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Abstract
<p>This report summarizes the technical basis and state-of-the-art designs of the calciner technology used in the cement industry. The calcination reaction, which form the basis of considerations, and its requirements on the technology concerning temperature level etc. are described. In the course of development different calciner types have been developed and are applied with respect to the plant-specific requirements. The main groups are: extended duct, separate line and in-line calciners. As basis for the oxyfuel investigation typical operational values are listed. Previous theoretical investigations and some practical experience on the calcination behavior during oxyfuel mode and the expected impact on the operation are described.</p>

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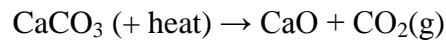
1 INTRODUCTION

First introduced in 1966 initially as air-through riser-firing the calciner technology nowadays increases clinker production and ensures that the rotary kiln keeps a manageable dimension. In 1973 IHI of Japan first developed and introduced the flash calciner process with tertiary air duct by employing an additional vessel between the bottom and second lowest cyclone stage. In addition to the kiln main burner the calciner is equipped with at least one burner firing up to 60% of the total fuel input and thermal heat demand. Combustion air for the calciner is taken directly from the clinker cooler via a separate duct, so called tertiary air duct. The material for the calcining process is discharged from the 2nd lowest cyclone stage while the reaction product is collected in the lowest cyclone. After a few seconds the material is then discharged to the kiln inlet. At this point in time apparently over 90 per cent of the material supplied is already calcined including some recycled kiln dust which is formerly a 100 per cent calcined. In contrast to a preheater kiln (without precalciner) the kiln speed is about 3.5 rpm or higher which is much faster than the preheater kilns. Together with the high degree of precalcination in front of the kiln and the higher rotational speed freshly-calcined material is rapidly transported into the burning zone preventing it becoming dead-burnt and unreactive. Due to the reduced degree of calcination required in the rotary kiln, kiln size could be reduced by about 50 per cent and higher output can be reached with the same kiln size. This milestone has generated a lot of benefits e.g. maintaining linings in large-diameter kilns and allowance of scale up of production by factor of two or more while fuel efficiency is barely improved. Due to the lower required temperature level of 900°C alternative (waste based) fuels with lower heating values than fossil fuels can be applied. Beside of that in combination with precalciner-facilities low NO_x designs are made possible due to the lower amount of thermal NO_x on the basis of a lower temperature precalciner combustion regime which also allows using low grade fuels. For these reasons the precalciner kiln has been virtually the only type installed over the decades.

2 TECHNICAL BASICS

In a cement kiln calcination is the process of decomposition of limestone in solid calcium oxide and gaseous carbon dioxide (44 % of the mass). This reaction is required before the clinkerization process forming the relevant minerals like alite and belite takes place.

The main chemical reaction of limestone calcination, a decomposition process, is



About 2/3 of total thermal energy consumption of the clinker production process is spent on this reaction which can be split into 3 partial steps:

1. Heat transfer from the surrounding to the particle and heat conduction through the particle to the temperature required for the decomposition
2. Chemical reaction
3. Diffusion of the CO_2 through the particle and material transfer from particle surface to the surrounding

Calcination rates in excess of 90% are achieved with approximately 60% of the total fuel directed to the calciner. As combustion and calcination of raw material take place outside of the rotary kiln, the heat loading of the kiln size can be reduced.

Figure 2-1 shows the equilibrium pressure of CaCO_3 against the reciprocal temperature. With increasing temperature equilibrium pressure is increasing, too. The temperature at which limestone yields calcium oxide in the calciner of a cement plant is usually given as 825 °C, but stating an absolute threshold is misleading. Calcium carbonate exists in equilibrium with calcium oxide and carbon dioxide at any temperature. At each temperature there is a partial pressure of carbon dioxide that is in equilibrium with calcium carbonate. During the process of heating up raw meal decarbonation starts at about 600-700°C and ends between 900-1000°C, following a so-called “S-curve”. Since the available retention time is very short (a few seconds) and due to heat transfer processes the required temperature is about 20°C above theoretical values.

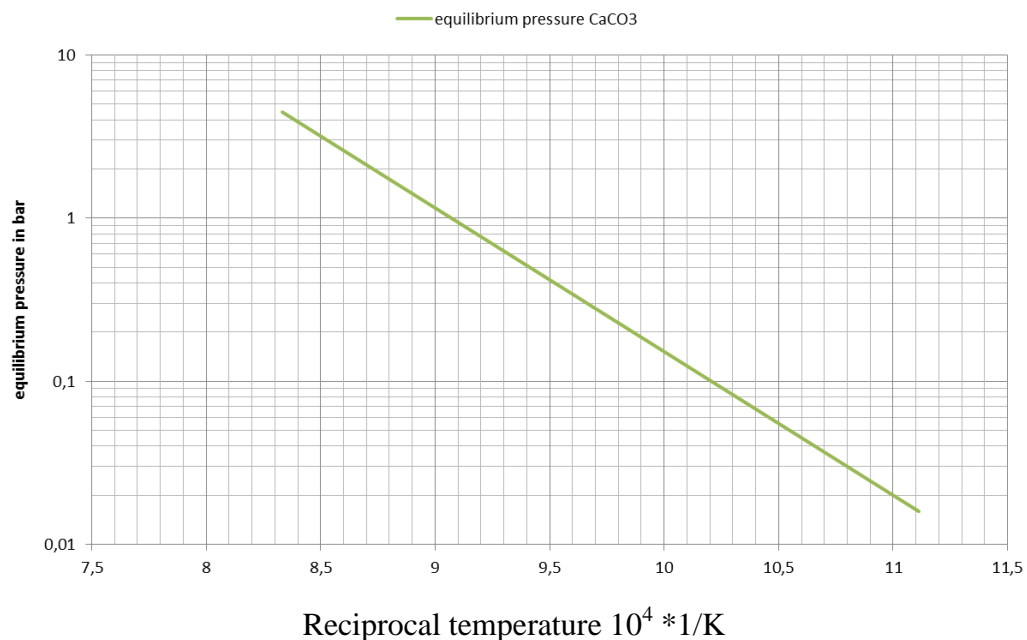
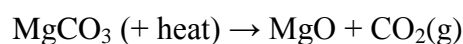


Figure 2-1 Equilibrium Pressure of CaCO₃

Due to the optimum temperature range between 830 to 950°C inside of the calciner this technique allows a flexible usage of for example alternative fuels with lower heat values or coals with a high amount of ash. In addition a staged combustion inside of the calciner is allowing a significant reduction of kilns NO_x-emission and has therefore become state-of-the-art. Different calciner types basically differ in their location of fuel distribution, their kind of fuel, material and tertiary air distribution and lastly their geometrical form.

