LATEST DEVELOPMENTS IN CIRCULATING FLUID BED CALCINATION BASED ON OPERATING EXPERIENCE OF LARGE CALCINERS

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1. Introduction

The world's two largest single CFB-calciners with capacities of more than 3.000 tpd have been in operation since 1999/2000.

The operational results of these calciners are being utilised for the design of future calciners. This generation of 3000 tpd calciners represents the latest status of the CFB-technology. For practical operation regarding flexibility and integration into the Bayer refinery with respect to hydrate inventory and product flow, a calciner capacity of 3000 tpd appears to be an optimum choice for an alumina refinery above 2 million tons per annum.

Although the energy efficiency of these plants has achieved an almost optimum level, it will be demonstrated that further improvements are possible.

In the alumina cooling section of the calcination process energy is still available which can be utilised for pre-drying of the feed-hydrate.

For this additional energy recovery, well proven equipment and technology is applied. The methodology has continued by tests in a pilot plant.

2. Operating Results and Experience

2.1 *The plants*

In 1998 Nabalco Pty Ltd ordered a new CFB Calciner with a rated capacity of 2700 tpd for their existing alumina refinery at Gove, NT Australia, as part of their ongoing programme of plant modernisation. This calciner was installed to partly replace the 4 rotary kilns which until that time provided the complete alumina calcination capacity [5].

Nine months later — end of 1999 — Worsley Alumina Pty Ltd ordered a new CFB calciner with a capacity of 3100 tpd. This calciner is the fifth for the West-Australian Alumina Refinery.

In 1983 the Worsley refinery started its operation with Lurgi calciners when 3 units with a capacity of 1500 tpd each were installed. In 1990 the refinery was expanded and a new calciner with a rated capacity of 1850 tpd was installed.

Alumina calcination in this refinery has been provided by Lurgi CFB calciners since the first day.

In figure 1 the process flow diagram for both calciner — Nabalco and Worsley — is shown. The flow diagram represents the improved concept with hydrate by-pass and two cooling cyclones to achieve lower energy consumption [5]. Both calciners are the first industrial units where these modifications together are implemented.

2.2 *Operating Results of the Nabalco Calciner*

The design criteria for the Nabalco calciner are as follows:

Capacity	2700 tpd
Capacity with spare blower	3000 tpd
Fuel	Heavy fuel oil
Stack temperature	min. 170°C
Dust emission	max. 40mg/Nm^3
Specific surface area	75 m^2/g BET

During the commissioning of the Nabalco Calciner some problems had to be resolved.

First of all, operators had to become familiar with the large mass flow of this 3000 tpd calciner which was at this time the world's largest single calcination unit.

The significant mass-flow of 235 t/h hydrate had to be monitored very precisely to avoid hidden blockages and/or storage in the system, with quick actions needed to prevent unforeseen shut-downs.

- Unstable operation during start-up and some times at partial load caused by insufficient sealing between air and water cooled sections. This could be resolved by modified start-up procedure and operation mode.
- In some areas of the calciner, such as the crossover duct in lift pipes refractory damages occurred. An investigation of the problem demonstrated that the damages were caused by incorrect anchor and refractory installation.
- Due the unusual high number of start-up and shut-downs cracks were observed in some areas of the calciner such as primary air header and lift ducts. Design modifications were made to resolve the problem.

Together with the usual start-up problems the commissioning period was extended to about 4 months.

After a period of approx. 4 months the operating personnel had gathered enough experience to operate the facility with the same stable conditions as for all previous CFB-calciners with smaller capacity.

Today the unit is operating satisfactorily and shows the following performance figures:

Average	Availability	Energy Consumption	Breakage 44 µ	Breakage 20 µ	Alumina < 44 µ	Alumina < 20 µ
2002	95%	2.86 GJ/t	4.5%	0.9%	6.4%	1.7%

