

1 Final publishable summary report

1.1 Executive summary

The CaOling project is the largest and most ambitious project in the world to test the concept of post-combustion CO₂ capture by calcium looping using flue gases from a commercial power plant. This is a promising CCS technology in terms of lower CO₂ capture cost and efficiency penalties.

This project focuses on the experimental pilot testing and scaling up of the process at scales in the MW range. A 1,7 MWt calcium looping pilot has been built in the Hunosa 50 MWe CFB coal power plant of “La Pereda”, using a side stream of flue gases of the commercial plant. This flexible pilot is able to operate in a wide range of conditions close to future commercial conditions, providing valuable data for scaling-up purposes. The Pereda pilot plant has been operated for more than 1900 hours with coal combustion in the calciner and working in a dual fluidized bed mode with solids circulating continuously between both reactors. During this period, around 380 hours in CO₂ capture mode have been achieved with capture efficiencies up to 95%.

To support the experimental work developed in the project, one and three dimensional fluid dynamic and process models for both carbonator and calcinator have been developed and validated with experimental data from lab scale plants and the 1,7MWt pilot. These tools have been used to support the conceptual design of a 20MWt calcium looping plant.

Results of the project have been used to develop a technoeconomic feasibility study for the technology which shows very competitive cost of the CO₂ captured with this process, in the range of 20-25 €/tonCO₂ captured, with additional opportunities for a substantial reduction in this figure if some innovations on the basic Calcium looping scheme succeed in the future .

A parallel research program has been developed along the activities linked to the design, commissioning and operation of the 1.7 MWth pilot, including research activities at lab-scale and fundamental knowledge on sorbent properties. Experiments focused on the reactivation of Ca-based sorbents have been conducted in the project. These experiments allowed us to better understand the fundamental knowledge on sorbent properties and helped to propose some simple and inexpensive methods for the enhancement of the reactivity of CaO-based sorbents. This parallel scientific program has been developed by the scientific leaders in the development of this technology worldwide, and has helped with the design of the pilot to a better understanding of the results generated from the pilot.

The scope of this project is a necessary step towards a possible pre-industrial demo plant (10s of MWt scale) in Europe .This is clearly in line with the expected level of development for this technology outlined in the European Technology Platform for Zero Emission Fossil Fuel Power Plants R&D plans and also with wider targets to accelerate the development of breakthrough technologies for CO₂ capture under the EU Strategic Energy Technology Plan

1.2 Summary description of project context and objectives

The aim of the CaOling project was the demonstration of a rapid scaling-up of one of the most promising concepts for CO₂ capture from coal power plants: postcombustion calcium looping systems.

The main example of a full postcombustion calcium looping capture system, and the main focus of work in this project, is represented in Figure 1.

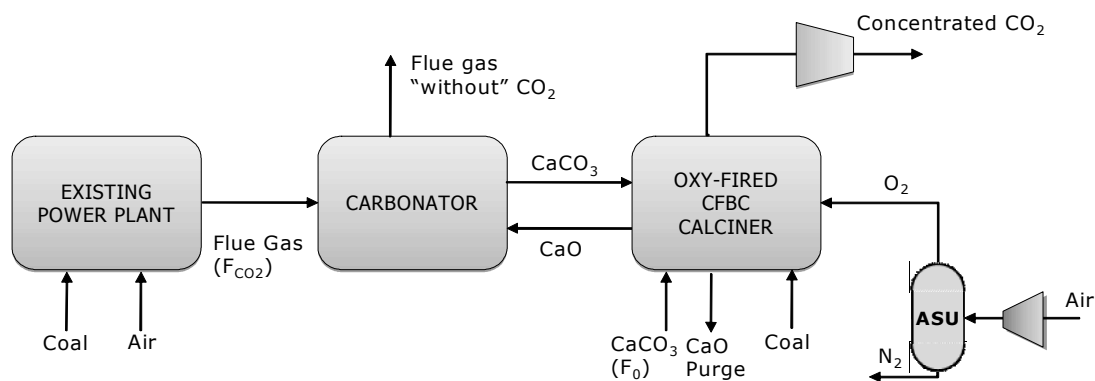


Figure 1. Schematic of a calcium looping system to capture CO₂ from an existing power plant, that generates more power when the oxy-fired circulating fluidized bed combustor acts as calciner.

CO₂ is captured from the combustion flue gas of an existing power plant in a circulating fluidized bed carbonator operating between 600-700°C. The solids leaving the carbonator (with a certain conversion of CaO to CaCO₃) are directed to a second fluidized bed where calcination/regeneration takes place. Coal burns in the calciner in an atmosphere of O₂/CO₂ (oxycombustion) at temperatures above 900°C to produce the heat necessary to calcine the CaCO₃ back to CaO and CO₂. This second fluidized bed calciner operates with oxygen supplied by an air separation unit. The CO₂ captured from the flue gases as CaCO₃, and the CO₂ resulting from the oxy-fired combustion of coal in the calciner, are recovered in concentrated form in the calciner gas, and are suitable for final purification and compression.

The efficiency penalty for the capture of CO₂ is drastically reduced with this concept due to the possibility of producing new extra power from the recovery of the heat generated in both reactors. Furthermore, due to the fact that the purged material from the system is mainly CaO, the integration of the Ca looping cycle with cement manufacture holds the promise of decarbonising both industries at even lower net efficiency penalties.

Calcium looping has a set of important advantages:

- The overall efficiency penalty of the combined process is very low.

- It produces additional power due to the high quality heat sources available.
- Limestone is a cheap and widely available CO₂ sorbent.
- Fuel flexibility both on the existing power plant and in the oxycalciner.
- The flue gas does not require pretreatment, such as SO₂ removal.
- CO₂ produced in the calciner has very high concentration (around 95 vol. % after water condensation).
- Purge material, with its high CaO content, can be used in cement production, reducing the operating costs for the looping cycle and de-carbonising both the power and cement industries.

1.2.1 Objectives

This project focused on the detailed design, building, commissioning and experimental pilot testing of the process at 1,7 MW_t scale. It was also within the scope of this project to take all the necessary step towards the scaling up and construction of a pre-industrial demo plant (20 MW_t scale). This is clearly in line with the expected level of development for this technology outlined in the European Technology Platform for Zero Emission Fossil Fuel Power Plants and also with wider targets to accelerate the development of breakthrough technologies for CO₂ capture under the EU Strategic Energy Technology Plan.

The main objectives of the project were:

- To advance in the experimental validation of the calcium looping cycle (at a 1,7 MW_t pilot size) and demonstrate that this is a low cost, highly energy efficient CO₂ capture technology, suitable for retrofitting coal combustion power plants.
- To gain the necessary design data and experience for rapid scale-up of the technology, building experimentally validated models from the careful interpretation of results produced from lab-scale prototypes and from the experimental campaigns in a 1,7 MW_t test facility.
- To evaluate and optimise the concept in operating conditions equivalent to large-scale industrial units and integrated in a commercial plant.
- To analyse the controllability and stability of the process though the results obtained in a large test campaign in a 1,7 MW_t facility.
- To find the optimum set of operating conditions to minimize sorbent make-up flow cost (calcination temperatures, O₂/CO₂ ratios in the calciner, requirement for steam in the carbonator and calciner, best suitable approach for SO₂ capture, etc).

- To measure the effects of initial particle size and attrition phenomena in the continuous circulating fluidized bed system.
- To analyse the effects of the flue gas impurities in the sorbent performance and the effects of coal ash and sulphur content.
- To improve the average sorbent activity by new reactivation and preactivation approaches.
- To develop fluid dynamic models and reaction models to aid in the understanding and simulation of the process.
- To advance the conceptual designs of alternative and more innovative calcium looping concepts with even lower efficiency penalties and cost.
- To find uses for the purged deactivated material for cement manufacture or flue gas desulphurization.
- To generate a conceptual design for a 20 MWt calcium looping demonstration plant that prepares the next development step of the technology.
- To incorporate the results obtained in the tests into a feasibility study of the process at full-scale application.

