Thermal Energy Consumption in Gas Suspension Calciners

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Abstract

Since the first unit were designed in 1984, Gas Suspension Calciners (GSC) were equipped with four (4) stages of direct heat recovery from hot alumina. This is the most energy efficient heat recovery flow sheet compared to other stationary calciner flowsheets, resulting in an average specific heat consumption as low as 2722 KJ(NHV)/kg SGA, reported for GSC units operating with a Calciner Furnace temperature of 1025 - 1050 °C and a few seconds of solids retention time. This paper reports the analysis of the thermal energy consumption of operating GSC units equipped with a fluidized holding vessel providing a few minutes of additional retention time, and operating at 100 - 150°C lower temperature in the Calcination Furnace. The analysis confirms even lower thermal energy consumption than previously reported, with significant potential for further improvements with an optimized design for refractory lining. The retrofit of a coal gas fired GSC with a fluidized holding vessel is also analyzed with respect to increased thermal energy efficiency and productivity.

Keywords: Alumina, calciners, thermal energy consumption.

1. Introduction to Gas Suspension Calciner with 4 Direct Heat Recovery Stages

The first oil fired Gas Suspension Calciner (GSC) unit equipped with an Electrostatic Precipitator was successfully commissioned at Hindalco, India in 1986, replacing three rotary kilns [1].

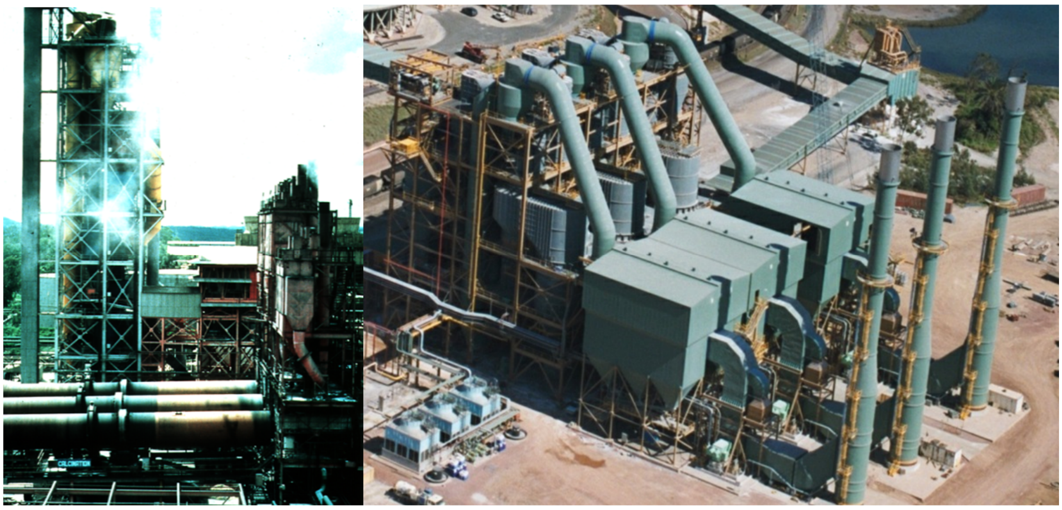


Figure 1. 850 tpd GSC, Hindalco, and 3 x 4500 tpd units Queensland Alumina.

Over time, FLSmidth Minerals has supplied and started up many more GSC units, including the world's largest Natural Gas fired stationary calciners: 3 x 4500 tpd GSC units equipped with Fabric Filters (Bag House) at Queensland Alumina Limited, Australia in 2003 - 2004 [2]. The new GSC units were replacing nine (9) rotary kilns.

Gas Suspension Calciners (GSC) are designed with four (4) stages (CO1 - CO4) of direct heat recovery from hot alumina since the first units were designed in 1984 as seen from the GSC flow sheet in Figure 2 below.

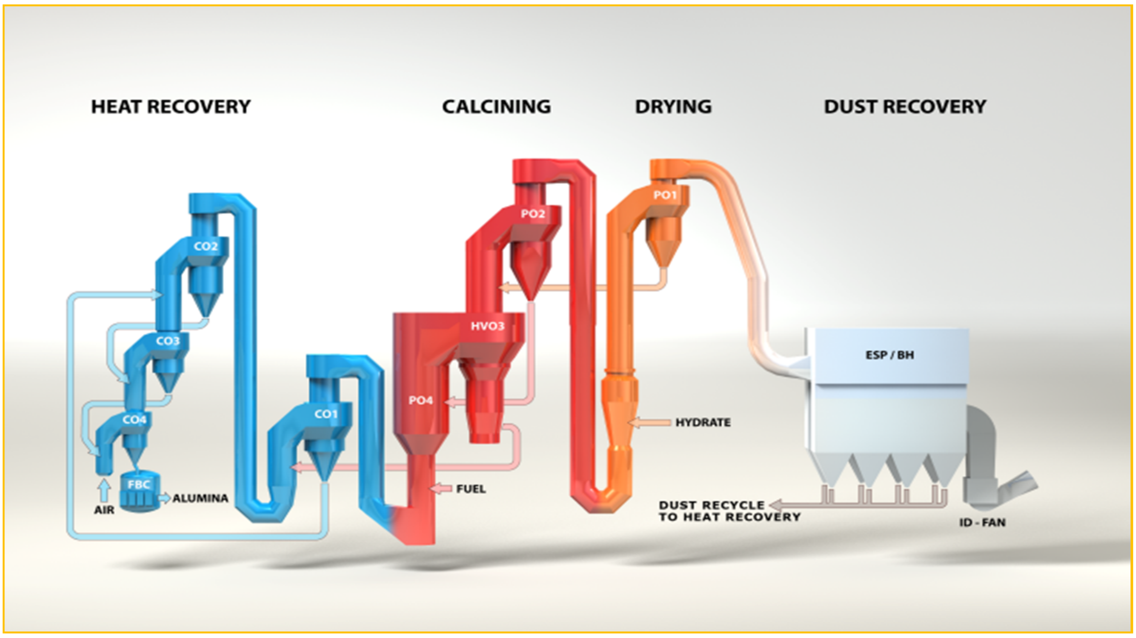


Figure 2. GSC flowsheet with 4 direct heat recovery stages (CO1-CO4).