Raw material also contains a small amount of MgCO₃ which also releases CO₂. The reaction is described as



Magnesium carbonate is like calcium carbonate an inorganic salt that is a white solid. The most common magnesium carbonate forms are the anhydrous salt called magnesite. By calcination magnesium carbonate forms magnesium oxide, so-called magnesia, while releasing gaseous carbon dioxide. Both described reactions occur simultaneously while the reaction of calcium carbonate predominates.

3 DECARBONATION REACTION INDICATORS

For characterization of decarbonation reaction there are numerous written references on several reaction indicators. Some of them are listed below:

Degree of Decarbonation ε

$$\varepsilon = \frac{CO_{2,0} - CO_{2,\tau}}{CO_{2,0}}$$

with $CO_{2,0}$ as the initial amount of bound CO₂ and $CO_{2,\tau}$ as the actual amount of driven out CO₂.

Degree of Loss of Ignition

$$\varepsilon_{LOI} = \frac{LOI_0 - LOI_\tau}{LOI_0}$$

with LOI_0 as the initial loss of ignition and LOI_τ as the actual loss of ignition.

4 STATE OF THE ART

Most modern calciners are based on a modular system. Systems can be designed according to the fuel used or other specific requirements. In general the following design groups of calciners may be distinguished:

1. Extended Riser Duct
2. Separate line calciner
3. Inline calciner with separate Combustion Chamber
4. Inline calciner (with staged combustion)

Within these design groups equipment suppliers offer different structural solutions based on the specific boundary conditions (such as fuel type, raw materials, emission control etc.) of the cement plant. In the following the designs are shown on a general basis in order to provide a reliable investigation basis without giving any bias to a certain supplier design and avoid conflicts of interest.

4.1 Extended riser duct design

The simplest “calciner” is the extended riser duct without tertiary air duct also called “air through calciner” (Figure 4-1). The combustion air required for the calciner fuel is passing the kiln. Advantage is that the thermal load of the kiln is reduced. Such arrangement allows about 20% firing rate and is used for modifications only.

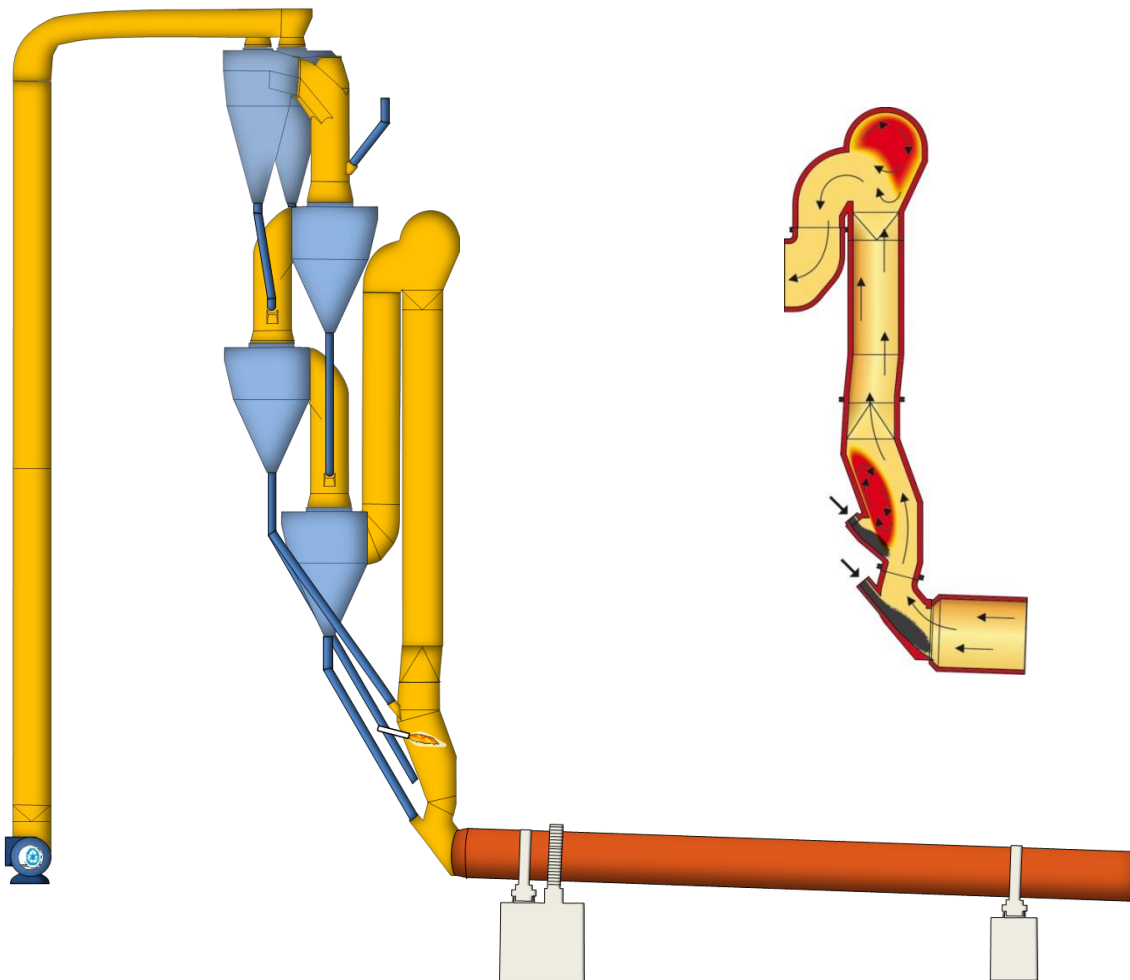


Figure 4-1 Extended Riser Duct

4.2 Separate line calciner design

The separate line calciners have a calciner connected to the tertiary air only (Figure 4-2). The kiln gas is routed to the bottom cyclone directly. All meal is going along the calciner before entering the kiln. This solution was often realized to boost the output of a preheater kiln by installing a second preheater with this type of calciner and installing a new cooler. Drawback is that separate line calciners are restricted to primary fuels and risk of tertiary air duct elbow blockage cannot be denied. A typical sketch is enclosed. Various arrangements in regard to meal routing are available.

