

## Industrial Prospects and Operational Experience

with 32 mtpd Stationary Alumina Calciner

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A new simple approach based on a flowing gas/solids suspension with a few seconds' retention time, and utilizing simultaneous combustion and calcining in the reactor volume at 1150-1450°C has been developed by F.L. Smidth & Co.

This paper explains the FLS Calciner principles, and test experience obtained from a 32 mtpd pilot plant is reported.

Preliminary construction and operational principles of a 1000 mtpd plant layout are presented, suggesting the following advantages compared to other stationary calciners:

- i) Reduced maintenance costs from simple design
- ii) Estimated specific power consumption of about 15 kWh per ton alumina.
- iii) Small investment costs owing to simple design.

These advantages are obtained with a competitive heat economy of about 3180 MJ per ton alumina, and without losing generally accepted merits of fluid-bed technology compared with rotary kilns.

Particle breakdown caused by using a very fragile hydrate and excessive gas velocities has been unacceptable, and new test runs are in progress in order to verify a solution. Other requirements as to reduction grade alumina are met, and can be controlled by the temperature level in the calciner.

1. Introduction

The final step in the Bayer process is calcining of aluminium hydroxide to alumina.



The overall reaction is strongly endothermic and requires addition of heat.

Today the calcining process is conducted industrially in rotary kilns and in stationary calciners utilizing fluid-bed technique.

Conventional rotary kilns comprise a long kiln equipped with a planetary cooler for primary cooling. A water cooled fluid-bed cooler is used as secondary cooler. In fact, this cooler represents one of the earliest applications of fluid-bed technique within the alumina industry. A large number of more than 60 rotary kilns supplied by F. L. Smidth & Co., are equipped with this type of cooler. The latest rotary kiln made by FLS was started up in 1978, but new calcining capacity and replacement of old rotary kiln installations will no doubt call for stationary calciners.

The reasons for this are attributed to the following main advantages of stationary calciners (1, 2):

- i) Specific heat consumption is 3140-3350 MJ per ton alumina compared with 4190-5020 MJ per ton for rotary kilns due to smaller process heat losses.
- ii) Investment costs are lower.
- iii) Stationary calciners require less space area due to a more compact design.
- iv) Maintenance costs are lower, mainly due to a prolonged life of the kiln lining.

Along with these main advantages the following drawbacks are widely accepted:

- i) Application of fluid-bed technology up till now has led to higher specific power consumption: 21-23 kWh per ton alumina, compared with 13-15 kWh per ton for rotary kilns.
- ii) Less product flexibility with respect to the product quality produced, as hard burnt flourey alumina cannot be obtained.
- iii) Degree of particle breakdown was and still is generally higher in stationary calciners compared to long rotary kilns, owing to a more gentle handling in rotary kilns.

The larger power consumption has been accepted by the alumina refiner in view of the substantial savings in fuel consumption, maintenance, and capital costs.

As smelters are generally changing their demands from fine, hard burnt flourey alumina to coarse, sandy alumina (3), the only serious drawback to stationary calciners is the problem of particle breakdown (4). Particle breakdown or attrition is, in fact, an old and well known problem from appliactions of fluid-bed technique to catalytic cracking in the oil industry (5, 6).

The term attrition (7) means all mechanisms of solid size reduction except those caused by chemical reaction alone. Therefore, attrition covers (1) decrepitation, (2) thermal shock shattering, (3) bursting apart or fragmentation during reaction, (4) abrasion among solid particles, bed walls, and internals, (5) distributor jet impingements, and (6) abrasion in solid circulation ducts (including cyclones) to and from the bed.

Very few studies have been reported on the attrition of solids in fluidized beds (4, 8, 9). Only the studies by Blinichev et. al. (9) appear to give some insight into fluid-bed design parameters which contribute to attrition. The following empirical correlation based on dimensional anlysis and evaluation of experimental data was established:

$$W = K_1 \times K_2 \times v^{5.8} \times t_{\text{mean}}^{0.6} \quad (2)$$

where W = ratio of dust generation to material in the fluid-bed. (Kg/ton).

K<sub>1</sub> = constant accounting for material properties including mechanical strength.

K<sub>2</sub> = constant accounting for constructional features of the gas distribution grid.

v = linear air velocity in empty tower cross section m/sec.

t<sub>mean</sub> = mean retention time of particles in the apparatus.

The full correlation has been checked in a 100 ton/h industrial fluid-bed drier, drying potassium chloride, and a good agreement is claimed.

The important deduction made from the stated correlation is that attrition or particle breakdown in fluidized beds depends on (1) material properties including mechanical strength of material, design of gas distributor, and on the fact that (2) particle breakdown increases with gas velocity and mean particle retention time in the bed. Blinichev et al. have not included a Bayer hydrate in their investigations. Several Bayer hydrates were compared in a recent publication (4) by Schmidt et al. emphasizing the importance of producing a mechanically strong hydrate.

The importance of gas distributor design also seems verified (4), as each opening in the grid may operate like an element of a jet mill, as pointed out by Blinichev et al.

The effect of the linear gas velocity in the bed itself is more doubtful even though there is a violent agitation of the solids in the bed.

Zenz has investigated cracking catalyst and found (8) that high velocity particle impact in cyclone collection systems is the main effect in fluid-bed systems as higher gas velocities are applied in cyclone inlets than in the fluid-bed itself.