1.2.2 Workplan

The work plan was divided in four main closely interrelated conceptual areas:

- Fundamental knowledge. This area aims to improve the basic knowledge of the process (lab-testing (WP1) and sorbent performance analysis (WP2)). It gives very valuable inputs to the design and construction of the 1,7 MWt facility, to the development of the models of WP3, and also to the experimental validation. The final objective of this area has been to improve the knowledge of the process, to obtain a better design of the 1,7 MWt plant and reduce the experimental time needed in this facility to validate the technology.
- Design and construction. The detailed design of the 1,7 MWt facility (WP4) is based on the previous knowledge of the system from previous projects, the initial results obtained in the WP1 (lab testing) and also in the initial works developed in the WP3 (reactor models).
- Validation of the technology: The experimental campaign of WP5 in the 1,7 MWt plant is the core of the work program. Models developed in WP3 and validated with the experimental tests, has improved the analysis of the results and the definition of the required new tests
- Scale-up. Based on the experimental results of WP5 and also in the simulation tools developed and validated in WP3, the program proposes a final activity for preparing the next validation and demonstration step of the technology in a large pilot 20 MWt demo plant.

1.2.3 Innovative impact

The innovative contributions of CaOling Project are:

- It has demonstrated the feasibility at 1,7 MWt size of a CO₂ technology with low efficiency penalty, suitable for the retrofitting of any type of coal combustion power plant.
- The project has researched new techniques for improving sorbent performance, and the use of new sorbents.
- The project has created new simulation tools, and models, that could be used for scale-up of the process towards a large demonstration.
- The interconnection with an existing power plant, and the use of real combustion gases, will help to consolidate the operation procedures under real conditions (start-up, shut down, stable operation...).
- The project has contributed to the detailed design and operation procedures of interconnected fluidised bed reactors that could be used also by other technologies.
- It contribute to develop a technology feasible for other industrial sectors with high CO₂ emissions.

1.3 Description of the main S&T results/foregrounds

WP1: Lab-Scale Pilot testing

The activities in this workpackage were designed to provide experimental results in a wide range of conditions from small scale experimental facilities testing Ca-looping which range from 10 to 75 kW available in Spain (CSIC), Germany (USTUTT-IFK) and Canada (CANMET/U Ottawa). The information obtained in these pilots was the base for the design of the 1.7 MW pilot (WP4) and for the model development exercise for scaling up purposes (WP3).

The work started with a comparative analysis of the methodologies and type of information obtained in the three lab-scale fluidized bed test facilities. The main difference between this facilities is that the CANMET/U Ottawa pilot is in semi-continuous mode while the operation of IFK and INCAR-CSIC rigs can be in continuous solid circulation mode between carbonator and calciner reactors.

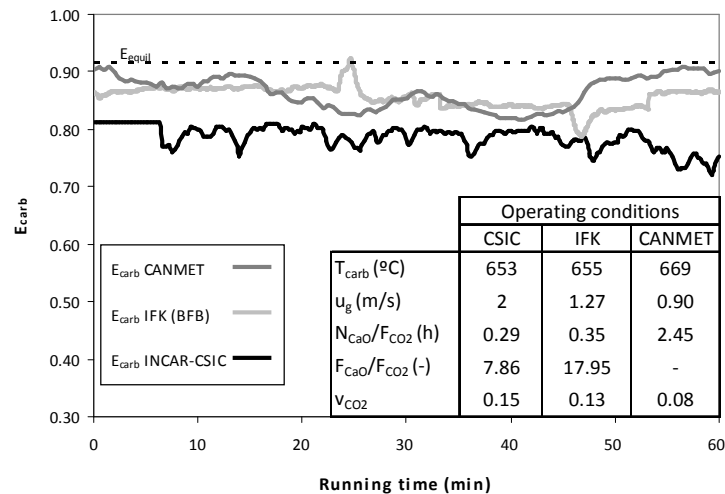


Figure 2. Example of experimental capture efficiencies (E_{carb}) in the three rigs at CSIC, IFK and CANMET for different conditions and reactors set-ups. (T_{carb} : average carbonation temperature; u_g : average gas velocity; N_{CaO}/F_{CO_2} : space time; F_{CaO}/F_{CO_2} : calcium looping ratio between reactors; v_{CO_2} : CO_2 inlet volume fraction). (Rodríguez et al. Energy Procedia 2010)

A more detailed comparison of results obtained in continuous operation using circulating fluidized bed reactors at CSIC and IFK has been possible (see Charitos et al. I&ECR 2011 for more details). These facilities are able to operate under steady-state conditions with a continuous supply of CaO to the carbonator. This allows to validate the experimental information with a closure of the mass balance of the CO_2 that disappears from the gas phase and the $CaCO_3$ formed in the circulating stream of solids. Both installations are able to report reliable CO_2 capture efficiency values measured from the gas phase as the CO_2 disappearing can be measured continuously by gas analyzers. The other term of this CO_2 mass balance can also be calculated by measuring the solid circulation rate between reactors and the increment of carbonate between the inlet and the outlet in the carbonator

from the analysis of the samples taken from the reactors. A good agreement between both terms has been found during most of the steady states analyzed.

Another way to express the quality of the closure of this CO₂ mass balance is to consider the amount of active CaO entering the carbonator and the CO₂ fed to the carbonator. In view of this CO₂ mass balance, a necessary condition in the carbonator is that the active flow of CaO supplied to the carbonator should be greater than the molar flow of CO₂ being captured. This is shown in Figure 3a where the carbonation efficiency is plotted against the ratio between the active flow of CaO and the CO₂ fed to the carbonator. Most of the experimental points lie close to the solid line which represent modes of operation where the active molar flow of CaO entering the carbonator matches the CO₂ fed to the carbonator. However, some experimental points lie on the right on that line. This corresponds to steady states where there is insufficient inventory in the carbonator or where gas solid contacting limitations become significant (e.g. situations where large gas volumes bypass the carbonator bed).

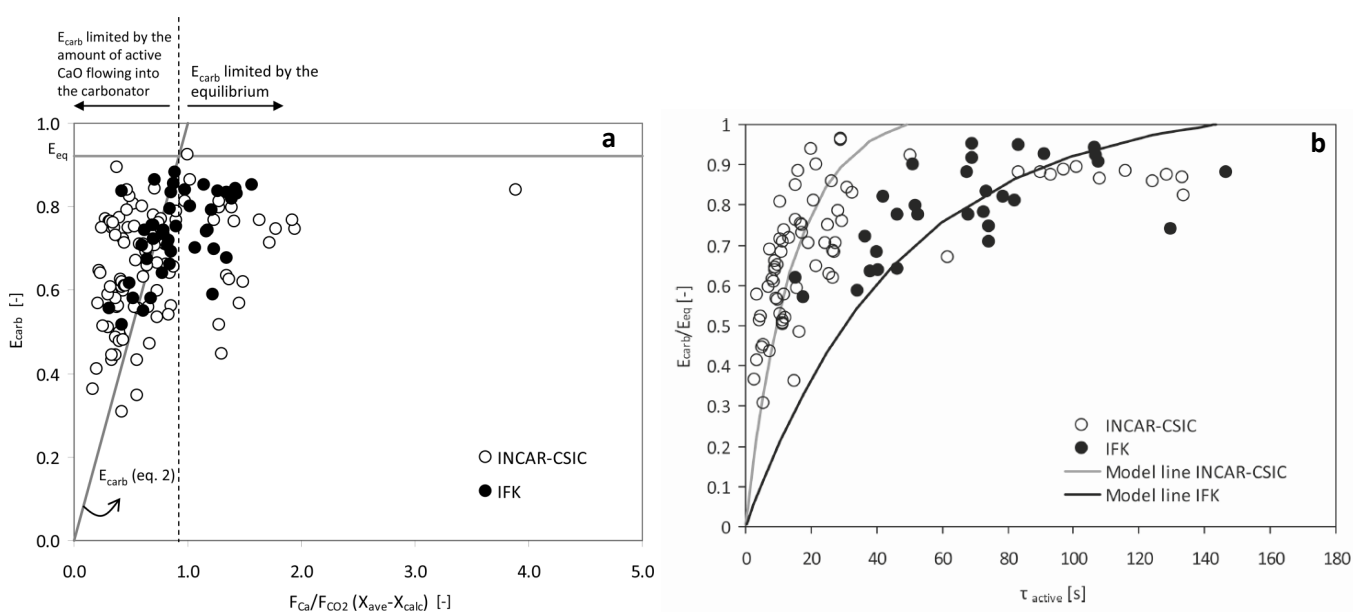


Figure 3. a) The CO₂ capture efficiency vs the active flow of CaO circulating into the carbonator. b) Normalized carbonation efficiency vs active space time (experimental data from CSIC and IFK facilities) (see Charitos et al AICHE 2011 for details)

