european cement research academy

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Technical Report

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ECRA CCS Project: Report on Phase IV.A

ECRA CCS Project: Report on Phase IV.A

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1 Introduction

1.1 ECRA's CCS project

Climate protection is regarded as one of the most important items on the worldwide agenda. All states are facing the challenge of reducing their CO_2 emissions significantly, focussing on the major CO_2 sources in order to contribute to the reduction targets being worked out in many countries and on a global level. To reach these targets, a portfolio of different reduction measures is seen to be required, including CCS/CCR, the capture of carbon dioxide and its geological storage, often referred to as "carbon capture and storage" (CCS) or "carbon capture and reuse" (CCR).

Different roadmaps including in particular those on CCS have been published. They include the requirement that the cement industry has to contribute to CO_2 mitigation by a set of measures, among which CCS is seen as indispensable. Consequently, these road maps point out the need for demonstration projects, implying that after 2020, carbon capture technologies are expected to be commercially applicable at least in the OECD countries, and to a significant extent also in emerging countries like China and India.[1]

Against this background ECRA has decided to examine the capture of carbon dioxide as a prerequisite for the safe geological storage of CO_2. ECRA's goal is to examine the technical and economic feasibility of this technology as a potential application in the cement industry. To initiate this research, ECRA established the structure for a long-term research project on CCS, comprising seven phases.

In the first two phases, general studies on CCS technology were conducted and suitable techniques for carbon capture in the cement industry were identified. As the oxyfuel process and post-combustion carbon capture were found to be principally applicable solutions, these two concepts were studied in detail during phase II [2]. In the third phase, the technical details of both principle capture technologies were studied in depth. The main overall conclusion of phase III[3] was that oxyfuel technology might have a higher potential than previously expected to be applied to existing kilns. Even if many questions needed to be examined in more detail, it became clear that under these circumstances the focus of phase IV should clearly be on oxyfuel.

On this basis the objectives of phase IV.A are the further optimisation of this technology and investigations on remaining challenges. The second major topic is the concept study of an oxyfuel industrial kiln.

1.2 Current state of the oxyfuel project

Oxyfuel technology involves the combustion of fuels with pure oxygen in combination with the recycling of flue gas to moderate the temperature profile. As a consequence, both the material conversion in the kiln system and the operational specifications of the overall process are different from those in conventional kiln operation.

In phase II a process design with regard to the application of full oxyfuel technology was developed which included issues like waste heat recovery, the oxygen supply and flue gas

[1] IEA: Cement Technology Roadmap 2009 – Carbon emissions reductions up to 2050.
[2] ECRA CCS Project - Report about Phase II (TR-ECRA-106/2009), http://www.ecra-online.org
[3] ECRA CCS Project – Report about Phase III (TR-ECRA-117/2011); http://www.ecra-online.org

conditioning. The principal configuration of an oxyfuel cement plant remains almost unchanged compared to conventional technology and is illustrated in **Figure 1-1**. In terms of retrofitting an existing kiln plant, the initial geometry of the rotary kiln, preheater tower and calciner is kept unaltered. For the reason of gas tightness during oxyfuel operation the sealings have to be optimised compared to during conventional operation.

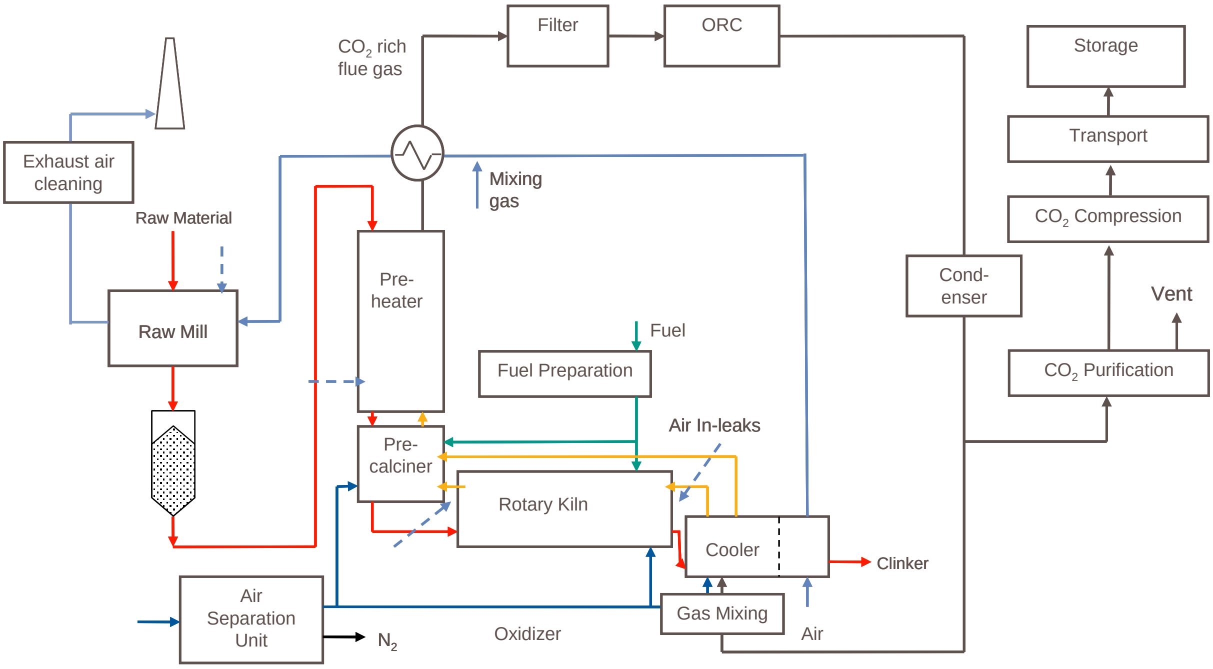


Figure 1-1 Configuration of a full oxyfuel cement plant

The grate cooler is fully implemented into the gas flow in order to avoid a separate preheating of the recycled flue gas and to ensure a high degree of heat recuperation. To prevent CO_2 emissions with the cooler waste air/gas the cooler has to be split into two stages. The first cooler stage is operated with recycled flue gas, which is needed in the burning process. As this would result in clinker temperatures which are still too high, a second cooler stage which is operated with ambient air was considered. This air leaves the cooler as exhaust air and can be used for raw material drying or fuel preparation. No CO_2 from stage 1 should leak into stage 2. A major advantage of using cooler exhaust air for drying purposes is certainly the fact that the raw mill does not have to be operated under airtight conditions. In order to increase the drying potential of the cooler exhaust air, additional heat can be recuperated from the kiln exhaust gas by means of a heat displacement system.

Still containing significant energy, the flue gas - after being deducted - is supplied to a waste heat recovery unit. The efficiency of waste heat recovery units is limited by the flue gas temperature. The Organic Ranking Cycle (ORC) and KALINA technologies use organic substances or NH_3 as cycling media which evaporate at lower temperatures. Compared to steam turbines, these systems can work more efficiently. By using these systems in conventional operation up to 12 $kWh/t_{clinker}$ electrical power can be produced on average. Preferably, the remaining flue gas enthalpy should be available for the raw material drying. Therefore, the heat recovery operates more efficiently for lower raw material moisture, meaning higher remaining flue gas energy.

An oxygen-enriched flow - the so-called oxidizer - is provided for the precalciner and the kiln firing process (as primary “air”) as well as for the premixing of cooling gas by a common air separation unit. For a medium-size cement plant with a kiln capacity of 3,000 tpd the oxygen demand is estimated to be around 25 to 30 tph. For logistical reasons, such amounts of oxygen can only be provided by an on-site oxygen supply system. Depending on the production capacity, oxidizer purity, separation technology and oxygen pressure, the commercially used air separation processes require between 0.25 and 1 MWh/t O_2. Cryogenic oxygen production is currently the most power-efficient state-of-the-art technology for the production of large quantities of oxygen. The economical optimum O_2 purity is between 95 and 97.5 vol %. Therefore, the oxidizer purity was fixed at 95 vol % O_2, 3.5 vol % Ar and 1.5 vol % N_2 as a basis for the process modelling.

In phase III this research was deepened through different work packages on oxyfuel technology with the aim of achieving a better understanding of it. The focus was placed on optimised sealings and refractories, burner- and cooler design, and the CO_2-purification-unit (CPU) design. Finally, the impact of oxyfuel operation on the clinker and cement quality was investigated.

A main result was a reliable and sophisticated concept for an oxyfuel cement plant, which could also be adopted for the retrofitting of an existing plant. It emerged that most operational problems could be handled and that negative impacts on the clinker produced and the refractory lining seemed to be negligible.

1.3 Scope of the study

If at any time a carbon capture plant is to be built in the cement industry, commercialisation can only occur based on sufficient experience gained from industrial testing. How such a test kiln must be dimensioned and where and how it could best be built is examined in this study.

The whole work package is subdivided into six smaller packages covering specific tasks. Separate reports are available for each task and are summarised in this final report. The following topics were planned:

Table 1-1 List of work packages in phase IV.A

No.	Task	Executed by
B1	Identification of a suitable plant capacity	Aixergee
B2	Design principle	Aixergee
B3	Plant dimensioning	Cinar / Fives FCB / Research Institute
B4	Concept for control and safety devices	postponed
B5	Cost estimation and time schedule	ECRA expert group
B6	Concept for the reuse of the plant	ECRA expert group

2 Identification of a suitable plant capacity

2.1 Plant capacity limitations and scaling aspects

The range of production capacity for the pilot plant has been discussed within the limits of 10 t/d and 1000 t/d as this seems to be the appropriate size range to allow both the proof of concept for the oxyfuel clinker production process and a sufficient size similarity to industrial production lines. Within this range, scaling aspects are taken into account which can have an impact on the physical similarity of the processes and the mechanical design of the equipment used in a future industrial plant.

Basically, the minimum capacity is limited by the following scaling aspects:

- Influence of false air more significant in smaller-sized plants than in large.scale plants due to due to similar sizes of gaps at the sealing locations
- Impact of radiation on sintering and calcination too small as the surfaces involved with the heat transfer are smaller with respect to the production throughput
- Flame size and shape in the kiln may not be geometrically similar due to mechanical design reasons
- Retention time in the kiln and calciner too short as the original particle sizes remain unchanged, which generates a different flow pattern and heat transfer
- Heat transfer in the clinker cooler affected as the size of the clinker nodules is comparable to those in bigger plants.

The upper capacity limit is mainly set by the cost for equipment and operation and the complexity of materials handling.

2.1.1 Scaling effects on transfer processes

The miniaturisation of process equipment through scale-down or its magnification through scale-up generates some similarity dilemma, especially concerning the size of the meal, clinker and fuel-particles and the related heat transfer. The material size cannot linearly follow the geometrical scaling of the process equipment influencing all transport phenomena:

<u>Momentum transfer:</u>
The size ratio of meal particles to equipment size differs considerably if a lab-scale pilot plant or an industrial one is chosen. This influences the momentum transfer, especially in pneumatic conveying processes in the preheater and the calciner. With a given meal particle size and similar equipment geometry the retention time in the calciner can only remain unchanged by adapting the viscosity of the gas. As this is not a variable, inconsistencies in small-size equipment occur affecting the retention time, heat transfer and chemical reactions such as the decarbonisation.

<u>Heat transfer:</u>
Temperature levels depend on the equipment size for various reasons. Due to the higher relative surface of smaller plants, their heat loss to the environment is higher. A higher combustion rate has to compensate this heat loss, which again increases the specific gas flow through the plant and all impulse-related phenomena. Furthermore, the ratio between the

convective heat transfer at the outside of a flowing particle and its conductive heat transfer inside will change with the scale-up factor for the same size particle. In the cooler the heat transfer in the packed clinker bed changes with the equipment size for identical clinker particles sizes and specific cooler geometries. Geometrical similarity (length to width ratio) leads to a lower number of granule layers if the size of the granules is identical to conventional ones. Consequently, the contact surface and time for the heat transfer is reduced, for which reason the temperature of the recuperated air is decreased in a small-sized plant compared to in a large-sized plant.

<u>Combustion:</u>
The combustion (ignition and burn-out) of coal and fuel particles strongly depends on their size, which causes scaling effects at a constant fuel particle size similar to those described above. Assuming smaller equipment sizes, this could lead to flames which are longer with respect to the kiln length.

2.1.2 Scaling aspects of individual equipment parts

Different aspects concerning the flow pattern but also the mechanical feasibility limit the scaling of individual equipment parts:

<u>Kiln burner:</u>
Assuming a typical coal (fuel)-particle size, the particles are still more than 2 magnitudes smaller than the reference kiln diameter in a smaller kiln. No scaling issues are expected with respect to the flame placement inside the kiln if the following burner parameters are kept within the reasonable ranges:

- Principle channel burner type with comparable primary air rate (12%)
- Assumed burner area for primary air flow: 40%
- Assumed maximum primary air velocity: 40 m/s

The minimum requirement on the burner diameter is defined by the primary air velocity and a mechanical limit of a standardised DN 200 tube size. The upper level is determined by the maximum permissible gas velocity of the secondary air at the burner tip of 16 m/s.

<u>Rotary kiln:</u>
The dimensioning of the rotary kiln is usually defined by the fulfilment of the following limiting conditions:

- Material load per volume of 5 t/(m^3*d),
- Gas velocity in the kiln inlet of 8 m/s
- Heat stress in the sintering zone.
- Assumed kiln L/D-ratio of 15

Based on these limitations, **Figure 2-1** and **Figure 2-2** show the sizing option for the rotary kiln and the kiln burner as a function of the production capacity which allows equipment layouts without violating the principle rules of similarity. One sizing parameter for the kiln diameter is the maximum permissible gas velocity in the inlet chamber, which is different for plants with a tertiary air duct (TAD) and for plants without. The minimum required kiln diameter for plants without a tertiary air duct is shown in **Figure 2-1** by the blue and red lines. The requirement for plants with a tertiary air duct is shown by the blue line in **Figure 2-2**. In the case of no TAD the criteria is controlled by volumetric loading of the kiln at a low production rate and changes to maximum permissible gas velocity in the kiln inlet chamber above 500 t/d. The permissible burner diameters do not state a limitation in any case.

