

# SMELTER GRADE ALUMINA QUALITY IN 40+ YEAR PERSPECTIVE – WHERE TO FROM HERE?

B.E. Raahauge

*Alumina & Pyro Technology, FLSmidth Minerals, Copenhagen, Denmark*

*Corresponding author: Raar@flsmidth.com*

## ABSTRACT

All commercial Smelter Grade Alumina (SGA) is produced by Calcination of Aluminium Hydroxide in Alumina Refineries. The Aluminium Hydroxide, or Hydrate, is precipitated from a super saturated solution and classified into fine seed and coarse product for Calcination in rotary kilns or stationary calciners.

Over the past 40+ years significant technology changes/shifts has taken place in Refineries and Aluminium Smelters, the only customer for SGA. The drivers for these changes were and still are Economy of Scale, Energy Efficiency and/or improved Environmental performance.

Following an introductory review of the major technology changes/shifts that has taken place, the associated challenges in SGA quality is reviewed. Emphasis will be on Alumina properties from Calcination in Rotary Kilns and Gas Suspension Calciners (GSC) at low or high temperature onwards into Smelting, such as: Particle Size, Strength and Dust Generation & Management, HF Generation and Capture, Alumina Feeding and Dissolution. Finally the impact of major chemical impurities will be discussed. The current SGA quality requirements beyond consistency are summarized with a view to optimization of the available Hydrate Precipitation and Calcination technology. Future SGA quality requirements are discussed from a Calcination point of view as the intermediate processing step between the Refiner and the Smelter, including its potential constraints or drawbacks.

## 1. INTRODUCTION

Significant technology changes/shifts has taken place in Refineries and Aluminium Smelters over the past 40+ years. The drivers for these changes were and still are Economy of Scale, Energy Efficiency and/or improved Environmental Performance.

The major technology changes/shifts that have taken place and the associated challenges in SGA quality are reviewed.

Alumina properties from Calcination in Rotary Kilns and Gas Suspension Calciners (GSC) at low or high temperature calcination will be presented and discussed.

Future SGA quality requirements are discussed from a Calcination point of view as the intermediate processing step between the Refiner and the Smelter, including its potential constraints or drawbacks.

## 2. MAJOR TECHNOLOGY CHANGES

### 2.1 Alumina Refinery / Bayer Process

The major technology changes to the Bayer process are summarized in Table 1 below.

The increase in **refinery capacity** has been and is driven by economy of scale outside China. In

China replication of approximately 500,000 tpy standardized trains is the preferred business model (Kjaer, 2011).

The reduction in refinery **energy consumption** is mainly caused by indirect slurry heating by steam in digestion, introduction of co-generation power plants and introduction of stationary calciners replacing rotary kilns.

In **digestion** the tube digestion technology has been introduced treating bauxite with a mixture of gibbsite and boehmite, or diasporic type bauxite (Kelly et al, 2006).

The introduction of flocculants, feed slurry dilution and optimized feed well design has resulted in the introduction of much smaller tank sizes for Deep Cone Thickeners and Pressure Decanter by Rio Tinto Alcan and High Rate **Thickeners and Washers** by FLSmidth (Laros, 2009).

Following the **Red Mud Pond** incident in Hungary in October 2010 has highlighted the need for a change in design of red mud disposal areas from Dry Stacking introduced by British Aluminium (World Aluminium, 2013) to more Dry options named Dry Storage. This design change involves the replacement of Vacuum Drum Filters with Automatic Pressure Filters and transportation to the pond area with modern material handling equipment (Bach & Weston, 2011).

Parameter/Year	~ 1970	~ 2010
<b>Refinery Capacity</b>	600 -800 ktpy	1.6 - 2.4 Mtpy
<b>Energy Consumption</b> GJ/ton	Bayer Refinery: 13 – 14	Bayer Refinery: 8 – 9
<b>Digestion</b>	Autoclaves	Tube Digestion
<b>Precipitation</b>	Batch / Continuous	Continuous
<b>Classification</b>	Thickeners	Hydro-clones
<b>Settlers ,</b> Footprint (Flocculants) Effluent	1.8 - 2.3 <sup>2</sup> m /tpd 100-200 ppm O/F	0.5 <sup>2</sup> m /tpd < 50 ppm O/F
<b>Washers,</b> Footprint (Flocculants) Effluent	0.9 - 1.1 <sup>2</sup> m /tpd 300-400 gpl U/F	0.25 <sup>2</sup> m /tpd > 500 gpl U/F
<b>Red Mud Residue</b>	Dry Stacking British Alum.	Dry Stacking/ Storage
<b>Calcination</b>	Rotary Kilns & Fluid Flash	Stationary calciners
<b>Particulate Emissions</b> mg/Nm <sup>3</sup> (dry)	Electrostatic Precipitators 50	Fabric / Bag Filters 20
<b>SGA Quality</b>	Floury & Sandy	Sandy with Low Alpha

**Table 1: Technology Shifts - Bayer Refinery**

## 2.2 Calcination Technology

In 1935 the first **rotary kiln** for calcination of Aluminium Hydroxide with a capacity of 110 tpd alumina was installed.



**Figure 1: 3 x Ø 3.95 x 107 m Roatry Kilns**

Until about 1972 where the largest Ø 4.3 x 122 m rotary kiln was contracted with a capacity of 1400 tpd sandy alumina, this technology dominated the industry (Figure 1).

However, in 1952, Alcoa commissioned their first 300 tpd Fluid-bed calciner producing sandy alumina (Fish, 1974). Since the oil crisis in 1972 only stationary calciners have been installed (Figure 2) owing to about 25% less fuel consumption per ton alumina.



**Figure 2: 3 x 4500TPD GSC Units & Bag Filters**

Almost all rotary kilns has now been replaced with **stationary calciners** and many of the latest Gas Suspension Calciner (GSC) units are now equipped with **Bag House/Fabric Filters** (Figure 2) instead of Electrostatic Precipitators (ESP).

The major reason being that ESP units are not an absolute filter because a plume of alumina dust is emitted in case of a power failure. In addition hereto Bag House/Fabric Filters offer a lower dust emission at lower Capex.

Around 1970 Smelters were required to lower their emission of HF by installing **Dry Scrubbers** requiring **sandy alumina** with SSA of 50–80 m<sup>2</sup>/g.