Another formulation of the CO₂ mass balance more interesting from the design point of view compares the CO₂ disappearing from the gas phase and the CO₂ reacting with the CaO in the bed. This reaction balance takes into account the inventory of CaO per flow of CO₂ in the carbonator, the active fraction of CaO that is reacting in the fast reaction regime and the average reaction rate of these solids in the reactor. To solve it, it has been assumed that the carbonator reactor behaves as an instantaneous and perfect mixing reactor for the solid phase and plug flow for the gas phase. For the reaction rate, it has been considered that the CaO particles attain the maximum conversion at a constant rate and after that the reaction rate becomes zero. This is an oversimplification of the reaction rate at particle level but it is consistent with results reported in the literature and from investigations in WP2. According to this assumption, the fraction of active particles in the carbonator bed corresponds to those that have not yet fully reached their maximum possible conversion. The particle residence time distribution in the carbonator allows to calculate the fraction of particles with a residence time lower than the time needed to increase the carbonate content of the particles entering the carbonator up to the maximum value. In addition, a gas-solid contacting factor has been

included in the reaction rate to take into account the diffusional resistances to the carbonation process. Values close to one has been found showing the good gas-solid contacting of the CFB carbonator

From this analysis a key parameter called active space time (τ_{active}) has been derived which links all the operating parameters in the carbonator with the CO₂ capture efficiency. Figure 3b represents the experimental normalized CO₂ capture efficiencies obtained in the CSIC and IFK facilities against the active space time. As can be seen there is a reasonable correlation between both parameters. The differences of the solids lines corresponding to each installation are mainly due to the different apparent reaction constant rates of the limestones used (0.20s^{-1} for IFK tests and 0.43s^{-1} for CSIC test). From the results shown in Figure 3b, it can be estimated that values of the active space time around 30-90 s are needed to achieve normalized CO₂ capture efficiencies above 0.9.

Another aspect analysed in this workpackage was the evolution of sorbent activity at CSIC and IFK facilities. Many experimental samples have been obtained and analysed from experiments that have been carried out starting with the calcination of an initial batch of limestone. As the experiments advances in time, there is a progressive drop in the average activity of the material due to the increasing average number of carbonation-calcination cycles that particles experience in the system. Due to experimental procedure in the small facilities to calcine the initial batch of limestone, it has been impossible to conduct experiments with highly active CaO. However, the positive results from these tests is that the residual activity attained by the solids as the number of cycles increases is remarkably close to the one measured in TG test for the limestones used in these facilities. This value also remains very stable with the number of cycles and is capable of maintaining a sufficient CO₂ capture level as long as the circulation rate of solids is sufficiently large to compensate for the activity.

On the other hand, the desulfurization capacity of CFB carbonators has been tested at CSIC facility. The experimental procedure for these trials was similar to that used for CO₂ capture tests, but using a simulated flue gas containing SO₂ (see Arias et al. IECR 2013for details). In order to obtain useful information for model validation, most of these tests were carried out under experimental conditions that lead to SO₂ capture efficiencies well below 1, even though in these conditions, CO₂ capture efficiencies were unacceptably low. To analyze SO₂ capture in the carbonator, a similar approach to that used for CO₂ was followed by closing the sulphur mass balances during these tests. In the CSIC facility, there is no addition of fresh limestone to the system. Then, the SO₂ disappearing from the gas phase and reacting irreversibly with CaO results in an increase in CaSO₄ content of the solids circulating in the system. The evolution of the CaSO₄ with time during experiments with a stable inventory of material in the system can be calculated and compared with that measured from samples taken from the reactors at different periods of time. A good agreement was found between the experimental and calculated values indicating the adequate mass balance closure (see Arias et al. IECR 2013 for details).

Regarding the balance between the SO₂ removed from the gas phase and that reacting with the CaO in the carbonator bed, it was assumed that the fraction of partially carbonated CaO does not play an active role in SO₂ capture since it is unlikely that SO₂ will react directly with CaCO₃ to produce CaSO₄ at carbonator conditions. Moreover, the particles that are covered by the layer of CaCO₃ cannot be considered active as the low diffusivity of SO₂ through the carbonate layer at the relatively low temperature of the carbonator would prevent any conversion of the inner CaO inside the particles. On the basis of these assumptions, it is inferred that only those particles that do not reach their maximum CO₂ carrying capacity in the carbonator bed are active for SO₂ retention. Thus, the active inventory of CaO for SO₂ capture is the same as that reacting with CO₂.

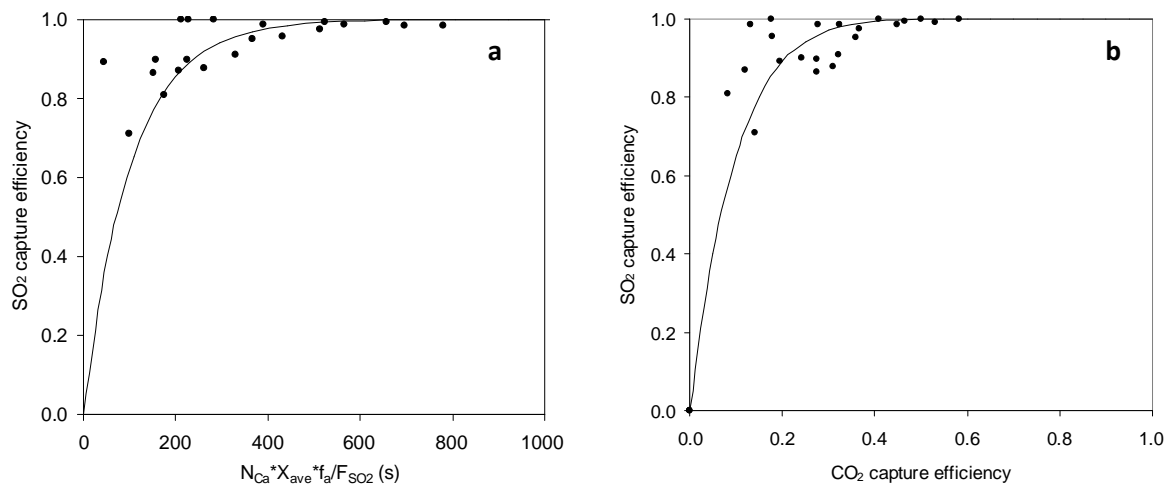


Figure 4. a) Comparison of experimental SO₂ capture efficiency vs. active space time for SO₂. b) Comparison between the experimental results of SO₂ capture efficiency and CO₂ capture efficiency

As for the analysis of the carbonation process, an active space for the SO₂ has been proposed. As can be seen in Figure 4a, the SO₂ capture efficiency correlates adequately with an active space time defined for SO₂ (see Figure 4a). We have also analyze the relationships between CO₂ capture efficiency and SO₂ capture efficiency arising from the assumption that both are reacting with the same active inventory. This can be seen in Figure 4b, where the CO₂ capture efficiency is plotted against SO₂ capture efficiency. These results have shown that when the carbonator is operating under experimental conditions that allow a typical CO₂ capture efficiency of above 0.8, virtually all the SO₂ fed into this reactor is removed from the flue gas. Only when the operating conditions in the carbonator yield a CO₂ capture efficiency that is well below 0.5, will the SO₂ capture efficiency be lower than 1.

Main conclusions

Despite the large differences between the facilities involved in this workpackage, it has been possible to identify conditions in the three rigs at which very high CO₂ capture efficiencies (close to the maximum allowed by the equilibrium) were achieved. Closure of carbon balances between reactors is shown to be satisfactory for conditions achieved at IFK and CSIC with continuous solid circulation and stationary state carbonation and calcination conditions. This closure of the mass balance has been achieved even with specific flow rates of flue gases close to the expected in large scale applications (6-10 molCO₂/m²s).