This is the most energy efficient heat recovery flow sheet compared to other stationary calciner flowsheets, resulting in a yearly average specific heat consumption as low as 2722 kJ (NHV) / kg SGA, reported for two GSC units operating with a Calciner Furnace temperature of 1025 - 1050 °C and few seconds of solids retention time [3]. The availability, or Operating Factor was 98.8 % for both GSC units in the 32 month periods between the 27 day overhaul period for maintenance [3].

GSC units equipped with a fluidized holding vessel provide a few minutes of additional retention time, operating at 100 - 150 °C lower in the Calcination Furnace. See Figure 3 (rhs) and Table 1.

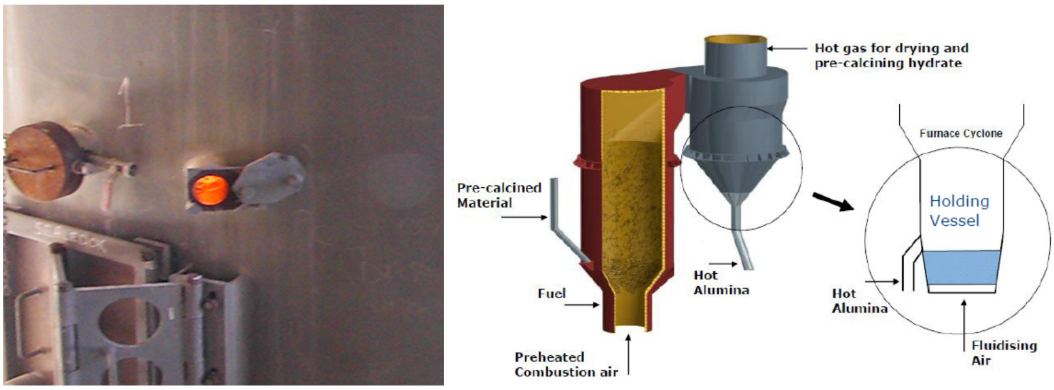


Figure 3. Combustion below Atmospheric Pressure - GSC with/without Fluidized H.V.

The impact on GSC operation of a Holding Vessel is to lower the temperature profile in the GSC unit and thus the specific thermal heat consumption as seen in Table 1, showing the average 2012 thermal energy consumption for each GSC unit in Plant A.

Table 1. Holding Vessel impact on calcination energy consumption.

GSC Units - Plant A	1	2	3	4
Holding Vessel	No	No	Yes	Yes
GJ (GHV*/NHV) / ton	3.30 / 2.98	3.35 / 3.03	3.10 / 2.80	3.14 / 2.84

*GHV = Gross Heating Value.

2. Gas Suspension Calcination in Theory and Practice

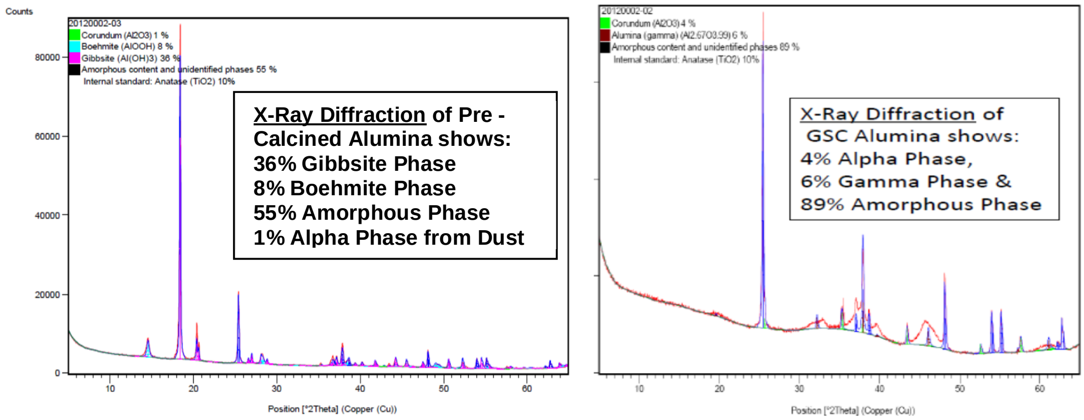
Due to the GSC design with part of the operating pressure below atmospheric pressure, it is possible to safely observe the initial combustion in the duct below the Calciner Furnace, PO4. See Figure 3 (lhs).

Samples of pre-calcined alumina from the underflow (U/F) of cyclone PO2 feeding the Calciner Furnace, PO4, can be extracted safely for analytical purposes.

In its simplest form the calcination of aluminium hydroxide to aluminium oxide or alumina can be formulated:

$$2Al(OH)_3\ \mathbf{(s)} + \Delta H_{Calcin.} \rightarrow Al_2O_3\ \mathbf{(s)} + 3H_2O\ \mathbf{(g)} \qquad (1)$$

Where $\Delta H_{Calcination}$ = 2011 kJ/kg Alumina or 480 kcal/kg Alumina is the standard theoretical endothermal energy of calcination at 25°C for this simplified calcination reaction.



X-Ray Diffraction of Pre - Calcined Alumina shows:
36% Gibbsite Phase
8% Boehmite Phase
55% Amorphous Phase
1% Alpha Phase from Dust

X-Ray Diffraction of
GSC Alumina shows:
4% Alpha Phase,
6% Gamma Phase &
89% Amorphous Phase

Figure 4. Phase composition of pre-calcined Alumina (lhs) and S.G.A. (rhs).

From X-ray diffraction analysis of pre-calcined alumina samples from the underflow of cyclone PO2, it becomes evident that the pre-calcination reaction forms only some Boehmite and large amounts of X-ray amorphous alumina phase as seen in Figure 4 (lhs) and Table 2 below.

Table 2. Typical GSC operating data with alumina phases formed.

GSC Unit	Design Capacity TPD	Holding Vessel HVO3	Temperatures °C		Gibbsite	Boehmite	X-ray Amorph.		"γ"	"α"
			PO2 U/F	PO4 O/F	Pre-Calcination - PO2			Smelter Grade Alumina		
Plt. A	3200	Yes	299	910	36%	8%	56%*	89%	6%	4%

(*) Including 1% Alpha Phase from Dust.

