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Methods to Reduce Operating Costs in Circulating Fluidized Bed Calcination

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Abstract

The calcination of gibbsite or hydrate to alumina is one of the most energy consuming process steps in every alumina refinery based on the Bayer process. Approximately 30% of the thermal energy input is used for the calcination process step.

Over the years new technologies such as Circulating Fluidized Bed (CFB) Calciners by Outotec (formerly Lurgi) have reduced the energy consumption for the calcination step significantly (see also recent Energy Efficiency Award, given in 2010 by the German Energy Agency). The CFB technology was introduced as early as 1961. Up till then rotary kilns were the standard technology for the calcination of alumina. This innovation reduced the consumption of fuel by up to 30%. Since then, CFB calciners have been constantly improved, and methods have been developed to reduce fuel consumption even further.

Not all, but some, of the methods to reduce fuel consumption have resulted in increasing process complexity and operator and maintenance demand. However, a number of measures have also been introduced to mitigate the negative effects of the increased process complexity and to even improve on operability and maintainability.

In this paper the different options and methods introduced for reduction of fuel consumption, but also to increase of operability and maintainability are compared and evaluated with regard to their effects on installation and operating costs.

Introduction

Calcination of hydrate to alumina is the last step in the production of smelter grade alumina from bauxite in an alumina refinery. The energy consumption for the production of alumina varies between 6.5 and 18 GJ/t alumina for the whole refinery. Depending on the technology, approx. 30% of this is consumed in the calcination step.

Until stationary calciners, such as flash and CFB calciners were introduced, rotary kilns were used. Rotary kilns have a thermal energy consumption of around 4.5 to 5.5 GJ/t alumina based on lower calorific heating value. With the introduction of stationary calciners and the first CFB calciner by Outotec (at that time known as Lurgi) in 1961, the consumption dropped instantly to approx. 3.3 GJ/t. Thus operating costs dropped significantly, which gave a huge incentive to invest in this new technology.

Since then the technology has matured and improved with current energy consumptions of 2.79 GJ/t for the state of the art flowsheet as shown in figure 1.

The CFB calcination process utilizes maximum heat recovery by having air and furnace off-gas flowing in a countercurrent scheme to the hydrate feed or alumina product, respectively.

In the current flowsheet the hydrate is preheated in two stages with the off-gas of the CFB furnace. In the first preheating stage the free moisture of the hydrate, with typical values of 6 to 8% moisture, is evaporated and the hydrate is dried.

The dried hydrate is then separated from the gas in an electrostatic precipitator and fed to the second preheating stage, where it is mixed with the 950°C hot off-gas of the furnace. Here a significant step of the reaction is already taking place with approx. two thirds of the hydrate water released. The solids are separated again from the gas and transported to the CFB furnace.

Finally under the high solids density and homogeneous mixing and temperature field in the CFB furnace, the precalcined material is calcined to the specified alumina quality. The good mixing and high solids density, with a sufficient residence time of several minutes, allows very homogeneous calcination of the material and the optimum product quality, with very low alpha alumina and no residual gibbsite, as well as low particle breakage.

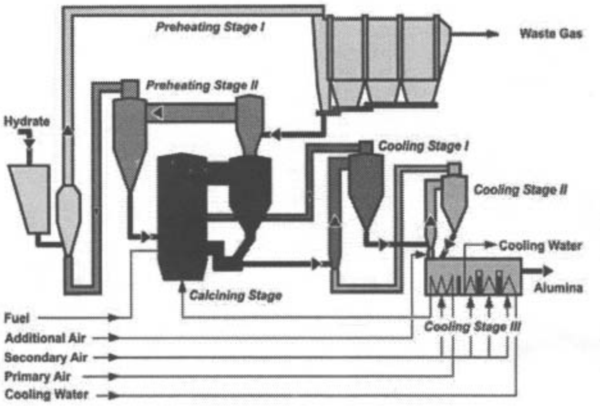

Figure 1: CFB Calciner Flowsheet of recently installed calcination plants

The final product is discharged from the CFB calcining stage and then cooled down in two direct gas solids contact cooling stages and one indirect cooling stage, which is the fluid bed cooler.