It can be confirmed that high capture efficiencies are obtained when there is sufficient active CaO in the carbonator per incoming CO₂ flow and this is ensured by a combination of operational parameters. These include the circulation of CaO between the reactors, carbonator inventory and sorbent make-up flow to maintain sorbent activity. The results obtained during SO₂ capture tests show that CFB carbonators are excellent desulfurization units and SO₂ capture efficiencies of above 0.95 can be achieved, even with low inventories of solids in the reacting bed.

The information from this workpackage, together with WP2, has been used to improve models in WP3 and support the definition of the most valuable experimental methodology to be followed during the imminent testing campaign with the 1,7 MWt pilot (WP5) in la Pereda.

WP2: Sorbent properties

Upon carbonation and calcination with a CaO-based sorbent, it has been found that the CO₂ carrying capacity of the sorbent falls with increasing number of cycles. Depletion of the reactivity of the sorbent is influenced by: sorbent sintering that causes an increase of density of sorbent particles; pore closure/loss and reduction of the reacting surface area (Manovic 2008); competing reactions of the sorbent with sulphurous compounds; and ash fouling (Manovic 2009). Approaches to reduce the impact of the loss of sorbent carrying capacity include: thermal preactivation (Sun 2008); water and steam reactivation (Li 2008, Dennis 2009); the production of synthetic sorbents (Salvador 2003, Reddy 2004, Fennell 2007); and doping.

Experiments focused on the reactivation of Ca-based sorbents were conducted at IC, CSIC and Ottawa University. These experiments allowed us to understand the fundamental knowledge on sorbent properties and working on the development of novel, simple and inexpensive methods for the enhancement of the reactivity of CaO-based sorbents.

Doping: Doping can improve the reactivity of natural sorbent materials by reducing the rate of decay of carrying capacity and/or enhancing the residual carrying capacity. Some previous work has related the activation of CaO-based sorbent to the influence of foreign cations in a dopant, in particular the presence of magnesium ions (Megaritis 1998). At Imperial College, experiments were performed in a laboratory scale atmospheric pressure fluidised bed reactor (FBR). Full description of the reactor and the experimental procedure has been described in detail elsewhere (Al-Jeboori 2012).

IC has carried out a systematic investigation to test the effect of a range of inorganic salts MgCl₂, CaCl₂, Mg(NO₃)₂ and the Grignard reagent-isopropylmagnesium chloride dopants on the enhancement of the reactivity of a CaO-based sorbent for CO₂ capture. Our results showed that the anion of the dopant is the key factor for enhancing reactivity for doped samples (Al-Jeboori 2012). IC has also explored the role of a range of halogen mineral acids (HCl, HBr, HI), and also HNO₃ for comparative purposes on the enhancement of the reactivity of a range of limestones. The focus of the research was doping with metal-free dopants, and varying the type and size of the anions. In this work the concentration of the mineral acid required to significantly enhance the residual activity is very low. Cycling experiments showed that 0.167 mole-% HBr, as a dopant improved significantly the carrying capacity of examined limestones. However, the improvement of carrying capacity was related to the type and origin of limestone (Al-Jeboori 2013).

The behaviour of an undoped Longcliffe limestone and the doped Longcliffe with 0.167 mole-% HBr over 50 cycles was studied. The doped Longcliffe with HBr exhibited an increase in residual reactivity from ~ 0.05 to ~ 0.13 as derived from Grasa equation (Grasa 2006), Figure 5.

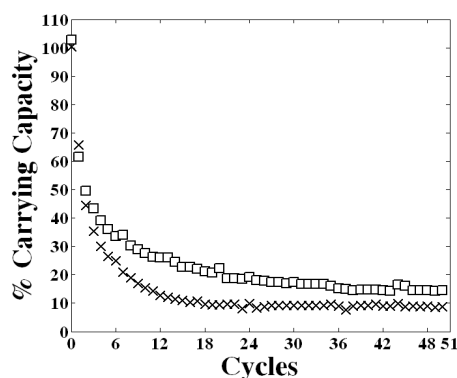


Figure 5. Carrying capacity (normalised) for Longcliffe limestone, plotted against the number of cycles; (x) undoped, (□) 0.167 mol-% HBr.

On other front, IC team has investigated the role of sea water on the enhancement of the reactivity of a CaO-based sorbent for CO₂ capture. This work includes the use of undoped and sea water doped Longcliffe limestone. A series of experiments using various concentrations, from 80 μL to 150 μL, of sea water were conducted. The doping concentrations demonstrating the greatest enhancement for Longcliffe were in the range of 120 – 130 μL of sea water mixed with deionised water (DI), making 2 mL of the total volume.

Addition of 10 % steam upon cycling of doped limestone with the best dopant concentrations, 0.167 mole-% HBr for Schwabian Alb, Cadomin, Compostilla and Purbeck; 130 μL of sea water for Longcliffe, showed a significant improvement in long-term carrying capacity and an additive effects have been noted between doping sorbent with 0.167 mole-%HBr or 130 μL of sea water and the effects of 10% steam in long-term carrying capacity upon cycling (Figure 6).

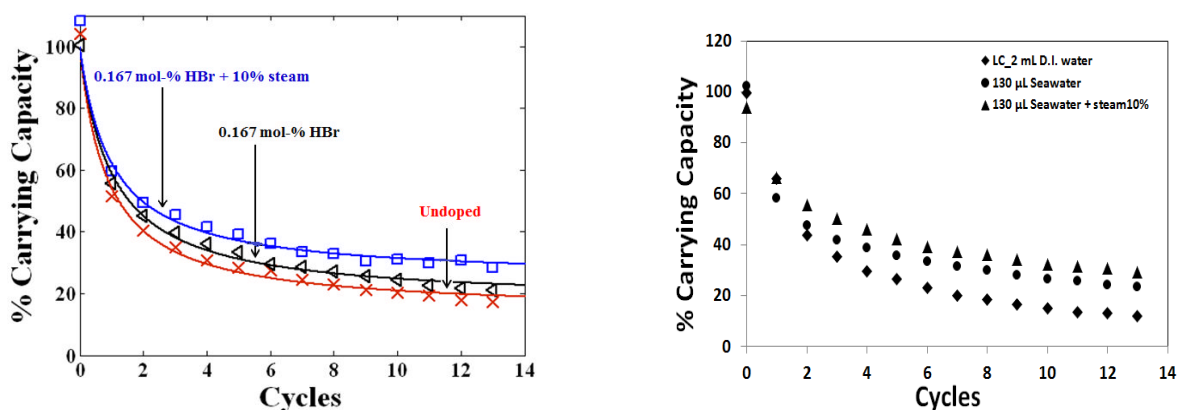


Figure 6. a) Carrying capacity (normalised) for Schwabian Alb limestone, plotted against the limestone number of cycles. b): Carrying capacity (normalised) for Longcliffe, plotted against the number of cycles.

Recarbonation method: The recarbonation method investigated by CSIC aims to maintain the activity of the sorbent by forcing the CaCO₃ particles to increase their conversion above the maximum CO₂ carrying capacity (Arias et al Energy & Environmental Science 2012 for details). This can be carried out in an additional recarbonator reactor working in a continuous mode and treating the stream of solid leaving the carbonator. In this reactor, the operation conditions (high temperature and high CO₂ partial pressure) allow to operate with short reaction times. TGA tests carried out have shown that the addition of a recarbonation step substantially improves the activity of the sorbent, with residual activities almost two times higher after 100 cycles. A conceptual design of

the process has been done by solving energy and mass balances to illustrate the benefits of this process which can reduce greatly the limestones consumption in Ca-looping systems.

Hydration: The work conducted at CSIC related with sorbent hydration has been carried out using a TGA (Martínez et al Energy and Fuels 2011). The effect of the hydration degree on sorbent reactivation has been tested. The results obtained have confirmed that hydration can improve the sorbent the CO₂ carrying capacity. However, the modification of textural properties reduces the mechanical properties of the particles, in agreement with previous findings in the literature. Results obtained have shown that a reasonable strategy to moderate steam consumption in the hydration process would be to hydrate only a fraction of the solids circulating between the reactors in the Ca looping cycles, to the maximum level allowed by the constraint imposed by the mechanical strength of the reactivated material. This strategy has been confirmed in a complementary modeling work carried that analyzes the integration of a reactivation process in a Ca-looping cycle by means of a hydration reactor (Arias et al. Chemical Engineering Journal 2010).