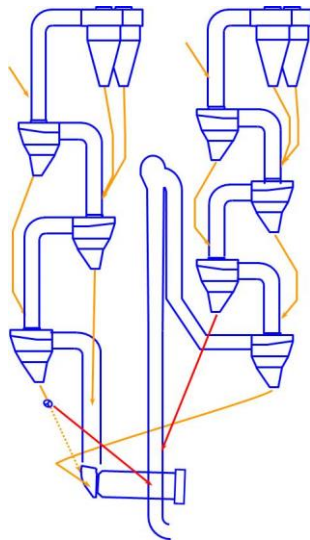


Figure 4-2 Separate Line Calciner

4.3 Inline calciner design with separate combustion chamber

At the very beginning calciner consisting of a separate combustion chamber and a short inline riser duct were used (Figure 4-3). Tertiary air is introduced tangentially into a chamber, gas is moving in a spiral downstream, and connected to calcining channel/riser duct. The fuel is introduced from top of the chamber into the center. The meal is introduced close to the tertiary air inlet creating a high meal concentration along the wall and acts as “meal curtain”. In the center the meal concentration is low and fuel dried and ignites in pure hot air. This arrangement is favorable for fuels which are hardly to ignite since the ignition and burn out process is more intensive due to the high oxygen concentration. Burning coarse alternative fuels became possible because ignition and start of burning results in decreasing the density and such fuels became flyable. In the calcining channel the final fuel burn out takes place. With longer calciner channels burning alternative fuels become common practice.

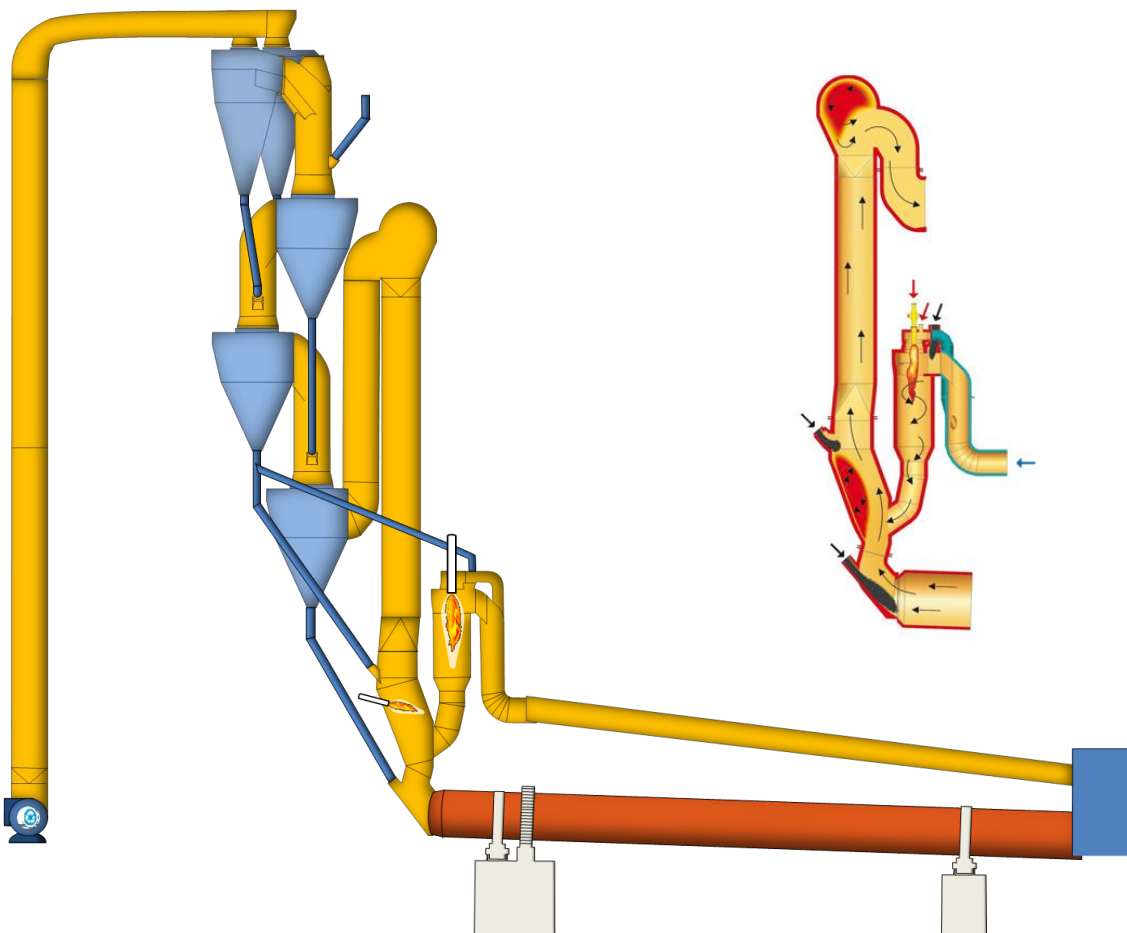


Figure 4-3 Inline calciner with separate Combustion Chamber

4.4 Inline calciner design (with staged combustion)

The inline calciner is the most often realized and up to date solution (Figure 4-4). In general kiln gas and tertiary air entering the calciner and meal from the second lowest stage and the fuel is added. Main fields for developments were:

- Elongation of the calciner channel to prolong the gas retention time to improve the fuel burn out
- Installing an upper deflection point of the calciner in such a way that the gas is not only turning from upstream to downstream rather mixed to eliminate gas strands. By that measure burn out could be improved
- Introduction of meal at different levels to influence the temperature to avoid overheating and improve fuel ignition conditions
- Introduction of tertiary air at different levels to operate the calciner with less oxidizing or even reducing conditions within a certain volume (staged combustion)