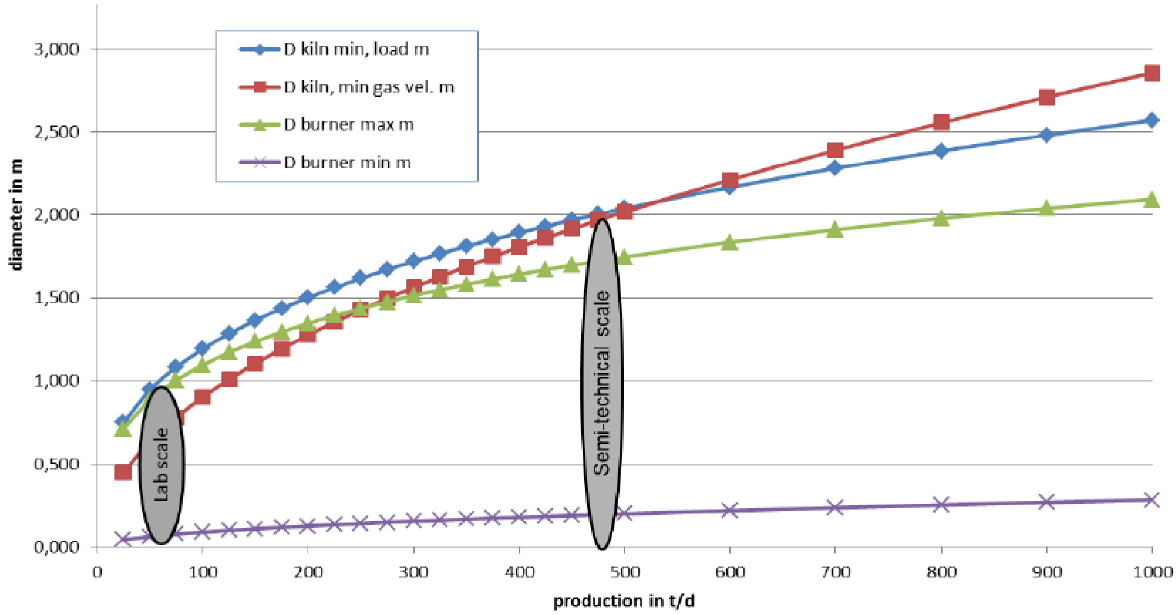


Figure 2-1 Sizing options for a rotary kiln and burner operated without tertiary air duct (TAD)

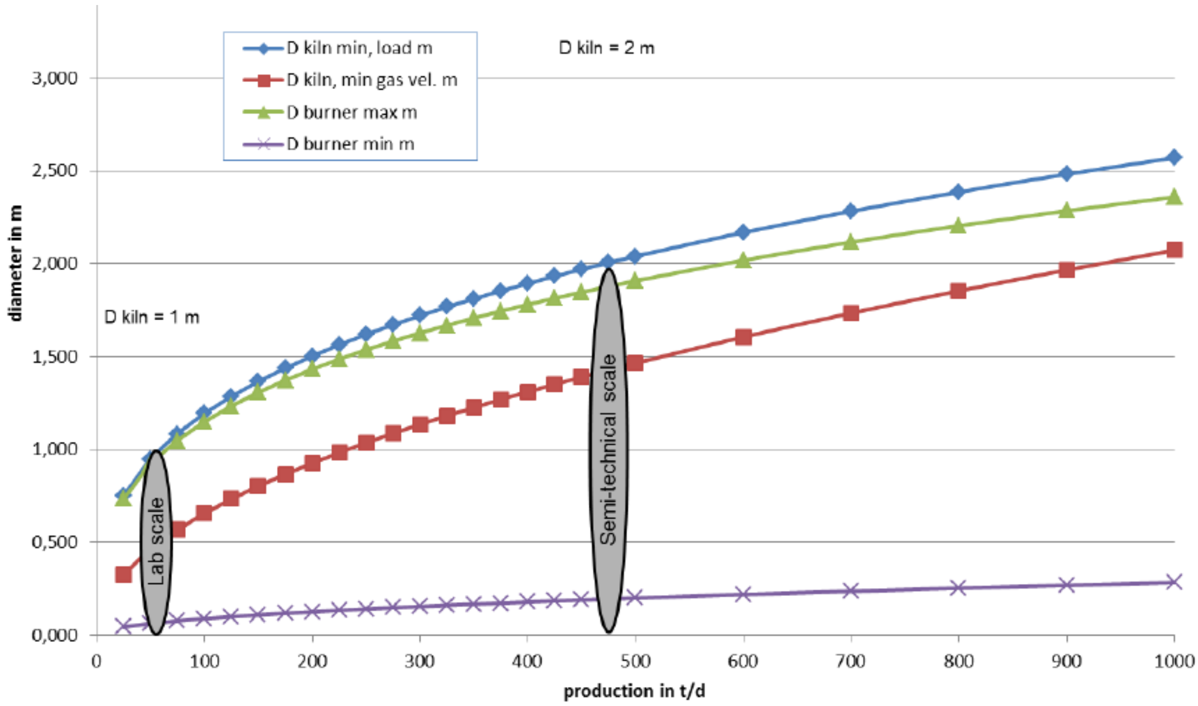


Figure 2-2 Sizing options for a rotary kiln and burner operated with tertiary air duct (TAD)

In summary, the sizing of the rotary kiln is not ruled by constraints from similarity and can be decided upon according to other factors, like for instance the availability of existing equipment.

Clinker cooler:

The nodularisation of the clinker granules is mainly dependent on the viscosity of the sintering material and temperature properties in the kiln. Hence the clinker nodules are likely to form typical particles of usual sizes irrespective of the size of the plant equipment. On the grate cooler these clinker particles form the bed as an assembly of a certain number of layers, finally setting the bed height. But total geometrical similarity would require smaller size clinker nodules, which is not feasible for the above mentioned reasons. Therefore, the clinker particles' active surface for the heat transfer is reduced in smaller coolers. In order to keep the heat transfer similar, the clinker bed height has to be increased by narrowing the cooler.

Preheater:

Besides heat and momentum transfer the scaling effect on the preheater is mainly driven by false air intrusion. The false air flows into the process through inevitable gaps in the sealing locations, which are not scaled up linearly to plant size. A reduction of the gap size is technically difficult below a certain level, which makes their relative contribution to the total plant's surface bigger in smaller plants. Extensive process parameters, especially pressure differentials at sealing locations are independent of plant sizes, so that the influence of false air gains in importance for smaller plants.

2.2 Conclusion

The core equipment of the oxyfuel plant, the rotary kiln, can be sized for any capacity within the considered range between 10 t/d and 500 t/d. The diameter of the kiln should accommodate the requirements of the maximum permissible kiln loading. Thus any capacity bigger than 500 t/d is possible without jeopardizing scale-up. However, if not equipped with a tertiary air duct, the kiln has to be bigger in diameter in order to keep the gas velocity at the kiln inlet below the permissible threshold value.

All other parts of the equipment have to be adapted to the chosen target capacity anyway in order to maintain the necessary gas velocities and heat transfer rates. Accordingly, the sizing of the testing facility can be chosen to satisfy other important parameters, like for instance economic aspects, without jeopardizing the technical validity of the oxyfuel testing. The validity of the industrial testing as well as the scaling ability favour a pilot plant size of 500 t/d or above.

3 Plant design

Designing an industrial oxyfuel testing facility includes the evaluation of several aspects in order to determine an optimal design principle, such as:

- Configuration including all relevant plant components and auxiliary devices
- Concept for the supply of oxygen (tank system or on-site generation) depending on the plant capacity
- Location: Synergies with existing infrastructure, availability of staff and process materials (raw material, fuel etc.), further use of the plant, accessibility

3.1 Influencing aspects on the design principle

As a minimum requirement for the design concept, the applied process has to fulfil the requirements for the proof of concept of oxyfuel-operation:

Plant size
As shown above, a plant capacity between 10 t/h and 1000 t/h can fulfil the task without generating any scaling problems. With respect to increasing operational test costs (e.g. consumed oxygen) and costs for the equipment to be installed with increasing plant capacity it is obvious that the size of the plant plays an important role in the overall considerations. Equipment costs can be reduced by using existing plant equipment, which would then define the size of the plant.

Availability of plant sites
Theoretically, the testing facility could be built to any size of choice, but practically would however have to be adapted according to the individual case. Different concepts were developed to form the basis for a testing facility.

- Greenfield-concept

  In this study a greenfield concept is understood as consisting of a new production line to be installed at a completely new site. There will be no infrastructure or environment to be reutilized for the operation of the production line. Apart from the input materials like limestone and fuels, the missing staff to operate such a facility poses a problem. Another greenfield scenario could be, for instance, a newly installed technical-scale clinker production line (~ 10 t/d) at a cement producer's research centre. Existing lab facilities could potentially be used, but at least auxiliary devices for oxyfuel operation such as recirculation, oxygen supply etc. would have to be added. Due to the small size of a lab kiln the result transferability to real sized kilns is limited.

- Brownfield-concept

  A brownfield concept is defined as a new oxyfuel clinker production line in an existing site environment, probably reutilizing sections of the existing site infrastructure, such as raw mill capacities, clinker handling capacities, fuel supply, power supply, auxiliaries supply and disposal facilities. Therefore, the existing cement production site has to provide sufficient available space as well as surplus raw meal and a clinker handling facility.

Compared to the greenfield-concept probably less new equipment due to the existing infrastructure would have to be installed, but the kiln plant equipment itself would be bigger than laboratory-scale. **Figure 3-1** shows the general scheme for a brownfield concept.

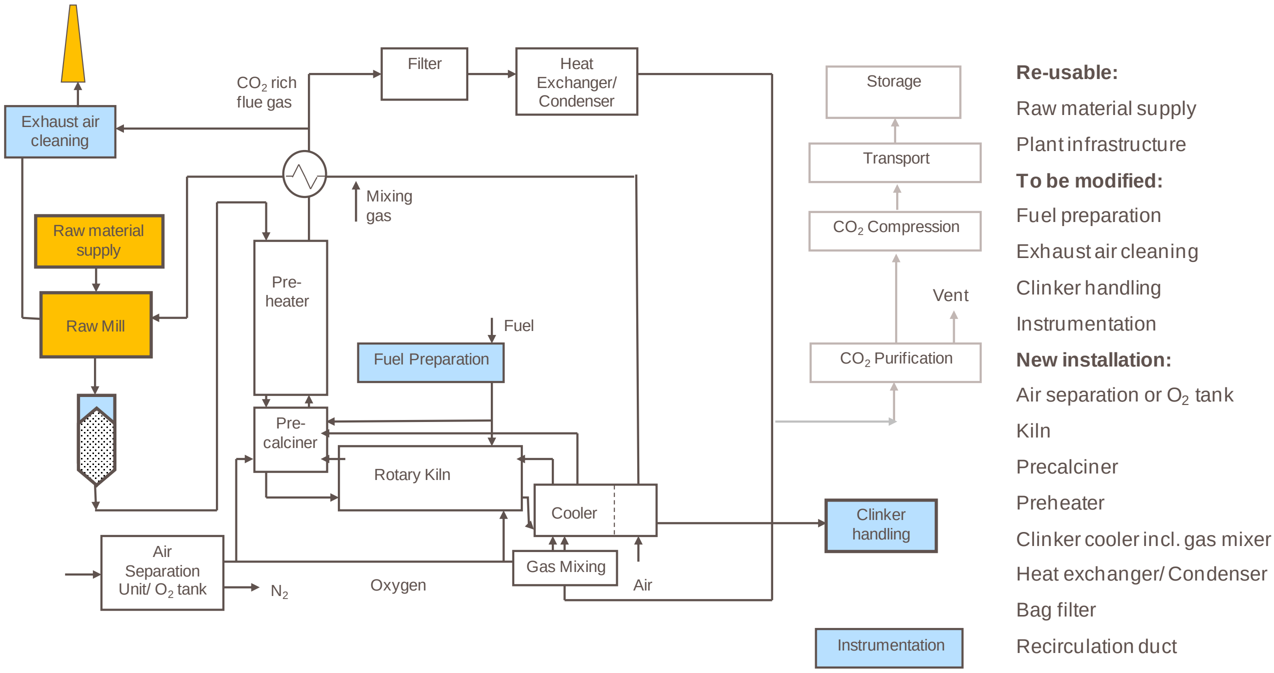


Figure 3-1 General scheme for a brownfield concept: black: to be installed new, blue: to be utilized from existing plant, yellow: to be modified, grey: not needed for proof of concept

- Blackfield-concept

The blackfield concept is defined as the modification of an existing clinker production line to suit the requirements of the oxyfuel process. Contrary to the brownfield concept, the new pilot plant does not run in addition to the existing kilns, but substitutes the capacity of the kiln which it is based on. A typical blackfield scenario could be, for instance, that a kiln plant out of operation could serve as a basis for the new oxyfuel kiln. The kiln line would be retrofitted with whatever is required to allow the operation of the oxyfuel process. Typically, the total project costs would be further decreased in comparison to the two other concepts by the reutilization of the rotary kiln and the preheater, but design compromises would have to be accepted. **Figure 3-2** shows the general scheme for a brownfield concept.

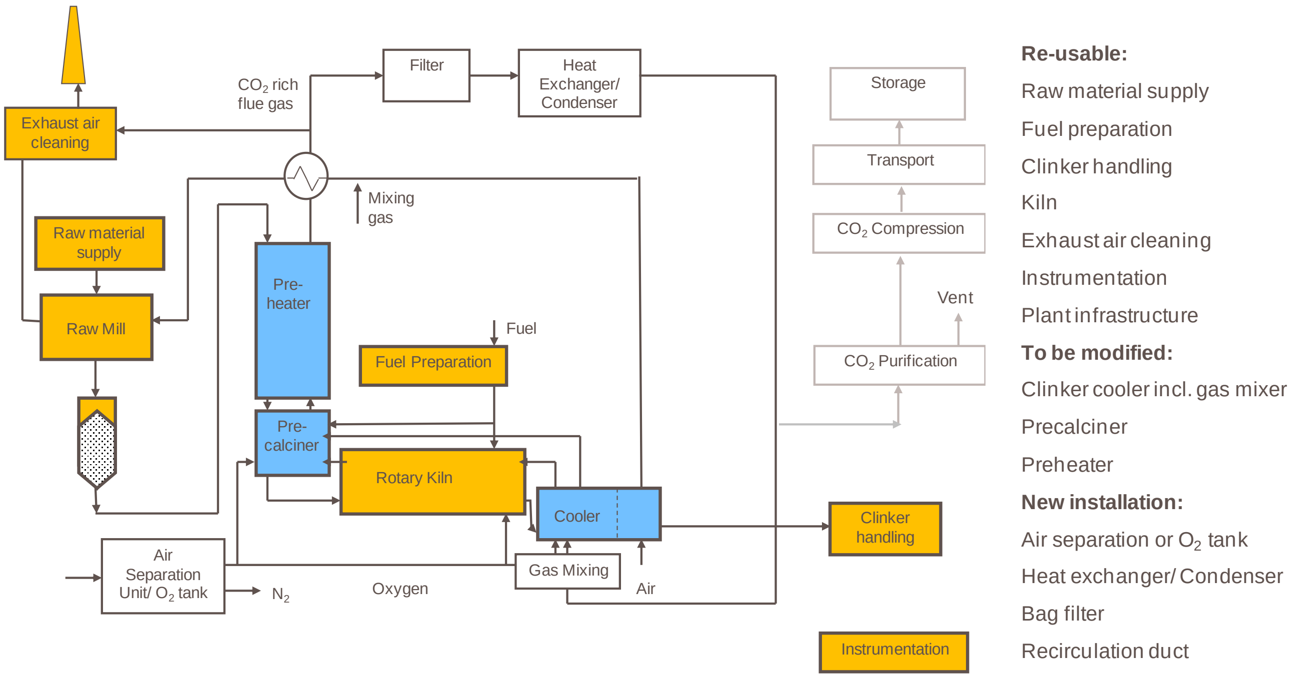


Figure 3-2 General scheme for a blackfield concept: black: to be installed new, blue: to be utilised from existing plant, yellow: to be modified, grey: not needed for proof of concept

In each concept different aspects have to be considered as illustrated in **Figure 3-3**, which shows the disadvantages and advantages of the respective concepts (**Table 3-1**).