Thermal pre-treatment method: A collaborative work between CSIC and Ottawa University (Arias et al. Energy & Fuels 2011) has analysed the fundamentals of the thermal pre-treatment method and the self-reactivation phenomena detected by CANMET in an earlier works. This phenomenon has been modelled as the result of a dynamic equilibrium between the loss of activity in one cycle and the accumulated gain of activity by extended carbonation times. The proposed model is able to predict reasonably well the evolution CO₂ carrying capacity of pre-treated sorbent with the number of cycles. This study has shown that self-reactivation may not be expected under typical reaction conditions of a circulating fluidized-bed carbonator at atmospheric pressure.

Particle reaction rates: Calcination kinetics with typical cycled solids in CaL systems under rich CO₂ atmospheres has been studied using a TGA by CSIC (see Martínez et al. Energy & Fuels 2012). It has been experimentally observed that the calcination reaction is chemically controlled because internal mass-transfer resistance is negligible up to 300 µm particle size, even in the presence of CO₂ in the reaction atmosphere. In this way, the calcination reaction depends upon the calcination temperature and CO₂ partial pressure, whereas the CaCO₃ content and/or particle lifetime do not affect the reaction rate. A particle reaction model based on a grain model has been used and the kinetic constants of the reaction have been determined. On the basis of the results obtained, it is shown that calcination temperatures between 880 and 920 °C could be sufficient to achieve nearly complete calcination conversion at a typical solids residence time of CFB calciner reactors (2-3 min). The particle reaction model proposed is able to predict the calcination conversion of the limestone.

Effect of steam on the reaction rate: Ottawa University and CSIC have investigated the effect of steam on the reaction rate between CaO and CO₂ (Arias et al I&ECR 2011). This work was based on an early experiments carried out at CANMET/ UNIOTT where they observed large effects of steam on the carbonation reaction under the difussional stage. The effect of steam on carbonation kinetics has been studied using two different limestones under differential conditions in two TG analyzers. Reaction rates have been determined under typical carbonation conditions with and without steam. It has been shown that steam has no influence on the reaction rate constant and similar values were calculated with and without the presence of steam (20% v) in the reacting gas. Values obtained for each limestone are in agreement with those found previously in the literature in absence of steam.

Pelletisation: Research group at Ottawa University conducted a larger scale production of pellets suitable for testing in other pilot plant units. In collaboration with IC, pellets have been tested in their small bubbling bed reactor. Preliminary results do suggest an improvement in attrition behaviour over 20 or more cycles, and more tests will be carried out as part of joint research with Imperial

College. Pellet preparation procedures are now being standardized and about 30 kg of pellets have been prepared for the University of British Columbia, for continuous (as opposed to batch test) pilot plant investigations and a similar amount will be prepared for use with the redesigned Calcium looping unit at CanmetENERGY. Further, the effects of the bromine enhancement method developed by IC (Fennell 2012), in a TGA using pellets has shown a significant impact on the reactivity of the pellets. Experimental data resulted in a publication with IC (Manovic 2013).

Attrition: Experiments at Imperial College supported previous work regarding the positive effect of doping on the attrition of limestone upon cycling. As was proposed earlier, the positive effect of doping on the friability of the sorbent is likely to be revealed in a diminishing mass loss with increasing doping concentration. For Havelock limestone, the mass loss monotonically decreases with increasing doping concentration. A clear tendency can be seen for dopants that the mass loss decreases as soon as the dopant is introduced and is further reduced if a higher amount of dopant is added. For Purbeck however, this trend is not exhibited. In general, a slight increase in mass loss was observed. This could be due to the fact, that the mass loss for this limestone is initially much lower than for undoped Havelock, which loses more than 20 % of its mass after 260 min of cycling. On the other hand, Purbeck only loses 3 % of its mass over the same time period. In comparison, a limestone would lose more than 80 % of its reactivity from sintering over the same period.

At CSIC, the attrition of limestones on post-combustion Ca-looping applications has been studied using the 30 kW test facility of INCAR-CSIC. During extended attrition tests, it was observed that initial calcination particle size was rapidly reduced. However, after this initial period, the resulting material was no longer prone to attrition and its average particle size remains almost constant at value around 90 μm for up to 140 hours of continuous circulation and/or carbonation or calcination.

Effect of Sulphur: IC has studied the effect of $[\text{SO}_2]$ on the reactivity of the limestone in air. This was to test the influence of the presence of SO_2 on the reactivity of the samples in a more realistic condition. Results showed the SO_2 has reduced the carrying capacity of all samples during repeated cycles. Experiments of Longcliffe with HBr doped samples of 0.167 to 0.205 mole-%, in air and SO_2 concentrations from 555 to 1100 ppm have been carried out. HBr doped samples showed a marginally higher carrying capacity at the 13th cycle. The results obtained for carrying capacity over 13 cycles, upon sulphation using 555 ppm, of doped Longcliffe limestone with 0.167 mole-% HBr showed a 12% increase in the carrying capacity compared with the sulphation of undoped limestone. XRF results obtained for doped Longcliffe with HBr after 13 cycles in presence of SO_2 showed no bromine left in sorbent. This could be related to the fact that; at high temperature, SO_3 reacts with the HBr-doped limestone to produce CaSO_4 and releasing Br_2 and also through direct oxidation of HBr to form (volatile) HOBrO_3 . However, in previous work the XRF analyses for cycled samples of doped limestones with HBr acid revealed that the composition of the sorbent remains basically unaltered upon cycling except that there is $\sim 0.045 - 0.06$ mole-% of the bromide (assumed to be present as an anion) for Havelock and Longcliffe and that of Purbeck around 0.05 mole-% left in the sorbent after 13 cycles (Al-Jeboori 2013). Addition of 1-2 % steam upon sulphation, using 555 ppm SO_2 , of Longcliffe limestone showed a marginal improvement in long-term carrying capacity and a marginally additive effects have been noted between doping sorbent with 0.167 mole-% HBr and the effects of 1-2% steam in long-term carrying capacity upon sulphation.

Regarding the effect of SO_2 , CSIC have focused on the investigation of SO_2 capture and sulphation rates under the carbonation conditions (see Arias et al. AIChE 2011). Sulphation tests in TGA have been carried out at 650 $^\circ\text{C}$ using sorbent with different number of cycles. The experimental data obtained during these tests has been interpreted using a random pore model. The results obtained indicate that cycled sorbents (submitted to at least 20 cycles of carbonation and calcination) did not

undergo pore plugging during sulphation even for CaSO_4 conversions over 20%. The high sulphation rates measured with highly cycled (carbonation-calcination) particles seem to indicate that post-combustion Ca-looping carbonator reactors are very effective reactors for capturing SO_2 from flue gases. These results have been confirmed during the experiments carried out in the small facilities (WP1) and in the pilot plant (WP5).

WP3: Reactor modelling and scale-up tools

The goal of workpackage 3 is to develop and apply modelling tools for the calcium looping process. Several modelling approaches have been developed during the CaOling project. Modelling approaches range from 0-D process models to more intrinsic a 1-D dynamic process model and a 3-D steady-state reactor model. Experimental data and experience gained from other workpackages have been exploited in the construction of these models. Vice versa, the models have been utilized in process efficiency analysis, pilot design and experimental interpretation and scale-up studies of the calcium looping process. Developed models provide valuable tools for the future evolution and evaluation of the calcium looping process.