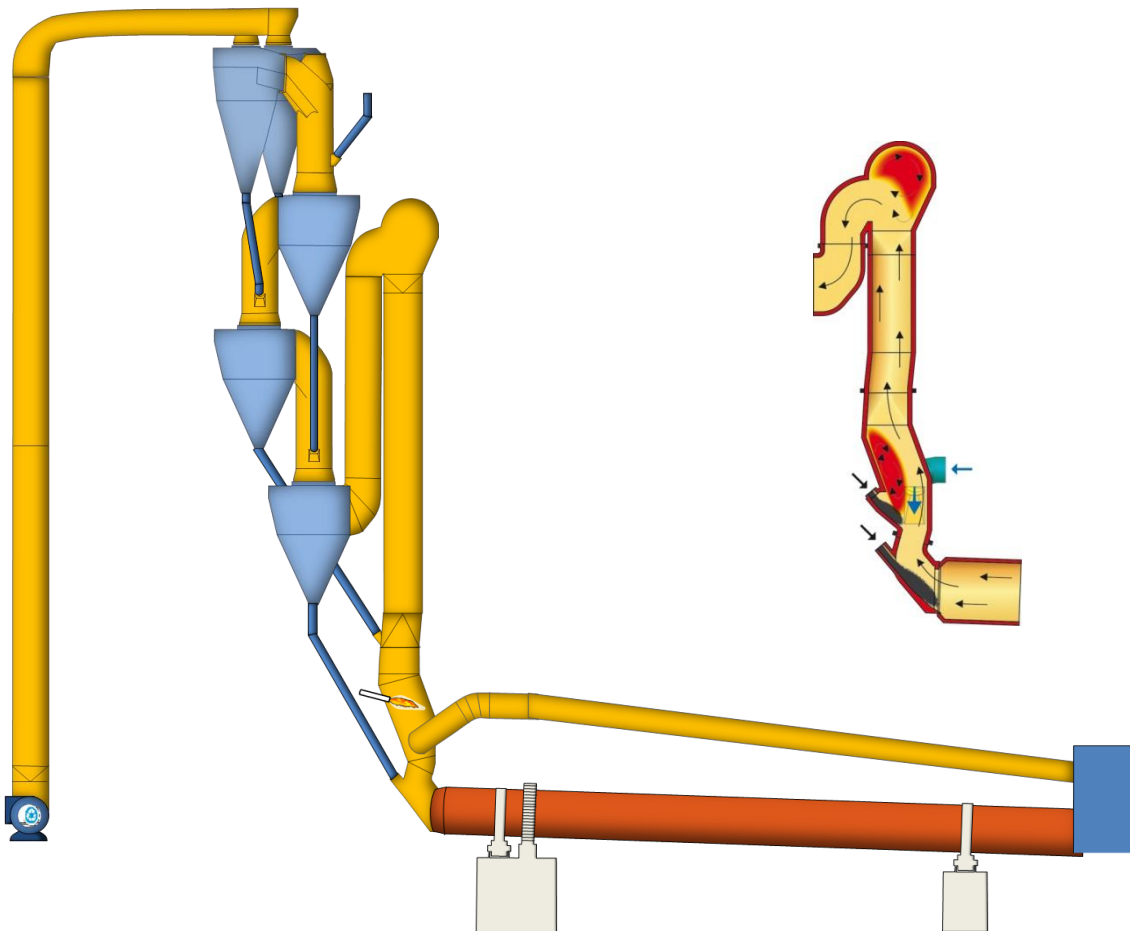


Figure 4-4 Inline calciner

Moreover low-NO_x calciners are designed as a measure for NO_x reduction that is integral with the process. This variant is characterized by staged combustion in the calciner (Figure 4-5). Fuel is introduced in the kiln gas between kiln inlet chamber and tertiary air inlet to create a reducing zone to produced deliberately amount of CO to achieve a NO_x-reducing effect. Precondition is a low oxygen concentration in the kiln gas and easy to ignite fuels. The final burn-out of the calciner fuel then takes place in the burn-out zone operated with excess air.