In the first two cooling stages the hot alumina is mixed with air and the air heated up as the alumina is consequently cooled down. In each stage the alumina is separated from the heated air by means of a gas cyclone and transported further to the next stage, whereas the air is flowing in direction of the CFB furnace to serve as heated combustion air.

At last in the fluid bed cooler all residual heat, which is not easily recovered in the calcining process, is extracted from the alumina to enable discharge with safe alumina temperatures for handling. The fluid bed cooler consist of several sequential compartments, where primary fluidizing air is preheated for the nozzle grate of the CFB furnace and finally cooling water lowers the alumina temperature to safe handling levels.

Key Consideration for Fuel Efficient Calciner Design

Essential for the success of the CFB calcination flowsheet for optimized performance is not only the design of the Fluid Beds (CFB and Fluid Bed Cooler [1]), but also the cyclones in the system. In the presented calcination process heat recovery is done by counter current solid gas heat transfer. The means of separating solids and gas are the cyclones. Any amount of solids in the cyclones off-gas is counteracting the counter current heat recovery scheme. Further, the fines in the off-gas of one cyclone will be fed to the cyclone in the upstream recovery stage and subsequently reduce the separation efficiency as well.

The cyclone design is not only critical for best separation performance and thus for the process efficiency, but also for the amount of fines in the system. The fines are both fed to the process with the hydrate feed, and also generated in the process by particle breakage. A major part of the particle breakage is generated in the cyclones. This needs to be taken into account in the process and specifically cyclone design [2]. Cyclone design, e.g. geometry, velocity profile, etc. needs to be carefully optimized to have max. separation efficiency not only for one cyclone, but also for the overall process as breakage and fines influences all of the cyclones at the same time. This phenomenon is not only known for alumina calcination [3]

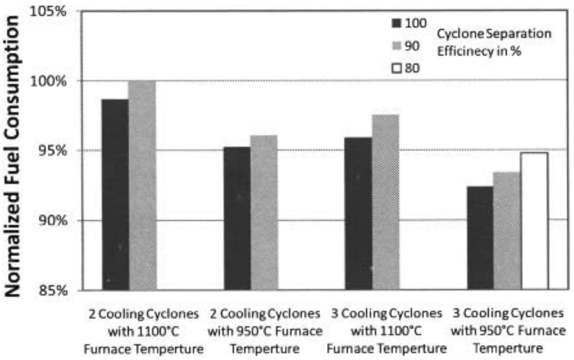

Figure 2: Fuel Consumption depending on Cyclone Efficiency and Calcination Temperature

Besides velocity and solids concentration in the cyclone inlet, the operating temperature also has a strong influence on the performance [4]. With increase of gas temperature, the gas viscosity increases and the drag on the particles inside the cyclones increases. The higher wall friction also causes reduced tangential velocities for the centrifugal field. As a result the separation efficiency decreases for a cyclone with constant inlet velocity. In order to compensate the inlet velocity needs to be increased, which will then increase breakage and generation of fines. As described above this again will have an impact on the overall process performance.

In figure 2 the influence of cyclone performance on the fuel consumption of a stationary calciner is shown. For comparison two calciners one with 2 and one with 3 cooling cyclones and two different calcination temperatures are assessed. CFB calciners are typically built with two cooling cyclones and operated at 950°C in the furnace. Stationary calciners are often equipped with 3 cooling cyclones and operated at 1100°C [5]. The furnace temperature is thus defining the general temperature level in the cooling stages and consequently has an influence as described on the performance of the cooling cyclones.

The process calculation results show that in fact with a reduced furnace temperature, the fuel consumption of the calciner is less than with a higher furnace temperature. This can then only be compensated by more cooling stages. Not surprisingly reduced cyclone separation efficiencies also lead to increased fuel consumption.