At CSIC the contribution to WP3 has included research in the basic understanding of the process and modelling of fundamental reactions present in the process. From the experimental information obtained in WP2, CSIC has completed modelling work at particle level on SO_2 absorption in the carbonator and a preliminary model for SO_2 absorption in the carbonator reactor has also been defined. In what refers to modelling at particle level, a semi-empirical sulfation reaction model suitable for carbonation reaction conditions (see report on WP2) has been completed. The pore blocking mechanism typically observed during the sulfation reaction of fresh calcined limestones is not observed in the case of highly cycled sorbents ($N > 20$) and the low values of sulfation conversion characteristic of the sorbent in the Ca-looping system. The random pore model (RPM) is able to predict reasonably well the CaO conversion to CaSO_4 taking into account the evolution of the pore structure during the calcination/carbonation cycles. The sulfation proceeds through an initial chemically controlled step followed by second period where chemical reaction and diffusion through the product layer are the controlling resistances. Sulfation has been found to be a first reaction order with respect to SO_2 under the experimental conditions tested. The rate constants for surface reaction (k_s) between $4.32 \cdot 10^{-9}$ and $5.63 \cdot 10^{-9}$ $\text{m}^4/\text{mol s}$ were calculated at 650 °C for the three limestones used. The calculated values of effective product layer diffusivity (D) range from $2.43 \cdot 10^{-12}$ to $4.88 \cdot 10^{-12}$ m^2/s . The intrinsic reaction parameters derived for chemical and diffusion controlled regimes are in agreement with those found in the literature. A publication has been submitted to AIChE J with the main results.

In parallel with the work described in WP1 between IFK and CSIC, a reactor model has been developed in this WP to provide a preliminary interpretation of available results at the small pilot scale units used in the experiments. The model assumes perfect mixing for solids, plug flow for gas and carbonation reaction rates proportional to fraction of active CaO in the bed. All these parameters are conveniently grouped in the parameter of active space time. Two approaches have been utilized in order to find a suitable expression for this parameter. The first (approach A) assumed that all the particles in the carbonator bed react in the fast reaction regime and that their reaction rate is proportional to the difference between their average CO_2 carrying capacity (X_{ave}) and average carbonate content (X_{carb}). The second (approach B) assumed that a fraction of the particles react in the fast reaction regime which is dependent on a simple residence time distribution in the carbonator riser. Approach A has shown to fit only INCAR-CSIC data well, while it has given unacceptable results for the IFK data sets. Main reason for this is the large residence time of solids in the IFK

carbonator, leading to carbonate contents that are very close to the CO_2 carrying capacity of a particle. Approach B fits the data sets of both installations well and is considered more general, since it links the sorbent activity with a particle residence time distribution of particles in the carbonator. Depending on the inlet CO_2 concentration of the flue gas entering the carbonator, the active space time required in order for CO_2 capture efficiency higher than 90 % of the equilibrium value to be achieved is in the range of 30-92 s. The apparent reaction constants have been found to be equal to 0.43 s^{-1} and 0.20 s^{-1} for the INCAR-CSIC and IFK case, respectively. Therefore, the gas-solid contacting factor ϕ has been found to be in the range of 0.8-1.3 for both installations, thus demonstrating the excellent gas-solid contacting of the CFB carbonator reactor, independent of their operating regime, i.e. turbulent fluidization (INCAR-CSIC facility) and fast fluidization (IFK facility). A joint publication (Charitos et al, 2011) has been submitted to Ind. Engng. Chem. Res.

A valuable modelling tool constructed at CSIC is the 0-D mass and energy balance solver for the calcium loop, figure 7. This tool was widely used to design and aid simulations for the 1-D and 3-D models in the work package 3. Basis for the 0-D balance is fully mixed reactors and continuum for mass and energy. Average carrying capacity is solved iteratively from the make-up flow and overall looping ratio. The final result of the balances is the energy required in the calciner.

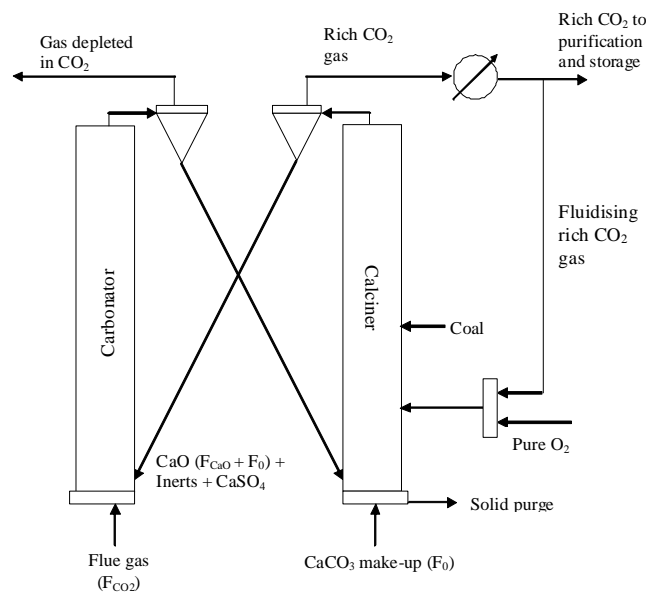


Figure 7. 0-D calcium looping process model balance

Second modelling approach introduced at CSIC is the Aspen Hysys[®] steam cycle model developed for the analysis of overall system efficiency. Optimal integration of the steam cycle to the high quality heat sources of the calcium loop play a key role in the realization of the process. Main findings from the steam cycle integration studies were that with state of the art components, efficiency penalties can be as low as 7.5 percentage points of net efficiency when working with low CaCO_3 make up flows and integrating the CO_2 capture system with a cement industry using the solid purge. Energy penalties are mainly related to the energy consumption of the air separation unit to feed oxygen to the calciner and the CO_2 compressor and auxiliaries. Results of this work were presented at GHGT10 conference in Amsterdam and in AIChE Journal .

Third modelling tool is the 1-D dynamic process model constructed at LUT, figure 8. The model incorporates the main components of the calcium loop to a spatially and time discretized mass and energy balance solver. Model has been created in practise to the Matlab[®] Simulink environment making use of the built-in ordinary differential equation solvers. During the course of the project, the 1-D model has been extended from single reactor carbonator model to an interconnected CFB system with the ability to describe several fluidized bed phenomena present in the calcium looping process. The model has been outfitted to model oxy-combustion in the calciner with flue gas recirculation, calculate sulfation, carbonation and calcination. Research done by project partners CSIC and IFK regarding calcium looping reactivity and solid material carrying capacity has been utilized in the model. Gas-solid interactions have been described by semi-empirical models acquired from literature and retrofitted to the model frame by means of validation. Major solid fractions, CaO, CaCO₃, CaSO₄ and ash, of the calcium looping are calculated enabling solid make-up and purge design. Turbulent mixing and core-annulus effects are taken into account in the solution of the temperature profiles and solid mixing. Energy balance and heat transfer modelling enables pre-design of heat transfer surfaces for the calcium loop. Solid circulation and loop seal modelling can be used to examine the efficiency and economics of the loop. The model has been validated in three occasions: with results from the 30 kW laboratory scale unit, by comparing the 1-D calciner results to 3-D calciner simulation results and finally with experimental data from the 1 MWt pilot. The 1-D dynamic model has also benefitted the scale-up studies done by FWES designing a 20 MWt demonstration unit. Overall description of the 1-D dynamic model frame and calcium looping modelling studies can be found from Ylätaalo et al. 2102 IJGHG 5. 130-135 and Ylätaalo et al. Modelling of the oxy-combustion calciner in the post-combustion calcium looping process. 2012 Fuel.