The progress of the calcination reaction with temperature, expressed by the measured LOI is shown in Figure 5 below (lhs). In the initial part of the calcination reaction the heat transfer rate from the gas phase to the solid particles is rate determining, but as the calcination reaction progresses it is the chemical phase changes that becomes rate limiting.

With the above observed data, the below *endothermic calcination reaction* scheme (2 - 4) can be proposed:

$$Al_2O_3{\cdot}3H_2O\ (\mathbf{c}) \overset{k_{GX}}{\longrightarrow} Al_2O_3\ (\mathbf{a}) + 3H_2O\ (\mathbf{g}) \qquad (2)$$

$$Al_2O_3{\cdot}3H_2O\ (\mathbf{c}) \overset{k_{GB}}{\longrightarrow} Al_2O_3{\cdot}H_2O\ (\mathbf{c}) + 2H_2O\ (\mathbf{g}) \overset{k_{BX}}{\longrightarrow} Al_2O_3\ (\mathbf{a}) + H_2O\ (\mathbf{g}) \qquad (3)$$

Where: **c, a, s, g** = crystal, amorphous, solid and gas phase;
k_{GX} = Phase transformation rate constant: Gibbsite to X-ray amorphous phase;
k_{GB} = Phase transformation rate constant: Gibbsite to Boehmite phase;
k_{BX} = Phase transformation rate constant: Boehmite to X-ray amorphous phase;

Reaction (2) shows the direct irreversible and *endothermic* dehydration of Gibbsite into X-ray amorphous alumina starting at the pre-calcination stage (Riser duct to cyclone PO2) at about 245 °C.

Reaction (3) takes place in the pre-calcination stage at around 300 °C and shows the direct irreversible and *endothermic* dehydration of Gibbsite via Boehmite into X-ray amorphous alumina.

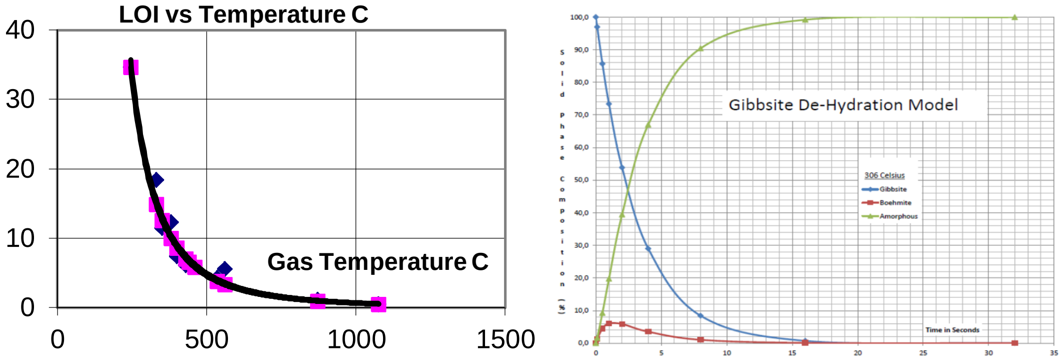


Figure 5. LOI versus temperature. Gibbsite dehydration model.

Assuming the reaction model proposed by Wang [4] applies with unchanged activation energies, and adjusted to plant data, the below *endothermic* Gibbsite Dehydration Model shown in Figure 5 (rhs), can be developed to match plant data.

$$Al_2O_3{\cdot}H_2O\ (\mathbf{c}) \ \rightarrow x\ \gamma\text{-}Al_2O_3(\mathbf{c}) + (1\text{-}x)\ \text{"}Al_2O_3{\cdot}yH_2O\text{"}(\mathbf{s}) + (1\text{-}x)\ (1\text{-}y)\ H_2O\ (\mathbf{g}) \qquad (4)$$

Reaction (4) illustrates the irreversible *endothermic* dehydration of Boehmite into a mixture of X-ray amorphous alumina phases, crystalline γ-alumina and non-stoichiometric hydroxyl-alumina compounds (i.e. $Al_{10}O_{16}H_2$ [5]) that takes place in the Calcination Furnace, PO4, at temperatures above 300 °C.

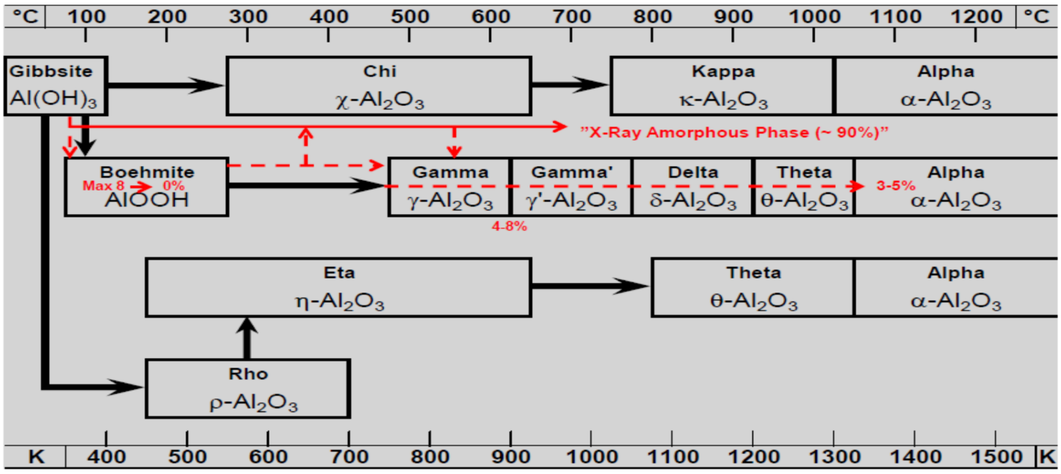
$$Al_2O_3\ (\mathbf{a}) \rightarrow \boldsymbol{\gamma}\text{-}\ ,\ \boldsymbol{\rho}\text{-},\ \boldsymbol{\chi}\text{-},\ \boldsymbol{\gamma}\text{-},\ \boldsymbol{\delta}\text{-},\ \boldsymbol{\Theta}\text{-}Al_2O_3 \qquad (5)$$

Reaction (5) illustrates the transition of some of the X-ray amorphous alumina phase into different, but poorly crystalline solid transition phases.