Table 3-1 Comparison of the concepts

Issue	Greenfield	Brownfield	Blackfield
Design limitations	Free design	Free design within the limits of the existing plant's arrangement	Design compromises
Size	Lab scale ~ 10 tpd up to industrial scale < 500 tpd	Lab scale ~ 10 tpd, more likely industrial scale < 500 tpd	Industrial scale, likely > 500 tpd – 1000 tpd
Transferability of results	Limited due to small size	Scientific rigor given	Scientific rigor limited
Investment costs	Very high (due to missing infrastructure)	High	Maybe lower depending on plant condition
Operational costs	High (due to infrastructure operation)	Maybe lower	Very high
Public access	Free	Limited due to the loca-tion at a competitor's plant site	Limited due to the loca-tion at a competitor's plant site
Installation	All equipment necessary; Infrastructure necessary	All equipment necessary; Infrastructure available	Parts of equipment nec-essary; Infrastructure available
Staff	not available	available	depending on plant site, but most likely not availa-ble
Materials	not available	available	available

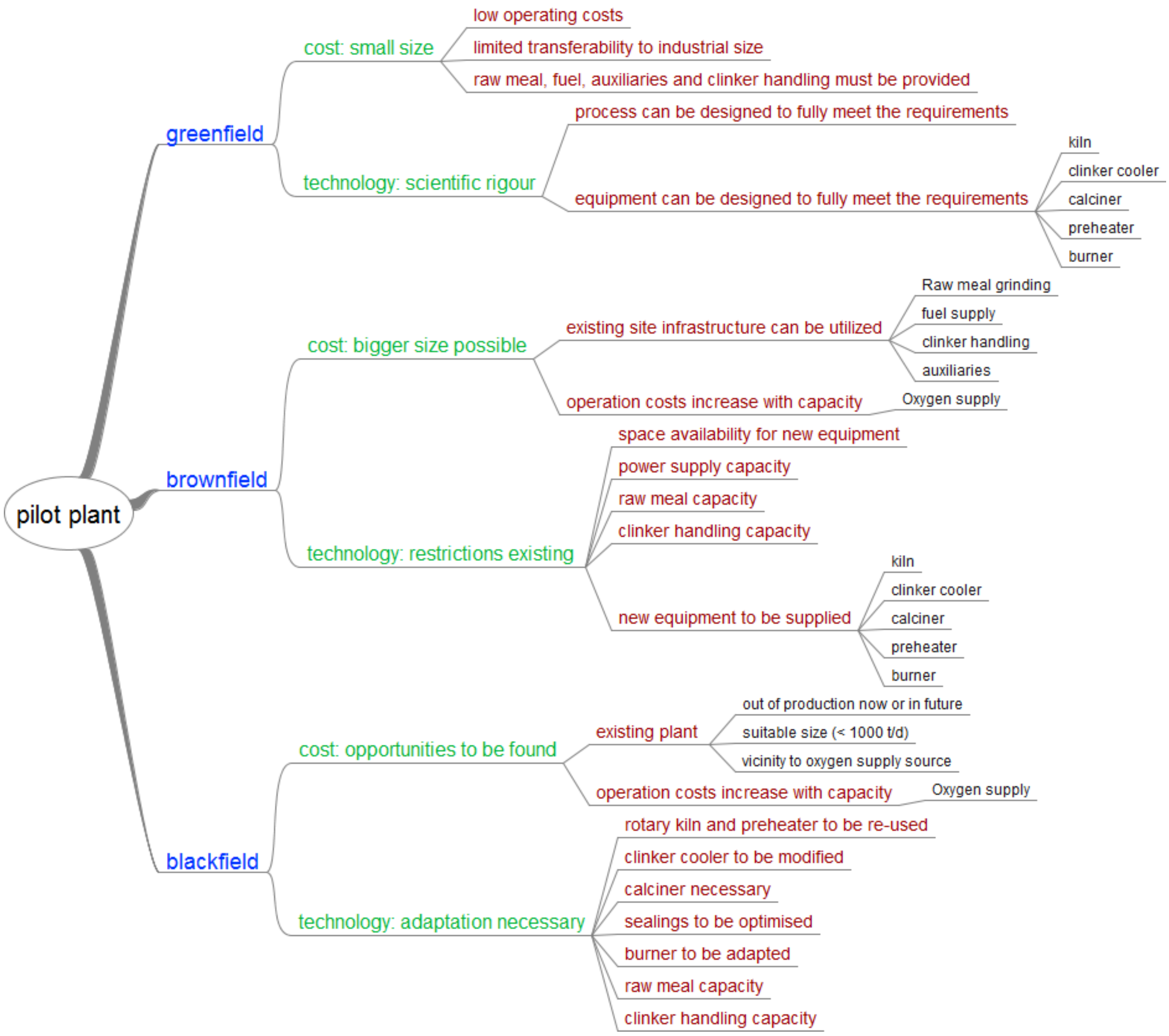


Figure 3-3 Mind map comparing the concepts

Based on these considerations the ECRA Technical Advisory Board decided to focus on brown- and blackfield concepts. The greenfield option, even as a lab-scale plant, was considered too costly to prove only the concept of oxyfuel operation. It would show if clinker is produced, without the ability to transfer operational findings to industrial size. Brownfield and blackfield options were therefore both to be rated with respect to site-specific criteria:

- Availability of kiln equipment (kiln, calciner, preheater, (cooler), mills, filter etc.)
- Availability of space for retrofit equipment
- Availability of materials and power infrastructure and capacities
- Proximity to oxygen supply (nearby OGP with available capacity)
- Accessibility of regional funding
- Favorability of environmental, local authority and neighborhood situation

3.2 Process options

3.2.1 Scope of the testing facility

To prove the concept of oxyfuel operation the kiln line should be designed as shown in **Figure 1-1** but does not require a CO_2 compression unit as these are already commercially available and the interface connection to the clinker production line is regarded to be trouble- free and controllable. Similar statements as for the CPU are also valid for the air separation unit. In both of the illustrated options for flue gas routing, materials like raw meal and fuels are available, thus no flue gas is needed to dry these materials (**Figure 3-4**). The extracted part has to be treated together with the exhaust gas from the existing line (option 1) or separately (option 2) if no spare capacity exists. The latter option envisages the dedusting and cooling within the recirculation path as the exhaust gas to the stack is extracted behind the filter. In both options the condenser can therefore be dimensioned smaller.

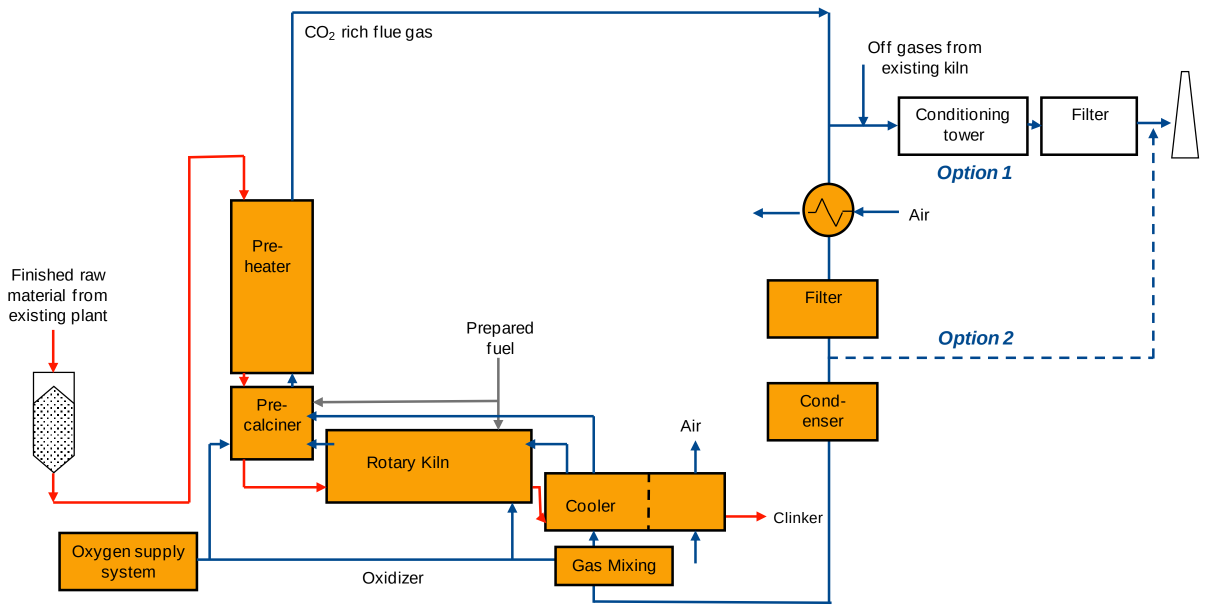


Figure 3-4 Layout of the integration of a brownfield plant (yellow: new installation)

The blackfield option might require the integration of the raw material production, if no running line exists next to the testing facility (**Figure 3-5**). For this purpose the flue gas can partly (< 50 %) be led through the raw mill before entering the stack.

In any case flue gas will be released via the stack. For future research a switch to a potential CO_2 reuse facility could be included to use some of the flue gas. The process scope is therefore limited to the core kiln equipment, which has to be newly installed or adapted to oxyfuel operation. Moreover, additional units of the recirculation path (heat exchanger, filter, condenser, gas mixing, piping) have to be installed.

Furthermore, the cooler exhaust air from a two-stage cooler is not necessarily used, as the brownfield design does not include a raw mill which has to be supplied with hot gas. For this reason the choice of a future design of the pilot cooler is free, for example, to include a gas recirculating one-chamber oxyfuel cooler with no exhaust gas.

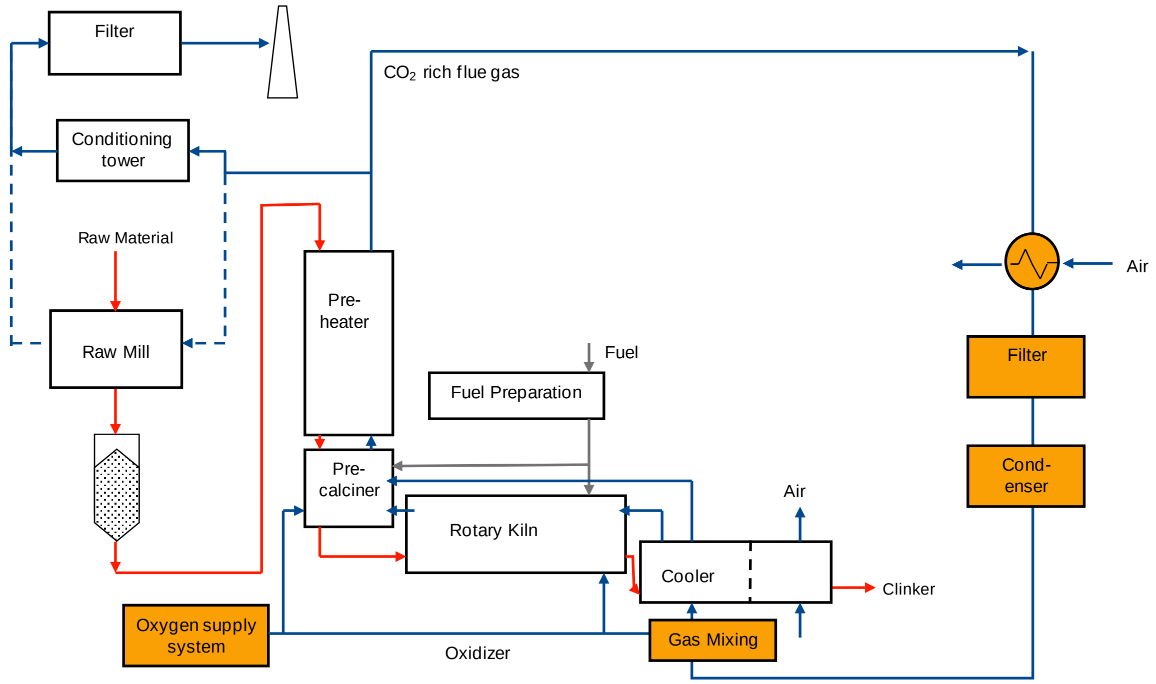


Figure 3-5 Layout of the integration of a blackfield plant (yellow: new installation)

In order to prove the concept some critical parts have been identified, such as:

- Sealings: Previous theoretical and practical studies showed that the performance of improved maintenace on a common sealing location is sufficient to run an oxyfuel kiln. However, lower levels of false air < 8 % are disirable to operate the CPU efficiently and to reduce the power demand. Therefore, available sealings, which are also used in other industries, shall be evaluated.
- Condenser: Operating a condenser is a completely new topic for the cement industry. Large-scale condensers are already used in other industries like the chemical industry, but the conditions and requirements of cement flue gas are different. The suitability of available systems needs to be investigated.
- Heat exchanger: The dust load in particular will be a critical issue for the operation of the gas-to-gas heat exchanger. Therefore, experience from existing systems will be summarised. A general layout and a rough sizing is presented in **Figure 3-6**, which already shows the huge dimensions of such a system.
- ID fan: The performance of a gas-tight fan handling gases of higher density compared to the conventional operation is a further topic to be investigated.