This conclusion also seems verified with respect to Bayer hydrate (4) calcined in a circulating fluid-bed calciner.

From the viewpoint of the alumina refiner, complete advantage of introducing stationary calciners may require expensive modifications in the hydrate cirquit of the Bayer process in order to meet the demand of a mechanically strong, coarse sandy alumina.

## 2. Principle of FLS Stationary Calciner for Alumina

In view of the change in calciner technology and F.L. Smith's long experience in the field of gas suspension calciner principles from the cement industry (10 to 13) it was natural to attempt an adaption of this technology to alumina.

The development work of the FLS stationary Calciner has now progressed so far that it is a viable, economically attractive alternative to fluid-bed technology.

The FLS calciner principle is seen from fig. 1 below.

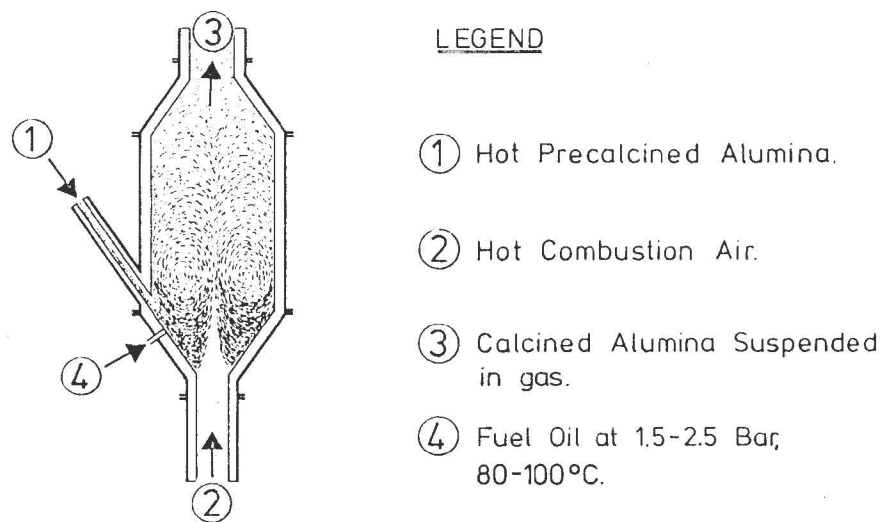


Fig. 1: Flow Pattern in the FLS Calciner.

The calcining furnace comprises a cylindrical vessel with conical bottom. The preheated combustion air coming from the cooler is introduced at the bottom of the furnace with a high velocity and a temperature of 850-1000°C. The 300-400°C precalcined alumina is discharged into the reactor volume in a direction paralleling the conical bottom of the furnace, and just above the inlet nozzle for fuel oil. The calcined alumina leaves the furnace at the top with a temperature of 1150-1450°C, dependent on the product specifications desired.

The reactor is not equipped with any sort of gas distributor plate etc., but relies on a torus shaped vortex zone which is formed in the lower conical part. The mixing of hot material and preheated fuel, combined with recirculating hot combustion products from the upper part of the vortex, causes a partial vaporization and cracking of the fuel. The combustion proceeds flamelessly and uniformly upwards through the furnace with hot particles surrounded by burning gas. The temperature of the gas is only slightly above the temperature of solids at this stage, due to a very high heat transfer rate between gas and particles. A very high combustion intensity of the order of 60 (kcal/sec-m<sup>3</sup>) has been realized in commercial cement plants (13).

There are two states of solids-flow in the furnace. One for the lower part of the calciner where the vortex zone persist, and one for the rest of the calciner volume. Solids-flow in the vortex zone is in a state of back-mixing. Solids-flow above the vortex zone is in a state of plug-flow as in continuous pneumatic transport. This means that the larger particles, owing to their higher terminal velocities, will have longer retention time than the smaller ones. Thereby a uniform calcination, regardless of particle size, results as seen from table I below.

Table I: Calcining Uniformity by Particle Size

Properties	Alpha alumina content (%)				B.E.T. Surface area (m <sup>2</sup> /gr)
	+74 μ	+63 - 74 μ	+44 - 63 μ	-44 μ	
Hydrate 1	75.2	75.7	76.2	79.0	15.5
Sample No.	16.7	14.5	14.5	13.6	48.1

It is also to be noted that the calcination is uniform both at high and low degrees of calcination.

In a way the FLS Calciner is selfregulating in the sense that solids with a wide size distribution will also have a wide retention time distribution. The retention time of the solid particles in the FLS calciner is of the order 1-5 seconds.

The retention time can be varied within certain limits as long as a certain minimum of gas velocity is ensured in order to prevent the collapse of the gas/solid suspension.

The reaction principles outlined above ensure an even temperature profile along the calciner axis as seen from figure 2, showing a typical temperature profile.