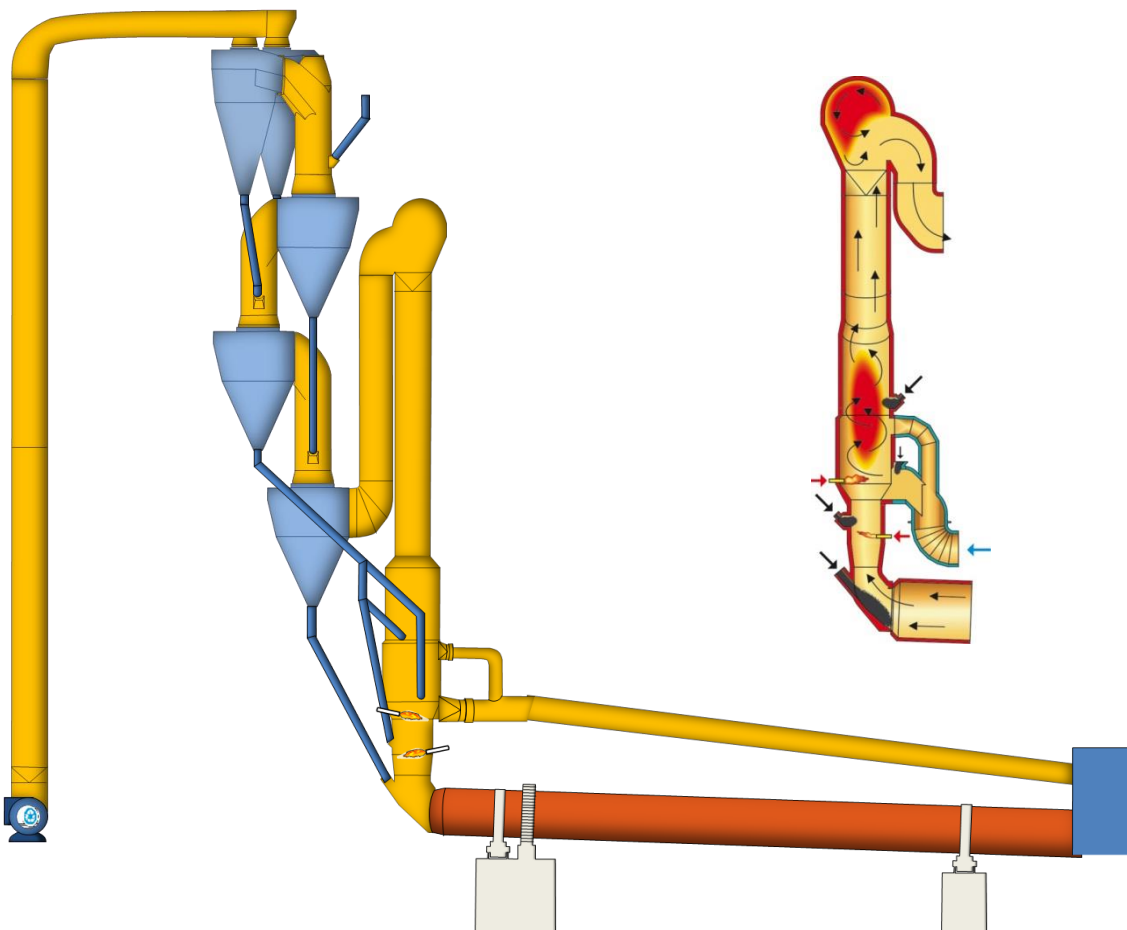


Figure 4-5 Inline calciner with staged combustion

4.5 Summary

The main function of the calciner is the calcination process. All state of the art calciners from various suppliers follow the above mentioned principles of a modern inline calciner.

While primary fuels like natural gas, heavy fuel oil or good quality hard coal was fired the required retention time for burn out and calcination was in the same range, about 3 seconds. Replacing above mentioned fuels by lower grade fuels like local coals, petcoke and nowadays flyable alternative fuels like plastic foils and bio fuels the gas retention time and ignition properties dominate the process. The calciner fuel amount is controlled by the temperature at the calciner end or after the bottom stage. The set point for this temperature is adjusted regularly on basis of the precalcination rate of the hot meal entering the kiln. As long as the composition of the fuel and the heat value is constant the entire system is running smoothly. Any disturbance requires stronger interaction. Controlling the calciner with poor quality and inhomogeneous fuels is a challenge. Since the combustion gas amount increases with poor fuel quality the fuel split, temperature after calciner and calcination rate must be carefully adjusted. Too much fuel results in overheating of the system which increases the overall heat demand and results in cyclone blockages since overheated material becomes sticky. Not enough fuel reduces the precalcination rate which requires more fuel to the main burner which increases the thermal kiln

load. Furthermore at the calciner lower quality fuel can be fired compared to the kiln, why the target is to fire as much as possible fuel to calciner to reduce cost. Longer calciners with approx. 7 seconds gas retention time and functional zones to improve the fuel ignitions conditions are state of the art. Separate line calciners are outdated and not suitable to burn coarse alternative fuels. Calciners with combustion chambers allow burning limited flyable alternative fuels but react sensitive to the moisture of the fuels.

Table 4-1 Advantages and disadvantages of calciner technologies

Design	Advantages	Drawback
Extended riser duct	<ul style="list-style-type: none"> Reduction of thermal kiln load 	<ul style="list-style-type: none"> Limited to about 20% calciner fuel
Separate line calciner	<ul style="list-style-type: none"> easy to install solution to increase the plant capacity by adding calciner and 2nd preheater 	<ul style="list-style-type: none"> not suitable for alternative fuels due to risk of material fall through
Separate Combustion Chamber off-line Calciner	<ul style="list-style-type: none"> the fuel is ignited in the pure tertiary air combustion chamber provides additional retention time allows firing of coarse alternative fuels because while fuels ignite the density reduces 	<ul style="list-style-type: none"> high thermal stress of refractory due to open flame in the core pressure restricted to low fuel moisture otherwise the combustion chamber becomes a drying chamber only
Simple inline calciner	<ul style="list-style-type: none"> simple, tertiary air inlet at one level, one meal inlet 	<ul style="list-style-type: none"> suitable for good quality fuels only
State of the art inline calciner	<ul style="list-style-type: none"> high flexibility due to introduction of fuel, meal and tertiary air at various levels 	<ul style="list-style-type: none"> more difficult to adjust more equipment to maintain

5 OPERATIONAL PARAMETERS

As mentioned in the above evaluation of the different calciner designs, the layout can vary depending on the intended purpose (e.g. maximizing alternative fuel rate, for emission control etc.). Nevertheless the following parameters are valid for each of the designs (Table 5-1).

Table 5-1 Typical operational parameters of existing calciners

Operational Parameter	Value
Gas velocity	10 to 20 m/s (with constrictions 35 m/s)
Residence time	3-9 s
Gas composition	20 – 30 vol.% CO ₂
Temperature level	850 - 900°C at calciner end, up to 1100°C in functional zones
Degree of calcination of material entering the calciner	10 to 20 %
Degree of calcination of material leaving the calciner	Up to 95 % (with TAD)
Raw material particle size	~ 12 % above 90 µm 1- 1.5 % above 200 µm

Within the course of the clinker burning process the chemical composition of the material changes, due to the drying, calcination, implementation of fuel ashes and remix with counter-current flowing dust. Table 5.2 shows exemplarily the chemical composition of the material entering and leaving the calciner as well as the main clinker phases. It becomes clear, that the CO₂ content is reduced and intermediate clinker phases are generated after the calciner.