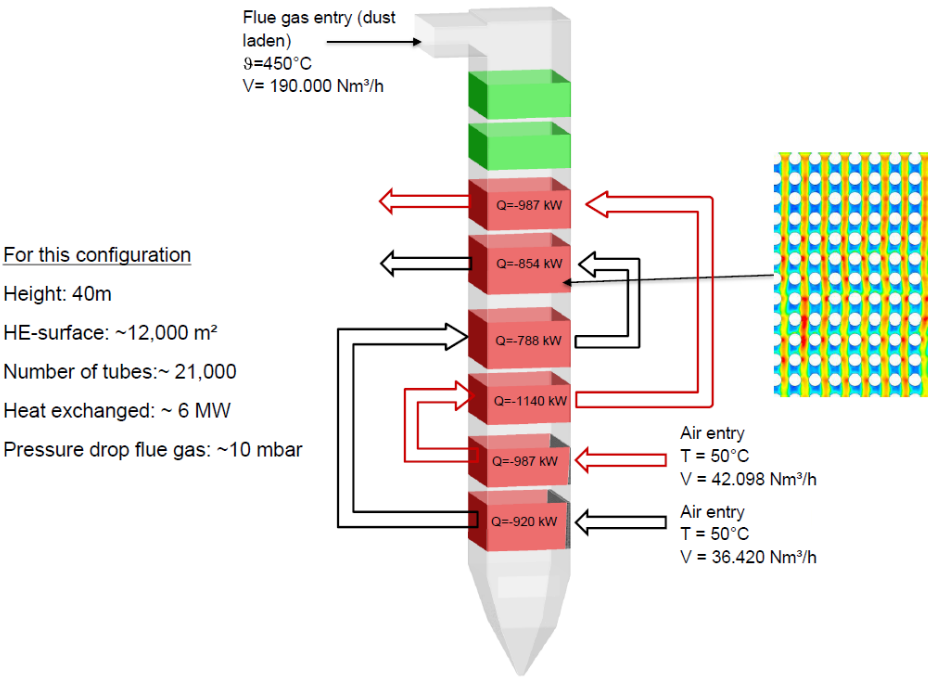


Figure 3-6 Concept for a gas-to-gas heat exchanger

The further investigation of the critical parts of the process will be addressed in a study which will be carried out in 2015. The evaluation of different sealing methods (covers etc.) and the development of procedures to reduce false air ingress during cleaning intervals are objectives of another work package in phase IV.B.

The adaptation of essential plant units has already been investigated in theory during the last research phases, for example:

- Two-stage clinker cooler: Suitable layouts have been developed by IKN using dynamic and static parts for the gas separation.
- Burner: Due to the different gas properties the burner has to be adapted in order to generate a similar flame characteristic compared to a conventional air flame. Aixergee presented a generic burner design in phase III.
- Refractories: The suitability of conventional refractory brickwork was proven in lab-scale.
- Precalciner: Laboratory tests showed a temperature shift of the calcination equilibrium temperature of up to 80 °C due to the high CO_2 concentration. Based on this and the still unknown reaction behaviour in practice, the calciner shall be dimensioned to include a certain offset to increase at least retention time.

The cooler, burner and precalciner design will be investigated as separate prototypes within the framework of the EU funded CEMCAP project in 2015-2017.

3.2.2 Retrofitting an existing plant at reduced capacity

Oxygen costs will probably be the biggest cost contributor to the operating cost. Therefore, an oxyfuel kin should be large enough in size to allow a proper proof of concept on an industrial scale. At the same time the kiln should be operated at the least possible capacity to keep the oxygen cost low. The final dimension depends on the existing plant`s situation, but capacities above 1000 t/d seem to be too high with respect to oxygen costs.

For a kiln with capacities exceeding 1000 t/d an adaption of the capacity might be needed. In particular, those pieces of equipment whose function is dominated by pneumatic transports and critical retention times, e.g. the cyclone preheater and the calciner need to be adapted for the lower gas flow. To meet this purpose, two possibilities are sketched in the following:

- Adaptation of the preheater

  In the case of reduced production, the exhaust gas volume flow can drop below a critical level of required gas velocities in the cyclone preheater. To ensure proper suspension at the meal boxes and to prevent meal from dropping out of the suspension flow, the gas ducts, meal boxes and cyclone inlets of the plant might have to be modified.

- Increase of preheater gas flow by cooler vent gases

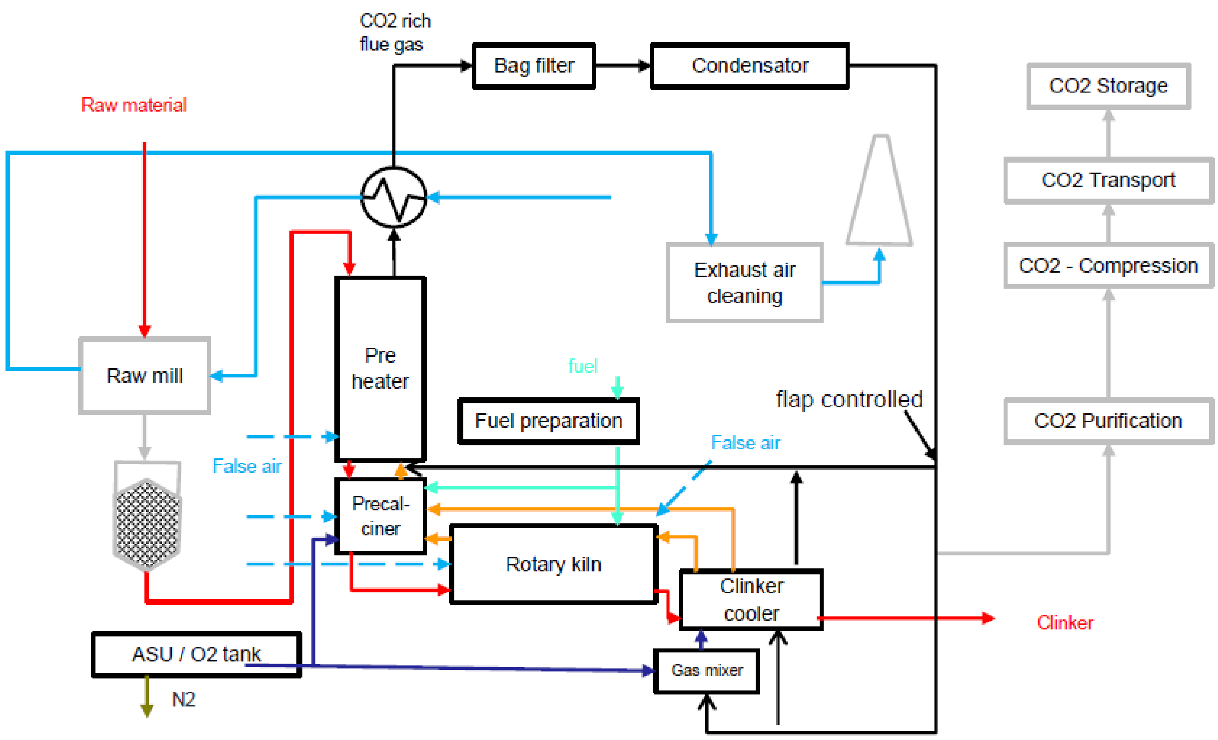


Figure 3-7 Adapted process scheme to significantly reduce the throughput of a given plant geometry

Figure 3-7 shows a variation of the general process scheme to increase the volume flow in the preheater by routing the cooler exhaust gases into the preheater in between the calciner and the cyclone preheater. This measure allows an operation with lower throughput than the preheater tower's pneumatic transport was designed for. The calciner and cooler have to be replaced. The rotary kiln can be run with reduced throughput without any problem as the process stability does not depend on pneumatic transport processes. By suitably designing the kiln burner, the heat transfer in the sintering zone of the kiln can be adapted to the needs of this reduced capacity operation.

Table 3-2 Comparison of both options

	PH modification	Increase of PH gas flow
Investment costs	Cost for replacement of cyclone parts	Additional duct from cooler vent to calciner incl. control device Bigger heat exchanger behind pre-heater
Operation costs	Fuel consumption expected to be lower due to better meal/gas-ratio at preheater	Higher fuel consumption

3.2.3 Conclusion

Within this section the different design options have been discussed in order to establish a basis for the evaluation of potential plant locations. It was concluded that a greenfield plant is not a reasonable option due to the missing infrastructure, e.g. the raw material supply. The brownfield option design allows a high degree of freedom with regard to design aspects and plant size at higher investment costs. Whilst the blackfield option includes less necessary investment, design compromises have to be accepted. For this reason the scientific evaluation might be limited as process optimisations are restricted by the existing equipment. The production capacity in particular is likely to be higher than in the case of a brownfield plant, which strongly influences the operational costs. A concept to significantly reduce the capacity in the existing kiln equipment was presented, which would again modify the process and therefore jeopardize the proof of concept.

Nevertheless, in both favoured options, brownfield and blackfield, the scope of the testing is restricted to the core kiln plant. The real capture of CO_2 is excluded from the proof of concept, which implies that no CO_2 conditioning unit is installed and flue gas is released via the stack to the environment. Depending on the specific plant conditions, different layout options e.g. including or excluding the raw mill could be chosen. Moreover, critical plant units were identified, which will be the subject of studies within phase IV.B.

4 Plant dimensioning

Based on the identification of a suitable production capacity for the pilot kiln and in interrelation to the development of a design principle, the process parameters and plant components were dimensioned. A brownfield plant with 500 t/d and a potential 1150 t/d (at reduced capacity to 1000 t/d) blackfield plant using coal as regular fuel served as the basis for the analysis. This evaluation provides initial conclusions about intricate operational issues, but has to be performed for each specific plant separately.

4.1 Methods

Different methods were applied to dimension the oxyfuel testing kiln. Detailed descriptions are depicted in separate reports.

4.1.1 Models

- Process modelling:

  A process engineering model, which was developed by the Research Institute of the Cement Industry, was used as a basis. At its core it describes the process from the kiln meal feed to the output of the clinker from the cooler and is made up of individual models for the plant components preheater, calciner, bypass, rotary kiln and grate cooler. All the individual models can be linked mathematically with one another, which makes it possible to determine a steady-state condition for the entire rotary kiln plant. The individual plant sections can be defined geometrically so that different plant sizes can be simulated. The calculations themselves cover the energy and material balances for the flows of fuel, dust and gas. Not only the combustion calculations for the fuels and the heat transfer but also the relevant chemical and mineralogical solid state reactions, the gas phase reactions and the gas-solids reactions are taken into account.

- Sizing model:

  Apart from the process model described above, Fives FCB set up a model for the main purpose of sizing, which is based on validated performance criteria of the kiln equipment, such as cyclones efficiency and heat losses, calcination rates, cooler heat recovery efficiency, kiln heat exchange curves and false air inlets. Compared to the process model, the FIVES FCB model could be used to design cyclones using the stability of meal lifting and transport in the cyclones riser ducts as a criterion.

- CFD (Computal Fluid Dynamics) modelling:

  Via CFD modelling other parameters and assumptions can be determined or proven. Cinar applied its in-house-developed MI-CFD (mineral interactive computational fluid dynamics) to model calciner kilns operating under oxyfuel conditions (higher CO_2 and oxygen enrichment conditions). The MI-CFD model was required to take into account the heat extraction and mass additions into the combustion products during the calcination reactions where endothermic reactions take place. The release of CO_2 from $CaCO_3$ continuously alters the temperature field and gas species concentrations. For the present study, a modified particle reaction model was implemented in the MI-CFD model. The model predicts laboratory experimental data well, as shown in previous research projects. The model has already been optimised and validated against over 100 calciners with different configurations and operating conditions.

4.1.2 Lab-scale testing

A series of small-scale tests were carried out to evaluate the burning kinetics and the heat transfer coefficients in a realistic and continuous burning operation. Both the conventional air-fired and the oxyfuel case modes were tested and compared. With regard to the varying recirculation rate in oxyfuel mode, three different combustion gas compositions were set (**Table 4-1**).

Table 4-1 Test matrix for the combustion gas mixture

Trial	CO_2 content	O_2 content	Mode
Reference	0 vol.%	21 vol.%	Air-fired
1	66 vol.%	34 vol.%	Oxyfuel
2	56 vol.%	44 vol.%	Oxyfuel
3	49 vol.%	51 vol.%	Oxyfuel

The testing facility is mainly composed of a small rotary kiln (Ø 0,6 m x 4 m) equipped with a propane burner, a clinker cooler, meal feeding, gas cleaning and all necessary instruments (**Figure 4-1** and **Figure 4-2**). The inner diameter of the kiln is 200 mm, thus giving a ratio L/D = 20. For the testing 30 kg/h of raw meal provided by the CCB Gaurain plant (ITALCEMENTI group) were processed. The residence time in the kiln is about 30 minutes with a 3.1 rpm rotation speed and 3.9 % slope.