Measures for Further Reduced Fuel Consumption

In the above sections the basic flowsheet of Outotec's CFB calcination technology is described and assessed. As the technology is now well proven and lot's of research and development has been undertaken in recent years, there are several developments and options available in this flowsheet, which further improve energy consumption, availability and operability. Which option is best suitable depends on client's requirements, fuel prices and project structure such as investment budget and predicted plant life time and life cycle.

In the following sections, the different options available for reduced fuel consumption and improved plant performance are described and the impact on investment and operating costs are discussed. The discussions further distinguish between options for greenfield plants or revamps and upgrades of existing units.

Based on the above basic flowsheet several further options and variations of the flowsheet have been developed to reduce fuel consumption and hence reduce operating costs. Naturally these options will have an impact on the investment costs as well.

There are several options available to reduce fuel consumption of the calcination process. The available measures are targeting the optimization of the heat recovery and energy utilization within the calcination process inside the battery limits of the calciner. The two main options possible are discussed in this publication.

The available options are:

- Hydrate bypass, where preheated material is bypassed around the CFB furnace and calcined with heat from CFB discharge alumina
- Pre-drying of hydrate feed with waste heat from fluid bed cooler

In figure 3 the first option the hydrate bypass is shown. This option has been developed in the 1990s by Outotec [6] and [7].

In the bypass flow sheet a part of the dried hydrate from the first preheating stage is taken and directed to the discharge of the CFB furnace. The bypassed solids are mixed with the hot alumina from the CFB discharge and allowed to react in a so-called mixing pot, which will provide sufficient residence time for the reaction. With this method, heat from the CFB alumina discharge is utilized for the calcination reaction of the bypassed hydrate, rather than for heating up of combustion air. Hence the heat is utilized more efficiently and secondly the amount of hydrate or alumina respectively, which needs to be heated to the reaction temperature

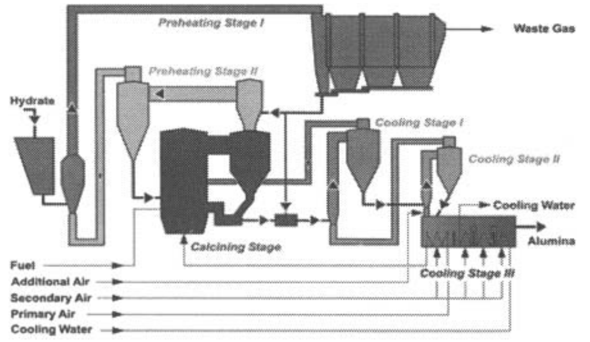

Figure 3: Flowsheet with the Hydrate Bypass implemented.

of the CFB of more than 950°C is reduced. This reduces the required amount of fuel significantly. The Bypass was developed in the 1990s and has already been installed in more than 6 plants worldwide since then.

With the hydrate bypass in operation, the temperature profile in the cooling stages changes, as the solids enter the first cooling stage with a lower temperature than before. Furthermore, due to the reduced fuel consumption, less air is required and also less combustion off-gases are produced. This allows the design of the preheating stages, CFB stage and cooling stages to be smaller, with less construction material for the same nameplate capacity.

To further reduce the fuel consumption of the calcination process, a closer analysis of the process is required. A thermodynamic heat and mass balance analysis of the calcination flowsheet shows that CFB with its fixed and homogenous temperature profile is a "pinch point" for the heat flow and exchange within the process. This means that with the existing counter current heat exchange and recovery scheme it is very difficult to transport heat from the back end of the process through the CFB stage into the front end of the process.

One way could be to install more preheating and more cooling stages to increase heat recovery at both ends. However this would mean more investment costs for very big and refractory lined vessels.

Another possibility is to collect the heat in one end and transport it around the "pinch point" to the other end to utilize it there. A method for this approach has been developed by Outotec in the past [8].