Rotary Kiln Alumina	Floury	Sandy
<b>SSA (m<sup>2</sup>/g)</b>	5	50-80
<b>Alpha Alumina (%)</b>	80-90	10-25
<b>% Alumina &lt; 45 µm</b>	45-60	Max. 10
<b>Angle of Repose</b>	45	30-32
<b>Bulk Density, Loose (kg/L)</b>	0.8	0.95
<b>Precipitation Yield, g Al<sub>2</sub>O<sub>3</sub>/L</b>	70	50 - 80+

**Table 2: SGA from Rotary Kilns**

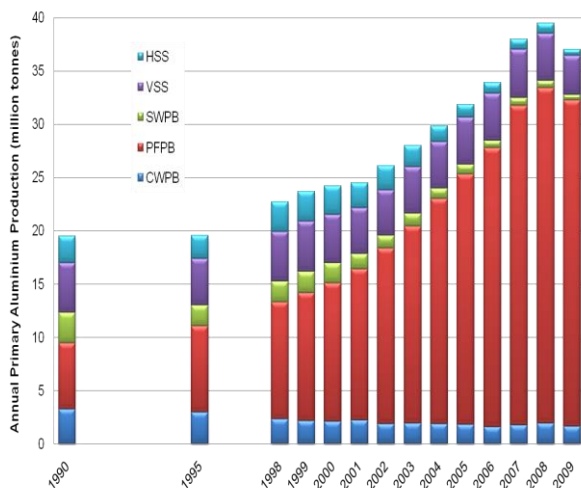
The change in alumina quality to Sandy made **Floury alumina** (Table 2) unacceptable to most Smelters overnight. and became a significant challenge for the Bayer Process operator to change their precipitation circuit to produce a coarser Sandy alumina particle, while maintaining the high precipitation yield of producing Floury alumina (Tschamper, 1981). Today most modern precipitation circuit designs comprises an agglomeration section followed by a growth sections followed by a three-step classification of hydrate into coarse product for calcination, coarse

seed subject to growth and fine seed subject to agglomeration.

However, the change to sandy alumina increased the rotary kiln capacity with about 25% and lowered the specific heat consumption from about 4,315 KJ/kg flourey alumina to 3,935 KJ/ kg sandy alumina.

### 2.3 Smelter Technology

The basic smelting process is taking place in the reduction cells of which a few basic designs has evolved over time (Figure 3).



**Figure 3: Technology Shift in Reduction Cells**

It is clear that the **Point Feeding Pre-Baked (PFPB)** anode cell has gained a dominating market position compared to the **Vertical Stub Soderberg (VSS)** cell, the **Centre Worked Pre-Baked (CWPB)** anode cell, the **Horizontal Stub Soderberg (HSS)** cell and the **Side Worked Pre-Baked (SWPB)** anode cell.

The development of the Key Performance Indicators of the reduction cell performance is summarized above (Table 3).

Parameter	1945	1995	2009-10
Cell Amperage, kA	25 - 50	300	200 - 500
Cell Voltage, Volt	5.0	4.1	3.85 - 4.20
Current Efficiency, %	80 - 58	92 - 95	91 - 96
Power Consumption, kWh/kgAl	20 - 25	13	12.5 - 13.5
AnodeCons. Kg/ton Al	< 776	420-550	<400

**Table 3: Technology Shift in Reduction Cells**

The change in basic cell design to PFPB cells and their size has been the major smelter technology

change together with its associated modification of anode and cathode designs.

The **pot/cell size** and production have increased as indicated by the increase in cell current, while cell voltage, current efficiency and power consumption has shown only minor improvements since 1995.

With respect to the **environmental performance** significant improvements have taking place with respect to reduced anode consumption, number of anode effect per cell/pot per day < 0.08-0.1 #/p/d, total fluoride emission < 0.35-1.0 Kg/ton Al, and reduced average PCF gas emissions to ≤ 0.6 ton CO<sub>2</sub> equivalent/ton Al (IAI,2008).

### 3. SMELTER GRADE ALUMINA CHALLENGES

Smelter Grade Alumina (SGA) quality parameters and their importance are summarized below:

Property	Importance (Welch&Kuschel, 2007)
Purity	Na and Ca for AlF <sub>3</sub> consumption, other impurities.
Loss on Ignition	0 – 300°C adsorbed readily releases volatiles (can form “volcanos” for point feeders). 300-1000°C predominantly Al-O-H compounds that are major contributors to HF evolution.
Bulk Density	Accuracy of correlation between volumetric feed and assured mass.
Angle of Repose	Covering quality if using only Al <sub>2</sub> O <sub>3</sub> , flow properties, ability to fill storage vessels actual volumetric feed transfer
Particle Size Distribution (PSD)	Consistency of properties, tendency to release dust (fines), cell performance perception of impact on solubility (coarse).
Surface Area, BET	Potential capacity for HF gas absorption.
Alpha Content	Measure of conversion of the calcined Al(OH) <sub>3</sub> to the most stable alumina phase, secondary measure of surface area for a given calcination method, crusting tendency.
Attrition Index	Tendency for the poly-granular alumina to disintegrate to a finer PSD.
Flow Funnel Test	The flow ability of alumina.
Dust Index	Qualitative assessment of ability to become suspended in the cell environment and be transported away.

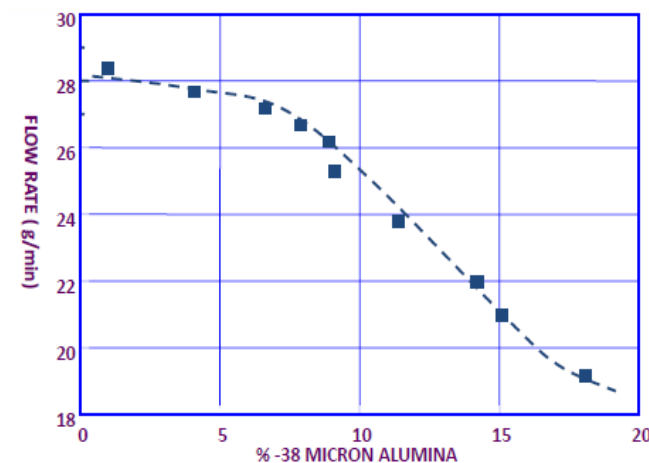
**Table 4: Smelter Grade Alumina (SGA)**

The understanding and adequacy of the above SGA parameters are still under debate (Perander

et.al.,2011 and Lindsay,2005) and new more sophisticated parameters maybe introduced in the near future. It is however important to bear in mind that variation in SGA properties from the refinery is changed, when the SGA is handled and transported, and passes dry scrubbers (Taylor, Chandrashekar et. al. and Lindsay, 2005). In addition hereto improper pot operation and alumina feeding may cause more variation than the variation in SGA properties from the refinery (Kuschel & Welch, 1991). All these factors make it difficult to correctly interpret changes in pot operation from changes in alumina properties (Meyer et al, 2012). The following discussion will focus on the most important aspects of SGA quality parameters and their perceived impact on smelter cell operation.