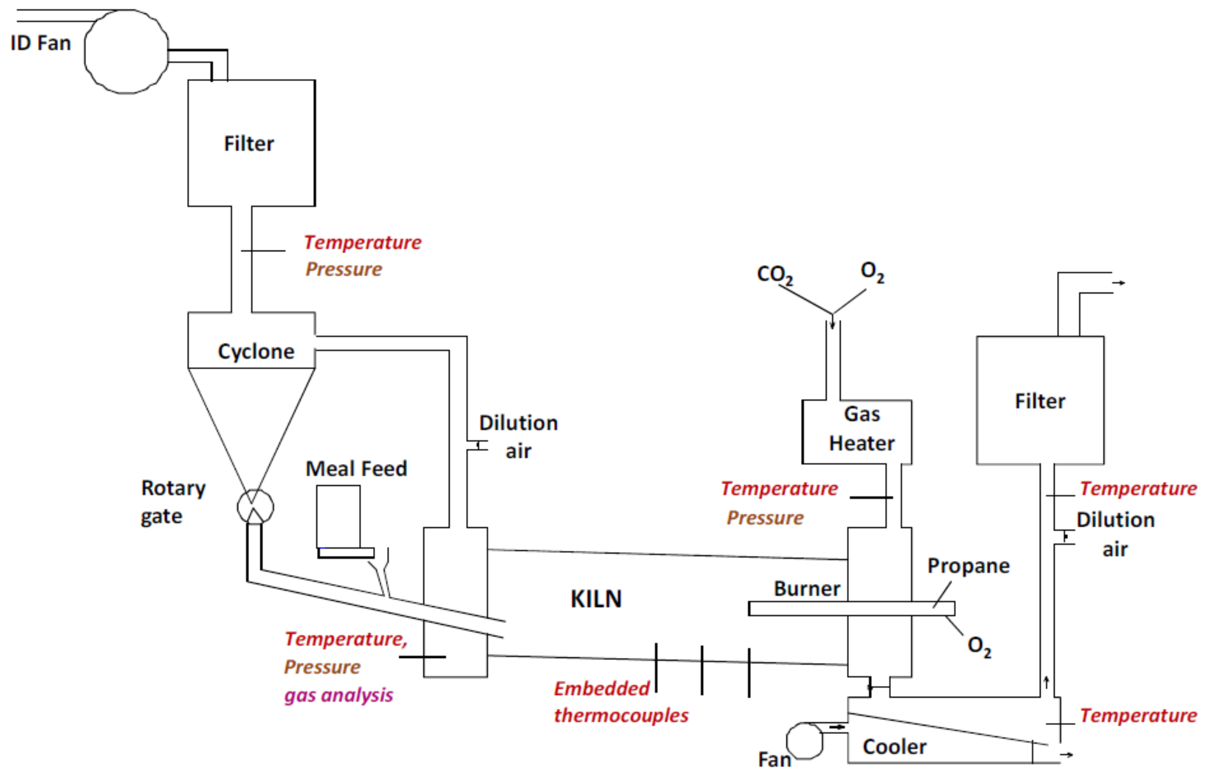


Figure 4-1 Scheme of the lab kiln in FCB's facilities

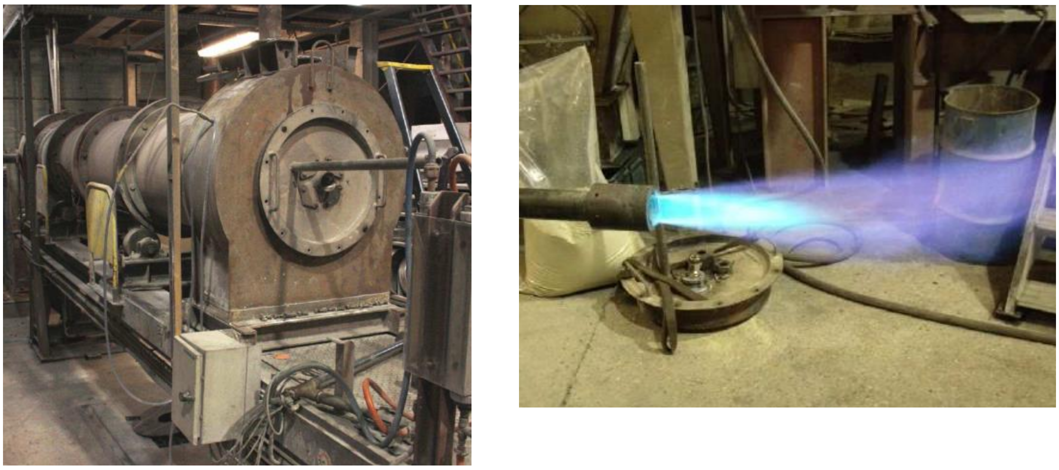


Figure 4-2 Photo of the lab kiln and burner in FCB's facilities

The kiln is heated with a propane burner with a maximum capacity of 125 kW. In order to have a very short flame, which is necessary in such a small-scale kiln, a burner with gas pre-mixing was chosen (nozzle diameter 50 mm). An attempt to use a burner designed especially for oxyfuel burning failed. The fuel is commercially available propane, with a specific mass of 1.98 kg/Nm³ and a LCV of 91,400 kJ/Nm³.

The clinker cooler only allows a quick cooling of the clinker. No hot air from the cooler is used as secondary air in the combustion zone. However, a resistor gas heater is implemented to preheat the combustion air or the CO_2-O_2 mix. A unit including a CO_2 tank, oxygen cylinders, and the necessary gantry was supplied to the FCB research facility by AIR LIQUIDE. It feeds both the burner and the burning zone.

Table 4-2 Operational modes

	Reference operation	Oxyfuel operation
Clinker outlet	9.5 kg/h	15.7 kg/h
Fuel energy input	70.4 kW	67.9 kW
Kiln and hood heat losses	20.9 kW	19.4 kW
Flue gas enthalpy	35.5 kW	23.4 kW

Due to the different supply and preheating of combustion gases and varying experimental settings (different clinker production) caused by lab kiln instabilities the operational modes are not directly comparable with each other (**Table 4-2**). However, under oxyfuel conditions (CO_2 -O_2 mix) a proper operation could be achieved. The comparison of the development of the kiln inner temperatures does not allow to differentiate between the profiles with air and those with oxyfuel burning. With the change to oxyfuel burning, all relevant parameters had to be newly adjusted each time. Once during the adjustment time temperatures in the burning zone were too high and even led to the melting of the refractory.

By optimising the process parameters a suitable temperature profile could be adjusted to produce clinker granules of high quality (**Figure 4-3**). Due to the gas firing no ashes were introduced to the clinker. As raw material from an existing kiln line was used without adaptation to the missing ashes, free lime contents in the lab clinker were higher than usual due to higher LSF.

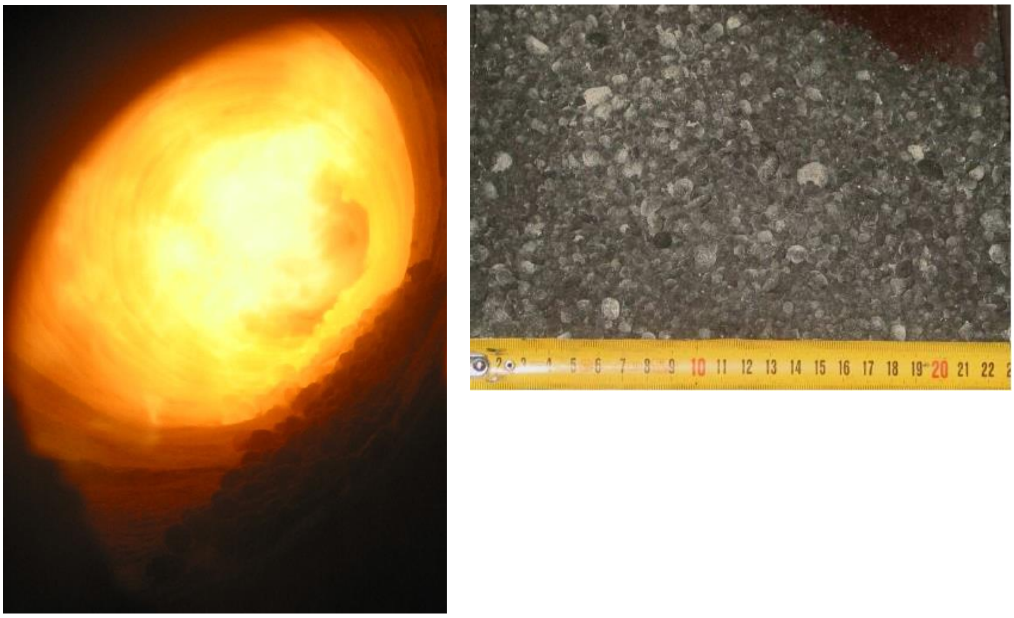


Figure 4-3 Clinker granules within the kiln and produced clinker

Similar clinker qualities could be produced under air and oxyfuel-fired conditions (**Table 4-3**). Apart from kinetic parameters this lab-scale testing proved that chemical reactions like the calcination and mineralogical conversion of the material are feasible under oxyfuel conditions.

Table 4-3 Clinker composition

Parameter in wt.%	Reference	Oxyfuel
Alite	62.9	65.4
Belite	18.3	17.4
C_3A cub	7.2	5.3
C_3A orth	< 0.5	< 0.5
$C_2(A,F)$	7.0	7.3
Free lime	3.3	2.4
Periclase	0.8	0.5

4.2 Equipment dimensions

A simplified layout of an oxyfuel cement plant was used as a simulation basis (acc. **Figure 1-1**). The plant is based on BAT-standard, meaning a 5-stage cyclone preheater, calciner with tertiary air duct, rotary kiln and two-stage grate cooler. Auxiliaries like exhaust air cleaning and the utilisation of waste heat are combined with an existing infrastructure. The treatment of the discharged CO_2-rich flue gas stream is not included.

Coal is taken as a reference fuel (**Table 4-4**). A typical composition for the coal and raw material was chosen (**Table 4-5**).

Table 4-4 Coal composition

Parameter	Value	Ash composition	
Specific heat capacity	2.50 kJ/(kg*K)	SiO_2	41.40 wt.%
Heat value	27,000 kJ/kg	Al_2O_3	27.30 wt.%
Composition		TiO_2	1.40 wt.%
C	69.00 wt.%	P_2O_5	1.10 wt.%
H	4.00 wt.%	Fe_2O_3	4.00 wt.%
O	9.00 wt.%	Mn_2O_3	0.10 wt.%
N	0.48 wt.%	CaO	18.20 wt.%
S	0.50 wt.%	MgO	1.70 wt.%
Cl	0.02 wt.%	SO_3	4.00 wt.%
Moisture	0.50 wt.%	K_2O	0.60 wt.%
Ash	16.5 wt.%	Na_2O	0.20 wt.%

Table 4-5 Raw material composition

Parameter	Value	Parameter	Value
SiO_2	13.80 wt.%	SO_3	0.34 wt.%
Al_2O_3	3.25 wt.%	Sulfid	0.05 wt.%
TiO_2	0.06 wt.%	K_2O	0.55 wt.%
P_2O_5	0.04 wt.%	Na_2O	0.12 wt.%
Fe_2O_3	1.96 wt.%	chlorine	0.01 wt.%
Mn_2O_3	0.05 wt.%	TOC	0.10 wt.%
CaO	43.22 wt.%	H_2O	1.00 wt.%
MgO	0.71 wt.%	LOI	35.74 wt.%
CO_2	34.74 wt.%		

Based on the assumption that the plant should also be able to be operated under conventional conditions (e.g. for retrofitting an existing plant), a recirculation rate of 0.54 (at an oxygen level in the combustion gas of 22.4 vol.%) was determined.

4.2.1 Preheater

Based on the assumption of an optimal gas velocity of 15 to 20 m/s in the cyclone entrance and the separation efficiency as shown in **Table 4-6**, the cyclones were dimensioned (**Table 4-7**).

Table 4-6 Cyclone preheater separation efficiency

	Cyclone 1	Cyclone 2	Cyclone 3	Cyclone 4	Cyclone 5
Efficiency in %	96	88	86	83	78

Table 4-7 Cyclone preheater sizing

	500 t/d		1000 t/d	
	Diameter in m	Pressure drop in daPa	Diameter in m	Pressure drop in daPa
Cyclone 1	2.8	155.3	4.2	150.5
Cyclone 2	2.8	75.1	4.1	66.3
Cyclone 3	2.9	74.5	4.2	70.0
Cyclone 4	2.9	70.7	4.3	69.9
Cyclone 5	3.0	38.4	4.4	34.5

4.2.2 Calciner

In the previous phases it was observed in the small-scale experiments that calcination temperatures increase by 80 K due to higher CO_2 partial pressure conditions under oxyfuel operation. However, under the complex thermo-fluid dynamic conditions of a calciner, it would require detailed analysis using a model with combustion and mineral interactions to establish the calciner exit temperatures as well as the degree of calcination of the injected meal.

To evaluate the calciner performance the different steps of the calcination reaction have to be examined:

1. Heat transfer from the surrounding gas atmosphere to the particle and heat conduction through the particle
2. Chemical reaction
3. Diffusion of the CO_2 through the particle and material transfer from the particle surface to the surrounding gas atmosphere

In typical calciners, where raw material particles are introduced as meal, the chemical reaction is the determining step. The degree of calcination is therefore strongly dependent on temperature, CO_2 content in the surrounding gas phase and residence time. However, in practice the heat transfer and therefore the fuel burnout in the calciner also have to be considered.

4.2.2.1 Pre-examination

In order to assess the behaviour of a calciner under enriched CO_2 conditions, an existing calciner configuration of a clinker production line of 5,600 t/d and a fuel split of 60/40 % (calciner/main burner) were used as a pre-examination. The calciner was supplied with 12.5 t/h pet-coke and a residence time of 6 seconds was envisaged. Apart from the composition of tertiary and kiln exhaust gases, process parameters related to the performance, like fuel and material distribution, injection points for material and gas phases etc. were kept constant. Moreover, in the oxyfuel case the primary air was replaced by oxygen.

The 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 fuel burns 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 the oxyfuel case.

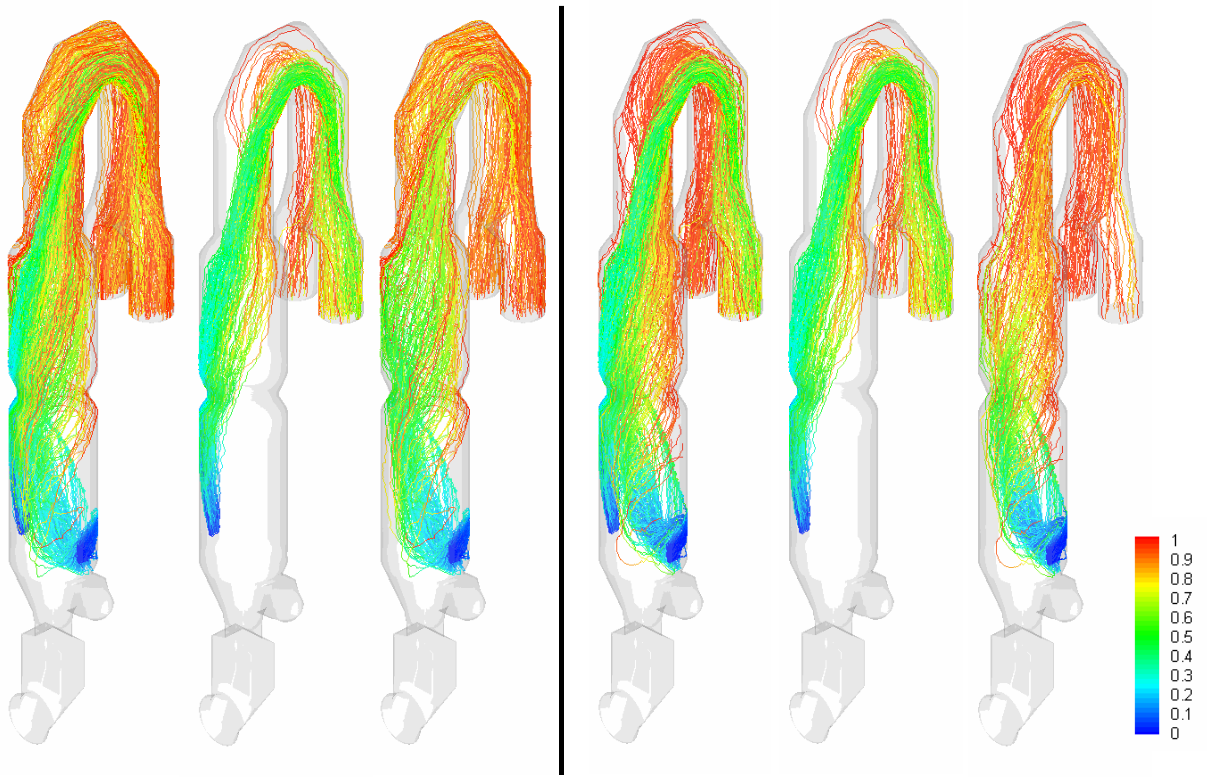
- Chemical reaction:

  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 cases exhibit different velocities, which result in a slightly higher residence time in the oxyfuel case.