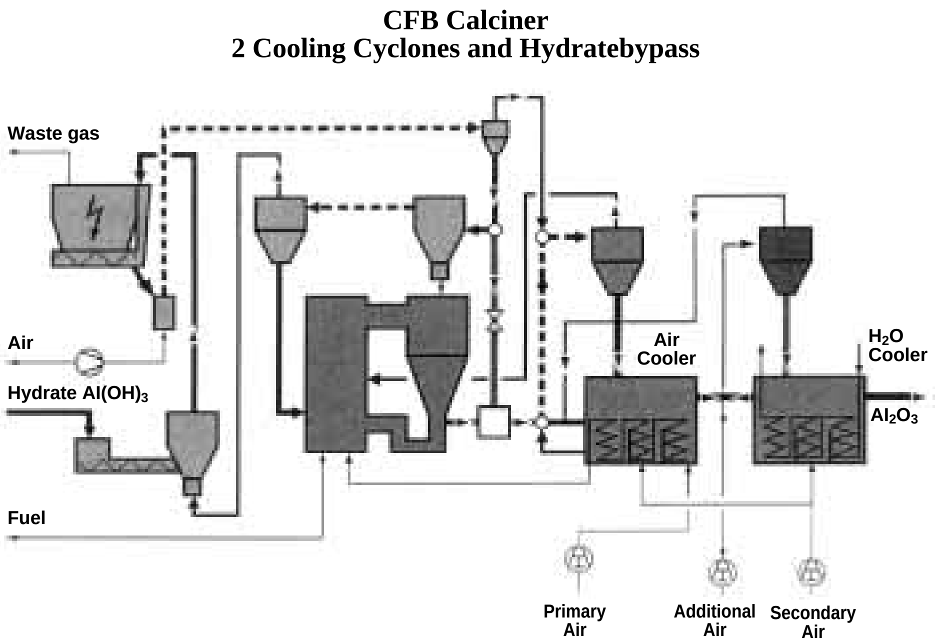

Figure 1 — Process flow diagram

The principle of the circulating fluid bed is that it has the ability to "swallow" variable product flow rates peaks, etc, especially at high mass flowrates due to its material inventory. The operation therefore remains steady and stable with a uniform temperature distribution in the CFB reactor resulting in uniform product quality [6, 7, 9].

A quick look at the breakage figures of the system shows that the average breakage in 2002 at –20 µ with 0.9% is extremely low. During 2001 the average breakdown figure was 0.6%.

Figure 2 — Nabalco calciner, Gove NT, Australia

2.3 *Operation Results of the Worsley Calciner*

The design criteria for the Worsley calciner are as follows:

Capacity	3100 tpd
Fuel	natural gas
Dust emission	50 mg/Nm^3
Specific surface area	75 m^2/g BET

The Worsley No. 5 calciner was commissioned about 6 months after the Nabalco calciner was put on line.

The results and the experience from the Nabalco commissioning were utilised at Worsley and avoided similar troubles during the plant commissioning phase.

The Worsley unit also had the advantage of 4 smaller "sister" calciners being in operation for almost 20 years. In addition the two refineries exchanged their experience with the large unit and Worsley delegated personal to the Nabalco unit during the whole start-up phase.

This very open policy of experience and data exchange was certainly a significant contribution to the successful commissioning of the two large units.

The performance data of the Worsley calciner are shown in the following table:

Average	Availability	Energy consumption	Breakage 44 µ	Breakage 20 µ
2002	97%	2.75 GJ/t	4.5%	1.0%

It should be noted that the Worsley calciner has been operated since commissioning with the Hydrate Bypass. This explains the lower heat consumption figures [1, 2, 8]. It was also shown by others that particle breakage is mainly influenced by product quality of the feed hydrate and with a minor influence only by the calciner design [12].

Nabalco has chosen not to utilise the by-pass from the beginning and therefore had a higher energy consumption.

3. Further Utilisation of Thermal Energy in the CFB Calcination Process

3.1 *Basic recovery of calcination energy in the CFB process*

As described above the present state of the art CFB alumina calcination process already represents a high degree of utilisation of the energy input [1, 2, 3, 4]. The modern state of the art Lurgi fluid bed calciners represents a considerable improvement over the rotary kiln technology and other fluid bed technologies with a typical specific heat consumption of approx. 2900 kJ/kg Al_2O_3.

31% of the world production of alumina is still calcined in rotary kilns which need up to 60% more energy than a Lurgi calciner (using Lurgi calciner as the base).

The basic principle of the heat recovery system of the Lurgi fluidbed process can be shown on the block diagram of the process in figure 4.

The heart of the calcination system is the circulating fluidbed reactor. The fuel needed for the calcination process is burned at a temperature of approx. 950°C. Two basic streams leave the fluidbed reactor:

a) The combustion waste gas together with the water originating from the final dehydration which takes place in the fluid bed reactor.
b) The calcined alumina product.

Both streams leave the fluid bed reactor with the calcining temperature of approx. 950°C and constitute a considerable flow of energy.

In the interest of minimising the required thermal energy for the process, both energy flows should be utilised to the highest possible degree.

