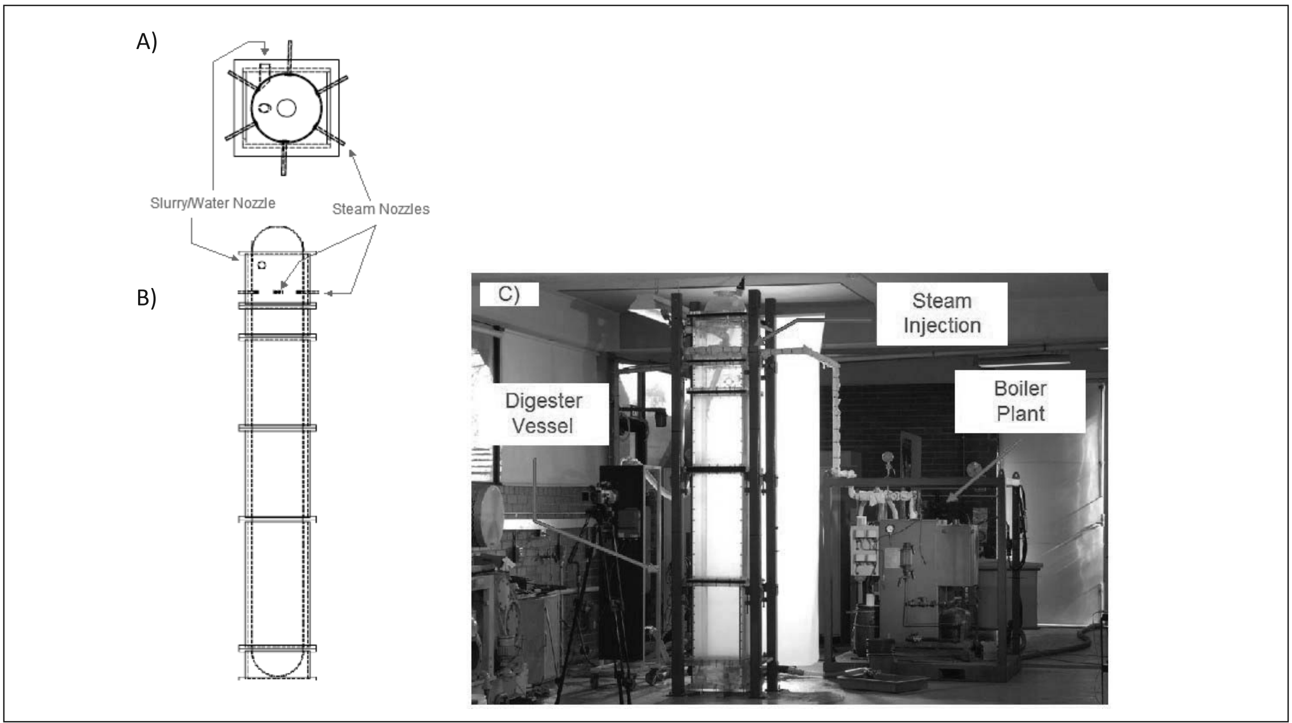

Figure 1: Schematic Diagram of the Digester, top view A), Side View B) and Photograph of digester Perspex model and steam boiler plant C).

A dimensional analysis of all known physical and operational related variables needed to be taken into consideration. The resultant analysis yielded the following :

<u>Geometric similarity</u>: The geometric aspects of the model must reflect that of the full scale version, diameter, height, location and size of slurry and steam nozzles.

<u>Reynolds Number similarity</u>

$\frac{\rho_L V_L D}{\mu}$, A fair assumption is that the bulk flow characteristics will change from laminar to turbulent flow, the latter is being maintained in the model, hence the criticality of the Reynolds number may be neglected.

<u>Froude Number similarity.</u>

$\frac{V_L^{2}}{gD}$ considerations must be made on the use of live steam and its buoyant effects in a moving fluid. Froude number similarity dictates whether the likelihood of steam bubbles rising rather than being dragged with the bulk fluid occurs. Froude number similarity between the full scale facility and model was achieved. Visual observations of the steam in the model indicated that the buoyancy effects of the steam bubbles were deemed negligible.

<u>Liquid phase to steam phase velocity ratio.</u>

$\frac{V_L}{V_S}$, The slurry flow/steam flow velocity ratio in the full-scale digester was maintained with the physical model. This scaling criteria ensures a similar vapour gas hold-up between full scale and model. The liquid phase-steam phase boundary is more affected by the gas hold-up, significantly less by the gas phase density, implying that a velocity ratio is more suitable than momentum ratio $\frac{\rho_L V_L^{2}}{\rho_S V_S^{2}} \frac{D_L^{2}}{D_S^{2}}$. (CSIRO, 2011).