Reaction (6) shows the final and *exothermal* formation of some α-phase alumina, the only thermodynamic stable alumina phase.

$$\gamma\text{-}Al_2O_3 \rightarrow \boldsymbol{\rho}\text{-}, \boldsymbol{\chi}\text{-}, \boldsymbol{\gamma}\text{-}, \boldsymbol{\delta}\text{-}, \boldsymbol{\Theta}\text{-}, \alpha\text{-}Al_2O_3 \quad (6)$$

In view of the above it is perceivable that the Phase Transformation path of Gibbsite dehydration follows the "red-lined" path indicated in Figure 6.



[Adapted from Wefers and Misra, Alcoa 1987]

Figure 6. Phase transformation path of Gibbsite dehydration in GSCs.

Consequently, the previously 100 % Gibbsite dehydration reaction path via Boehmite assumed by FLSmidth for many years, is no longer valid. Since most of the SGA formed is X-ray amorphous, by-passing the Boehmite phase, this will also have impact on the endothermal energy of calcination [6].

SGA from a GSC Unit comprises mainly X-ray amorphous alumina phases (~ 90%), which is a highly disordered structure, beneficial for the dissolution process in the smelters [12].

3. Combustion Fundamentals - Net Heating Value and Air to Fuel Ratio

The heat for calcination is provided by the combustion of fuel in the Calciner Furnace, PO4. The fuel is introduced in the inlet duct to the calciner furnace through burner nozzles located on the diameter of the inlet duct. See Figure 7.

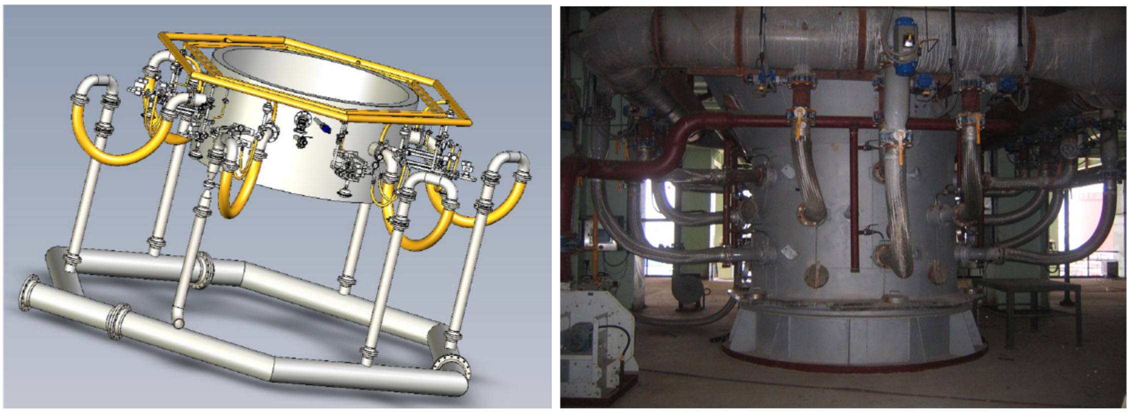


Figure 7. Arrangement of burner nozzles (lhs). Coal gas burner nozzles (rhs).

Production of alumina by calcination requires a relatively pure fuel with respect to solids ruling out almost all solid fuels. Combustion of Methane, CH_4, is discussed in some detail below to illustrate the general concept of the Low or Net Heating Value, ΔH_{NHV}, of the fuel being used:

$$CH_4\ (g) + 2O_2\ (g) = CO_2\ (g) + 2H_2O\ (g) + \Delta H_{NHV} \quad (7)$$

Since all calciners operate with a stack temperature in excess of 100 °C, the water formed during combustion (and calcination) leaves the stack as water vapor. This means that only the Low or Net Heating Value, ΔH_{NHV}, of the fuel is available for the calcination process.

$$2H_2O\ (g) = 2H_2O\ (l) + \Delta H_{Cond.} \qquad (8)$$

If the water vapor is condensed, then the heat of condensation, $\Delta H_{Cond.}$is released in addition to Net Heat Value, ΔH_{NHV}:

$$CH_4\ (g) + 2O_2\ (g) = CO_2\ (g) + 2H_2O\ (l) + \Delta H_{NHV} + \Delta H_{Cond.} \quad (9)$$

Adding reaction (7) and reaction (8), and thus adding the Low or Net Heat Value, ΔH_{NHV}, of the fuel and the heat of water vapor condensation, $\Delta H_{Cond.}$ the High or Gross Heating Value of the fuel is obtained:

$$\Delta H_{HHV} = \Delta H_{GHV} = \Delta H_{NHV} + \Delta H_{Cond.} \quad (10)$$

The heat recovered by condensation of the water vapor (Vol. - % H_2O) in the combustion products at 1 bar pressure and 25 °C is:

$$\Delta H_{Cond.} = 585 * \text{Vol - \% } H_2O * v_{oz} * 18/2240\ [\text{kcal/Nm}^3 \text{Fuel Gas}] \quad (11)$$

Table 3 shows some typical fuel data for selected fuel gasses and HFO [7], where the difference between Net and Gross Heating Values between HFO and Natural Gas are clearly demonstrated.

Table 3. Stoichiometric air demand, combustion products and heat of combustion.

Fuel	**v_{ol} (Air)**	**CO_2**	**H_2O**	**N_2**	**v_{oz} (Gas)**	**ΔH_{NHV}**	**ΔH_{GHV}**	**$\Delta H_G / \Delta H_N$**
Unit	Nm^3/Nm^3	Vol%	Vol%	Vol%	Nm^3/Nm^3	$kcal/Nm^3$	$kcal/Nm^3$	-
CH_4	9,57	9,5	18,9	71,6	10,57	8560	9499	1,110
Nat. Gas	9,71	10,1	18,3	71,6	10,76	8750	9676	1,106
Coal Gas	1,21	17,4	8,5	74,1	2,02	1375	1456	1,059
HFO	10,68 (*)	14,4	10,8	74,8	11,29 (*)	9680(**)	10253(**)	1,059

(*) Nm^3/kg, (**) Kcal/kg

Any modern and safe calcination system is equipped with a Burner Management System (BMS) to supervise safe combustion conditions in the plant by calculating and monitoring the minimum safe operating Air-to-Fuel Ratio (AFR) from measured flow rates of air for combustion and fuel flow [2].