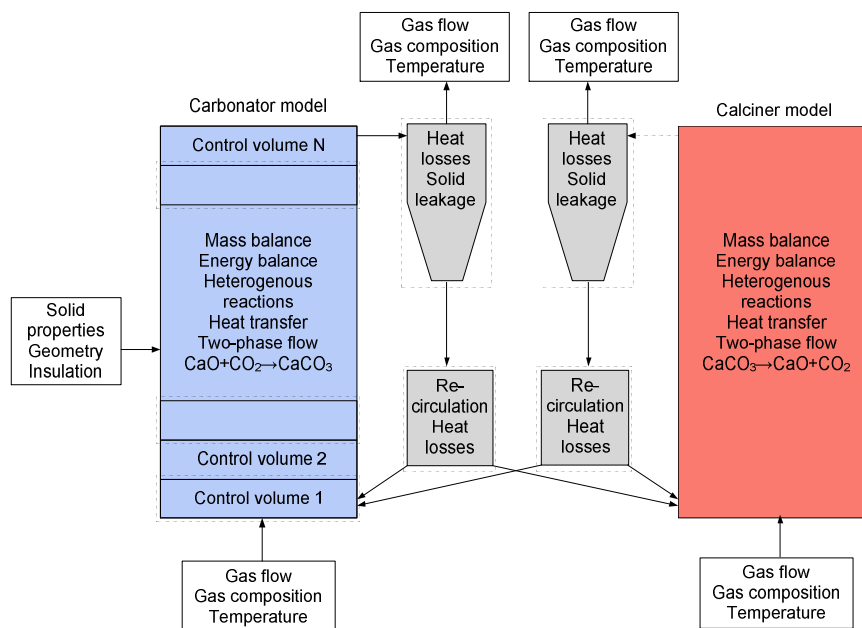


Figure 8. 1-D dynamic model frame

Second modelling approach introduced at LUT is the 3-D steady-state reactor model. CFB3D model has been used to create 3-D calciner model. CFB3D model is a three-dimensional, semi-empirical, steady state model for simulating the combustion, gasification and formation of emissions in CFB processes. The modelling tool can simulate all the essential sub-phenomena: fluid dynamics, reactions (combustion and sorbent), the attrition of particles and heat transfer. More information about the model can be found in Myöhänen K, Hyppänen T. A Three-Dimensional Model Frame for

Modelling Combustion and Gasification in Circulating Fluidized Bed Furnaces. International Journal of Chemical Reactor Engineering 2011; 9: Article A25. The general 3-D model calciner frame is illustrated in figure 9. It includes a 3-D description of the calciner reactor, which is linked to separate sub-models (separators, return legs and possible external heat exchangers). The model combines fundamental balance equations with empirical correlations, which enables practical calculation of full-scale CFB reactor. The sub-models include fluid dynamics of solids and gases, fuel combustion and limestone reactions, comminution of solid materials, homogeneous reactions, heat transfer, sub-models for separators and external heat exchangers and a post-solver for nitrogen oxides. The furnace of the CFB reactor is modelled three-dimensionally by applying a control volume method to discretize and solve the various balance equations in a steady state condition. The calculation mesh is structural with hexahedral calculation cells. The balance equations are solved by the first order upwind differencing scheme and the Gauss-Seidel method with successive overrelaxation. The solved 3-D balance equations include:

- total gas (continuity and momentum),
- total solids,
- fuel reactions and species (moisture, volatiles, char),
- sorbent reactions and species (CaCO_3 , CaO , CaSO_4 , CaS , inert),
- homogeneous reactions and gaseous species (O_2 , CO_2 , H_2O , SO_2 , CO , H_2 , CH_4 , C_2H_4 , C_g , H_2S , NO , N_2O , HCN , NH_3 , Ar , N_2),
- energy (heat transfer within suspension and to surfaces, temperature field).

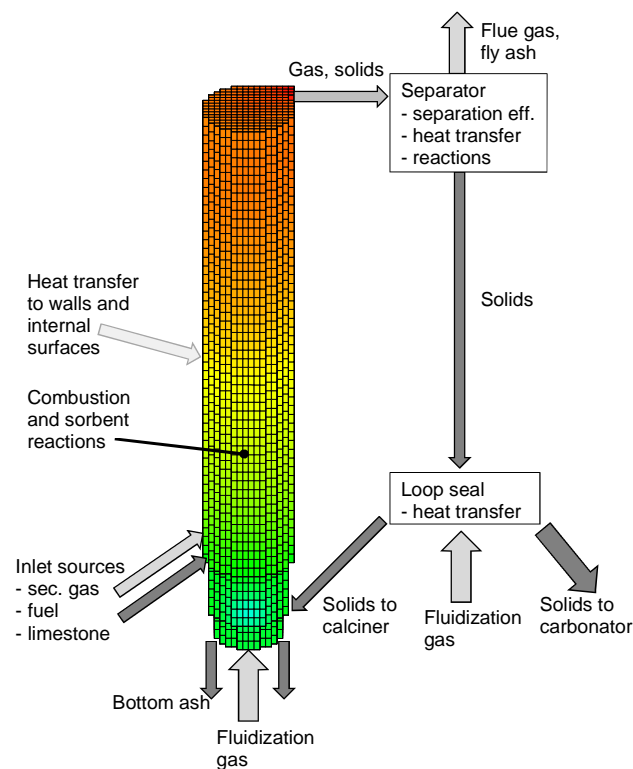


Figure 9. Steady-state calciner model frame.

The 3-D calciner has been successfully implemented to the 1,7 MWt pilot plant scale, compared with the 1-D calciner model with the conclusion that the operation of the pilot is fairly one dimensional. Also some initial scale-up calculations have been done in the 30 MWt demonstration plant scale.

WP4: Plant design and erection

The main objective of this workpackage was the design and construction of the 1,7MWt facility in La Pereda Power plant, the key element of the project. WP partners decided at the beginning of the project to increase the size of the plant and its flexibility, in order to have more representative results in the project. The plant was designed to treat a flue gases flow equivalent to 1,7 MWt, instead of the 1MWt that was originally considered in the project. Additionally the installation has increased the flexibility and is able to operate in a wider range of conditions.

For the basic design, CSIC provided to FWES all available information in WP1 and previous experience running the 30 kW lab test facility, to the detailed design of the 1,7 MWt rig. From that information a preliminary design was defined by FWES. The design was checked with a preliminary model by LUT. So, LUT estimated the basic flow dynamical parameters of the initially FWES designed calcinator and carbonator reactors, including the solid circulation rates, solid fluxes at risers and return leg systems, and operation velocities of cyclones. The solid circulation rate estimations were based on empirical correlations developed for CFB combustors. The determined operation parameters were evaluated based on earlier experience and some modifications were suggested. The parameters of the modified design were then rechecked.

The core of the CO₂ capture pilot plant consists of two interconnected circulating fluidized beds. Typical operation temperatures are around 650 °C in the carbonator and 920 °C in the calciner. Each reactor is equipped with a high efficiency cyclone (cut size 5 µm) and a double loop seal, which can divert the stream of solids coming from the stand pipe to the same reactor (internal circulation) or to the opposite reactor (external circulation). The design of the loop seal allows adjusting the solid flow between reactors. A dedicated high pressure fan is used to fluidize the loop seals and to control the solid circulation rate between reactors. Regarding the dimensions of the reactors, the diameter was fixed to achieve gas velocities similar to those encountered in CFB boilers (3-5 m/s). To ensure enough gas residence time and inventory of solids, the risers were designed with a height of 15 m.

The table below shows main inputs of the plant.

Flue gas flow to carbonator (kg/h)	680-2400
Maximum coal flow to calciner (kg/h)	325
Maximum fresh limestone flow (kg/h)	300
Oxygen flow to calciner (kg/h)	300-600
CO ₂ flow to calciner (kg/h)	700-2250
Air flow to calciner (kg/h)	600-2500

Table 1. Main input of La Pereda CCS plant.

Circulation rates inside the operation window can vary between 5-10 kg/m²s, which are in the typical range of CFB reactors. These solid circulations between reactors allows for effective CO₂ capture with a conservative value of calcium conversion to CaCO₃ in the carbonator.

Figure below shows a layout of the integration of the pilot plant with “La Pereda” existing power plant. A stream of flue gas from “La Pereda” is taken after the electrostatic precipitator and will be sent to the CO₂ capture pilot plant. A fan with an average gas flow of 1400 kg/h is used to increase the pressure of the flue gas before entering the carbonator. The gas leaving the carbonator at high temperature (approximately 650 °C) will be cooled down and returned to main flue gas stream of “La Pereda” power plant before the electrostatic precipitator.