On the waste gas side the thermal energy contained in the waste gas is utilised for drying, preheating and partial dehydration of the feed hydrate. The waste is cooled in two Venturi reactors to approx. 150°C depending on the sulphur content in the fuel oil. The waste gas temperature has to be limited to a temperature above the due point of sulphuric acid to avoid corrosion in the waste system. If natural gas is used as fuel, the minimum operating temperature of the waste gas system could be reduced to approx. 100°C. Accordingly the utilisation of the energy content of the waste gas is already close to the technically feasible limit.

On the calcined alumina product side the thermal energy contained in the product is utilised to a high degree by preheating the required combustion air for the fuel. The available energy in the alumina is the sensible heat, which is released during cooling the product from the calcination temperature of approx. 950°C to the required final product temperature of approx. 80°C. The amount of cooling air which is available for cooling the alumina and thereby for

Figure 3 — Worsley calciners No. 1–5, Collie WA, Australia

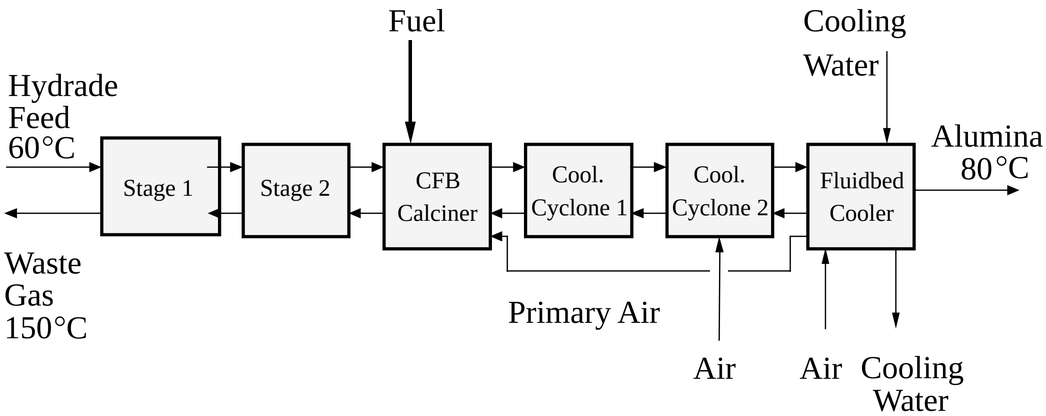

Figure 4 — Block Diagram of the CFB — Calcination Process

heat recovery is limited to the amount needed for the combustion of the fuel (plus a reasonable amount of excess air). The preheating of the combustion air and at the same time the cooling of the alumina is done in a series of cooling cyclones and fluidbed cooler stages. Unfortunately by this method, the utilisation of the energy content of alumina for the preheating of the combustion air in the calcination process itself is limited to an alumina discharge temperature of approx. 250°C. The residual energy content of the alumina is to be removed by an external heat carrier.

Normally the energy is just removed by cooling water and the energy content is lost! This represents a heat loss of 300 to 400 kJ/kg alumina.

3.2 *Utilisation of energy in other parts of the Bayer process*

Basically the major part of this energy can be utilised in parts of the Bayer process for low temperature applications. This has been introduced in some alumina plants for:

- preheating of wash water
- preheating of boiler feed water
- preheating of fresh process water.

The flow sheet in figure 5 shows such an application in which after the first part of the fluidbed cooler, the primary combustion air is preheated, a series of water cooled sections of a fluidbed cooler are arranged, before the alumina is finally cooled by cooling water. By this method up to 200 kJ/kg Al_2O_3 of the primary heat input can be recovered.

Unfortunately in many alumina plants, this type of heat recovery is difficult to use because:

- already sufficient “waste-energy” from other parts of the Bayer process is available
- integration of the energy transfer from the calcination plant to other parts of the alumina plant needs too many interfaces.

3.3 *Improved recovery of calcination energy in the CFB process*

Although this kind of energy recovery constitutes a significant credit in energy cost, a more elegant solution is to find a way for energy recovery in the calcination process itself. In this case a saving of primary energy would be achieved. An analysis of the heat flow diagram of the alumina calcination process shows the following relations:

The lower the specific heat consumption for the calcination becomes, the lower will be the quantity of combustion air. Combustion air is the only useful cooling medium for alumina. By operating the plant with higher excess air amount than needed for combustion the recovery of heat from the alumina would increase but on the other hand this would increase the waste gas volume and temperature. In total there would be no positive effect on specific heat consumption, but increased capital cost due to the increased air and waste gas flow rates.

On the other side of the process, for the hydrate drying and preheating, the demand for energy is still the same, whereas the amount of available heat for the purpose of drying and preheating of hydrate decreases with decreased specific fuel consumption, because the waste gas volume decreases.