The minimum safe AFR is defined as follows:

$$\text{Min. Safe AFR} > \text{Stoichiometric Air Requirement} + 5\% = 1.05 * v_{ol} = \text{AFR Trip Condition} \ (12)$$

$$\text{Minimum Operational AFR} > \text{AFR Trip Condition} + 5\% = \text{AFR Alarm Point} \qquad (13)$$

As an example, using the value for natural gas in Table 2 we find:

$$\text{Operational AFR} > 9.71 \times (1.05) \times (1.05) = 10.7\ (\text{Nm}^3\ \text{Air} / \text{Nm}^3\ \text{Natural Gas})$$

Due to the important safety function of the BMS, air and fuel flow into the calciner furnace, PO4, is measured continuously with the required accuracy and redundancy to safely guard against explosion risk at all times.

4. Factors Important for Low Specific Thermal Energy Consumption in Calciners

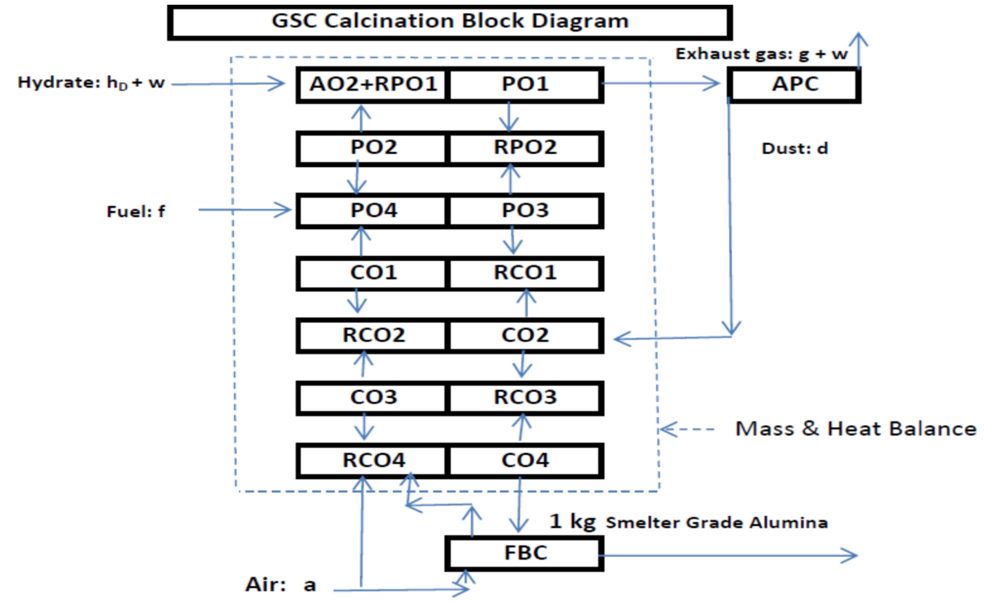
Thermal energy consumption in stationary calciners represents a significant part of the overall thermal energy consumption of any alumina refinery (e.g. 37–38 % at Alunorte, Brazil [8]).

Consequently, a basic understanding of the elements of the thermal energy balance for calciners and the significance of each element is vital for understanding and minimizing the thermal energy consumption of calcination. Below some of the major factors are discussed.

With reference to the list of symbols, the steady-state heat balance for the GSC can be formulated (see List of Symbols):

$$h_D\Delta h_{Hyd} + w\Delta h_{wl} + a\Delta h_{AiR} + K = g\Delta h_{Gas} + w\Delta H_{Cond.} + d\Delta h_{Dust} + \Delta h_{SGA} + \Delta H_{Calcin.} + q_{Loss} \qquad (14)$$

The heat balance is expressed per ton or kg of SGA produced as per equation (14), within the GSC battery limits shown in Figure 8. From Figure 8 it shall be noted that the heat recovered by the cooling water from indirect cooling of the alumina in the FBC, is *not included*.



NOTE: APC = Air Pollution Control by Electrostatic Precipitator or Bag House

Figure 8. Battery limits for calcination mass and heat balance calculations.

There are three major factors in minimizing thermal energy consumption in calciners:

(i) Steady-state operation;

(ii) Number of gas-to-solid / solid-to-gas Direct Heat Transfer (DHT) stages, and;

(iii) Design and installation of a cost efficient refractory lining.

The *first factor* of importance is to operate the calciner at steady state rather than in a non-steady state. In Figure 9, a steady state operating calciner shows little potential for reducing the specific energy consumption (SpEC) [2], when compared to un-steady state operation during commissioning of the GSC unit at Plant A.

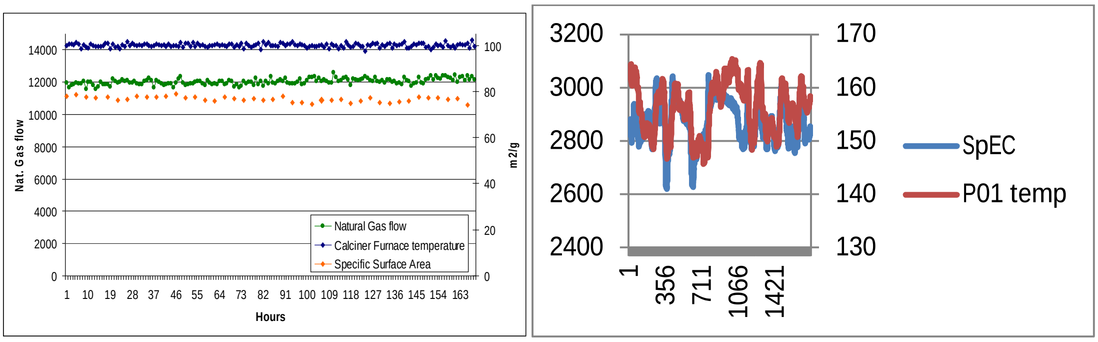


Figure 9. Steady state GSC operation. **Un-steady state GSC at commissioning.**

The *second factor* is to design with maximum economical number of gas-to-solid / solid-to-gas Direct Heat Transfer (DHT) stages for pre-calcination and heat recovery from hot alumina discharged from the Holding Vessel in the Furnace Cyclone, HVO3.