Balancing these factors the calcination degree drops from 93 to 91 % comparing the air-fired and the oxyfuel case (**Figure 4-4**). The temperature is increased from 914°C to 930°C at the calciner outlet. The slight reduction in the calcination levels could be addressed further, if needed, by optimising the oxyfuel burners, since a 6-second residence time is sufficient to ensure both higher fuel burnout and meal calcination levels.



(a) (b)

Figure 4-4 Comparison of meal calcinations (a) for air-fired case achieving 93% and (b) for oxyfuel case achieving 91%.

4.2.2.2 Sizing of the testing calciners

The calciner design for the testing facility should be a common in-line calciner, which has been scaled down. The calciner has two opposite burners, which are both located 1.5 m below the tertiary air injection level at a 30° downwards angle. Two multi-channel burners inject the fuel with a transport air velocity of 30 m/s. An axial sleeve injects the primary air at velocities of approx. 15 m/s.

The meal is introduced just above the burners located on the other two opposed walls in a direction perpendicular to that of the burners (see geometry in **Figure 4-5**). The calciner has a diameter of 1.98 m and a height of 37 m with two small restrictions after 8 and 11.3 m. This calciner geometry allows a residence time of approximately 3 seconds.

Figure 4-6 shows the input conditions of the gas composition for the oxyfuel calciner.

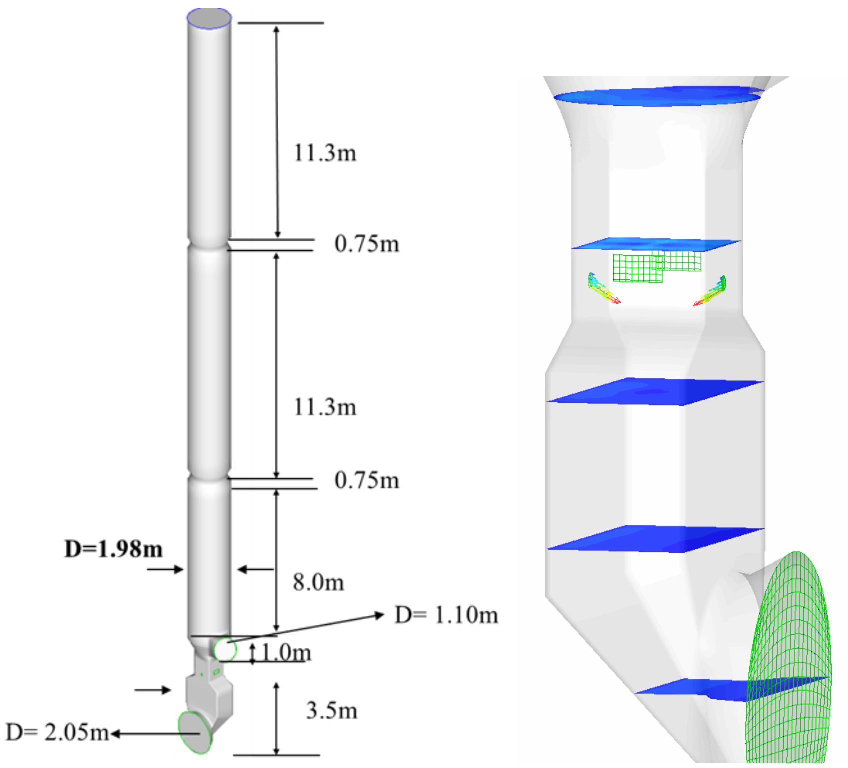


Figure 4-5 The geometry of a 500 t/d scaled-down testing calciner

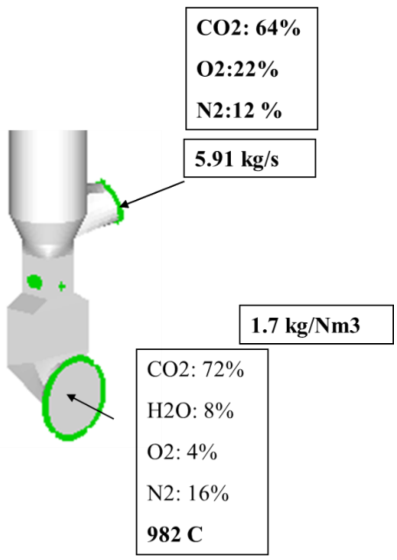


Figure 4-6 Input conditions of the oxyfuel calciner

Table 4-8 shows the performance of the calciner under both conditions, the air-fired and the oxyfuel case. Based on the impacts on the process, which have been identified in section 4.2.2.1, the degree of calcination drops by 6 % at a slight exit temperature increase under oxyfuel conditions. Using this operation as a starting point, the performance was optimised by adapting:

- Transport gas composition
- Meal supply

Table 4-8 Comparison of the calciner performance (500 t/d) in air-fired and oxyfuel case

Case	Degree of calcination reached in calciner	Total calcination (at kiln inlet) including the entering of pre-calcined material	Calciner exit temperature
Air-fired	93 %	94.5 %	904 °C
Oxyfuel	86 %	88.5 %	917 °C
Delta		6 %	13 °C

The adjustment of the transport gas composition had only a minor impact on the final calcination rate, as shown in **Table 4-9**. The degree of calcination was reduced slightly due to the lower temperatures achieved initially in the near burner region when the oxygen percentage was reduced.

Table 4-9 Variation of the transport gas composition

Transport gas composition	Burnout	Total calcination
7 % O_2 / 93 % CO_2	97 %	87 %
21 % O_2 / 79 % CO_2	97.4 %	88 %

If more meal is supplied above the tertiary gas injection (meal splitting) the temperatures near the burners will increase. For this reason the meal, which is supplied in the lower level, is calcined to a higher degree. Although the residence time for the meal supplied above the tertiary air injection level is slightly decreased, the total degree of calcination is marginally increased (**Table 4-10**).

Table 4-10 Variation of the meal split

Meal split	Burnout	Total calcination	Temperature
All meal below tertiary air injection	98 %	89 %	917 °C
30% meal input above tertiary air injection	97 %	89.5 %	914 °C
50% meal input above tertiary air injection	98 %	90 %	912 °C

Based on these findings the 500 t/d geometry (including burner position and meal split) was adapted to the second case of 1000 t/d production rate. A goose-neck calciner (looped tube without complex turbulence chamber) was considered to accommodate a residence time of around 3 seconds with a calciner vessel of a mean diameter of 2.8 m as shown in **Figure 4-7**.

For this geometry two variations were studied depending on the cooler specifications. The conditions would correspond to the following:

1) Short and wide cooler resulting in lower tertiary air temperatures (spec. 1)
2) Long and narrow cooler resulting in higher tertiary air temperatures and changed fuel split (spec. 2)

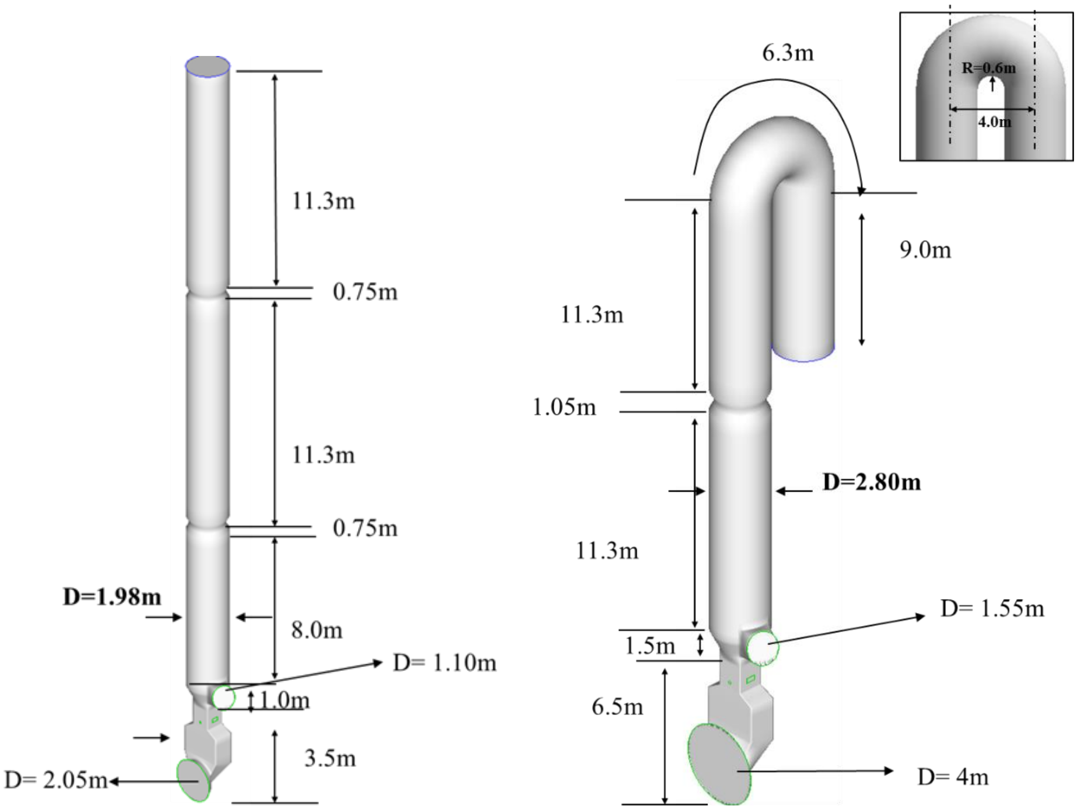


Figure 4-7 Geometry of the 500 and 1000 t/d kiln calciner

Table 4-11 Comparison of the calciner performance (1000 t/d) in air-fired and oxyfuel cases depending on the cooler specification

Case	Degree of calcination reached in calciner	Total calcination including the entering of pre-calcined material	Calciner exit temperature
Air-fired, cooler spec. 1	93 %	94 %	929 °C
Oxyfuel, cooler spec. 1	88 %	90 %	938 °C
Difference	5 %	4 %	9 °C
Air-fired, cooler spec. 2	84 %	87 %	897 °C
Oxyfuel, cooler spec. 2	80 %	83 %	920 °C
Difference	4 %	4 %	23 °C

Basically, similar trends to that of previous findings were observed with a reduction of calcination by 4 % in the oxyfuel case (**Table 4-11**). The effect of the two cooler types on the calciner is mainly driven by the different heat load supplied to the calciner. Although the wide/short cooler configuration generates lower tertiary gas temperatures, the heat input is higher due to higher volume flows and higher kiln exhaust gas temperatures. On this basis the first configuration allows a higher degree of calcination (difference of 8 %) than the narrow/long cooler.

The tertiary air duct is sized according to the kiln length and the requirement of ca. 25 m/s gas velocity to avoid a settling of dust in the tube (**Table 4-12**).

Table 4-12 Size of the tertiary air duct

Capacity	Diameter	Length
500 t/d	1.48 m	30 m
1000 t/d	2.10 m	58 m

4.2.3 Kiln

Table 4-13 lists the kiln dimensions of a newly constructed 500 t/d kiln (based on optimal specific construction values) and the fixed kiln dimensions of a 1000 t/d kiln, on the basis of which a retrofitted kiln is dimensioned.

Table 4-13 Kiln dimensions

	500 t/d	1000 t/d
Specific Load (according to designed capacity)	5 t/(d*m^3)	1.5 t/(d*m^3)
L/D ratio	15	14
Kiln inner diameter	2 m	4.15 m
Kiln length	30 m	58 m
Steel strength	0.02 m	0.02 m
Refractory	0.2 m	0.2 m

4.2.4 Cooler

The cooler was newly dimensioned for both capacities using a specific load of 40 t/(d*m^2) and a clinker bed velocity of 0.5 m/min as a basis (**Table 4-14**). In order to build up a sufficient clinker bed height to ensure a good heat transfer between the clinker and recuperation gas the cooler was quite narrowly dimensioned (specific width of 670 t/(d*m)).

Table 4-14 Cooler dimensions

	500 t/d	1000 t/d
Cooler area	12.6 m^2	28.75 m^2
Cooler width	0.75 m	1.6 m
Cooler length	16.8 m	18.0 m
First stage (CO_2), length	8.0 m	8.0 m
Second stage (Air), length	8.8 m	10.0 m

4.3 Process parameters

The following section shows the process parameters of the 500 t/d and 1000 t/d plant as well as the partial load operation originating from simulation (as proposed in **Figure 3-7**). The operation in partial load is based on an exemplary existing 1150 t/d kiln, but on reduced capacity to 500 t/d.

For all three operational modes the following data are listed:

- Mass flows (**Table 4-15**)
- Gas volume flows and dust contents (**Table 4-16**)
- Energy balance (**Table 4-17**)

Comparing the mass flows of the 500 t/d kiln and the kiln with partial load of 500 t/d, the raw meal input is nearly similar. Based on the higher energy demand, the absolute mass of fuel and the related ashes, which become part of the clinker, the clinker production is higher in the partial load kiln (~ 6 %).