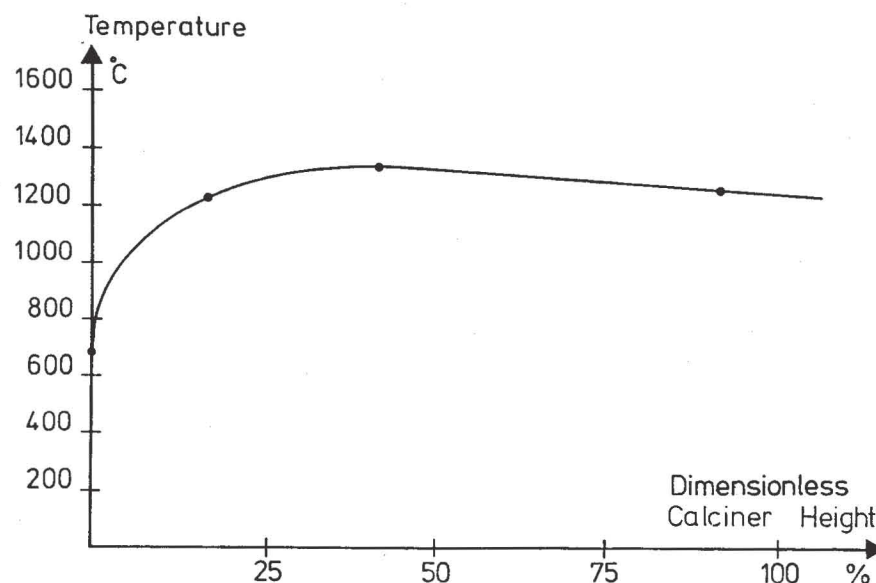


Fig. 2 Temperature Profile in FLS Calciner.

The loss on ignition of the material entering the calciner is about 5 to 10% so that the main part of the heat released during the flameless combustion is used for heating of the material to the desired reaction temperature at which the exothermal conversion of  $\gamma$ -Al<sub>2</sub>O<sub>3</sub> to  $\alpha$ -Al<sub>2</sub>O<sub>3</sub> proceeds very rapidly. From the shape of the temperature profile it can be seen that the main part of the combustion process is finished within the vortex zone of the calciner. Thus when leaving this zone each particle acts as a small auto-thermal batch reactor where the exothermal conversion to  $\alpha$ -Al<sub>2</sub>O<sub>3</sub> supplies the heat necessary for final calcination. The final calcination process is thus terminated within a few seconds compared with several minutes in other stationary calciners (1, 2).

3. Test of FLS Calciner Principles

The first experimental test was conducted in our small pilot plant 1 with a capacity of 200 kg/h of calcined alumina. This test was conducted in January 1978 with a rather fine Bayer hydrate with a granulometry as seen from table II.

Table II: Granulometry of Hydrate No. 1

Microns	128	64	32	16	8	4	2
% passing	97	73	26	9.5	5.0	3.3	2.7

From laboratory experiments it was known that the primary variables determining the amount of  $\alpha\text{-Al}_2\text{O}_3$ , and thus LOI and BET surface are the calciner temperature and retention time. These expectations were confirmed and, furthermore, it was demonstrated that large amounts of X-ray amorphous anhydrous alumina were formed during flash conditions above 400°C, resulting in an increased rate of calcination (14) and formation of  $\alpha\text{-Al}_2\text{O}_3$ . Thus it was possible to obtain any degree of  $\alpha\text{-Al}_2\text{O}_3$  up to approximately 80% with a gas retention time between 0.9-1.4 seconds.

Figure 3 shows the correlation obtained between  $\alpha\text{-Al}_2\text{O}_3$  content and BET area for hydrate 1 and 2.

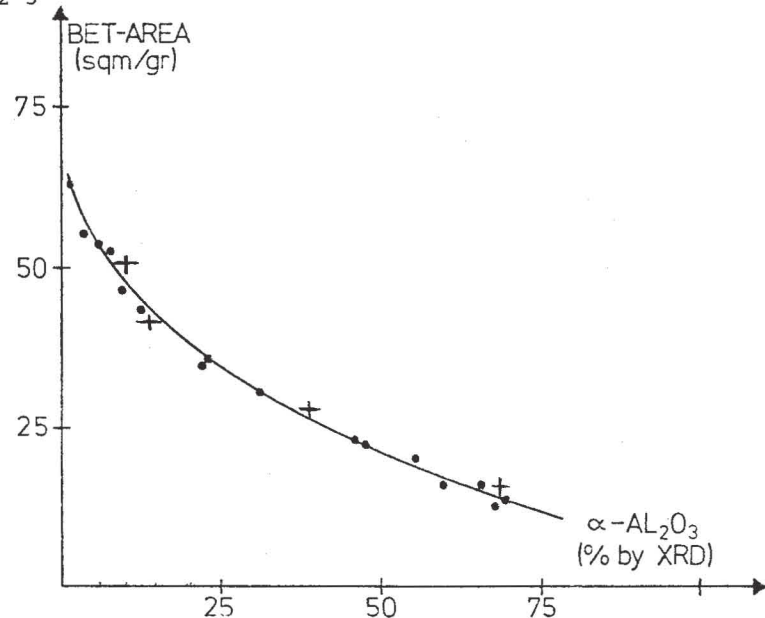


Fig. 3: BET Area versus  $\alpha\text{-Al}_2\text{O}_3$  Content by X-ray, Hydrate 1(+), 2 (.)

The correlation shows the usual properties of alumina from stationary plants compared with rotary kiln alumina (3). A lower content of  $\alpha\text{-Al}_2\text{O}_3$  for the same specific BET surface area than found in rotary kiln alumina.