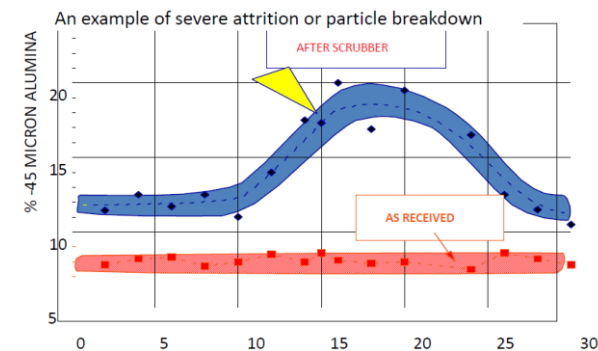
### 3.1 Particle Size, Strength and Dust Generation & Management

Increased content of **Fines** reduces the flow ability of alumina (Figure 4), subject to PSD and reduces the point feeding rate (Welch, 2010).



**Figure 4: Fines content versus Flow rate**

Increase in **Attrition Index** may lead to increased particle breakdown in Smelter. Alumina particle



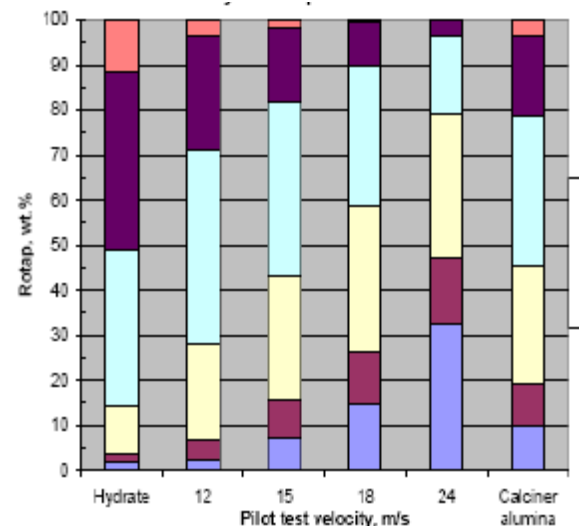
**Figure 5: Particle Breakdown from weak Alumina**

size, strength and breakdown determine the ultimate PSD fed to the reduction cell (Figure 5)

and thus its dust generation potential (Welch, 2010).

FLSmidth **R&D work** (Wind et al, 2010) has shown that larger particles break more than smaller particles when exposed to the same carrying gas velocity.

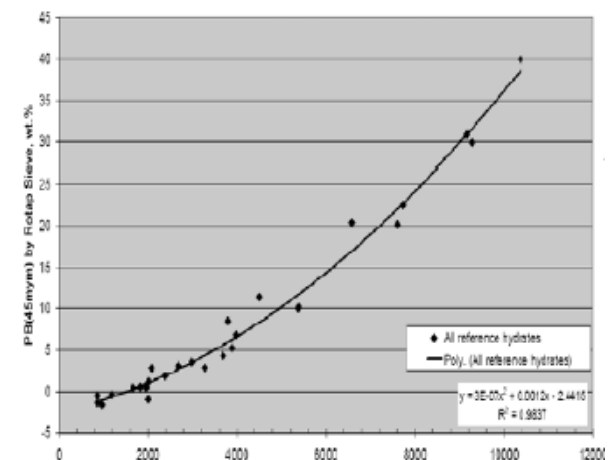
The largest particles break at 12 m/sec gas velocity in the pilot (Figure 6) to reach the same quantity of +150  $\mu$ m fractions as in alumina (RHS), while the +45 micron particles break at 15-18 m/sec to reach the same quantity of +45  $\mu$ m fractions as in alumina (RHS). This difference is understandable as the largest particles has the highest impact energy, but the grains making up the particles exhibit almost the same chemical bonding strength regardless of size.



**Figure 6: Coarse versus Fine Particle breakage**

The particle breakdown itself has been found to depend on the parameters AAI and U:

$$(1) \text{ Particle Breakdown (45 } \mu\text{m)} = f(\text{AAI} * U^2)$$



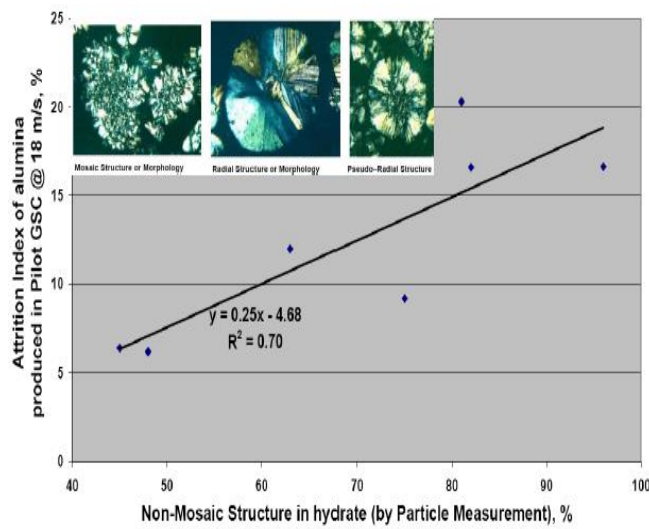
**Figure 7: Particle breakdown versus AAI x U<sup>2</sup>**



Where AAI is the Alumina Attrition Index at 45 micron (Alcoa method) and U is the gas velocity into the test cyclone.

The AAI itself depends on the morphology of the hydrate particles produced in the precipitation circuit (Figure 8) in the alumina refinery.

The **hydrate precipitation** parameters important for obtaining a strong alumina particle with low AAI (say < 10% on 45micron) has been thoroughly studied (Sang, 1986). Also the importance of avoiding over coarsening of the hydrate leading to increased particle breakdown rather than increase of the size fraction above 45 micron is of importance.



**Figure 8: AAI increases with Non-Mosaic Particle Structure measured optically on thin sections of hydrate particle.**

In conclusion it can be said that it make very little sense to produce a coarse but **weak alumina** particle (say, AAI > 20%) that breaks down relatively easily on its way to the reduction cells at the Smelter. In the alumina refinery **dust** is mainly generated in Calciners and collected in ESP's or Fabric Filters (Figure 9).



**Figure 9: Gas Suspension Calcliner**

Calcliner Dust is the primary source of **Fines** (< 45  $\mu\text{m}$ ) and **Superfines** (< 20  $\mu\text{m}$ ) in SGA shipped to the Smelter.