Heat can be recovered from the fluid bed cooler by means of a heat carrier, such as water under pressure or thermo oil. The heat carrier is then pumped to the front end of the calciner. There a part of the moist hydrate feed is taken and fed into a fluid bed hydrate dryer (for details see the section below). In the hydrate dryer the heat from the heat carrier is used to dry the hydrate before it is fed to the preheating stage I. Then the heat carrier is pumped back to the fluid bed cooler. Hence the waste heat from the cooler is used to evaporate the free surface moisture of the hydrate.

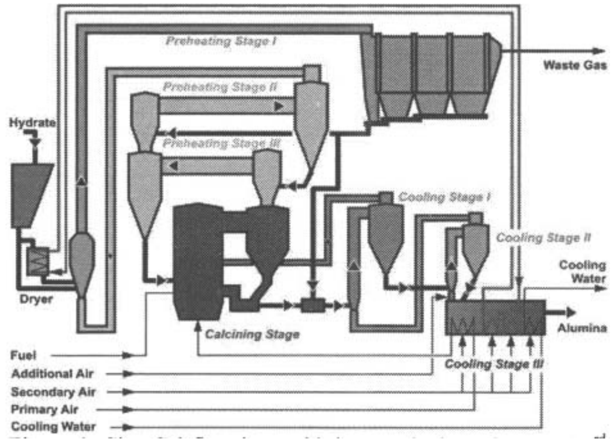

Figure 4: CircoCal flowsheet with bypass, hydrate dryer and 3rd preheating stage

As a result, less heat is utilized in the preheating stage 1 from the calciner off-gas. Consequently the off-gas temperature would rise. To utilize this heat, a third preheating stage can be installed, which then recovers the heat. Finally this means that less heat is rejected with the cooling water to atmosphere and there is also reduced heat loss with the calciner off-gas.

The flowsheet for this process is shown in figure 4 and forms the CircoCal process by Outotec. Advantages are that all extra equipment does not need to be refractory lined. Also there is only one more cyclone installed on the preheating side and no further cooling cyclone is needed to increase heat recovery. Cyclones are one of the major sources for particle breakage [2] and the most sensitive is the alumina on the product side (cf. section above). Therefore the impact of product quality is minimal for this approach, compared to the classical approach to expand the counter current heat recovery scheme.

Alternatively the dried hydrate can also be transported to a dry hydrate storage silo instead of feeding to the calciner if dry hydrate is desired as an extra product from the refinery. However this option is not investigated further in this publication.

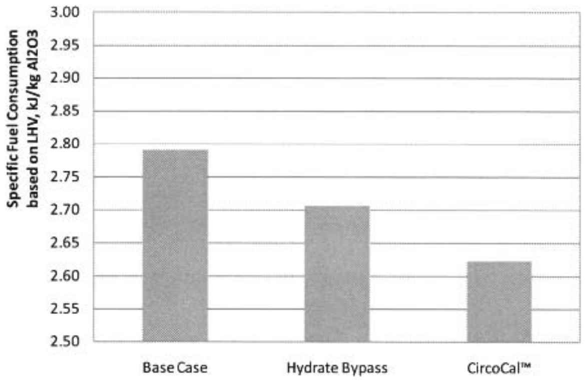

Figure 6: Specific energy consumption of different flowsheet variations

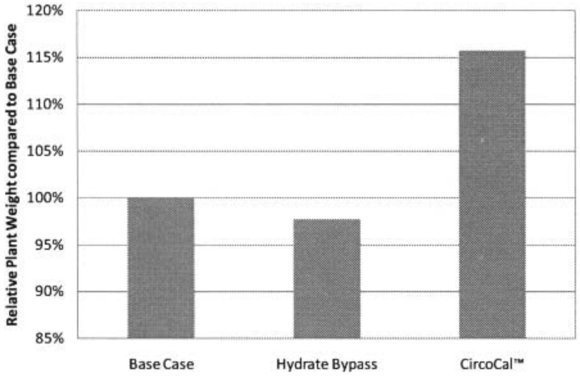

Figure 7: Impact on investment costs

Evaluation of Measures for Reduced Fuel Consumptions

In the previous section, different options are presented to further reduce the fuel consumption of CFB calciners. The base calcination flowsheet has a specific energy consumption of ~2.79 GJ/t with a feed hydrate moisture of around 7%. The results of the fuel consumption comparison are shown in figure 6.