From X-ray diffractograms of the alumina samples the following minerals were found:

Table III: Minerals in Samples from FLS Alumina Calciner

Content BET area sq. m/gr	Qualitative content of mineral phases		
	Large	Small	Traces
15.5	$\alpha\text{-Al}_2\text{O}_3$	-	$\gamma\text{-}$ , $\theta\text{-Al}_2\text{O}_3$
31.4	$\alpha\text{-Al}_2\text{O}_3$	$\gamma$ , $\delta\text{-}$ , $\theta\text{-Al}_2\text{O}_3$	-
48.1	$\gamma\text{-}$ , $\delta\text{-}$ , $\theta\text{-Al}_2\text{O}_3$	$\alpha\text{-Al}_2\text{O}_3$	-

As seen from the table the minerals found in these samples are quite normal which - in conjunction with the above shown correlation between  $\alpha\text{-Al}_2\text{O}_3$  and BET surface - suggests that dissolution properties in smelter bath will be similar to conventional alumina.

The test of the calciner principles proved that

- o The FLS Calciner principle works well on calcining aluminium hydroxide to alumina.
- o The alumina obtained from the FLS-Calciner is flexible with respect to specifications on smelter grade alumina.
- o The particles in the calciner require only a few seconds' retention time, regardless of the alumina specification desired.
- o Any degree of  $\alpha\text{-Al}_2\text{O}_3$  up to 80% can be achieved by operating the calciner at the proper temperature.

The promising result achieved by adaption of the FLS-Calciner principles to alumina calcining was followed by further test work.

4. Test on Large Scale 32 mtpd Pilot Plant

Plant Concept

New experiments were conducted in spring 1979 with the objective of testing the viability of a commercial plant designed around the FLS Calciner, and of evaluating operational parameters.

Figure 4 shows a flow sheet from our 32 mtpd pilot plant installation at the Dania works in Denmark.

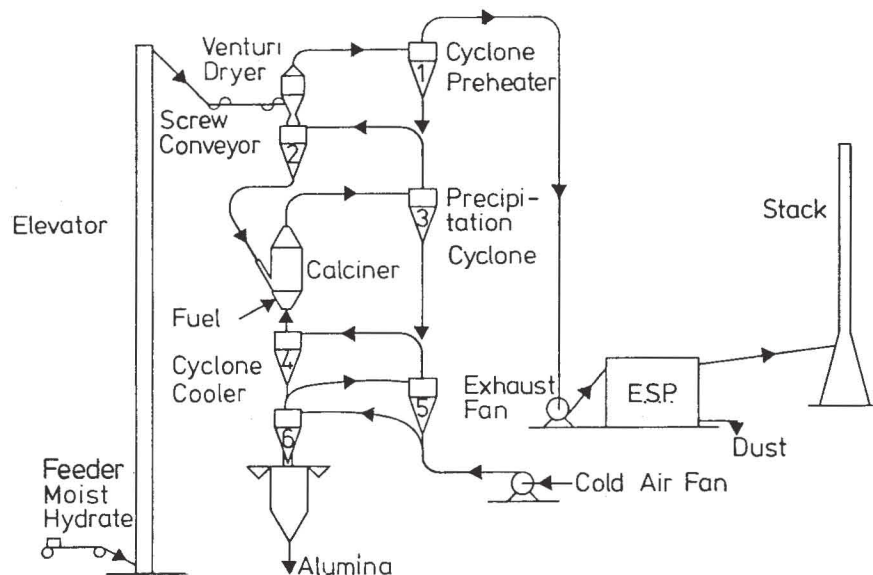


Fig. 4: 32 mtpd Pilot Plant for Alumina.

The plant concept comprises a 2-stage cyclone preheater including a venturi drier, the FLS Calciner with precipitation cyclone, and a 3-stage cyclone cooler. The installations were originally built for calcination of limesand, but was modified to accept a feed of moist Bayer hydrate of European origin.

Operational Experience from 32 mtpd Plant.

The plant was fired with heavy fuel oil and operated for nearly 3 months. The main problems encountered during the test run were caused by handling problems with moist hydrate, particularly due to ambient temperatures below 0°C most of the time.

The process itself performed in accordance with our best expectations, and it was proved that this very simple plant concept could be operated without problems.

The 8-12% free moisture from the wet hydrate is evaporated in the venturi drier by heat exchange with hot gases from the second preheater cyclone. Dried hydrate of approximately 165°C is discharged from the upper preheater cyclone into the very hot gases from the precipitation cyclone and flash calcined in a fraction of a second before entering the second preheater cyclone.

In this process, characterized by an extremely high rate of heat transfer, LOI of dry hydrate equal to 34.6% is reduced to approximately 5-7% for the partly calcined alumina entering the calciner at a temperature 320-400°C, depending on the calciner temperature. The exit temperature of the calciner has been varied between 1055°C and 1430°C, resulting in a specific internal BET surface area from about 100 sqm. per gramme down to 20 sqm. per gramme. The hot alumina leaves the calciner as a lean gas/solid suspension and enters the precipitation cyclone from where it is discharged into the multistage cyclone cooler.

The hydrate feed rate to the test plant has varied between 1.5 ton per hour up to 2.5 ton per hour, indicating a flexible throughput for this type of plant. The calorific consumption for the pilot plant has varied with the test conditions, and the consumption figures were generally higher than those calculated for a commercial scale unit, owing to the size of the pilot plant.