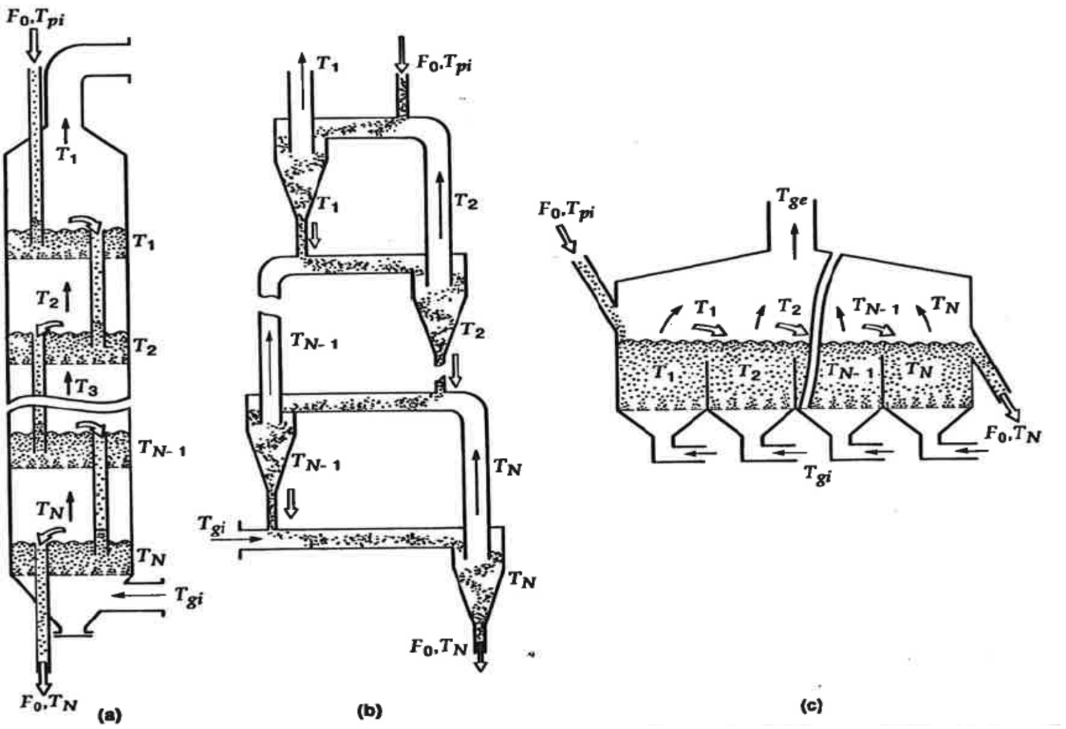
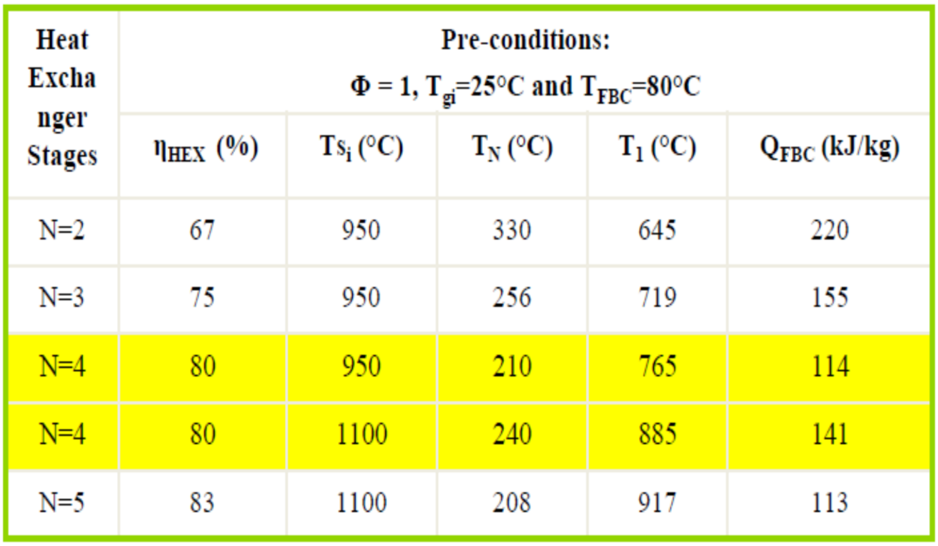


Figure 10. Four (4) direct heat recovery stages [13].

Of these, the latter design criteria is the most important for the cyclone cooler section owing to the large amount of heat of calcination consumed in the pre-calcination stage, PO2.

In Figure 10(b), the four (4) stage cyclone cooler, CO1 - CO4 is illustrated and in Table 4 below, the theoretical impact of number of DHT stages for heat recovery N, is calculated [9].

Table 4. Impact of number of DHR stages.



Heat Exchanger Stages	Pre-conditions: $\Phi = 1$, T_{gi}=25°C and T_{FBC}=80°C				
	η_{HEX} (%)	Ts_i (°C)	T_N (°C)	T_1 (°C)	Q_{FBC} (kJ/kg)
N=2	67	950	330	645	220
N=3	75	950	256	719	155
N=4	80	950	210	765	114
N=4	80	1100	240	885	141
N=5	83	1100	208	917	113

The theoretical thermal efficiency, η_{HEX} (%), increases with number of DHT stages N, in stage wise counter current flow of air and alumina, see (b) in Figure 10 [13].

Assuming a gas-solid heat-capacity-flow ratio, $\varphi = 1$, an inlet air temperature, $T_{gi} = 25$ °C, and an alumina discharge temperature from the fluid-bed cooler, $T_{FBC} = 80$ °C, the alumina inlet temperature to the fluid-bed cooler, $T_{N,}$ pre-heat temperature of the air for combustion, T_1, and heat loss to the fluid bed cooling water, Q_{FBC}, can be estimated in response to different cooler designs (N) and operating calcination temperature, T_{si}, in the Holding Vessel in the Furnace Cyclone, HVO3, assuming adiabatic condition.

For N = 4, it is seen that lowering the calcination temperature, T_{si}, with approximately 150 °C by introducing a Holding Vessel (Figure 3, rhs), lowers the heat loss from alumina to the fluid-bed cooling water by nearly (141 - 114) / 141 = 19 %. Compared to calciners designed with only 2 or 3 direct gas-solid heat recovery stages only, the theoretical heat loss with alumina to the fluid-bed cooling water is significantly higher.