Because of this situation (energy surplus on the alumina cooling side, demand for additional energy on the hydrate drying and preheating side) it is logical to shift energy from the alumina cooling system to the hydrate drying system.

The technical solution is shown on this flow sheet in figure 6.

Between the fluid bed cooler for primary air and the final water cooled fluid bed cooler, a fluid bed cooler is arranged for pre-heating a heat carrier like pressurised water or steam.

The heat carrier is used to dry hydrate in a fluidbed dryer.

Depending on the available energy from the intermediate fluid bed cooler, part or all of the hydrate feed is passed to the hydrate dryer. The rest of the hydrate feed goes straight to the venturi dryer.

As with this system part of the hydrate which is to cool the waste gas is already dried, the waste gas temperature will rise. Depending on the range of hydrate moisture, it may be more efficient to install one additional venturi preheater stage between the present feed venturi stage and the next venturi stage. By this concept the maximum utilisation of the waste gas energy is achieved. The overall savings in primary energy consumption of the combined application

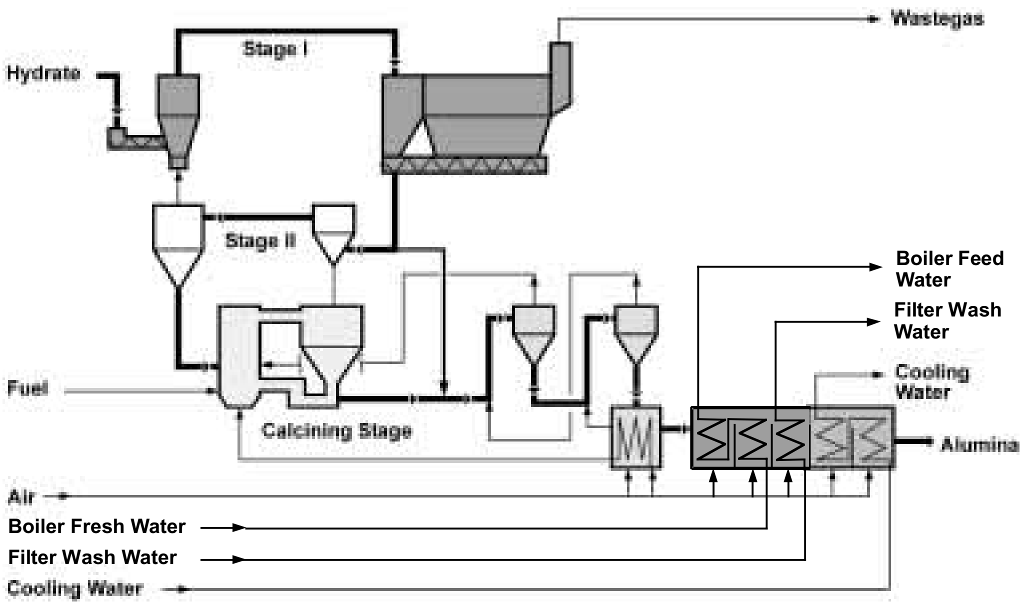

Figure 5 — Pre-heating of Filter Wash and Boiler Feed Water

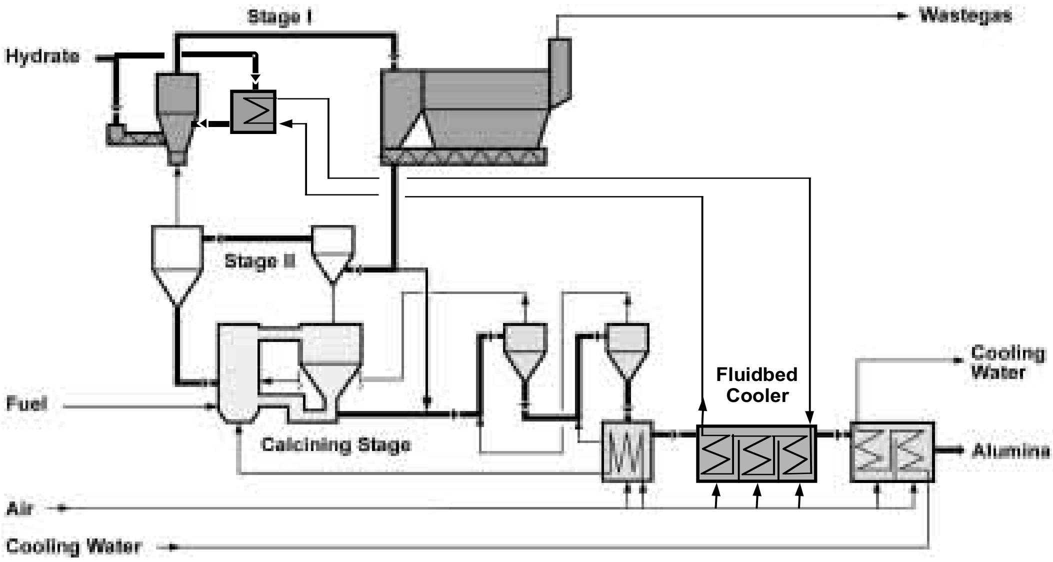

Figure 6 — Heat Transfer from FBC to Hydrate Dryer

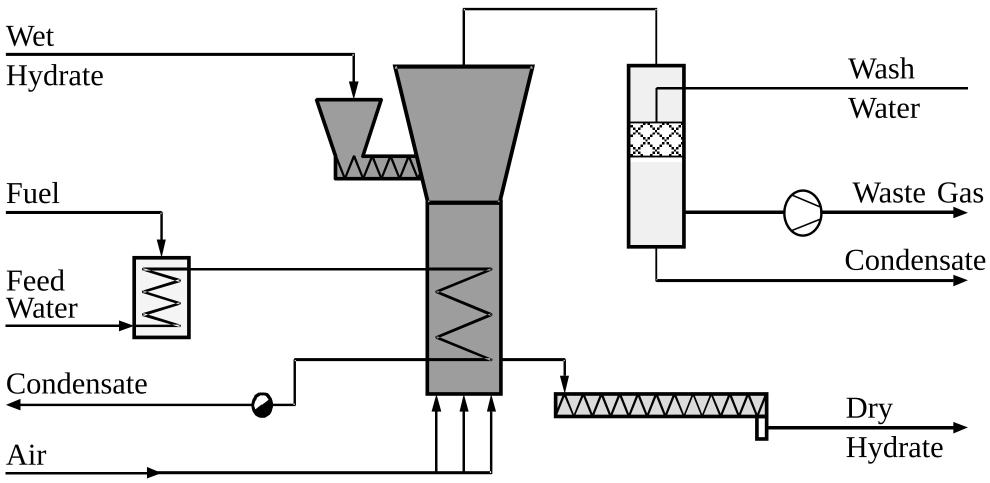

Figure 7 — Pilot Plant — Indirect Hydrate Dryer Flow Sheet, Capacity 1–2 t/h

of the hydrate dryer and the additional venturi stage is approx. 6 to 8% of the total energy!

3.4 *Pilot Plant for indirect drying*

The new process features proven equipment such as venturi dryers, cyclones and a fluid bed heat exchange system for the indirect drying and heat transfer as used in the fluidbed cooler. For testing the full range of operation of the dryer with regard to product quality, fluid dynamics and heat transfer, a pilot plant was installed which is capable of drying up to 2 t/h of hydrate and operated in a temperature range between 80 to 180°C. The simplified flowsheet of this pilot plant is shown in figure 7.