**Degree of calcination** of dust deviates significantly from the SGA produced in the Calciners (Table 5).

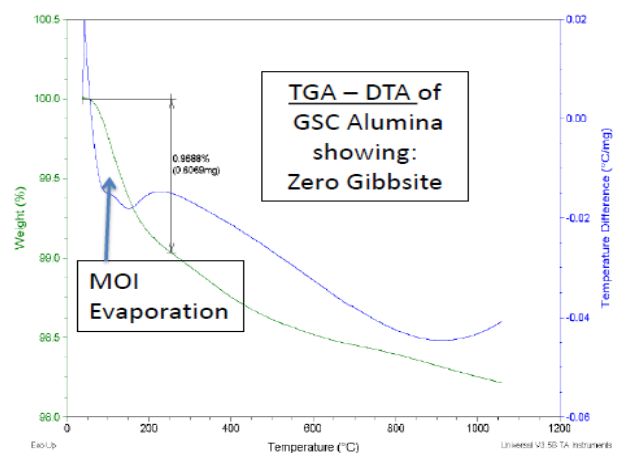
GSC Alumina	Gibb-Site wt%	LOI Wt%	Alpha %	SSA $\text{m}^2/\text{g}$
Dust	13.0	5.87	56	34.7
SGA	0	0.62	3	80.6

**Table 5: GSC Dust versus SGA incl. GSC Dust**

Smelters want to avoid dust to the extent possible and consequently it is of importance to consider what **Dust Management Options** is available?

- Recycle Dust to Precipitation (Chandra) or Digestion losing Refinery Productivity and increasing OPEX.
- Mix Dust with alumina from the Fluid-Bed Cooler producing SGA with Gibbsite content?
- Recover coarse dust by Dynamic Separation and Recycle mainly Superfines to precipitation or Digestion, whenever Precipitation experiences a Fines Generation Cycle?
- Back – Mixing Dust into the Hot Back-end of the Calcliner, calcining the Gibbsite and thus eliminate it from the SGA.

The last option is successfully practiced by FLSmidth producing SGA with no measurable **Gibbsite** (Figure 10) and low **Alpha alumina** phase (Figure 11) together with significant amounts of **amorphous phase**:



**Figure 10: SGA from GSC with 0% Gibbsite**

Some workers (Homsí, 2001) are of the opinion that amorphous alumina phase is an advantage over crystalline phases with respect to the dissolution of alumina (Figure 11).

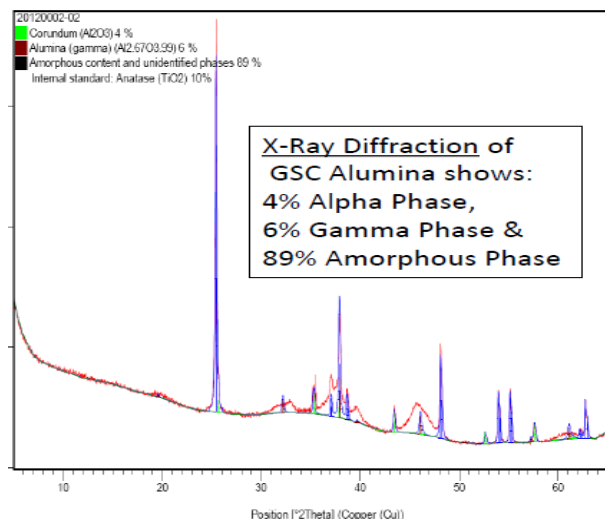


Figure 11: SGA from GSC low Alpha Phase

### 3.2 Generation and Capture of HF

The smelter grade alumina has a dual application:  
(1) **Removal of HF** from the smelter gasses in Dry Scrubbers/Gas Treatment Centres, before it is used as (2) **Feed stock** for production of primary aluminium in the electrolytic pot/cells.

This duality results in a conflict of specifications that can only be resolved with a compromise. The basic **Dry Scrubber/Gas Treatment Centre** flow sheet (Welch,2010) is shown below (Figure 12):

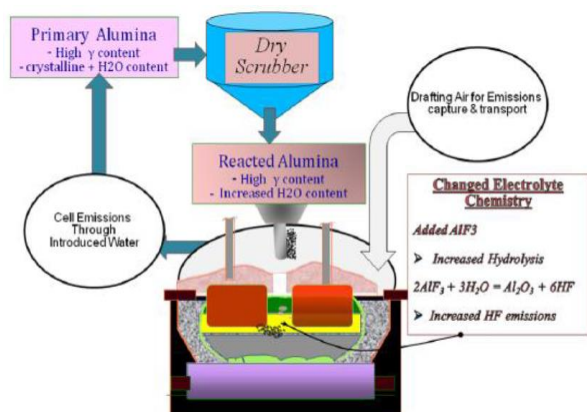


Figure 12: Alumina Flow through Dry Scrubber

The **Specific Surface Area (SSA)** is the primary physical property of the alumina specified for capturing HF gas in the Dry Scrubbing / Gas Treatment Centres emitted from the smelting pots.

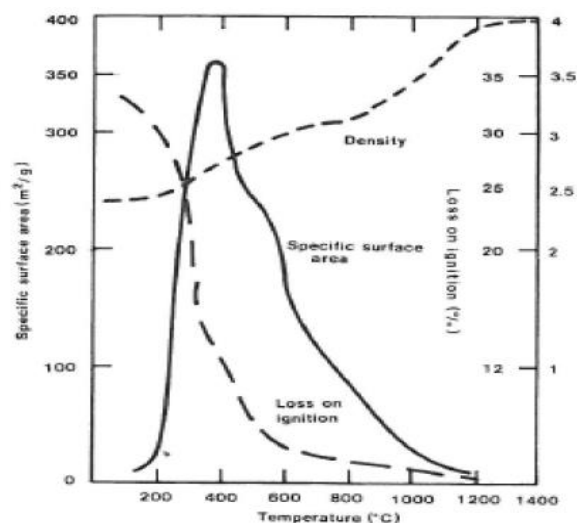


Figure 13: Degree of Calcination

The SSA increases sharply with the calcination temperature up to about 400°C, where after the SSA decreases to the 70-80 m²/g out of the calciners. At the same time LOI (300-1000°C) decreases from about 34.6% to below 1.0%.

It has been shown (Hyland, 2004) that **Residual Hydroxyl** groups account for more than 50-70% of the continuous water released during the smelting process (Figure 14).

Gibbsite, physically adsorbed moisture and some chemisorbed water molecules are released below 300°C (Metson,2009) during heat-up of the alumina fed to the reduction cell.