The *third factor* of importance is an economic refractory design to minimize heat loss to the surroundings. From an operational point of view, the heat balance of the refractory lining is always in a dynamic state. This is due to the changing ambient conditions over each 24 hour day, 365 days a year, regardless of the GSC process being operated at steady state itself. For this, and a number of additional reasons to be discussed below, it is impossible to design a refractory lining to the same minimum specific heat loss value at all times.

The surface heat loss from the surface of the calciner plant equipment, ducts and piping, q_{Loss}, comprises three (3) elements; Radiation, Forced and Natural Convection losses expressed in kJ per kg Alumina:

$$q_{Loss} = \Sigma q_{Radiation} + \Sigma q_{Forced\ Convection} + \Sigma q_{Natural\ Convection} \qquad (15)$$

The main complexities involved are:

i. Constant unsteady-state heat balance of all of the refractory lined vessel and duct elements;
ii. Complicated fluid-dynamics around the calciner vessels, ducts and pipes;

iii. Hot spots formed by refractory anchors in refractory lined vessels and ducts (Figure 11);
iv. Complicated geometry and radiation exchange pattern between the vessels and ducts;
v. Extended heat transfer surface area of vessels and ducts by plate stiffeners and connecting flanges;
vi. Use of stainless steel needles in selected refractory areas increasing the effective thermal conductivity of the refractory lining;
vii. Quality of refractory materials selected, their installation and dehydration before the application;

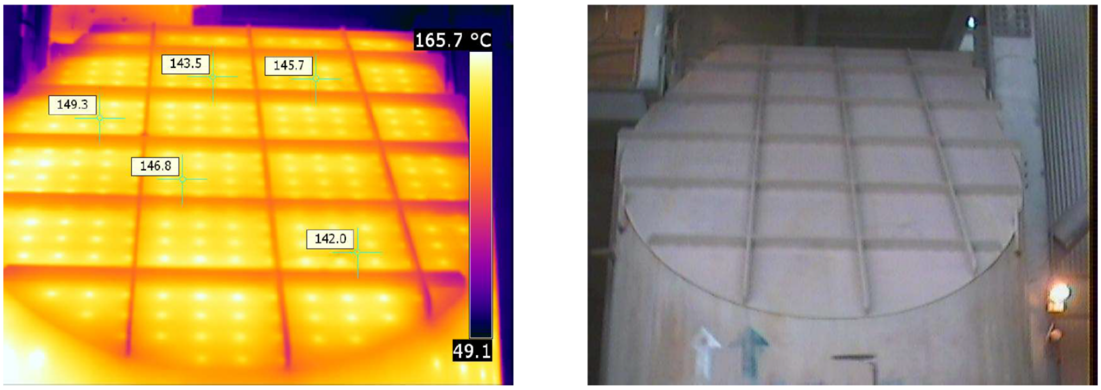


Figure 11. Infrared thermography survey showing a complex heat loss situation.

The calculation of each of the of the above heat loss contributions is well understood, but can only be done by using a number of very simplifying assumptions. Consequently, it makes little sense to do very detailed theoretical surface temperature calculations or make direct surface temperature measurements for heat loss estimation, rather than rely on experience.

5. Operational Experience and Thermal Energy Consumption Comparison

Table 5 below shows selected comparative mass and heat balance data for GSC plants A, B and C. The main columns for each plant are: GSC **Design**, reported operating GSC **Plant Data** and simulation of GSC Mass and Heat Balances (**M&H Bal.)** performed by FLSmidth.

Table 5: Comparative Mass & Heat Balance data for Gas Suspension Calciner units.

Parameter	Unit	Value	Value	Value	Value	Value	Value	Value	Value	Value
Source	MIPD	Design	Plant Data	M&H Bal.	Design	Plant Data	M&H Bal.	Plant Data	M&H Bal.	M&H Bal.
Data	-	A 10-12-2007	A 03-07-2015	A 03-03-2017	B 26-11-2007	B 01/03-04-2014	B 03-03-2014	C GSC Copy 2016	C GSC Copy 2017	C GSC Copy 2017
Alumina Capacity	TPD	3200	3192	3157	3500	3564	3695	1848	2047	2427
LOIA (300-1000C)	%	0,9	0,8	0,8	0,9	0,81	0,81	0,75	0,75	0,75
Hydrate Feed Moisture	%	6,0	6,1	6,1	7,0	4,2	4,2	6	6	6
PO1 O/F (Gas)	Celsius	161	150	150	155	145	145	177	177	147
PO2 Oxygen, Wet	%	2,5	3,2	3,2	2,6	2,5	2,5	NM	3	2,2
PO4 Outlet	Celsius	960	910	910	960	969	969	890 (+)	1093	905
CO4 U/F (Alumina)	Celsius	166	126	126	175	135	135	335	211	213
Specific Heat Cons. NG(LHV)	kJ/kg Ala	2783	2663	2662	2805	2718	2671	3369 (++)	3021	2845
Surface Loss + Heat of Evap & Calcination		2379	NM	2319	2410	NM	2353	NM	2399	2332
Total Sensible Heat Input	kJ/kg Ala	90	NC	89	96	NC	87	NM	15	16
Total Sensible Heat Output	kJ/kg Ala	495	NC	433	490	NC	407	NM	640	530
Total Sensible Heat Lost	kJ/kg Ala	405	NC	344	394	NC	320	NM	625	514

(+) Refractory Lining (++) Calculated from Main and Start Burner Consumption of Coal Gas.

The methodology for simulation/calculation of GSC Mass and Heat Balances was as follows:

1. *Plant Data* was maintained as reported by the GSC plant. This included; LOI of Alumina and free moisture of hydrate feed, fuel rate and specific heat consumption (if calculated), air intake rate (or measured Oxygen level) and measured temperature profile.

2. The calculated /simulated temperature profile from stage wise *Mass and Heat Balances*, were matched with measured plant temperatures by adjusting the specific surface heat lost from each stage of the GSC unit and the calcination capacity.

The production of alumina is seldom weighed directly, but calculated from a standard specific heat consumption applied to all calciners at the refinery, or calculated from the measured wet hydrate feed rate, moisture content and LOI of the smelter grade alumina. Very often the hydrate feed weigh belt has not been calibrated recently, so both of the above methods are rather uncertain. Neither has the specific energy consumption been determined for each individual calciner unit.