In the comparison the hydrate bypass reduces the specific energy consumption by ~3% and the CircoCal flowsheet has the lowest specific fuel consumption of all the options discussed. The reduction compared to the base case is about 6%.

For all options, the reduction in fuel consumption can be seen as direct reductions in operating costs. There will be an increase in labor and maintenance costs for the extra equipment in the plant. However these can be seen as marginal. None of the options include maintenance intensive equipment such as extra rotating equipment or huge refractory lined vessels. The extra fluidizing air requirements can be covered with the existing blower concepts and Outotec's technology for solids mass flow control does not require any moving parts inside the solids streams.

Investment costs of the plant will be impacted as presented in figure 7. It shows that an improvement in energy efficiency does not necessarily mean also an increase in investment costs.

Figure 7 shows the changes of plant weights normalized to the base case. Again here the boundary condition is that all plants have the same nameplate capacity. The changes in weight here are considering vessel plate work, refractory and steel structure. Instrumentation, piping and other equipment such as blowers, screws, etc. do not change significantly.

Due to the changed temperature profile in the hydrate bypass option the vessels can be reduced significantly in size as described above. This then leads to reduced overall plant weight and does in fact reduce investment cost for the plant when nameplate capacity is maintained as figure 7 demonstrates.

The reduced fuel consumption of the CircoCal flow sheet is offset by a potential increase of investment cost of about 15% if based on weight. This is the highest increase of investment of all options. On the other hand the savings in fuel can give a return of investment in 2 years based on an oil price of 80US$/barrel. The costs / benefit ratio can possibly be further improved if carbon projects are being included.

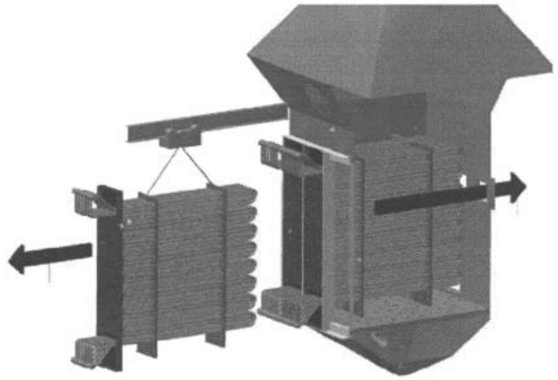

Figure 8: Schematic of fluid bed hydrate dryer design

Drying Hydrate in a Fluidized Bed

Key equipment of the CircoCal flowsheet is the hydrate dryer. There are a lot of refineries, which also produce dried hydrate. However in most cases direct heat transfer with hot off-gas from combustion is used to dry the hydrate. Here the waste heat from the fluid bed cooler is used. As the heat carrier, water or oil is used in a closed circuit (cf. figure 4). To transfer the heat to the wet hydrate from the hydrate filtration, indirect heat transfer via heat transfer surface is required.

Outotec has developed for this a hydrate dryer based on a bubbling fluidized bed similar to the fluid bed cooler. In the fluid bed dryer, heat transfer bundles are submerged in the fluidized solids. A schematic of the fluid bed dryer is shown in figure 8. In this fluid bed, large transfer areas can be installed in very small vessel volumes. For optimum bundle design the same design criteria are applied as for the fluid bed cooler with same success as reported in 2007 [1]. To ensure easy maintenance the bundles can be removed from the side and cleaned.

The implementation philosophy of the dryer into the overall calciner flowsheet is shown in figure 9.