The pressure drop across the plant was of the order of 500-800 mm H<sub>2</sub>O depending on gas flow. In fact, there is no pressure drop across the calciner due to the buoyancy effect.

The dust emission from the upper preheater cyclone was between 1-2% of the alumina production. In some of the test runs dust from the electrostatic precipitator was recycled to the elevator without causing operational problems.

Particle Breakdown

The granulometry of the test hydrate is seen from the right hand curve in figure 5.

The residue on 325 mesh screen of the hydrate has varied between 90.4% and 93.0% determined by Alpine classifier.

The residue of the alumina on 325 mesh screen was determined to 84% or less.

As minimum 90% residue is required for today's alumina specification a special test programme was set up to investigate the causes of the observed particle size reduction in further detail.

Offhand the following factors were considered for dominant influences:

- 1) The type of hydrate used
- 2) Gas/solids velocities in the calcining system, especially where abrupt changes in the flow direction of the lean gas/solid suspensions took place.
- 3) Thermal effects like vapour pressure, induced explosion or rapid thermal stresses, due to temperature increase in calciner.
- 4) Degree of calcination.

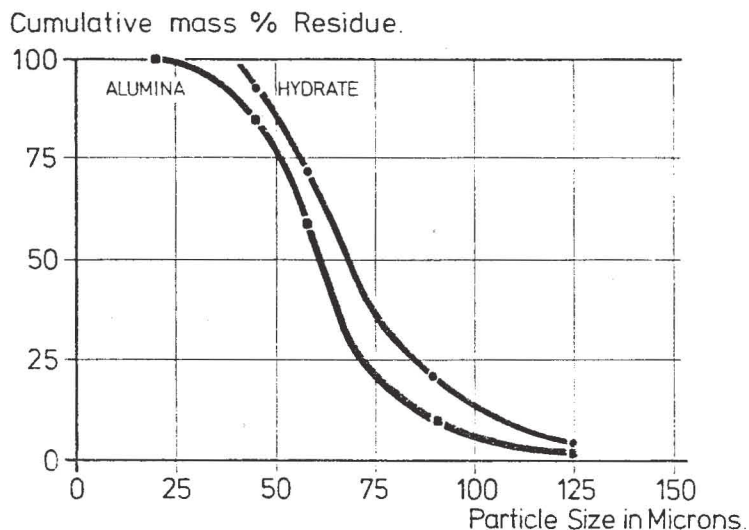


Fig. 5: Granulometry of Test Hydrate and Alumina.

Particle concentration and retention time is very small compared with fluid-bed systems and therefore this factor was given no consideration. The result of the investigation is summarized in table VII.

i) Influence of the Hydrate

Mechanical strength, size, structure, and form of any hydrate particle is laid down in the precipitation area of the Bayer plant. Many factors are decisive to the ultimate properties of the hydrate with respect to calcination, and their interaction is complex.

The hydrate used was indeed very fragile, and perhaps far from being representative for coarse Bayer hydrates in general. The fact that the hydrate was fragile is shown in table V below.

Table V: Hydrate Strength from Attrition Test

% Less than 44 microns	Initial	Change
Hydrate tested at Dania	6.7%	+14.2%
Another European hydrate untested	9.7%	+1.0%

It has not been possible to meet the requirements with respect to minimum 90% residue on 325 mesh with the hydrate tested in the pilot plant. On the other hand, application of a very fragile hydrate made it possible to localize and examine critical factors causing particle breakdown, which resulted in very useful design information.

ii) Influence of Gas/solid Velocities

No gas distributor or high velocity air nozzles or high material concentrations are present in the plant concept. Investigations of the influence from gas/solid velocities were confined to points of deflection in gas/solids carrying ducts, and cyclone inlets. It was found that abrupt deflections are very critical at too high gas velocities. This influence can, however, be almost eliminated by correct designing and sizing of ducts. Also the influence of cyclone inlets can be minimized by decreasing the inlet velocity down to the point where the cyclone performance as a separator becomes critical. It has not been possible to separate the influence from particle-wall and particle-particle collisions, but due to the low material concentration in the whole plant, particle-wall collisions are believed to dominate. In the pilot plant there will always tend to be a relatively large influence from small diameter cyclones, operating at velocities identical to full scale cyclones, because of a shorter deceleration distance.

iii) Influence of Thermal Effects

The intensive heat transfer rates prevailing in stationary calcining systems expose the hydrate to higher thermal stresses during calcination than is the case in long rotary kilns. In the FLS plant concept precalcination takes place at higher temperature gradient than in other stationary systems, due to the higher calciner temperature demanded by the very short retention time in this system.

Offhand it might be expected that a rapid increase in vapour pressure could lead to an explosionlike fragmentation of the hydrate particles, if the diffusion of water vapour is restricted too much.

Excessive vapour pressure may develop from two sources:

- a) By the flash calcination of chemically bound water, and
- b) By inclusions of liquor in the particles from the precipitation area.

At sufficiently small gas velocities in the preheater section, where the flash calcination takes place, no particle breakdown could be observed. This observation confirms previous findings (4).