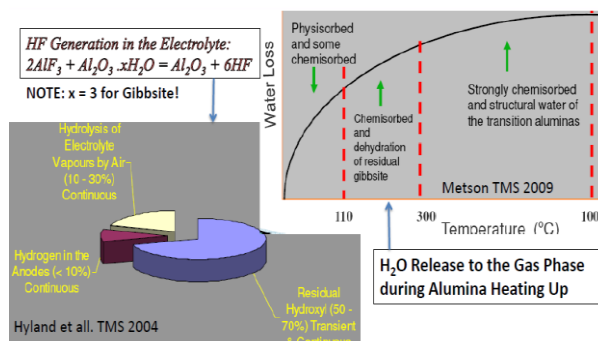
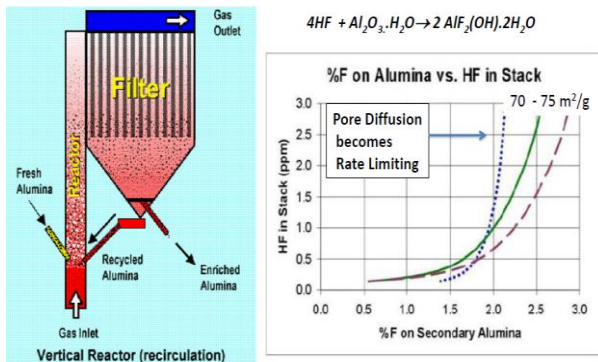


Figure 14: Water release from SGA.

In the Dry Scrubber / Gas Treatment Centre HF is chemisorbed on the SSA of the alumina (Lindsay,2009) as per the below reaction scheme (Figure 15).



**Figure 15: HF chemisorption on SGA.**

The rate of chemisorption of HF on the SSA of the SGA may be rate limited by **pore diffusion** subject to the duration of the contact time with SGA in the modern Dry Scrubber / Gas Treatment Centre.

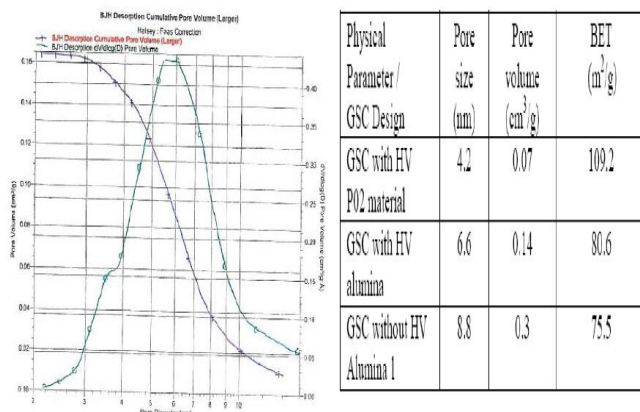
At low gas density or quite small pores, Knudsen diffusion (Satterfield) occurs and for a strict round pore,  $r_e$  [cm] the Knudsen diffusion coefficient is:

$$(2) D_K [\text{cm}^2/\text{sec}] = 9700 r_e (T/M)^{0.5}$$

Where T is the temperature in Kelvin and M is the molecular weight.

The development of **pore size** during Gas Suspension Calcination (GSC) is shown below (Figure 16):

The pore size distribution of alumina from GSC units is mono disperse with an average pore size of 6 – 9 nm. The range 6 - 8 nm is believed to be optimal (Meyer), though still not proven to be of importance for the HF adsorption capacity of alumina (Meyer, 2012, Perander, 2011).

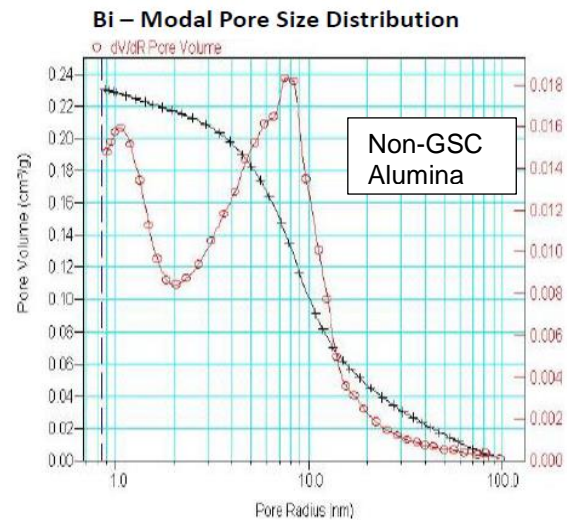


**Figure 16: Pore Size Development in GSC.**

Here it is worthwhile to remember that the molecular diameter of HF and H<sub>2</sub>O molecules respectively, is of the order of 0.1 nm (Jensen, 1965) and 0.2 - 0.3 nm, (Moore, 1962) only.

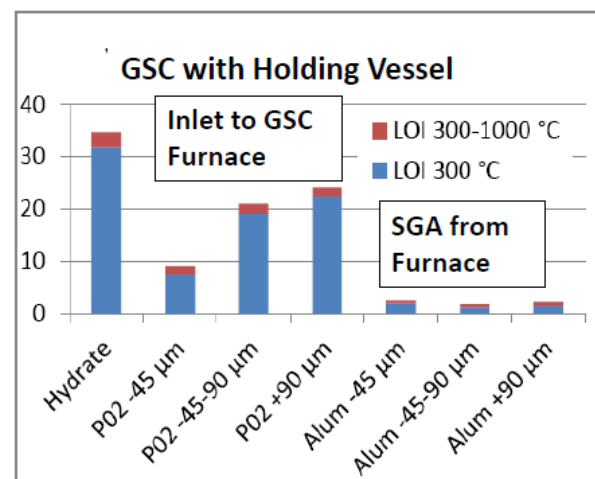
**HF adsorption** on plant samples is correlated with SSA, though not totally linear like for laboratory calcined samples from the same hydrate source [Coyne, 1989].

Anyway, maximizing SSA adsorption capacity can be achieved by producing alumina with a mono dispersed pore size distribution (Figure 16) by avoiding calcination using a hydrate, or pre-heated alumina by-pass (Figure 18) of the Calcination furnace.



**Figure 17: Calciner with Hydrate by-pass.**

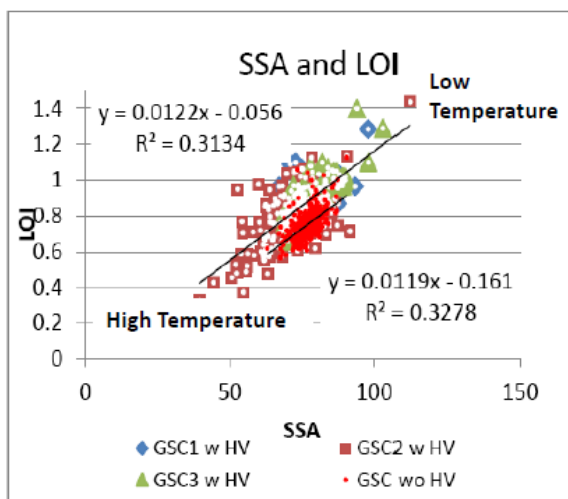
The reason for the bi-modal pore size distribution is due to mixing of under-calcined by-passed fraction of alumina (Figure 16) with small pores with an over-calcined fraction of alumina coming from the Calciner furnace with a large pore size, but relatively smaller SSA.