<u>Particle settling velocity/feed velocity ratio:</u>

$\frac{\sqrt{(4/3)((\rho_P - \rho_L)/\rho_L)(D_P/C_D)}}{V_L}$, Different size particles respond differently to vortex conditions, namely the size and strength of the vortex. The size and strength in this case is dictated by vessel geometry and localised velocity magnitude and direction. The liquid flow was laced with glass beads of a known particle size distribution. The D_{90} of the glass beads has the same particle settling velocity to vessel superficial velocity ratio as the bauxite in the full size digester (CSIRO, 2011).

<u>Liquid phase flow visualisation and residence time distribution:</u>

It can be observed that a strong swirl flow motion was produced due to the angular momentum in the feed flow; this is expected since the feed pipe is offset from the centre of the digester. The swirling flow is the dominant flow feature in the vessel and its presence aides in the suppression of scale formation. The reduction of the water inlet nozzle from 52 to 40 mm results in a 70 % increase in angular momentum. Figure 2 is a still frame of the mid section of the digester; the water was laced with polystyrene beads acting as a flow tracer. The stronger level of swirl experienced in Figure 2 B is highlighted by the fact that the polystyrene beads are concentrated around the centre axis of the digester. Another important flow feature of the digester that was observed was a counter rotating vortex towards the bottom section of the digester directing material upwards. Figure 3 is a schematic diagram of the counter rotating structure as it's reproducibility via photographic means makes it difficult to distinguish. The width and height of this particular vortex was reduced when the 40 mm nozzle is in place.

Figure 4 is an overall time dependant flow visualisation of the two studied cases, the digester on the left hand side of each frame is the 52 mm nozzle, and on the right is the 40 mm nozzle. It is quite apparent that there is little or no change in length wise (axial) mixing with a change of inlet nozzle. These results are confirmed with quantified liquid phase residence time distributions (Figure 5) which were determined by measuring the conductivity of the digester outlet of a single pulse of a saline solution fed to the digester.

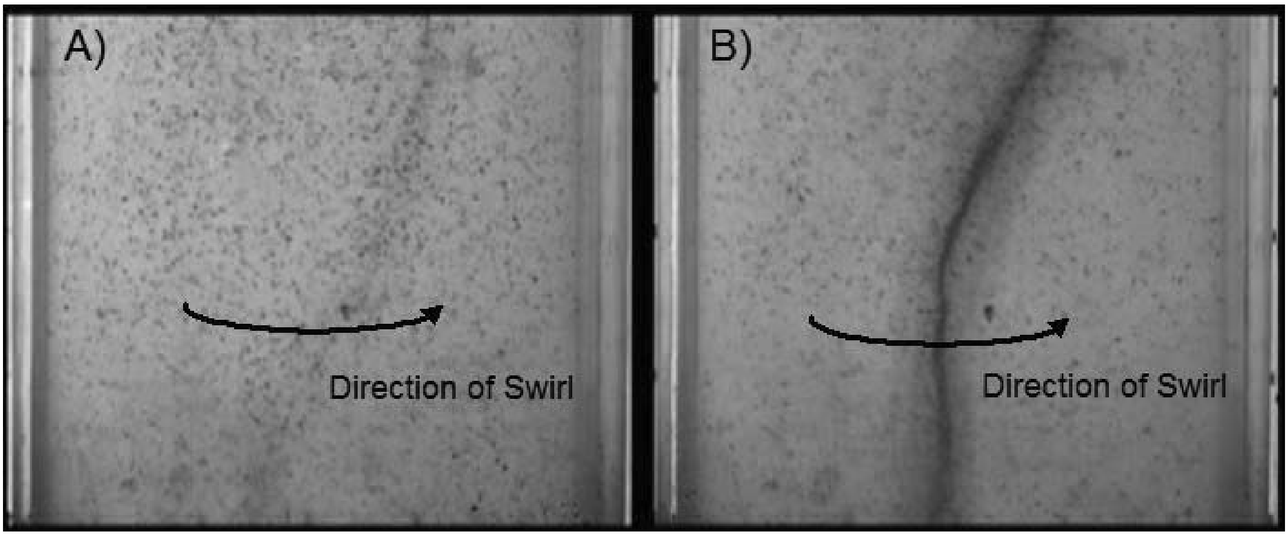

Figure 2: Flow visualization of mid section of digester, fluid laced with polystyrene beads A) 52 mm nozzle and B) the 40 mm nozzle.