Table 5-2 Material composition (in accordance with the CEMCAP framework reference case)

Parameter	Unit	Preheater inlet	Calciner inlet	Calciner outlet
SiO ₂	wt%	13.80	15.36	20.91
Al ₂ O ₃	wt%	3.25	3.69	5.33
TiO ₂	wt%	0.06	0.05	0.001
P ₂ O ₅	wt%	0.04	0.05	0.08
Fe ₂ O ₃	wt%	1.96	2.19	3.01
Mn ₂ O ₃	wt%	0.05	0.04	0.0004
CaO	wt%	43.22	47.75	63.61
MgO	wt%	0.71	0.81	1.23
CO ₂	wt%	34.74	28.62	2.81
SO ₃	wt%	0.34	0.44	0.73
Sulfide ¹⁾	wt%	0.05	0.003	0.00
K ₂ O	wt%	0.55	0.76	1.50
Na ₂ O	wt%	0.12	0.17	0.33
Cl	wt%	0.01	0.16	0.68
TOC	wt%	0.10	0.00	0.00
H ₂ O	wt%	1.00	0.00	0.00

Minerals	Unit	Preheater inlet	Calciner inlet	Calciner outlet
C3S	wt%	0.00	0.83	4.29
C2S	wt%	0.00	2.79	14.33
C3A	wt%	0.00	0.18	0.91
C4AF	wt%	0.00	0.14	0.74
CA	wt%	0.00	1.58	7.43
C2F	wt%	0.00	1.00	4.68

Due to the calcination and thus release of CO₂ the total mass of the raw material is reduced. Therefore the solid to gas ratio differs with the calciner's length. Moreover the amount of gas depends on different operational factors, e.g. most important: The oxygen (and therefore air intake) needed for the combustion is determined by the used fuels and the overall energy demand of the plant. Examples for that are given in Table 5-3.

Table 5-3 Solid to gas ratio at calciner inlet and outlet (in accordance with the CEMCAP framework reference case)

Scenario	S/G-ratio at calciner inlet	S/G-ratio at calciner outlet
Reference plant with 100 % coal	2.0 kg/Nm ³	1.0 kg/Nm ³
Reference plant with 60 % coal and 40 % alternatives (mix of RDF, animal meal, sewage sludge)	1.8 kg/Nm ³	0.9 kg/Nm ³
Reference plant with 100 % alternatives (mix of RDF, animal meal, sewage sludge)	1.4 kg/Nm ³	0.8 kg/Nm ³

5.1 Definition of a reference calciner

In order to keep the design for the evaluation as simple as possible the common inline calciner design as described in section 4.4 is chosen.

According to the reference case as defined in the CEMCAP preliminary framework document and in line with the above mentioned typical operational parameter (3 seconds, 15 m/s gas velocity) the reference calciner exhibit the following dimensions:

- Length: 45 m
- Inner diameter: 3.9 m

6 CALCINATION UNDER OXYFUEL CONDITIONS – OVERVIEW OF PREVIOUS INVESTIGATIONS

6.1 Influence of Oxyfuel conditions on calcination and limitation factors

As mentioned in section 2 there is a common understanding of calcination behavior of pure calcium carbonate. The calcination reaction is strongly depending on the temperature level and the CO₂ partial pressure in the surrounding. If the CO₂ partial pressure rises the equilibrium temperature has to be increased to overcome the reaction barrier and achieve similar degree of calcination. This circumstance might cause problems in the clinker burning process with:

- overheating the calciner walls
- causing increased coating formation
- limiting the use of low calorific alternative fuels.

Moreover the theoretical examination of the system of pure calcium carbonate shows that no complete calcination could take place in case of 100% CO₂ in the surrounding atmosphere in the relevant temperature range up to 1,000°C. In realistic cases only CO₂ levels up to 85 vol.% seem to be achievable due to dilution by ambient air, other volatiles from fuels, fraction of humidity and the excess oxygen. In static systems (under lab conditions) the CO₂ from dissociation leaving the inside of the particle could form a layer of highly concentrated CO₂ around the particle, which suppress further calcination reaction. In entrained flow reactors like calciner reactors it is expected that the gas flow will create a shearing force, which removes this layer.

Usually the calcination reaction takes places within seconds. For this reason today's calciners are designed for residence times from 3 to 9 seconds, mainly depending on the distances needed for complete fuel burn-out (especially with regard to alternative fuels). The higher reaction barrier due to the CO₂ rich surrounding could influence the kinetics and therefore make higher residence times necessary to achieve similar degree of calcination.

However, it is still not completely proven to which degree these factors will negatively influence the function of clinker burning process. In the following different pre-studies and other pilot tests will provide evidence that the named factors will be in a range which should not stop the application of oxyfuel in cement kilns, but might make adaptations necessary. Although even in case of achieving lower degree of calcination, counter measures like adapting the residence time in kiln or local dilution of the CO₂ rich gas could be implemented.

6.2 Previous projects and background information

Several investigations have been made on the dissociation of calcium carbonate with regard to the reaction kinetics on lab scale in the past. Still open is the up-scale of these results to cement raw meals and industrial scale calciners. Most recent experiences are listed below.

6.2.1 Experience from calcium looping processes

Calcium looping processes as post-combustion capture method especially for power plant applications have been intensively investigated in the last years. Within this process calcium oxide is used as sorbent for the CO_2 by forming CaCO_3 . In order to create a nearly pure CO_2 stream for transport/storage, the material is calcined under oxyfuel conditions in a subsequent step. Although the same reaction under similar conditions is used, the reactors are different. Current investigations are made in fluidized bed reactors allowing up to 2-3 minutes of residence times; thus it is expected to have reached thermodynamic equilibrium. On the other hand bigger particle sizes are used up to 200 -300 μm than in cement kilns which hinder the heat transfer to the particle and CO_2 diffusion to the outside of the particle.

For these reasons these investigated does not allow a direct transfer of results to cement kiln calciners, but could provide first evidence about, whether calcination takes place under oxyfuel conditions and at which temperature levels. Following experiences are reported:

- Lab tests: Complete calcination at temperature levels of 880-920°C. [12]
- Tests at 200 kW_{th} pilot plant: Complete calcination at temperature levels of 875 – 930°C under oxyfuel conditions. [13]
- Tests at 1.7 MW_{th} pilot plant: Calcination of limestone at temperatures of 920 – 950°C and under high CO_2 concentration in the surrounding atmosphere of up to 90 vol.% (compare Figure 6-1). As capture efficiencies are limited by carbonation step a complete calcination could be assumed. [14]