The influence of different heating rates of the material in the calciner was also tested, and in this case as well, particle breakdown could not be explained from this effect.

iv) Influence by Degree of Calcination

As calcination proceeds, surface texture changes, as well as the true density and pore volume (15, 16, 17). The BET specific area decreases from more than 200 sqm per gram at the inlet to the calciner down to 15-60 sqm per gram, dependent on desired properties of the alumina aimed at. Fissures develop in the single crystals forming each particle simultaneously with the dehydration. These fissures decrease the mechanical strength of the particles which then become more sensitive to breakdown by mechanical means.

In order to eliminate any mechanical effects from calciner operation a laboratory test was conducted for determination of the minimum expectable breakdown resulting from calcination and heating up rate alone.

Table VI: Particle Breakdown by Laboratory Calcination

Sample test conditions	Time	Temperature	+45 microns
1 Dry Al(OH) <sub>3</sub>		110°C	91.6%
2 Slow heating	12 hr.	1250°C	88.4%
3 Slow heating		500°C	
Quick heating	2 min.	1450°C	88.8%

From the figures in the table it appears that the same degree of particle breakdown is observed, regardless of heating rate and temperature level. From observation of scanning electron microscope pictures, cracks between the single crystals could be observed. These cracks were larger than fissures developed in each crystal during calcination. This observation suggests the conclusion that shrinkage of single crystals forming the particles during calcination, contribute to particle breakdown by developing stresses on the crystal interfaces in each particle.

In fact, similar observations were reported already in 1962 (17), where it was shown that particle strength decreases with an increasing degree of calcination. This again points to the importance of particle structure laid down in the hydrate plant, as large particles made up of a few, but large, crystals with a relatively small crystal interface area are likely to be more resistant to particle breakdown than particles of the same size which are composed of many, but small crystals with a large crystal interface area.

Table VII summarises the results of testing the influence of different factors on particle breakdown.

Table VII: Influences on Particle Breakdown

Potential Factor of Importance	Conclusion
1. Hydrate properties	Very important
2. Gas/solid velocities	Very important
3. Thermal effects	Negligible
4. Degree of calcination	Important

As a conclusion to the findings reported above, particle breakdown was confined to ducts connecting the calciner and precipitation cyclone and in the multistage cyclone cooler, due to excessive gas velocities in connection with abrupt changes in flow directions. Increased particle breakdown was observed with increasing degree of calcination.

Physical Properties of Alumina

The desired BET specific area of alumina from the FLS Calciner is obtained by adjusting the calciner temperature as seen from figure 6.

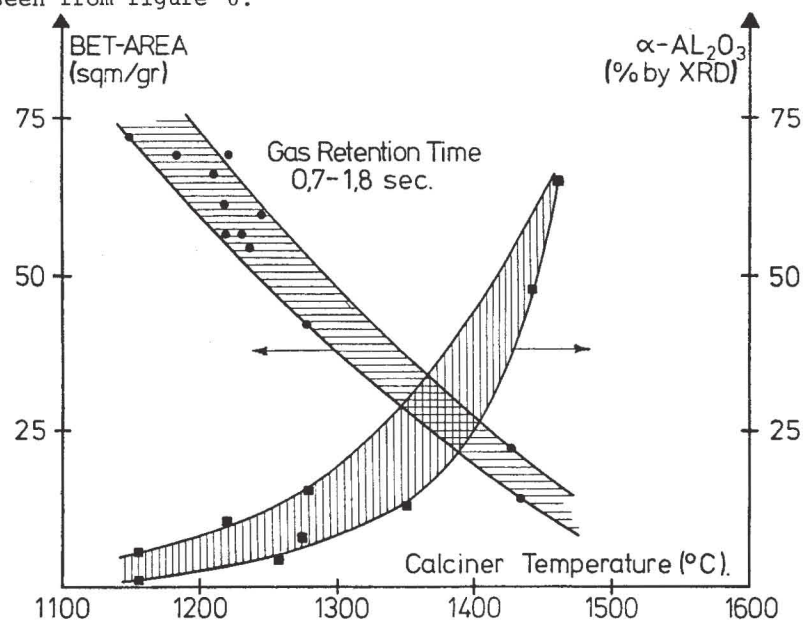


Fig. 6. BET Area and α-Al<sub>2</sub>O<sub>3</sub> Content.

The different values are obtained in different test runs with different values of gas retention time and calciner temperature.

The alumina from the FLS calciner is calcined very uniformly, regardless of particle size, as seen from table VIII.

Table VIII: Uniformity in Calcination

Hydrate No. 2		Size Fractions in Microns			
Sample	Property	-45 $\mu$	+45/-63 $\mu$	+63/-90 $\mu$	+90 $\mu$
1	BET (sqm/gr)	15,0	15,3	16,6	16,5
	LOI (%)	0,2	0,1	0,2	-
2	BET (sqm/gr)	44,0	44,4	47,4	45,3
	LOI (%)	0,4	0,3	0,3	0,4
3	BET (sqm/gr)	72,1	73,8	77,5	77,7
	LOI (%)	0,8	0,7	0,8	-

This is attributed to the working principles and the flameless combustion by which any local overheating is avoided. The uniformity is seen to be independent of the degree of calcination.