The plant performance was excellent. Extreme situations regarding fluidising velocities and feed rates were all covered during the operation of the plant. The ease of operation was well proven and demonstrates that this advanced plant concept represents a big step forward towards even higher energy efficiency of the Lurgi CFB calcination process. The results of the test are shown in the table of figure 8.

Test — Results	
Drying Temperature	90 to 110°C
Feed Moisture	5 to 7%
Moisture of Product	0.02 to 0.1%
H_2O-Content in Waste Gas	40 to 60%
Start-up/Shut-down Behaviour	Excellent, no blockages

Figure 8 — Pilot Plant Tests with Indirect Hydrate Dryer

A picture of the upper part of the pilot plant is shown in figure 9.

4. Conclusion

The operating results and performance of the two large calciners in Nabalco and Worsley are demonstrating that a 3000 tpd unit is an optimum solution for a modern large alumina refinery regarding stable operation, flexibility and consistency in product quality [9]. The scale up from 2000 tpd units was successful [10, 11] and resulted in a

Figure 9 — Pilot Plant — Indirect Hydrate Dryer

further 3000 tpd calciner presently under construction for Alunorte in Brazil. This calciner has in its design already incorporated the experiences resulting from the operation of the two Australian calciners. The start-up of the Alunorte calciner is scheduled for October 2002.

Further to these results it was possible to develop a concept in which the thermal energy of the CFB calcination process is further reduced to energy consumption figures below 2600 kJ/kg alumina. This concept is based on a combination of proven technologies and equipment. The performance of the indirect hydrate pre-drying could be successfully demonstrated in a pilot plant.

The small increase of capital investment can be justified by lower energy consumption within a short pay back period.

The presented flow sheet is the advanced concept for future calciners.

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