**Figure 18: LOI before and after GSC Furnace.**

Another way to increase the SSA sorption capacity is by lowering the **calcination temperature** (Figure 19):





**Figure 19: SSA versus LOI and temperature.**

The drawback or limitation of this modus operandi is that LOI increase with SSA and thus even more HF is generated during dissolution of the alumina requiring even more SSA.

In other words Mother Nature sets her boundaries.

### 3.3 Alumina Feeding, Dissolution and Crusting

Alumina **dissolution rate** in the liquid melt in the reduction cell is of primary importance for an economic (high current efficiency) and environmentally friendly (low frequency of anode effects) smelter operation (Wang, 2009).

Poor dissolution of alumina has many causes beyond the SGA quality coming out of the refinery (Table 6):

Cell Conditions	Impact on Dissolution of Alumina (Welch, 2007)
Mixing / Electrolyte Flow	More than a twofold impact on dissolution rate is observed by having good agitation that avoids aggregation – Poor mixing gives poor heat transfer and the combined effect gives extremely slow dissolution.
Alumina Concentration	With the normal operating range, increasing the alumina concentration can give 50% decrease in dissolution rate. However, at very low alumina concentrations, this impact is negligible – This is consistent with the basic dissolution equation and localized interfacial conditions.

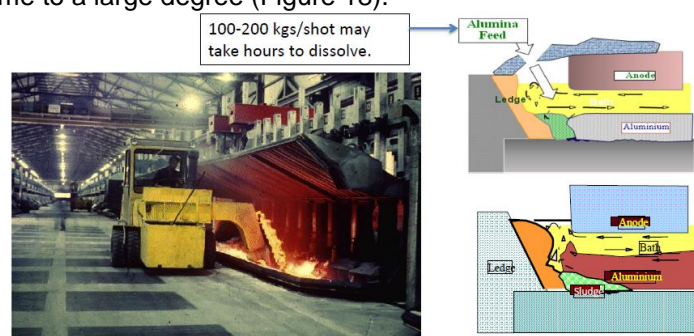
**Table 6: To be continued.**

AlF <sub>3</sub> Concentration	In the normal operating window of reduction cells (6-14%), this has been found to have 50% impact on the dissolution rate – The combination of low operating temperature and high AlF <sub>3</sub> concentration is worst.
Heat Transfer	Of similar importance to electrolyte agitation, poor heat transfer increases the tendency to form frozen aggregates. Worse for larger additions – High superheat helps reduce the extent of freezing.
Feeder Design	For fast dissolution it is important to have a steady, dispersing flow rather than “rapid” dump which can clump. – Keeping feeder holes open aids dissolution.
Mass of Feed	For a given accessible electrolyte volume, the smaller the the total mass of addition the faster the dissolution – Linked with both mixing and heat transfer.

**Table 6: Impact of Cell Conditions on Alumina Dissolution**

Both alumina handling and feeding system and Dry Scrubbr/Gas Treatment Centre may play a significant role (Lindsay, 2014) as well in changing the PSD of alumina arriving at the reduction cell.

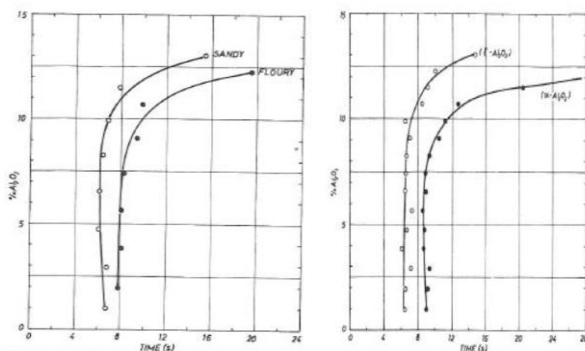
**Alumina feeding** method and rate or quantity per unit time depends on the design of the reduction cell (Welch, 2010) and determines the dissolution time to a large degree (Figure 18).



**Figure 18: Alumina feeding of SWPB cells.**

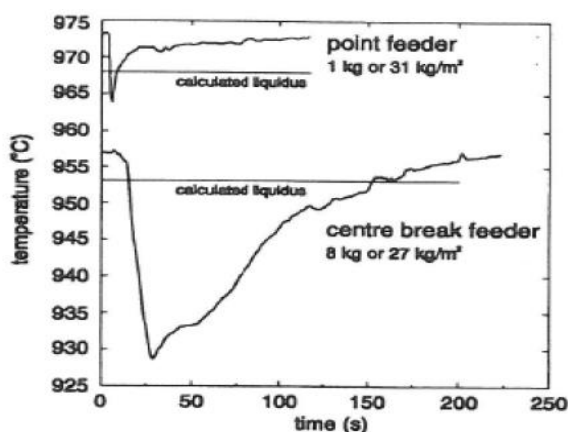
**Single particles** of alumina dissolve in seconds (Figure 19) regardless of degree of calcination i.e. Sandy or Floury, or Gamma or Alpha phase (Thonstad, 1972).





**Figure 19: Single Particle Dissolution Time**

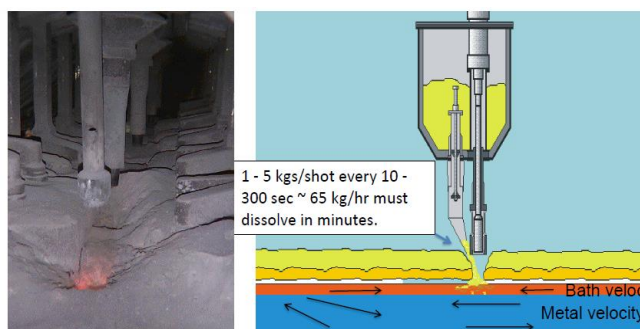
Below (Figure 20) is shown the effect on bath temperature for **point feeder** (Figure 21) on a PFPB reduction cell and a **centre break feeder** on a CWPB reduction cell (Walker et al, 1995).