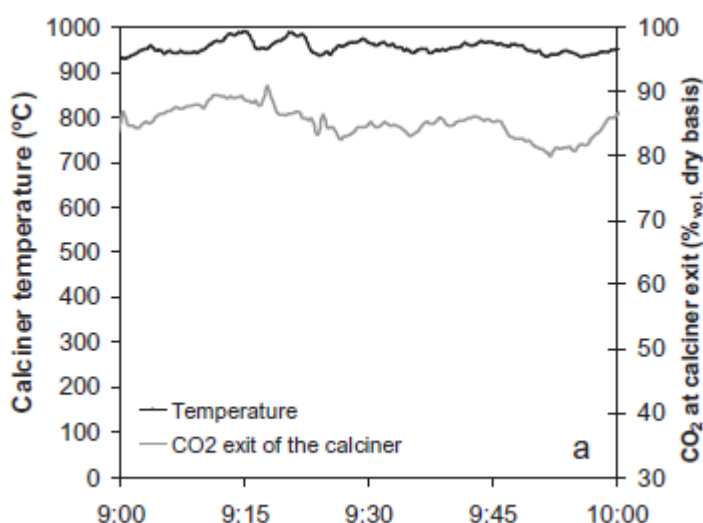


Figure 6-1 Calcination temperature in dependency of CO_2 concentration for calcium carbonate [14]

This practical experience proves the temperature shift, but also the feasibility of calcination under oxyfuel conditions. Temperature up to 920°C would be at least tolerable but not optimal for cement kiln lines. As these results are only valid for pure calcium carbonate systems, it is

expected that the peak temperatures are lower for cement clinker raw materials due to simultaneous reactions.

6.2.2 Theoretical investigations of ECRA

In contrast to pure calcium carbonate which is used in calcium looping processes, cement clinker raw meals contain silica, alumina and iron components as main constituents. These constituents and the respective reactions with each other at temperatures levels similar to the calcination equilibrium temperature influence the achievable degree of calcination.

Therefore laboratory test have been performed to evaluate the calcination behavior of different raw mixes under variable CO_2 concentrations by ECRA in its project phase II. Starting with 20 vol.% CO_2 in the surrounding atmosphere (conventional operation) the CO_2 concentration was increased up to nearly 100 vol.% (oxyfuel operation). The lab tests (Figure 6-2) proved a temperature shift of up to 80 K. But it was also proven that different raw materials showed different sensitivity to the CO_2 partial pressure based on its geological origin. Marl (as a natural grown structure of Ca and Si components) showed less temperature increase than limestone mixed with sand as more intermediate phase was formed, pushing locally the calcination reaction.

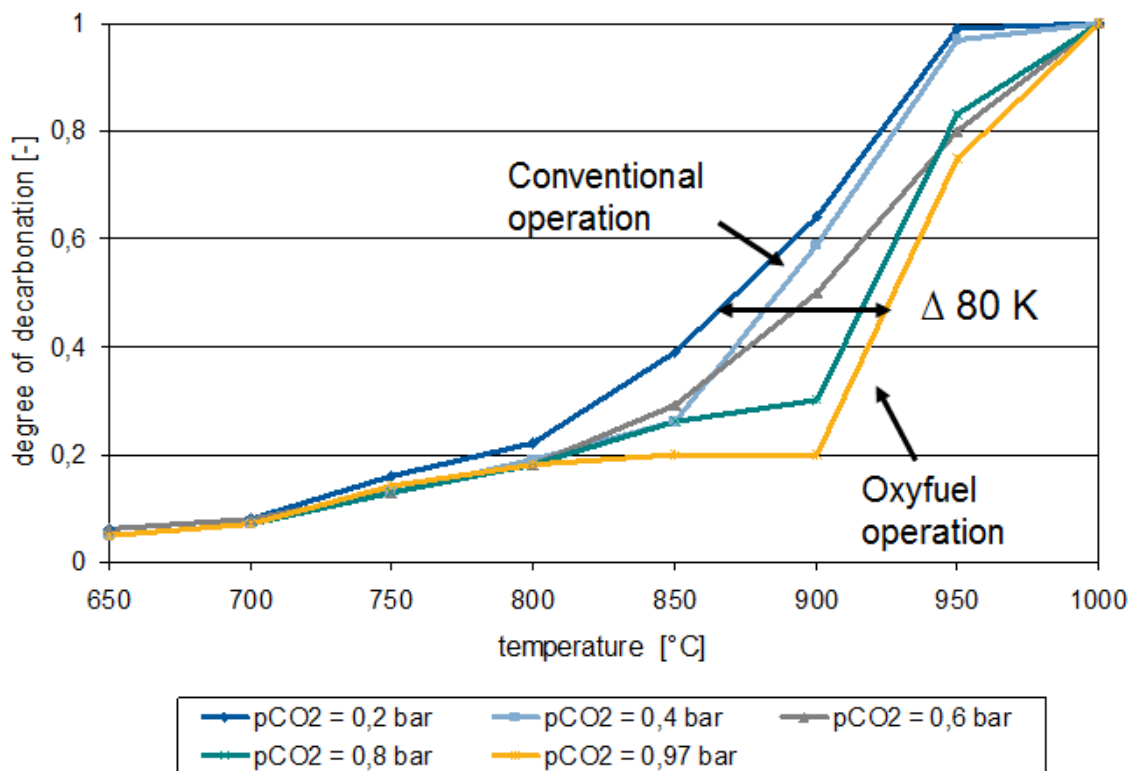


Figure 6-2 Lab tests: Degree of calcination of raw material in dependency of temperature and CO_2 partial pressure [11]

Based on this laboratory investigations, Cinar on behalf of ECRA made CFD simulations to evaluate the impact on the calcination shifting on the operation of the calciner itself [15].

Following observations on the above mentioned influencing factors were made:

- Fuel burnout and volatiles:

In the oxyfuel case the volatiles are quickly consumed and the petcoke is burning much faster (96 % burn out air-fired and 99 % oxyfuel case) due to the pure oxygen, which is directly applied to the stream of the fuel. This results in higher temperatures enhancing the calcination in the lower part of the calciner.

- Heat transfer:

Besides water vapour CO_2 is the major contributor to thermal radiation. For this reason the radiative heat transfer is increased in oxyfuel case.

- Chemical reaction:

In accordance with the lab-tests the chemical reaction is affected by the higher CO_2 concentration (between 76 and 86 vol.%) in the surrounding gas phase.