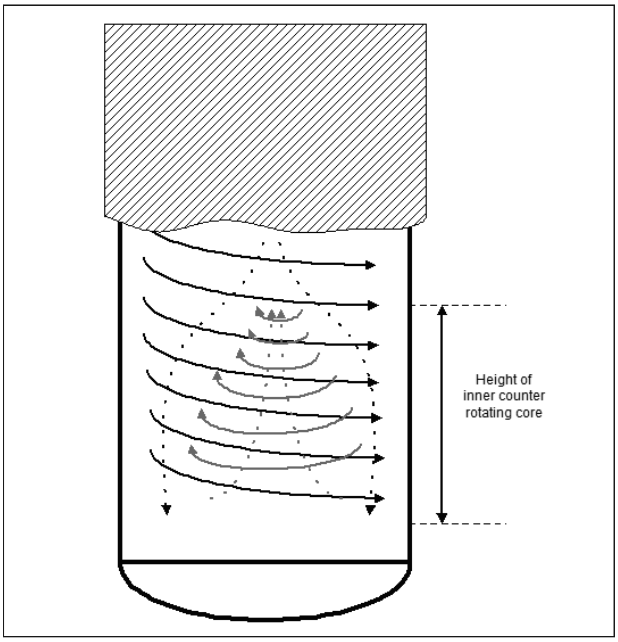

Figure 3: Schematic diagram of inner rotating counter vortex.

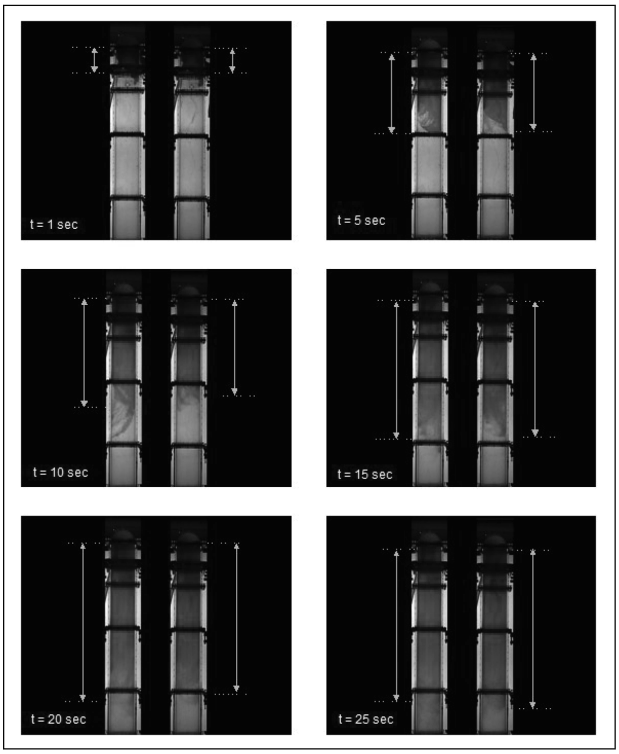

Figure 4: Dye tracing tests of the digester at selected time intervals. The left hand digester in each frame is the 52 mm nozzle case; the right hand side is the 40 mm case.

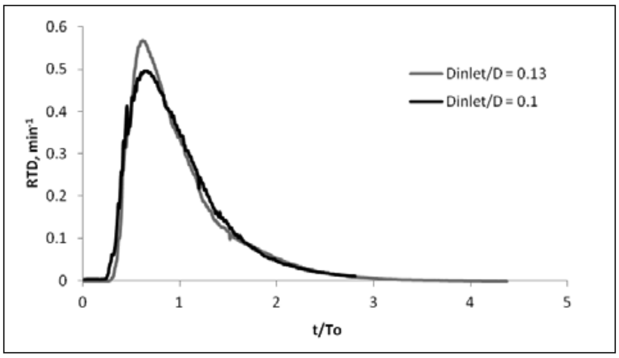

Figure 5: **Dimensionless residence time distribution of the digester**

<u>Solid phase residence time distribution</u>

The water being fed to the model digester simulating the bauxite slurry was laced with a pulse of glass beads of a known particle size distribution. The water solid mixture leaving the model was collected at known time intervals, dried and particle size measurements conducted resulting in residence time distribution per particle interval.

Figure 6 is the solid phase residence time distribution of selected particle sizes for both 52 and 40 mm inlet nozzle cases. The particle sizes present in Figure 6 have been reverse scaled and are the actual size for the full scale digester. As the particle size becomes progressively smaller the curve becomes less skewed, more widely distributed and on average retained for longer within the vessel. This observation is almost expected as smaller size particles will be more sensitive to the behaviour of the bulk flow field. Comparisons between the 52 mm and 40 mm cases show that particles of equivalent size on average will exit the digester earlier when more tangential momentum is enforced with the 40 mm nozzle.