**Figure 20: Temperature drop of Liquid Bath**

Consequently the design, operation and control of the feeding system are of paramount importance to obtain good dispersion of alumina to minimize the temperature drop and thus superheat requirements

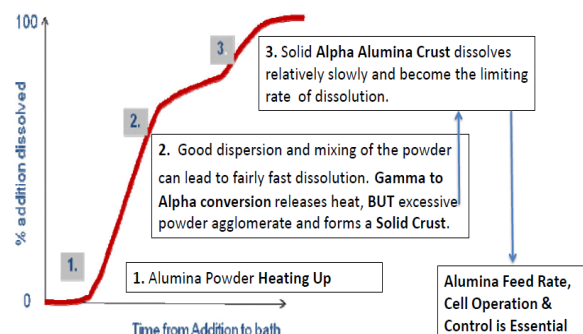
Alumina from Stationary Calciners with alpha phase content, say below 5-10% will require less **superheat** of the liquid melt than Sandy alumina from a rotary kiln with say 15-25% alpha content (Welch, 1983) at the same LOI.



**Figure 21: Alumina feeding of PFPB cells.**

**Dissolution** follows three distinct phases (Welch, 2010) as shown below (Figure 22):

- Alumina heating up by cooling the bath;
- Fast dissolution if good dispersion has been achieved.
- Slow dissolution of alpha crust formed owing to excessive feeding or poor dispersion.



**Figure 22: Dissolution Profile of Alumina.**

Before heat and mass transfer between alumina particle(s) and melted electrolyte can take place the alumina particle needs to be dispersed and wetted by the electrolyte bath. The **dispersion** is assisted by a relatively high LOI.

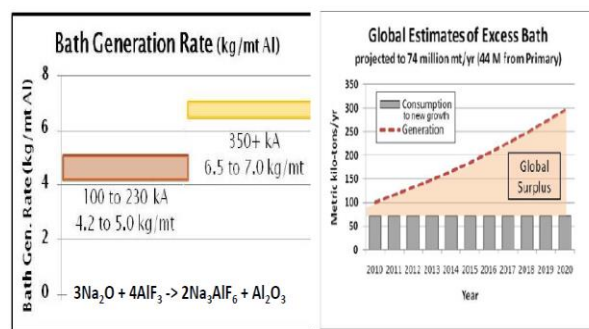
The amount of electrolyte bath to **wet** the whole surface per unit volume of alumina particle(s),  $m_v$ , increases with decreasing particle size:

$$(3) \quad m_v = 6 / d_p \text{ (units: } m^2/m^3 \text{)}$$

It is therefore understandable that ESP/Bag House dust with  $m_v = 883 \text{ (} m^2/dm^3 \text{)}$  requires much more liquid bath per unit volume or weight than product alumina including dust with  $m_v = 78 \text{ (} m^2/dm^3 \text{)}$  to be wetted assuming perfect dispersion into the bath.

The question then becomes, why not **control** alumina feed rate in proportion to  $m_v$  rather than mass (bulk density), as  $m_v$  is calculated by the laser analyser used for determining the PSD anyway (Yamamoto, 1973)?

### 3.4 Impact of major chemical impurities



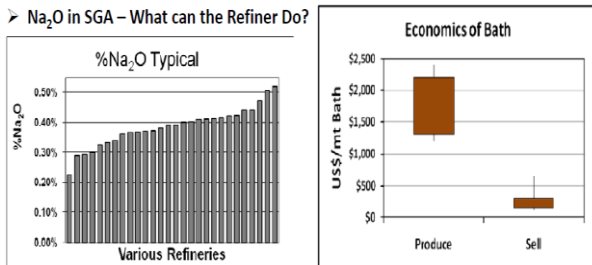
**Figure 23: Bath production from  $Na_2O$  in SGA**

**Sodium** from the SGA represents the largest chemical impurity introduced into the reduction cell

(Lindsay, 2012) and the bath generation rate increases with the cell size (kA) of the reduction cell (Figure 23).

The  $\text{Na}_2\text{O}$  content in SGA from various refineries averages about 0.40% and it is more expensive to produce excess  $\text{Na}_2\text{O}$  than it can be sold for (Figure 24).

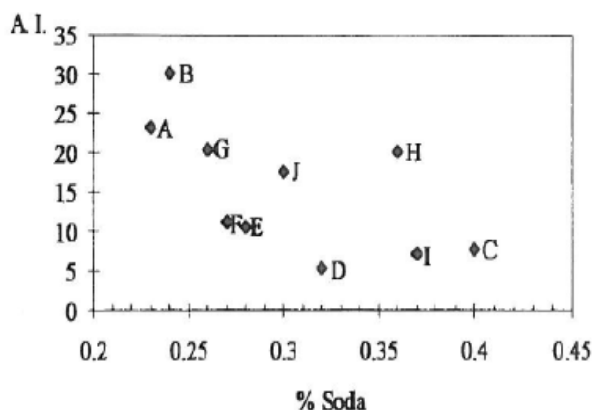
➤  $\text{Na}_2\text{O}$  in SGA – What can the Refiner Do?



**Figure 24:  $\text{Na}_2\text{O}$  in SGA and bath economics.**

What can the **refiner** do to mitigate this problem?

Produce a Hydrate Particle that, when calcined has alumina attrition index less than 15% and about 0.30 wt% soda (Armstrong, 1993), which seems to be the equilibrium value in modern pots (Homsí, 2001)?



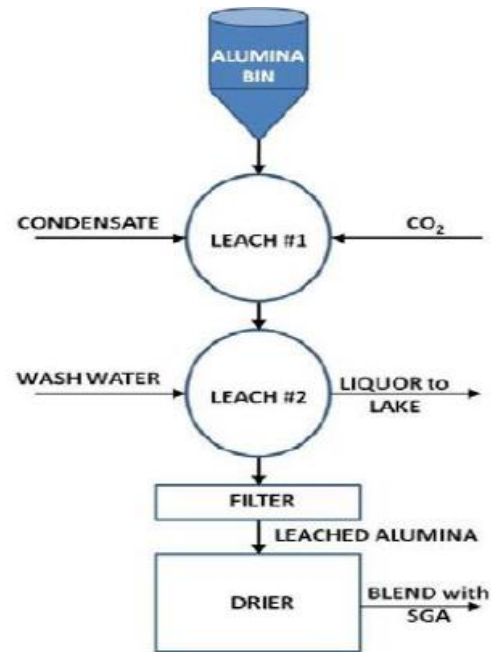
**Figure 25: Alumina AI versus %Soda in SGA**

**Lower Soda** content in the SGA, can be obtained from lower super saturation, but will reduce Refinery Yield & Productivity.