However, the measured temperatures are relatively accurate, subject to correct installation in the first place and regular maintenance over time.

As a result of the simulations the following can be observed:

Plant A Data vs M&H Balance: Higher calcination rate of 3192 tpd reported versus 3157 tpd calculated;

Plant B Data vs M&H Balance: Lower calcination rate was calculated based on applying an "average specific heat consumption" of 2718 kJ/kg SGA for all operating calciners in the refinery rather than the specific heat consumption of 2671 kJ/kg SGA calculated for the simulated GSC unit;

Plant A and B Design vs Plant Data: Very conservative (high) exit gas and alumina temperatures were accepted during design resulting in estimated excessive sensible heat losses with exhaust gas and alumina;

M&H Balance vs Design: Higher surface heat losses than used during design were partly compensated by lower sensible heat losses and heat of calcination from revised calcination model. The *Design* specific thermal energy consumption exceeds the realized values by 4-5%.

In view of the above, it can be concluded that further reduction in specific thermal energy consumption of GSC units can be achieved by improved refractory design and thus lower surface heat losses.

6. Capacity Upgrade with Reduction of Specific Thermal Energy Consumption

In 1991, FLSmidth Minerals was the first to start-up GSC units using coal gas as fuel at the Shanxi Aluminium Plant in China [10]. This was so successful that many different Chinese companies have built many more copies of GSC units [11] than the nine GSC units (820 - 1900 tpd), supplied by FLSmidth Minerals to China since 1987.

The *Plant Data* for *Plant C* in Table 5 is from a Chinese copied GSC unit fired with coal gas. However, the copied FLSmidth GSC unit (supplied to Zhengzhou, 1991, [10]) was designed for HFO, but copied without making appropriate design changes to critical vessel and duct dimensions.

Comparing *Plant Data* with *M&H Balance* data from *Plant C*, it can be seen that the calcination capacity achieved is 10 % less than predicted (1848 vs 2047) and the specific thermal energy consumption is 11 - 12 % higher than estimated (3369 vs 3021) for an FLSmidth designed GSC unit for Coal Gas firing.

Furthermore, there seems to be a significant potential for retrofitting the copied GSC units with a Holding Vessel and FD - Fan for adding more air for combustion and thus capacity.

If a capacity upgrade is decided, the calcination rate of the retrofitted GSC units can be increased from 1848 tpd to about 2427 tpd or about 31 %, while the specific thermal energy consumption can be reduced from 3369 KJ/kg SGA to 2845 KJ/kg SGA or a reduction of about 16 % on NHV basis.

7. Conclusion

In view of the above the below conclusions are suggested:

i. GSC Units designed with a fluidized Holding Vessel lower the Specific Thermal Energy Consumption without taking any credits for heat recovered by cooling water;
ii. Gas Suspension Calcination of Bayer Hydrate results in SGA with highly disordered, or X-ray amorphous alumina phases, beneficial for the dissolution process;
iii. Steady-state GSC operation, design with four stages of direct heat recovery and optimized refractory design points to a minimum Specific Thermal Energy Consumption;
iv. Retrofitting a GSC Unit with fluidized holding vessel suggests a significant increase in Alumina productivity at significant lower Specific Thermal Energy Consumption.

8. List of Symbols

a = Air intake rate (Nm^3/ Kg Al_2O_3);
d = Dust recycle rate (Kg/Kg Al_2O_3);
f = Fuel intake rate (Nm^3 Fuel Gas/ Kg Al_2O_3) or (Kg HFO/ Kg Al_2O_3);
g = Exhaust gas flow from cyclone PO1 (Nm^3/ Kg Al_2O_3);
h_D = Dry hydrate feed fate (Kg/Kg Al_2O_3);
K= f ($\Delta h_{Fuel}+\Delta H_{NHV}$), is the specific thermal energy consumption in KJ/Kg or MJ/Ton SGA;
$q_{Forced\ Convection}$ = Heat loss by forced convection (KJ/ Kg Al_2O_3);
q_{Loss} = Sum of heat losses by radiation, forced and natural convection (KJ/ Kg Al_2O_3);
$q_{Natural\ Convection}$ = Heat loss by natural convection (KJ/ Kg Al_2O_3);
$q_{Radiation}$ = Heat loss by radiation (KJ/ Kg Al_2O_3);
w = moisture in hydrate (Kg/Kg Al_2O_3);
Δh_{AiR} = Specific Enthalpy (sensible heat) of ambient air plus air from Fluid Bed Cooler (KJ/Nm^3);
ΔH_{Calcin} = Enthalpy of Calcination KJ/ Kg Al_2O_3 at 25°C and 1 bar absolute pressure;
$\Delta H_{Cond.}$= Heat of water vapor condensation/evaporation at 25°C and 1 bar absolute pressure;
Δh_{Dust} = Specific enthalpy (sensible heat) of dust recycled to cyclone CO2 (KJ/ Kg Al_2O_3);
Δh_{Fuel} = Specific enthalpy (sensible heat) of moist fuel in (KJ/Nm^3 Fuel Gas) or (KJ/Kg HFO);
Δh_{Gas} = Specific enthalpy (sensible heat) of moist exhaust gas (KJ/Nm^3 Gas);
Δh_{Hyd} = Specific enthalpy (sensible heat) of dry hydrate (KJ/ Kg Al_2O_3);
ΔH_{NHV} = Net Heating Value of fuel in (KJ/Nm^3 Fuel Gas) or (KJ/Kg HFO) at 25°C and 1 bar absolute pressure;
Δh_{SGA} = Specific enthalpy (sensible heat) of SGA discharged from cyclone CO4 underflow (KJ/ Kg Al_2O_3);
Δh_{wl} = Specific enthalpy (sensible heat) of water in dry hydrate (KJ/ Kg H_2O);

$\Delta h_x = \int C_{px}(T)dT$ from $T_o = 298^o$ Kelvin to T^o Kelvin, is the specific single phase enthalpy of a single component i.e.water (Δh_{wl}) or gas mixture, i.e. like air (Δh_{Air}) or gas products from combustion, evaporation and calcination, (Δh_{Gas}).

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