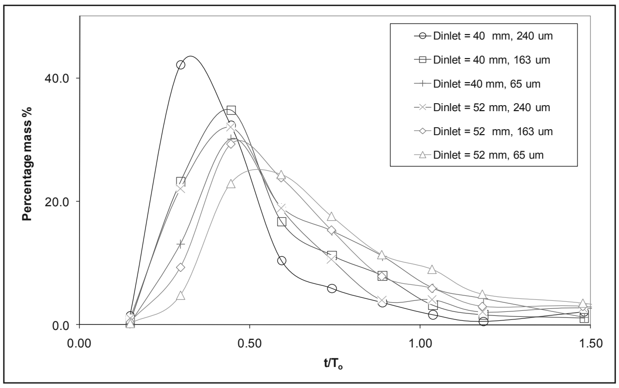

Figure 6: **Solid Phase residence time distribution**

2.2 Design considerations of steam injection nozzles

The steam injection nozzles can be simplified as jets. Jets are classified as fully separated flows and a free jet is classified as a fluid issuing from a nozzle into a stagnant or moving fluid. On a time averaged basis single phase jets experience three basic regions of differentiating flow:

- Potential core: Region of uniform velocity having the same velocity as the fluid of origin
- Mixing or shear layer region: A region where mass and momentum are transferred from the ambient fluid to the general directional flow of the jet
- Transitional region: This region represents the end of the potential core and non-uniform velocity profile
- Fully developed flow region: This region represents normalised centreline velocity whilst still growing due to entrainment.

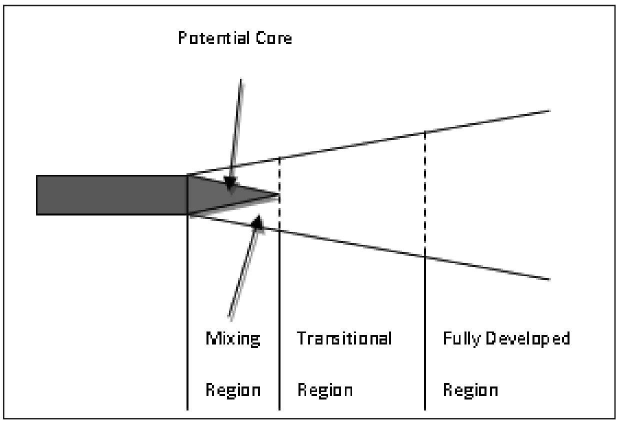

Figure 7: Flow regions of a turbulent free jet.

The length of the potential core region is normally 4-5 nozzle diameters (Rhine and Tucker, 1991).

Direct steam injection is sometimes a favourable option over indirect heat transfer due to the high heat transfer coefficients. The heat transfer coefficient is dependent on the actual flow regime of the steam interacting with the bulk liquid. Several design considerations must be made when selecting the steam delivery system, for this particular case as two items were of concern:

- Risk of chugging
- Risk of steam impingement on digester walls

Chugging occurs when there is insufficient velocity to force steam out continuously, resulting in steam jetting out as individual bursts, meanwhile in between each burst bulk fluid will occupy the steam nozzle, in the case of the digesters this is a highly undesirable outcome.

The literature suggests (Chan and Lee 1982, Aya and Nariai 1991, Petrovic *et al*., 2007) that the actual flow regime of condensing steam in a stagnant fluid is dependent on the mass flux of steam, the temperature gradient with surrounding fluid and nozzle diameter. Aya and Nariai (1991), suggest for low mass fluxes that the condensing steam can either be:

Chugging: Collapsing steam in intervalled bursts causes sub-cooled liquid to flood the nozzle.

Condensation Oscillation: Characterised by an oscillatory jet dependent on the enlargement and shrinking of steam bubbles. In this case the mass flux of the steam is sufficient so that flooding of the steam nozzle is non-existent.

Chan and Lee (1982) and Petrovic *et. al*., (2007) suggest that higher steam mass fluxes result in ellipsoidal or oscillatory cone jets.

According to Aya and Nariai (1991), the heat transfer coefficient of an oscillating condensing flow at a mass flux of 15 kgm^{-2}s^{-1} is approximately 100, 000 Wm^{-2}C, conversely Kozekia and Kuwabara (1972), suggest that heat transfer coefficient varies between 10^5 to 10^6 Wm^{-2}C, depending on both mass flux and temperature difference between steam and bulk fluid.