In addition hereto, decreasing the Soda content will also **increase alumina Attrition Index** (Figure 25) making a weaker alumina particle with a higher potential for particle breakdown on its route to the reduction cell.

Once more Mother Nature defines her limits.

However, Soda becomes Leachable after calcination and can be partly removed by a Leaching & Drying Process as shown (Figure 26).



**Figure 26. Soda reducing flow sheet**

Can or will the Smelter justify the higher alumina price to cover the additional Refinery cost for reducing the soda level in SGA?

#### 4. CONCLUSION

In view of the above the following conclusions and path forward from here is suggested:

- Refineries shall produce a smaller hydrate particle which upon calcination becomes strong with a **low Alumina Attrition Index**.
- Calciners shall be designed and operated with dust recycled to the hot part of the cooler system **minimizing hydrate by-pass**.
- Calciners shall be designed and operated to produce alumina with a **mono disperse pore size** distribution.
- SSA/LOI relationship** in Calciners shall be **optimized** to balance desirable SSA without excessive HF formation from too high LOI.
- Reduction cell **feed control** shall be based on the **specific particle surface**  $m_v$  ( $\text{m}^2/\text{m}^3$ ) rather than Bulk Density to match alumina feeding with wetting capacity of the liquid bath.
- Smelters consider **paying a premium** for SGA with equilibrium soda content and low attrition index to avoid excessive bath production and minimize particle breakdown at the Smelter.

#### 5. REFERENCES

Kjaer, A; (2010) "A case for replication of alumina plants", Light Metals, pp. 183-190.

Kelly,R; Edwards, M; Deboer, D; McIntosh, P; (2006) "New Technology for Bauxite Digestion",Light Metals, pp. 59-64.

Laros, T ,J, Baczek, F .A. (2009) "Selection of Sedimentation Equipment for the Bayer Process – An Overview of Past and Present Technology",Light Metals, pp.107-110.

World Aluminium (2013),"Bauxite Residue Management: Best Practises", European Aluminium Association, pp.1-32.

Bach,M; Weston, V; (2011),"Red Mud Filtration Test using AFP IV™ Automatic Filter Press", Travaux Vol.36 No. 40, pp.167-175.

Fish ,W .M;(1974) ,"Alumina Calcination in the Fluid-Flash Calciner",Light Metals, pp.673-682.

Tschamper ,O: (1981),"Improvements by the Alusuisse Process for Producing Coarse Aluminium Hydrate in the Bayer Process", Light Metals, pp.103-115

International Aluminium Institute;(2008), "Pioneering the Global Sectoral Approach",pp1-13.

Welch ,B .J and Kuschel ,G .I;(2007),"Crust and Alumina Powder Dissolution in Aluminium Smelting Electrolytes", Journal of Metals, pp.50-54.

Perander, L. M; Stam ,M .A; Hyland ,M. M and Metson, J. B; (2011),"Towards redefining the Alumina Specification Sheet-The Case of HF Emission", Light Metals, pp.285-290.

Lindsay, S.J; (2005),"SGA Properties and Value Stream Requirements", Proceedings of the 7<sup>th</sup> International Alumina Quality Workshop, pp.16-21

Chandrashekar,S; Jackson,S and Kisler,J.: (2005),"Alumina Fines' Journey from Cradle to Grave",Ibid,pp.5-9.

Taylor ,A; (2005),"Impacts of the Refinery Process on the Quality of Smelter Grade Alumina",Ibid pp. 103-107.

Lindsay, S.J; (2014),"Key Physical Properties of Smelter Grade Alumina", Light Metals,pp.597-601.

Kuschel,G .I and Welch,B .J; (1991),"Further Studies of Alumina Dissolution under Conditions Similar to Cell Operation", Essential Readings in Light Metals, Vol. 2: Aluminium Reduction Technology, pp.112-118.

Meyer, A.J. et al;(2012),"Examination of Drops in Bath Acidity Due to Change-Over of Alumina Qualities in the Sunndal Aluminium Smelter, Norway", Proceedings of the 9<sup>th</sup> International Alumina Quality Workshop, pp.316-321.

Welch,B.J.,Al Zarouni, Ali H.A.M., Lindsay,J.S.:(2010),"Modern smelting technology and its impact on alumina requirements",Travaux Vol. 35, No.39,pp.482-489.

Wind ,S; Jensen-Holm ,C;Raahauge, B. E, Raahauge: (2010)"Development of Particle Breakdown and Alumina Strength during Calcination",Light Metals p.17-24.

Sang ,J .V; (1987),"Factors Affecting the Alumina Products", Light Metals, pp.121-127. Chandrashekar,S; Jackson,S and Kisler,J; Ibid.

Homsy ,P; (2001),"Alumina requirements for Smelting", Proc. 7<sup>th</sup> Australasian Aluminium Smelting Technology Workshop, pp. 426-455.

Hyland, M. Patterson ,E. Welch,B.J. ;(2004), "Alumina structural hydroxyl as a continue source of HF", Light Metals, pp.361-366.

Satterfield, C .N;(1970)"Mass Transfer in Heterogeneous Catalysis", M.I.T.Press.

Perander, L.M; Ibid.

Meyer, A.J: et all; Ibid.

Jensen,K.A; (1965)"Almen Kemi I", Jul. Gjellerups Forlag, København.

Moore, W.J.; (1962),"Physical Chemistry",Longmans Green and Co. Ltd. London, UK.

Coyne,J.F., Wainwright,M. S.,Brungs, M. P., Bagshaw, A.N.:(1987),"The Influence of Physical and Chemical Properties of Alumina on Hydrogen Fluoride Adsorption", Light Metals,pp. 35-40.

Lindsay, S.J; (2009),"Dry Scrubbing for Modern Pre-Bake Cells", Light Metals, pp.275-280.

Thonstad ,J. Nordmo ,F. and Paulsen, J. B; (1972),"Dissolution of Alumina in Molten Cryolite", Met. Trans.6B,pp. 83-86.

Walker, D.I., Utigard,T.A.,Taylor, M: (1995) "Alumina Agglomerates in Aluminium Smelters",Light Metals, pp.425-434.

Jain, Tricklebank, S. B. Welch, B. J.; (1983)  
"Interaction of Aluminas with Aluminium Smelting  
Electrolyte", Light Metals, pp.609-622.

Lindsay, J.B.; (2012), "Customer Impacts of Na<sub>2</sub>O  
and CaO in Smelter Grade Alumina", Light Metals,  
pp. 163-167.

Armstrong, L. (1993), "Bound Soda Incorporation  
During Hydrate Precipitation", Proceedings of the  
3<sup>rd</sup> International Alumina Quality Workshop  
pp.282-292.

Homsy, P.; Ibid.