Table 4-15 Mass flows

Material flows	500 t/d		1000 t/d		Partial load	
	Flow rate kg/h	Specific flow rate, kg/kg_{cli}	Flow rate, kg/h	Specific flow rate, kg/kg_{cli}	Flow rate, kg/h	Specific flow rate, kg/kg_{cli}
Raw meal	33,850	1.6	78,672	1.6	35,200	1.6
Kiln feed	23,998	1.13	58,407	1.18	25,344	1.13
Fuel calciner	1,708	0.08	3,960	0.09	2,805	0.13
Fuel kiln	990	0.05	2,540	0.05	1,445	0.07
Clinker	21,200	1.0	49,341	1.0	22,448	1.0

Table 4-16 Gas volume flows and dust contents at standard conditions

Gas flows/ dust content	500 t/d		1000 t/d		Partial load	
	Flow rate, m^3/h	Specific flow rate, m^3/kg_{cli}	Flow rate, m^3/h	Specific flow rate, m^3/kg_{cli}	Flow rate, m^3/h	Specific flow rate, m^3/kg_{cli}
Cooling gas (recirculated gas)	17,928	0.85	42,024	0.85	41,221	1.84
Cooling air (second stage)	20,500	0.97	47,900	0.97	-	-
Secondary gas	4,966	0.23	11,641	0.24	8,656	0.39
Tertiary gas	12,908	0.61	30,257	0.61	18,137	0.81

Gas flows/ dust content	500 t/d		1000 t/d		Partial load	
	Flow rate, m^3/h	Specific flow rate, m^3/kg_{cli}	Flow rate, m^3/h	Specific flow rate, m^3/kg_{cli}	Flow rate, m^3/h	Specific flow rate, m^3/kg_{cli}
Cooler exhaust gas bypass to calciner					13,273	0.59
Cooler exhaust air	20,500	0.97	47,900	0.97	1,154	0.05
Cooler exhaust, dust content	10 g/m^3		10 g/m^3		10 g/m^3	
Kiln inlet	6,717	0.32	16,258	0.33	10,760	0.48
Kiln inlet, dust content	268 g/m^3		283 g/m^3		216 g/m^3	
Flue gas (preheater exit)	26,083	1.23	60,748	1.23	51,177	2.23
Flue gas dust content	60.7 g/m^3		60 g/m^3		32.3 g/m^3	
Oxidizer	3,740	0.18	8,935	0.18	6,200	0.28
Flue gas for recirculation	14,208	0.67	33,089	0.67	33,867	1.51
Flue gas for conditioning	12,103	0.57	28,186	0.57	17.838	0.8
False air, kiln plant	1,930	0.09	4150	0.09	4150	0.19
False air, recirculation	230	0.01	490	0.01	490	0.02
Primary gas, main burner	810	0.04	2060	0.04	2060	0.09
Primary gas, calciner	250	0.01	495	0.01	495	0.02

According to the energy balance the specific energy demand amounts to:

- 3,473 $kJ/kg_{clinker}$ in the case of 500 t/d capacity. Although the plant is designed to include best available techniques (e.g. calciner, grate cooler etc.) the energy demand is high compared to conventionally sized plants due to the higher specific wall heat losses.
- 3,576 $kJ/kg_{clinker}$ in the case of 1000 t/d capacity. The calculation is based on a real kiln designed for 1150 t/d, for which reason the specific energy demand is higher than the BAT level.
- 5,057 $kJ/kg_{clinker}$ in the case of partial load. Due to the strongly reduced capacity and the splitted gas management the energy demand is significantly increased.

Table 4-17 Energy balance of the oxyfuel testing kiln

Energy balance in kJ/kg $_{clinker}$	500 t/d	1000 t/d	Partial load
Input			
raw material	86.2	86.1	83.4
fuel (main burner), sensible heat	7.0	7.7	9.7
primary gas (main burner)	0.84	0.92	1.02
fuel (calciner), sensible heat	12.1	12.0	18.7
primary gas (calciner)	0.27	0.23	0.69
cooling air (cooler stage 2 input)	21.3	21.2	-
recirculated O_2 enriched gas (cooler stage 1 input)	43.7	45.3	108.2
false air (preheater)	0.73	0.91	1.97
false air (calciner)	0.38	0.26	0.56
false air (kiln inlet)	0.44	0.28	0.61
false air (kiln hood)	0.26	0.21	0.45
reaction enthalpy of fuels	3,454	3,556	5,029
sum of Input	3,627	3,731	5,254
Output			
CO_2 rich flue gas	996	1,058	2,302
flue gas gas dust	30	32	39
clinker incl. cooler off gas dust	110	82	142
cooler exhaust air	354	315	-
wall losses (preheater)	26	17	42
wall losses (calciner + tertiary duct)	160	129	263
wall losses (rotary kiln)	300	433	879
wall losses (cooler)	11	9	10
reaction enthalpy of kiln feed	1,640	1,656	1,777
incomplete combustion	0.0	0.0	0.0
sum of Output	3,627	3,731	5,254

Figure 4-8 to **Figure 4-10** illustrate the calculated material conversion in the kiln. In all cases adequate clinker is produced. At 500 t/d the intermediate clinker phase formation starts earlier in the kiln. Flame characteristics and temperature profiles differ with respect to the kiln length at an increased capacity of 1000 t/d. Consequently, the main material conversion is shifted to the sintering zone. Operating the 100 t/d kiln at partial load shows only slight deviations, as the material load in the kiln is much lower.

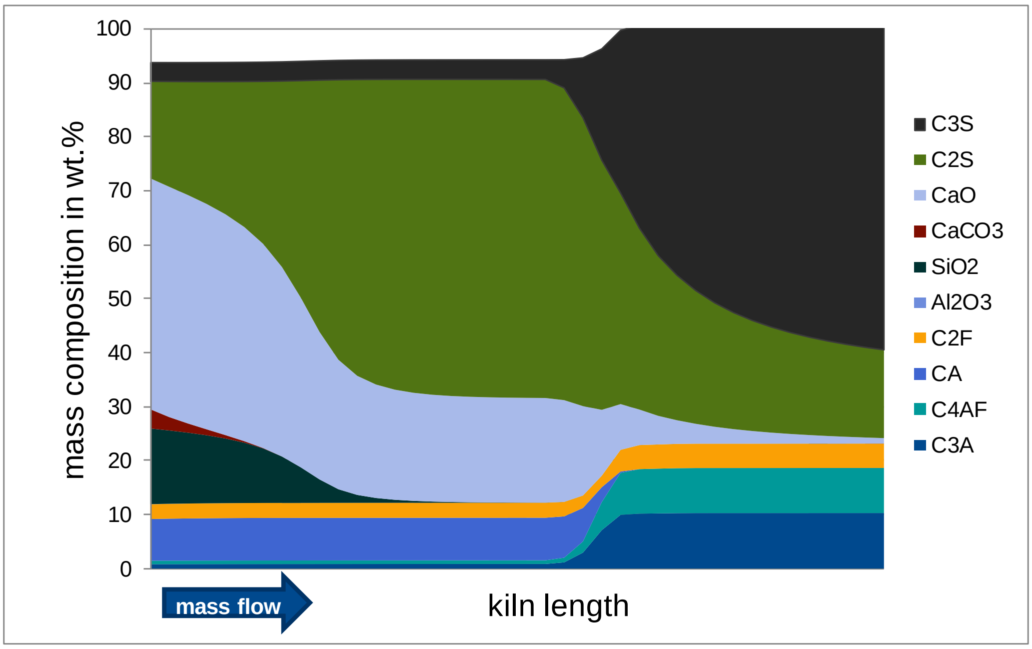


Figure 4-8 Material conversion in 500 t/d kiln

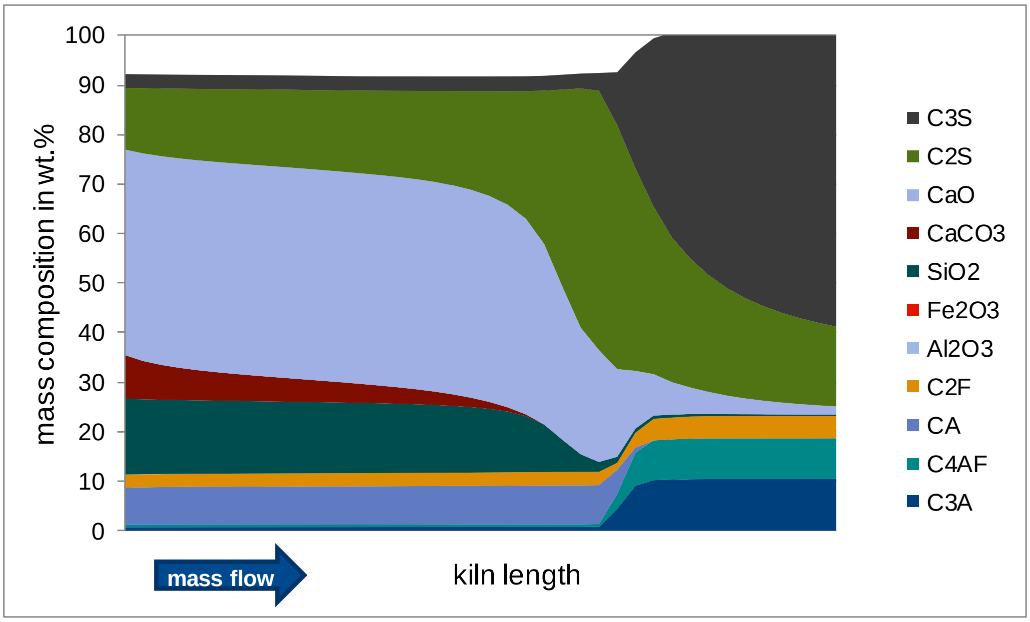


Figure 4-9 Material conversion in 1000 t/d kiln

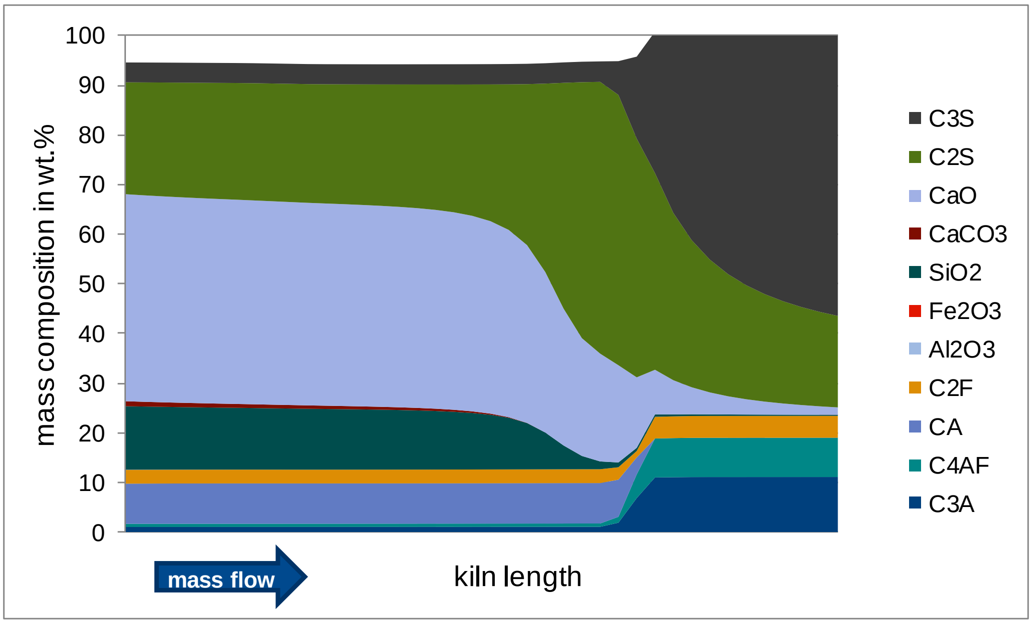


Figure 4-10 Material conversion in kiln, partial load

5 Proposal for operational set-up

The project to build and operate a testing facility is currently planned for a period of four years. The engineering, design and construction period is expected to require a minimum of two years. Thus the plant could be commissioned within the third year of the project. The testing period itself will last around two years with 3 months' operation per year. In between the two major testing campaigns the experimental set-up can be adapted depending on the previous results.

The operation of a pilot plant should verify the theoretical findings from previous project phases. These include:

- start-ups / shut downs or in general switching mode from conventional to oxyfuel operation
- operational mode and adjustment of the process parameters
- burner set-up
- clinker and cement quality
- long-term tests on refractories
- flue gas composition at different operational modes
- sealing aspect (improved maintenance) in long-term operation
- energy and mass balances based on experimental data
- impact of the changed burning atmosphere on volatiles (leading to incrustation and cycles)
- at a later stage: switch to alternative fuels (as an option)
- identification of plant-specific impacts (influenced by the size or surrounding aspects) and general statements

6 Cost estimation

The cost estimation is based on the described design and dimensioning of a potential testing facility. More precise costs can only be provided on the basis of a definite location. Therefore, the costs mentioned below are based on an uncertainty level of ± 25%.

A 500 t/d brownfield installation was chosen as the basis for evaluating the investment and operational costs. In the case of a blackfield plant the identification of investment costs is strongly related to the necessary modifications. In addition, the size of the kiln plant can significantly influence the operational costs. A precise plant location is therefore required to make a qualified statement about the overall costs.

6.1 Investment costs

The costs of the kiln equipment for a 500 t/d brownfield installation were estimated on a down-scaling of investment costs for existing kiln plants. The increase of specific costs per yearly production with decreasing size has been taken into account. However, only approx. 42 % of the equipment costs could be traced back to the kiln components themselves, the remaining costs are attributed to units related to the oxyfuel technology (**Table 6-1**), such as:

- Recirculation path: Many connections are necessary between the different new pieces of equipment (condenser, filter etc.), some of them internally, others externally insulated.
- Oxygen supply: Thr tank belongs to the gas supplier, costs occur only for the mounting and dismounting of the tank.
- Storage and supply chain: Connections from the existing plant for materials supply have to be installed, including all necessary dosing systems (solid fuel, raw meal, oxygen flow meter etc.)
- More efficient sealings: Conventional sealing of the plant is not sufficient for the oxyfuel process
- Advanced instrumentation: Detailed information will be gathered from gas analysers for e.g. oxygen and CO_2, and temperature and pressure probes. A dedicated control panel has to be installed, including a data acquisition system.

Table 6-1 Equipment costs

Position	Costs	Unit
Kiln plant	6.7	M€
Recirculation (fans, piping in stainless steel)	1.6	M€
Condenser, bag filter	3.5	M€
Oxygen tank system	0.2	M€
Storage and supply chain	1.5	M€
More efficient sealings	0.6	M€
Instrumentation	2	M€
Total equipment costs	16.1	M€

In order to estimate the total plant costs, charges for installation (civil steel works, erection), engineering, procurement, construction (EPC), fees and contingencies have to be added to the total equipment costs (**Table 6-2**).

Table 6-2 Total plant costs

Position	Costs	Unit
Total Equipment	16.1	M€
Installation costs	14.3	M€
EPC, fees	4.1	M€
Contingencies	3.5	M€
Total Plant costs (TPC)	**38**	**M€**

6.2 Operational costs

Before switching to oxyfuel conditions the plant has to be started up in conventional operation until stable conditions are reached. The calculation basis for the operational costs is therefore 80 days of conventional operation and 80 days of oxyfuel operation. **Table 6-3** and **Table 6-4** list the estimated operational costs. In total the operational costs amount to 4.43 M€ over the entire testing period.