Steam plume length as a result of direct steam injection into sub-cooled water has been investigated in a number of studies. De With (2009) showed using data from a number of previous studies that there is a correlation between the plume length with mass flux and condensation potential ($B=(c_{pL}/h_L)\Delta T$). The correlation showed that the largest plume lengths are seen at high mass flux and low condensation potentials and the shortest plume lengths at low mass flux and high condensation potentials. It is desirable at this stage that the steam plume length in the full scale digester from each nozzle be as short as practical.

Another important consideration is the effect on cross flow. De With (2007) suggests the plume length decreases in a flowing stream compared to that of a no-flow situation thus showing that the effect of increased condensation due to the constant supply of bulk fluid is much larger than the momentum effect. This tends to suggest that in a vessel that is well mixed the condensation is likely to happen very quickly resulting in small plume lengths unless very high mass fluxes of steam are experienced or that the condensation potential is low. New *et al.,* (2006) investigated the effect of momentum ratio of single phase round jets in cross flows, and concluded at a momentum ratio (transverse to jet flow) of 0.24 the jet starts to lose it's free stream trajectory and is skewed by the bulk flow.

The observations from the physical model revealed that the steam was condensing under a condensation oscillation regime, with no evidence of chugging. The mass flux of steam in the model is approximately 7 kgm^{-2}s^{-1}, whereas in the full scale digester it is in excess of 300 kgm^{-2}s^{-1}, thus guaranteeing no risk of chugging. Figure 8 is a snapshot of the steam injection nozzles highlighting the length and skewed nature of the steam plume, proving that the steam jets are overwhelmed by the momentum of the bulk fluid and indicating that there is little risk of steam impingement.

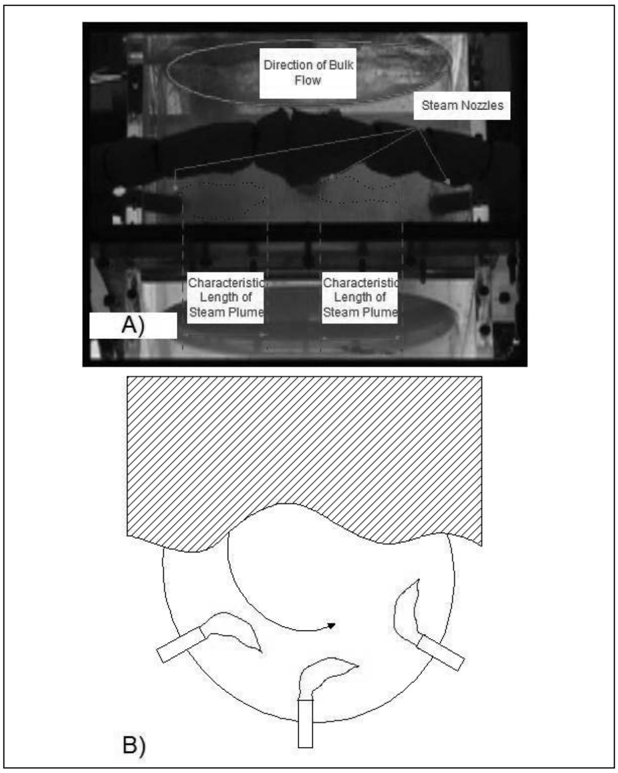

Figure 8: Close up and schematic depiction of steam nozzles showing A) length of steam plume and B) the skewed nature of the steam jets

3. Conclusion

Both qualitative and quantitative studies were conducted on a scaled down physical model of the digesters present at the Ras Al-Khair aluminum complex. The bulk flow field of the digester vessels is dominated by the semi tangential inlet of the water nozzle (slurry). A decrease in water nozzle diameter (increase in angular momentum) results in intensified swirl, with little or no change in axial mixing along the length of the digester. A secondary flow feature is the counter rotating vortex from the bottom of the vessel, the size and intensity of this structure increases with a drop of angular momentum or inlet nozzle size.

Liquid residence time distribution exhibits the semi-plug flow behaviour which was also similarly quantified by Woloshyn, and Oshinowo (2006). The liquid residence time distribution showed little variation with the change in inlet nozzle diameter. Conversely solid phase residence time distribution does exhibit dependence with the size of the inlet nozzle, an increase in nozzle size resulting in longer retention of smaller sized particles.

Physcial modelling verified the avoidance of chugging of the steam nozzles and using the criteria of Chan and Lee (1982) and Petrovic *et. al*., (2007), highlights the likelihood of conical steam plumes occurring in the full scale digester. The physical model also verified how the bulk fluid within the vessel dominates the flow field, resulting in the deviation of the steam plume from the geometric axis of the steam nozzles, aiding in the mitigation of steam impingement.

4. Acknowledgements

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