5. Industrial Prospects of FLS Calciner

Layout of Commercial Plant - Constructional Advantages

The FLS Calciner is an extremely simple reactor with no mechanical parts placed in the direction of solids and gas flow. Simplicity in plant design and reliability in operation are main design objectives. The design is based on our long experience with gas suspension calciner principle in the cement industry and application of fluid-bed technology in the alumina industry (18). The basic layout comprises a 2-stage cyclone preheater including a venturi drier, the FLS Calciner with precipitation cyclone, and a multistage cyclone cooler. Heat exchange in the single cooler stages takes place in a co-current flow of air and alumina. Good temperature equilibrium is achieved, due to the small particle size of the alumina. By arranging the cyclones stagewise, countercurrent heat exchange between alumina and combustion air is approximated.

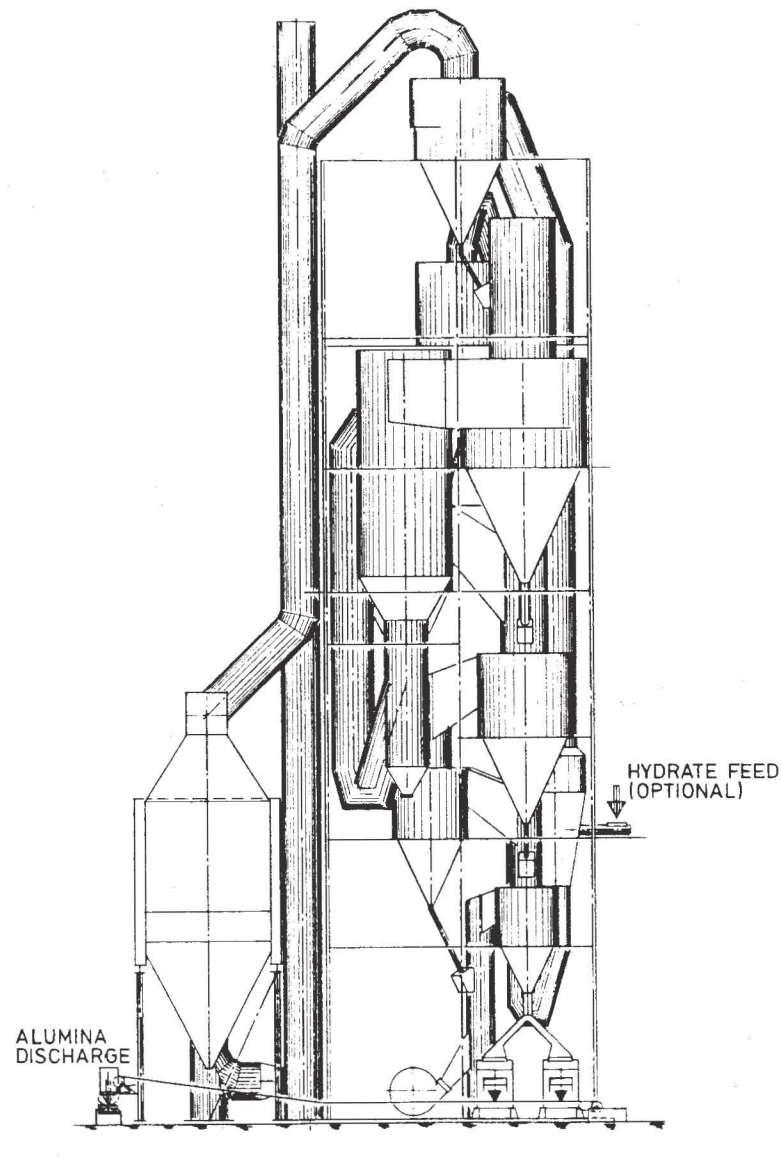


Fig. 7: Plant Layout of 1000 mtpd FLS Plant



With the current oil price levels we have found that a 4-stage cyclone cooler is about to be the most attractive solution seen from an economic point of view. In future, it is evident from the plant concept that the number of preheating and/or cooling stages can easily be increased without losing simplicity in design and operation so that overall economics with respect to minimum production costs can be complied with.

Figure 7 shows the plant layout in its current stage of development.

The height of the 1000 mtpd unit is 48 m and the ground floor area is 25 m x 20 m (500 sqm.). The altitude of hydrate inlet to the venturi drier is made flexible by using the relatively cold flue gas for transportation of the dry hydrate to the upper preheater cyclone. The inlet point can be varied from a minimum of 15 m above ground level and upwards. Thereby flexibility is obtained, ensuring easy implementation in new plants as well as in plants, where existing filter installations are to be used. The alumina from the cyclone cooler is discharged into conventional fluid-bed coolers and cooled down indirectly with water to a temperature which is acceptable to the conveyor system after the calcination plant.

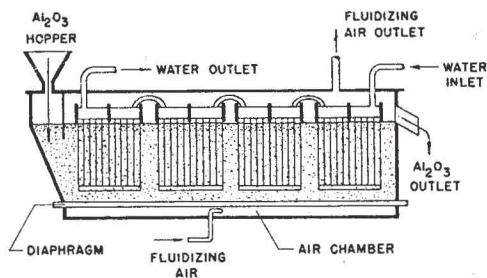


Fig. 8: Secondary Fluid-bed Cooler (18).

This type of water cooled secondary cooler has been used by F.L.Smith & Co. throughout the aluminium industry since its first appearance in the 50's and onwards. As the air requirements for fluidization are very small compared to the requirement for combustion, dimensions of the secondary coolers can be kept relatively small.