- Residence time:

Due to the different temperature dependency of the respective gas density the air-fired and oxyfuel case exhibit varying velocities, which results in slightly higher residence times in the oxyfuel case.

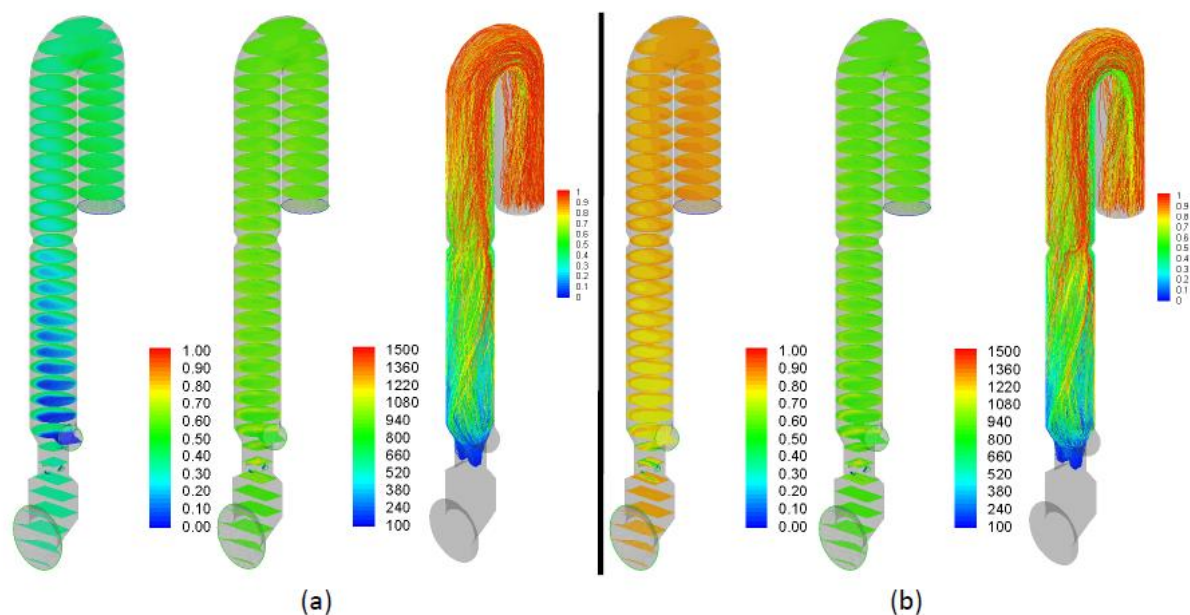


Figure 6-2 Results depicting contours of CO_2 content, temperature and calcination of meal particles for: (a) for air fired case achieving 93% calcination and (b) for oxyfuel case achieving 91% calcination [15]

Balancing those factors the calcination degree drops from 93 to 91 % comparing the air fired to the oxyfuel case (Figure 6-2). The temperature calculated within the CFD model increases from 914°C to 930°C at the calciner outlet. This temperature is above the temperature observed in real plants since clogging tendency would be so high that strong risk of cyclone blockage occurs. Depending on concentration of harmful components, introduced by raw materials and fuels, typical upper limit for temperature at calciner end is $880\text{--}900^\circ\text{C}$. The slight reduction in the

calcination levels could be addressed further, if needed, by optimizing the oxy-fuel burners, since 6-second residence time is sufficient in ensuring both higher petcoke burnout and meal calcination levels.

6.2.3 Pilot tests on partial oxyfuel applications

In 2009 Airliquide, FLSmidth and Lafarge initiated a joint research project on partial oxyfuel technology at cement plants including pilot tests at FLSmidth R&D center in Denmark. The pilot facility comprises the calciner and a four stage preheater with flue gas recirculation (2-3 tph raw meal feed). The rotary kiln was not part of the facility, thus input gas for the calciner was produced by a hot gas generator. The trial results enabled researchers to draw the following conclusions [16]:

- Stable operation of the pilot plant working in oxy-combustion precalciner mode possible.
- Up to 92% degree of calcination was achieved at a temperature increase of 60 to 70°C compared to air firing (which is in line with ECRA's lab tests).
- No hotspots or build-ups were formed when the calciner operated with appropriate burning configuration.
- Due to false air ingress 66 to 78 vol.% CO₂ concentration (wet basis) could be achieved at calciner entrance.
- XRD analysis of calcined raw meal shows no difference between air and oxycombustion - no impact on clinkering behaviour. Furthermore, burnability tests indicated no difference in burning behaviour.
- Results show that an existing preheater or calciner cement kiln line can be retrofitted to oxycombustion.

6.3 Conclusion and next steps

There is a common understanding of calcium carbonate dissociation, which indicates higher necessary equilibrium temperatures at higher CO₂ partial pressures. This circumstance could lead to impairment or even limitations of the calciner performance requiring the adaptation of the clinker burning process under oxyfuel conditions.

First conclusions about calcination under oxyfuel conditions in pilot scale could be drawn from Ca-looping tests. Due to the fact of different reactor types, pure CaCO₃ system, different particle sizes and residence times, only the temperature range from 880 to 930°C and the prove of calcination feasibility could be confirmed.

In cement clinker raw meals calcium is a main constituent next to silicium, alumina and iron. Within the temperature range of calcination, already intermediate mineral clinker phases are generated, which benefits the calcination reaction. This effect was shown in lab tests by ECRA,

however showing a temperature increase of up to 80°C depending on the type of raw material. Besides this limiting factor from chemical reaction a CFD modeling showed different influencing factors on the calcination, namely the enhanced fuel burn out and heat transfer and slightly higher residence times. This leads to comparable degree of calcination at tolerable temperature increase. Nevertheless, these theoretical results have to be proven in practice. First evidence could be drawn from the pilot tests of Airliquide/FLSmidth/Lafarge, who did report stable calcination operation at temperatures up to 915°C and degree of calcination of more than 90 %.

Against this background it is expected that the calcination is feasible under oxyfuel conditions to be of interest for industrial applications. In order to optimally design future calciners e.g. in terms of residence time, the CEMCAP prototype tests shall show the direct interference of degree of calcination, temperature and residence time for entrained flow calciners.

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