Table 6-3 Variable operational costs

Item	Unit prices		Demand		Absolute costs in €
Power	80	€/MWh	70	kWh/ $t_{clinker}$	450,000
Fuels	80	€/t	0.15	t_{fuel}/ $t_{clinker}$	960,000
Process water	0.2	€/m^3	0.2	m^3/ $_{clinker}$	5,000
Raw material	5	€/t	1.6	t_{RM}/ $t_{clinker}$	640,000
Oxygen	80	€/t	0.25	t O_2/ $t_{clinker}$	800,000
Sum					**2,855,000**

Table 6-4 Fixed operational costs

Item	Unit prices		Absolute costs in €
Maintenance	6.8	€/$t_{clinker}$	540,000
Labour	7.5	€/$t_{clinker}$	600,000
Administration	3.2	€/$t_{clinker}$	260,000
Insurance	1.1	€/$t_{clinker}$	90,000
Local taxes	1.1	€/$t_{clinker}$	90,000
Sum			**1,580,000**

6.2.1 Cost sensitivity

Consumable prices in particular are subject to local markets. The operational costs can therefore differ depending on the hosting country. As 30 to 60 % of the oxygen costs are due to transport costs, these costs also differ depending on the location.

A cost variation of the consumable prices shows that the operational costs are most sensitive to fuel and oxygen costs (**Figure 6-1**). However, fuel costs are currently quite stable (orhave even decreased in the recent years) in Europe, thus the oxygen costs are indeed the most influencing factor.

In summary, the maximum variation range of operational costs is 700,000 €, which is about 1.5 % of the total project costs. The major cost uncertainty can therefore be traced back to the investment costs.

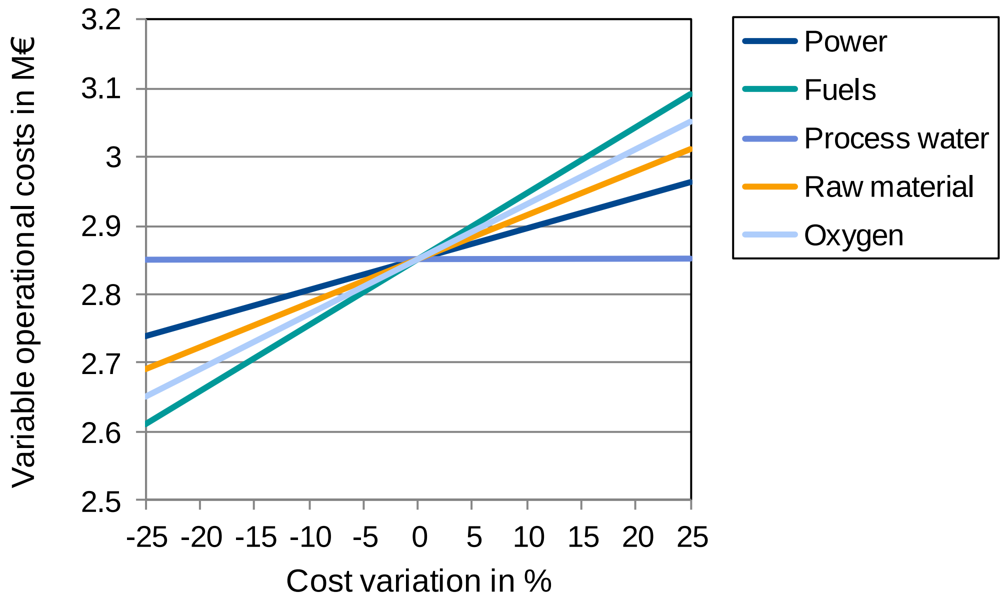


Figure 6-1 Sensitivity to variable operational cost fluctuation

6.3 Costs for scientific evaluation

The testing will be accompanied by comprehensive measurement campaigns. The measurement is estimated to cost 1 M€ including expenses for staff, transport, equipment, analysis and travelling. The scientific evaluation, coordination and dissemination is estimated at 0.5 M€.

6.4 Overall costs and financing scenarios

Including all costs described the project expenses amount to 44 M€ for a 500 t/d brownfield plant. Assuming a cost uncertainty of ± 25 % the project costs vary between 33 and 55 M€.

Although the investment for the kiln equipment might be lower in the case of a blackfield plant, the overall costs are not necessarily lower compared to the brownfield installation. Due to a higher production capacity of the existing facilities and due to the necessary plant modifications for a blackfield plant, not only the investment costs, but even more so the operational costs are site-specific and depend very much on e.g. the kiln size. A high production capacity would therefore increase operational costs, which then would counteract lower investment costs for retrofitting. In any case, the exact estimation of project costs for a blackfield pilot requires the definition of a specific location.

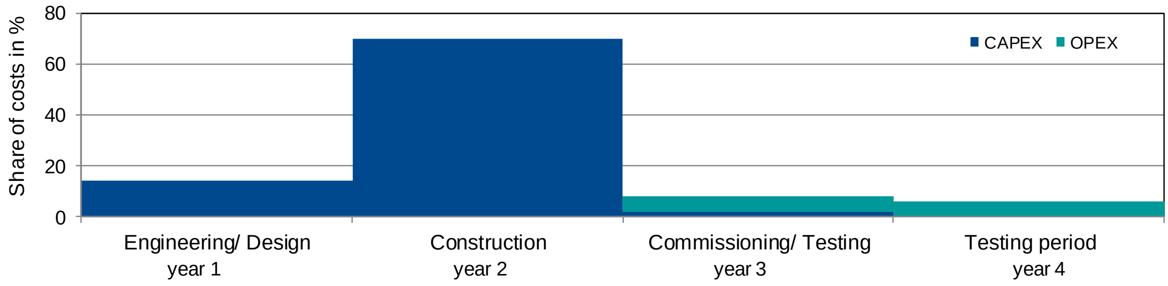


Figure 6-2 Distribution of costs per project phase

It is assumed that up to 90 % of the project costs will have to be spent in the first 2.5 years for engineering, construction and commissioning (**Figure 6-2**). Compared to conventional construction the cost fraction of the engineering requires a higher amount due to designing units regarded as critical parts (condenser, sealing etc.). The operational costs represent the last 10 % of the project costs for the testing period (year 3 and 4).

Basically, the end use of the plant as well as the use of the produced clinker is dependent on the terms and conditions of the funding organisation. In order to influence the project costs different business cases can be considered:

- No further use of the plant: Project costs remain unchanged (reference case)
- Further use of the plant for conventional purposes: Limited funding expected for conventional equipment parts
- Further use of the plant as a research centre: Limited funding expected for further used equipment.
- Selling the clinker at conventional production prices (no margin included)

In the case of a further use of the plant expected funding for the project could be reduced.

7 Organisational issues

7.1 Concept for plant reuse

To exploit the testing plant as much as possible and maximise the economic utilisation, the use of the plant after the testing phase would be worthwhile. The options depend on the location (e.g. raw material source, infrastructure), funding requirements, the local market and the plans of the hosting company. Potential options are:

- the reuse for other research projects
- the re-selling and reconstruction at another location
- operation for training aspects
- the production of special cements
- the disposal of special wastes.

In any case the optimal option depends on the site-specific aspects named above and shall therefore be evaluated for the final plant location.

7.2 Business case analysis

In order to maximise the economic benefit with respect to the long-term perspective, a business case for the pilot and its after-use was created. However, it is clear that there will be no case in which amortisation of the investment for the research equipment dedicated to the oxyfuel operation will be achieved.

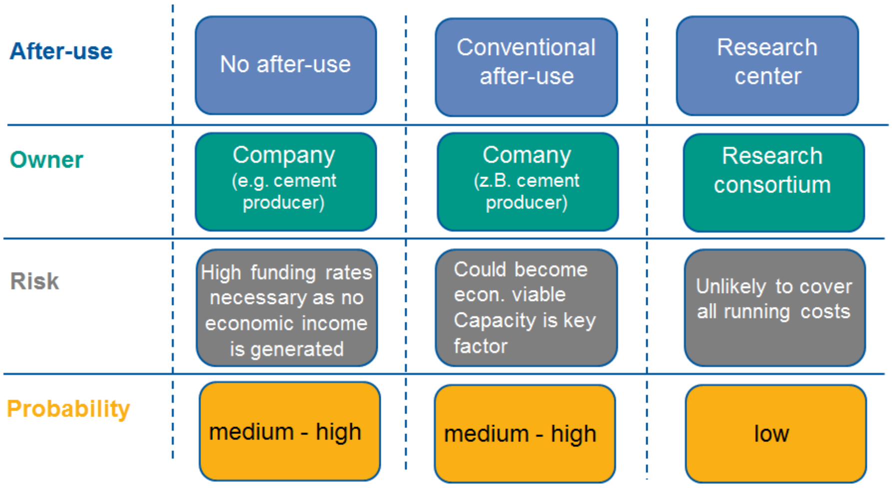


Figure 7-1 Evaluation matrix of business cases

Based on balancing the risks, benefits, probability and impact on costs, the scenarios can be ranked in terms of best available business cases (Figure 7-1). The highest economic benefit from a pilot installation could be achieved if the plant is used afterwards conventionally. But the envisaged capacity will probably be higher than 500 t/d, which influences the costs. Even if no income can be generated from a plant with no after-use, the scenario seems viable due to the lower investment (caused by shorter lifetime) and no risk for the future benefit. On the other hand, the operation of a research centre entails high risks for the future utilisation. During testing campaigns running costs accumulate (e.g. due to maintenance and minimum staff) and newly formed consortia may not possess the reserves to cover those costs.

7.3 Legal issues

After the technical and economic evaluation of a potential site several other aspects still need to be addressed and agreed upon between ECRA and the site owner. For example, these includes aspects regarding the operating and investment costs for the test campaigns, the accessibility to the plant for externals, the responsibilities regarding the modification of the kiln line and the running of the test campaigns (including all safety and environmental matters). However, it is clear that the hosting company from a legal point of view will be:

- the owner of the project
- the coordinator of the project on-site, incl. staff responsibility
- responsible for all safety aspects
- contractor to the sponsors (e.g. the EU)

The project management will have to be split into three task areas: project management, commissioning management and research management (**Figure 7-2**). The project and commissioning management is the same as for setting up a conventional kiln. This implies that the management can be done either by the hosting company or be subcontracted to the equipment supplier or another external company. A detailed memorandum of understanding will be developed by an ECRA task group.

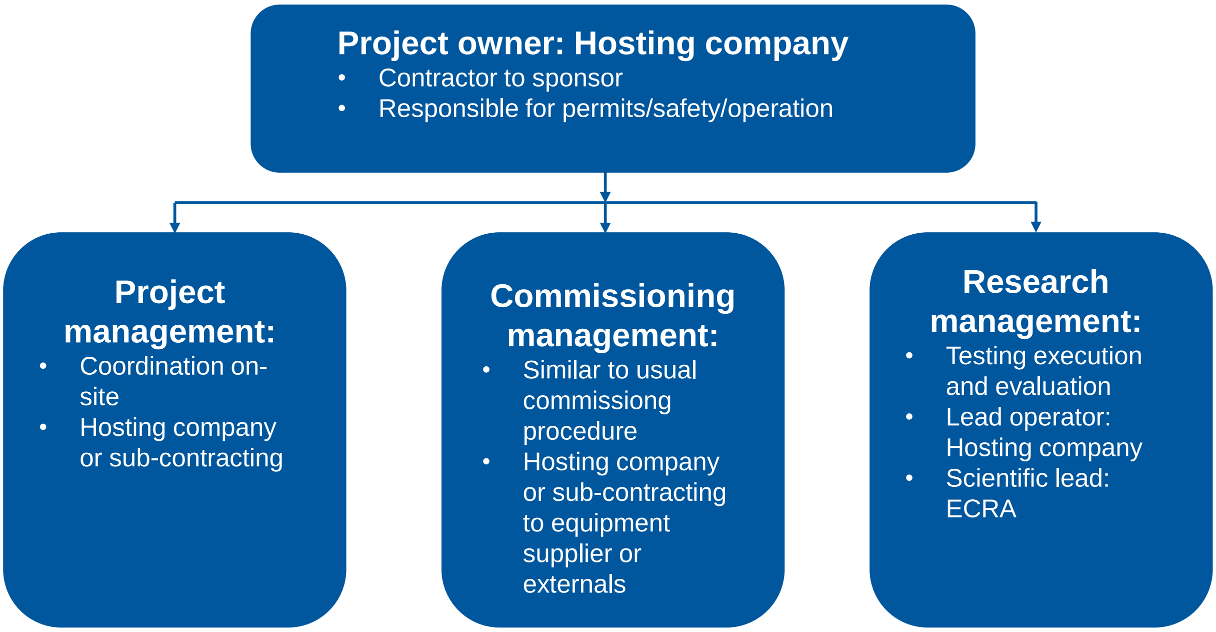


Figure 7-2 Project legal structure

8 Summary and outlook

Several roadmaps for CO_2 reduction underline the need for a significant share of cement plants worldwide to be equipped with carbon capture as a breakthrough technology. Applied to the European cement industry this share is expected to be around 60%.[4] To achieve this target carbon capture technologies should already have been proven today.[5]

Whereas post-combustion technologies are currently being investigated at pilot scale within the Norcem project in Norway, oxyfuel technology is still at lab-scale level. From today's perspective, oxyfuel technology seems to offer a high potential (e.g. with regard to costs) in comparison to other capture technologies and should therefore be further investigated in industrial surroundings. For this reason ECRA has developed a concept for an industrial testing facility which addresses the scale, design, dimensioning, costs estimation and further use of the facility.

The considerations about equipment sizes depending on the material and gas flow patterns lead to the conclusion that a production rate of 500 to 1000 t/d is required to allow the transferability of results to full industrial scale. Based on this, different options were developed. The industrial testing kiln could be realised either as a so-called "brownfield" or "blackfield" project. A "brownfield" site would provide the infrastructure for an oxyfuel pilot kiln, but the kiln line itself (preheater, calciner, rotary kiln, cooler) would have to be newly constructed. A "blackfield" site would take advantage of an existing kiln line and would require the modification of the kiln line. An operational installation running alongside the modified pilot kiln would be desirable in order to provide infrastructure and operating staff. The equipment and production capacity of each of these solutions were dimensioned by taking the changed operational mode (e.g. with regard to calcination) into account. The project costs including expenses for investment, operation and measurements of a 500 t/d brownfield plant were estimated at 44 M€ (± 25%). In the case of a blackfield plant the identification of investment costs is strongly related to the necessary modifications. In addition, the size of the kiln plant can significantly influence the operational costs. A precise plant location is therefore required to make a qualified statement about the overall costs.

Different scenarios concerning the end use of the plant or the product might influence the overall costs, which strongly depend on the final location of the plant. In any case, the project costs are too high to be borne solely by the industry itself, which makes funding essential. In order to apply for funding, a precise project outline will be created. In the case of the blackfield option in particular it has become clear that answers to some of the detailed questions (especially with regard to costs) can only be found on the basis of a specific location.

[4] International Energy Agency: Global Action to Advance Carbon Capture and Storage – A Focus on Industrial Applications. Annex to Tracking Clean Energy Progress 2013

[5] International Energy Agency: Cement Technology Roadmap 2009; Carbon emissions reductions up to 2050. December 2009