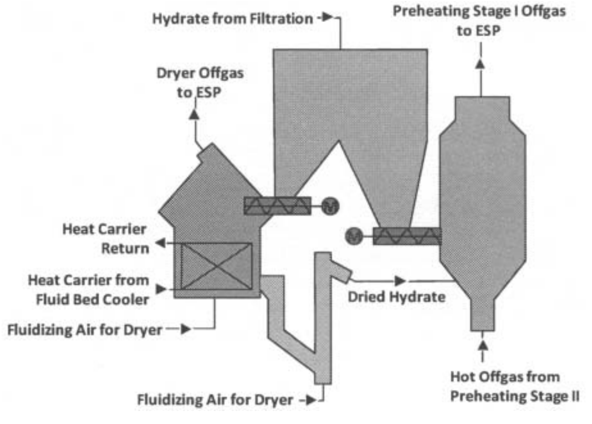

Figure 9: Schematic of hydrate dryer implementation in preheating stage I

The wet hydrate is fed into the calciner feed bin. The feed bin has two separate bottom discharges, which are connected to screw feeders. One of the screw feeders is feeding wet hydrate directly into the venturi of the preheating stage I. The second screw feeder is feeding into the fluid bed hydrate dryer. The dryer screw feeder is controlled by a temperature measurement inside the dryer fluid bed and ensures that wet hydrate is only fed into the dryer, when the temperature is above 100°C. This guarantees that the fluid bed is always dry and easily fluidizable. Secondly, the speed of the screw feeder is controlled to maintain stable dryer temperature and ensures that all the heat available from the heat carrier is used.

Furthermore the inventory of the fluid bed dryer needs to be kept stable to keep the bundles submerged during normal operation. This is achieved by means of a lifting seal pot in the same way as it is used to maintain the level of solids in the fluid bed cooler. The solids level in the fluid bed dryer is detected by means of a differential pressure measurement across the fluid bed height. The differential pressure is then controlling the fluidizing air to the fluid bed dryer seal pot, which discharges the dry hydrate directly into the venturi of the preheating stage I.

Finally in the preheating stage I both hydrate streams (wet and dry) are mixed with the hot off-gas of the preheating stage II. With the mixture of wet and dry hydrate, the utilization of the available heat in the heat carrier and thus in the fluid bed cooler is maximized.

The vapor from the hydrate dryer is then added to the preheating stage I off-gas and leaves the calciner together with the combustion off-gas through the ESP and stack. Alternatively the vapor can also used to recover water if required. The off-gas from the hydrate dryer will have very high amounts of water with small amounts of air.

Measures for Improved Operating Stability and Availability

Besides the energy optimized flow sheet and plant layout, stable operation of the process is critical for the achievement of low operating costs. It is not only important for the optimized performance with minimum consumption of utilities, but also for maximum lifetime of plant equipment and low maintenance requirements. For example the typical it experience has been that stable process operation and low frequencies of unplanned plant downtime is one key for long refractory life.

A further advantage for the refractory lifetime of the CFB system is the fact that it operates at significantly lower calciner reactor temperatures, compared to other stationary calciners while achieving similar product quality [9].

In the recent past, Outotec has frequently reported on the progress in plant operation and stability. With new technologies available, and with the experience of more than 60 plants built, the controls and automation has been developed to an extent that constant operation is achieved under nearly all conditions. This leads to a minimized number of unplanned plant downtimes as the available control philosophies now prevent plant drop outs by automated reaction of the system. Further the controls now assist in automated start-ups, change of plant loads and shut downs [9].

The base case includes 5 main control loops, which are:

- Hydrate feed
- Furnace temperature
- Furnace inventory
- Fluid bed cooler level
- Emergency water sprays

All these loops are tied together in a governing overall control scheme with smart feed forward and troubleshooting mechanisms. In the recent years the improvements of process control and automation have led to very easy operation and fast ramp up performance [10]. As result there is little difference anymore between the different calcination technologies available and the limitations for fast process changes, ramp ups and shut downs are only in the speed in which process equilibrium can be achieved. Figure 10 demonstrates the comparability in ramp up performance for different calcination technologies.