The simplicity in design and construction of the new FLS Calciner suggests the following advantages (1, 2):

- Addition of hot combustion air through one pipe without any air distribution grate or nozzles, giving low maintenance costs and simple air piping.

- No recirculation of solids or application of external combustion chamber, giving simple reactor design and simple separation equipment for the calcined product.
- Cooling of alumina in a multistage cyclone cooler by stagewise countercurrent preheating of combustion air is a simple and flexible solution.

The development of the optimal plant layout is still in progress in order to maximize simplicity and reliability and to minimize capital costs.

Process Principles - Operational Advantages.

The main problem in applying fluid-bed technique, when simultaneous combustion must take place within the bed, is that the air requirements for combustion are much larger than the air requirements for fluidization, when large amounts of fuel are required.

Different approaches have been used by Alcoa (2) and Lurgi-VAW (1) to overcome this problem which was initially a serious restriction to increase of plant throughputs above 250 ton per day (2).

The FLS-Calciner principles utilize simultaneous combustion of fuel oil and calcining of partially anhydrous alumina in a lean gas/solid suspension passing the calciner almost in plug-flow, resulting in very high gas throughput capacity and a short retention time of the solid.

By means of this approach a very high plant throughput can be achieved, and scaling-up can be kept reasonably simple. This is reflected in experience gained from the cement industry where plants sized up to 4000 mtpd cement have been put on stream (13).

The specific heat consumption for a 1000 mtpd unit is estimated in table IX.

Table IX: Specific Heat Consumption

Source	MJ/ton	%
Fuel oil	3160	97
Hydrate 65°C, 12% moisture	110	3
<b>Heat input</b>	<b>3270</b>	<b>100%</b>
Heat of reaction	2040	63
Evaporation of moisture	500	15
Surface loss	230	7
Combustion gas, 160°C	175	5
Water vapour + dust, 160°C	195	6
Product loss, 160°C	130	4
<b>Heat expenses</b>	<b>3270</b>	<b>100%</b>

The calculation above assumes an outlet temperature of 1200°C of the gas and solids leaving the precipitation cyclone (fig. 4), and a 4-stage cyclone cooler.

Some smelters prefer a hard burnt sandy alumina with a BET surface of 30 sqm per gramme. This specification can be met by increasing the temperature of the FLS Calciner, as the retention time is limited to a few seconds.

The associated change in specific heat consumption will not increase seriously because:

- 1) A higher calciner temperature results in a higher degree of precalcining of the feed to the calciner and thus a smaller endothermal reaction heat requirement in the calciner itself.
- 2) A higher combustion air temperature is achieved, due to a higher calciner outlet temperature.
- 3) A larger exothermal heat release, due to a higher degree of conversion to  $\alpha\text{-Al}_2\text{O}_3$ .

Therefore, a specific heat requirement not far from 3160 MJ per ton alumina is expected, regardless of the alumina specification aimed at.

This means that the calciner temperature is the only control parameter required in order to comply with varying product specifications.

Taken together with a small material hold-up in the whole calciner plant, operating conditions can be adjusted to new product specifications very easily without producing lots of material with undesired specifications.

Both the primary multistage cyclone cooler and the secondary fluid-bed cooler approximate countercurrent heat exchange principles resulting in high thermal efficiency. Power consumption is kept low, owing to the application of cyclones, where high gas capacities are needed and fluid-bed, where only very small air flows are needed.

As material hold-up in the calciner is very small, and as no air distribution grate is needed for fluidization purposes, pressure drops across the calciner are almost zero.

Specific power consumption is estimated to 15 kWh per ton alumina for a 1000 mtpd unit.

The production flexibility is governed by two limitations:

- i) Maximum throughput is determined by the allowable particle breakdown caused by the gas velocity at the desired degree of calcination.
- ii) Minimum throughput achieved is determined by proper functioning of the cyclones as separators for gas and solid.

The freefalling velocity of alumina is generally found to be below 1 m/sec. for the particle sizes of interest. Therefore, the principles of continuous flow of lean gas-solid suspension, as used all over in the FLS Calciner plant concept offers good flexibility with respect to variation of the gas flow and thus production level of the calciner.

The contact time between material and lining is much shorter than in plants utilizing fluid-bed technology, so that contamination of the alumina from the lining is likely to be minimized.

In addition to many constructional advantages mentioned above several advantages are expected from a process and operational point of view. These advantages are summarized below:

- Low specific power consumption due to small pressure drops.
- High conversion of solids and fuel per unit reactor volume, giving small investment costs.
- Very good uniformity of product regardless of particle size.
- Minimizing of process heat losses by application of cold air fluidization at the secondary cooler only where the material temperature is as low as possible.
- Close control of calciner temperature and product quality by simultaneous calcination and uniform flameless combustion.
- High thermal efficiency by approximated countercurrent heat exchange in preheater, primary, and secondary coolers.

Thus many potential advantages are viable from the FLS stationary Calciner principles.

Acknowledgement: The authors of this paper wish to thank: K. Theisen at the FLS Chemical Laboratory for her examination of physical and chemical properties of alumina from our pilot plants. J. Wiberg at the IED Engineering Department for working out plant layouts for commercial sized units. P. H. Hansen the IED R & D Section for computer evaluation of pilot plant data automatically logged by an on-line process computer during the experimental test period.

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