In 2008 the influences on availability and consequences from unplanned outages of CFB calciners have been discussed [11]. Although not the major reason for unplanned outages, the consequences of refractory failures are very time consuming and costly. With the latest installations of CFB calciners the impact of plant trips and subsequent cooling and reheating cycles on the refractory could be explored. The stable operation, as reported in [2] and [12], proved to have very positive impacts on refractory life, thus not only decreasing operating costs by reduction of fuel consumption, but also reducing maintenance costs and loss of production by increasing reliability and plant integrity.

The above mechanisms and approaches will also be applied to the options described above, to further reduce fuel consumption on CFB calcination. For the bypass option only one more control loop is required which is the mixing pot temperature.

For CircoCal the following control loops are needed:

- Hydrate feed
- Furnace temperature
- Furnace inventory
- Fluid bed cooler level
- Emergency water sprays
- Mixing pot temperature
- Hydrate Dryer temperature
- Hydrate Dryer level

With the base case, a thermal efficiency of ~88% is achieved. In order to get closer to the theoretical fuel consumption of 2.45 GJ/t more initiatives need to be taken. Therefore the complexity of the plant will likely also increase. For the operator, the complexity is only visible in the amount of information he has to process and the attention he has to pay to each individual unit.

In this case it can be argued that the complexity grows significantly from the base case from the Bypass options to the CircoCal process, when one looks at the number of control loops to supervise.

However the extra control loops are be integrated in the overall control philosophy to achieve same temperature stability as in the base case. The relationships between temperature changes and solids mass flows to different areas in the plant are well known, which makes the controls easily manageable.

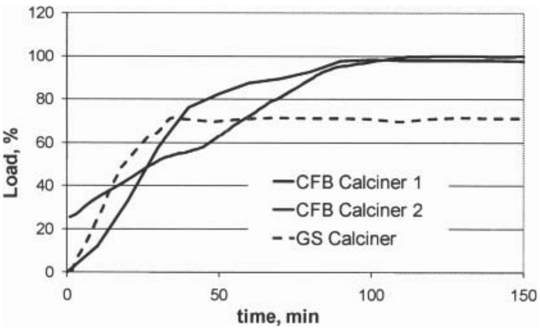

Figure 10: Ramp Up of stationary calciners. Data for GS Calciner performance adapted from [5]

Therefore it can be suggested that the increase of complexity lies only in the increased demand for instrumentation and control maintenance, which can be handled via the general refinery maintenance scheme and relieves the operator.

Retrofits to existing CFB Calciners

The measures presented for reduced operating costs and increased availability, can also be applied to existing units. The improved control schemes are implemented easily with only limited costs involved, depending on the existing instrumentation infrastructure.

Implementing a hydrate bypass or turning an existing unit into CircoCal operation is also feasible, but will involve installation of extra hardware. However this would not only reduce specific energy consumption and improve operability, etc. it also leads to an increase of plant production capacity. This is in contrast to the cases assessed above, where the plant capacity is deliberately kept constant. In a retrofit situation, the blower capacities and existing vessel and cyclone geometries are kept unchanged. This means that with changed temperature profiles, the throughput can be increased to maintain the velocity profile in the plant before and after the modifications. This also means that particle breakage can be kept constant or even improved depending on the existing unit.

Conclusions

In the above sections different options have been presented as to how to make CFB calciners more fuel efficient, how to reduce the operating costs and the impact on investment and process complexity and control.

With these different options, thermal efficiencies of up to 92% can be achieved and fuel consumptions as low as 2.65 GJ/t. The risks involved for particle breakage, product quality and process complexity is minimal and can easily be mitigated.

The options presented are not only of interest for new installations, but also for older ones and even viable for units of the very first generation of CFBs.

However for some options, the integration of the calciner in the overall refinery energy balance can give extra benefits. This needs to be the focus of future investigation on how to reduce overall operating costs in the production of smelter grade alumina from bauxite. These investigations shall not only focus on fuel consumption, but also on product quality such as fines in the final product. E.g. precipitation and calcination need to be looked at together in order to optimize investment and operating costs to achieve the required product specification.

Acknowledgement

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