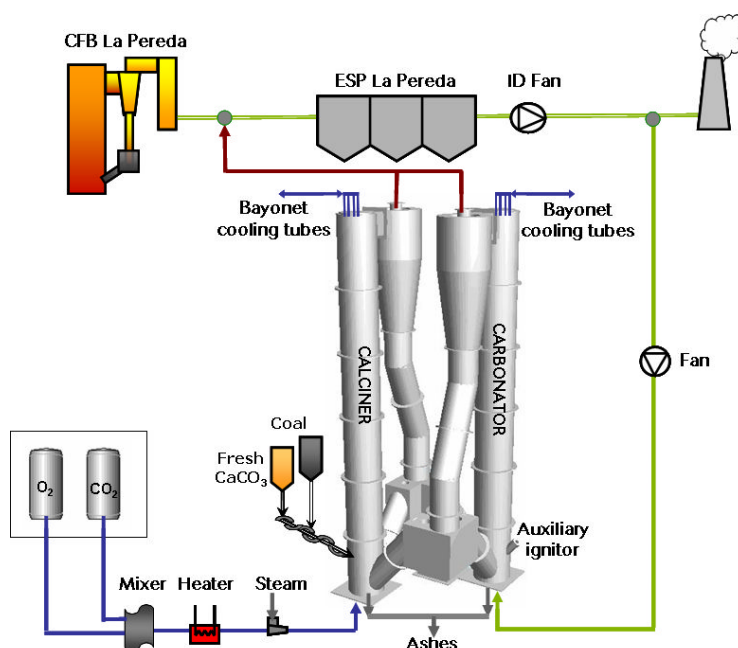


Figure 10. . Schematic configuration of La Pereda CCS plant

The typical temperature of flue gas from the calciner will be higher than 900 °C. No final purification of CO₂ was planned in this pilot. Therefore the gas stream from the calciner is mixed with the decarbonated flue gas leaving the carbonator before being cooled down. The coal feeding system was dimensioned to introduce a maximum flow rate of 325 kg/h. The system can use different types of coals or pre-treated solid fuels. The fuel is discharged in a feed hopper using a big bag handling system. From the feed hopper, coal is pneumatically transported to an intermediate bin. From this bin, coal is discharged to a common hopper where limestone and coal are mixed. A rotary feeder isolates the mixture feeding from furnace overpressure. Finally a screw feeder drives the solids into the calciner bed. As was indicated before, coal can be burned in the calciner using an oxy-fired or air-fired mode. It was decided to supply the O₂ and CO₂ by using tanks of liquefied gases as this is the more cost effective and flexible solution at this scale. The liquid O₂ and CO₂ pass through several atmospheric an electrical and vaporizers. O₂ and CO₂ flows are controlled and blended in a mixer skid. This facility enables mixing to produce a variable O₂ and CO₂ percentage. The temperature of this flow is increased in a gas heater. An additional steam line joins to the flow previous to the calciner inlet. This steam flow comes from “La Pereda” power plant and can be controlled to simulate different compositions of the gas, increasing the operational flexibility of the

plant. In order to work under air-firing mode, a fan will be installed to supply the air needed for the combustion of coal.

Limestone reception is discharged directly from trucks to a feed silo. From this silo, limestone is pneumatically transported to an intermediate bin before being mixed with coal. To remove the solids from the system, two pneumatically operated gates are located under the reactors. The calciner has an automatic purge removal system which consists of a water cooled screw conveyor. This screw conveyor discharges the cold ashes to a chain conveyor that transports the material to a container. The carbonator material purge can only be made manually and after stop the plant.

A water cooling systems using double-pipe cooling tubes is used to control the temperature of the reactors. This system was designed to dissipate the heat coming with the hot stream of solids from the calciner and the heat produced during carbonation reaction. The system consists in four retractable bayonet vertical tubes that permit the control of the reactor's temperature. Cooling water circulates through the bayonets and extracts the heat released within the system. A retractable system permits moving of the bayonet tube and modifies the heat exchange surface, regulating the total heat extracted in the reactor. This system gives a great flexibility and allows operation of the plant in a wide range of conditions.

For start-up sequences, each reactor is equipped with a start-up burner fired with propane. These burners are used to preheat the refractory of the reactors, increasing their temperature up to solid fuel ignition. The propane is supply by a liquid propane tank with all the associated equipment to gasify the fuel. A full skid of valves control the feeding process to the burners.

The instrumentation of the pilot plant was defined attending to control requirements and to obtain information to analyze the experimental results (inventory of solids, solid samples, gas composition, temperature...). There are different ports along the risers, stand pipes and loop seals to measure different parameters inside the reactors and to collect solid samples. Gas composition is analyzed at the exit of each reactor and along the risers.

The figure below shows the plant lay-out with the reactors and all the auxiliary systems:

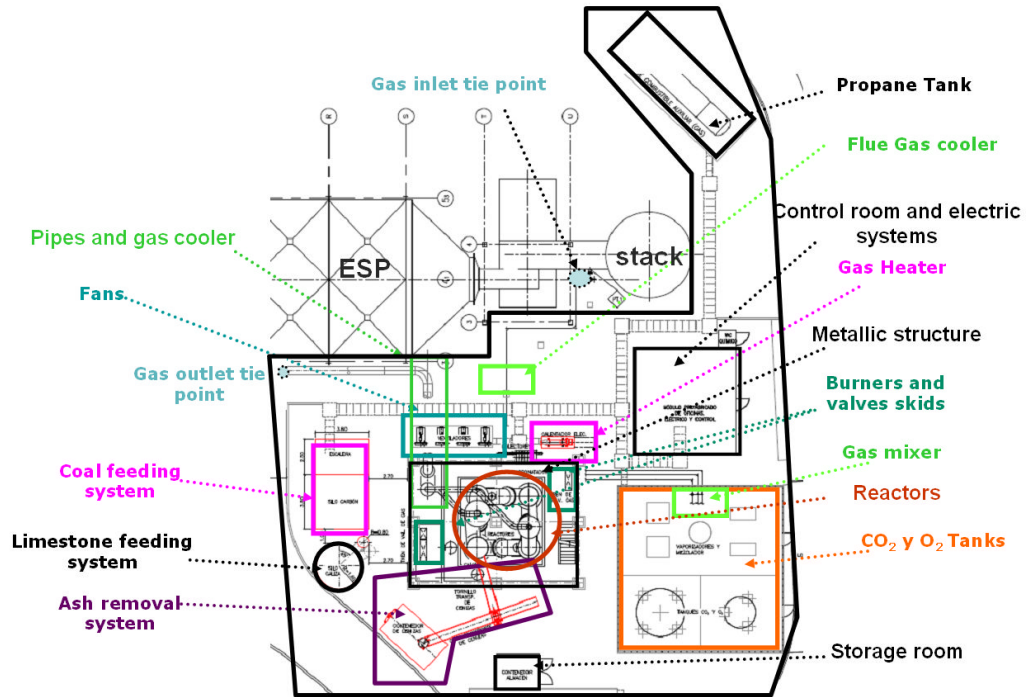


Figure 11. . Layout of La Pereda 1,7MWt plant

Civil works and field preparation started in November 2010. The construction of the plant took practically one year, up to September 2011. The commissioning of the plant was carried out in two phases between October 2011 to January 2012. By the end of January the pilot was prepared to start the operation. The following picture represent the current configuration of La Pereda 1,7 MWt calcium looping pilot.



Figure 12. La Pereda 1,7MWt Ca-looping pilot.

WP5: Experimental campaigns in the 1,7 MW test facility

The activities in this workpackage were aimed to validate the postcombustion Ca-looping technology at 1.7 MWt scale. Since the start up in February 2012, The “La Pereda” pilot plant has been operated for more than 1900 hours with coal combustion in the calciner and working in a dual fluidized bed mode with solids circulating continuously between both reactors. During this period, around 380 hours in CO₂ capture mode have been achieved (with capture efficiencies between 40-95%), including 170 hours with stable oxyfuel combustion of coal in the calciner.

Initial experimental campaigns were focused mainly on the carbonator reactor using the calciner in air combustion mode. The first experimental tests were carried to screen a wide range of conditions in dynamic tests and a number of more stable periods lasting from 1 to 15 hours under stationary state. After this phase, experimental campaigns using oxycombustion in the calciner were carried out at conditions expected in the future commercial application. Long-term tests (up to 80 hours duration) have been performed at oxycombustion conditions. Oxygen concentrations up to 35 %_{vol} have been tested during these experimental campaigns due to some limitations in the gas supply system. Regarding the sorbent behavior, no significant effects of oxy-calcination on sorbent activity have been observed during the experimental conditions tested so far. The main problems encountered during the operation of the pilot plant have been linked to auxiliary equipment malfunctions (due to the unusual scale of the equipment like the solid feeding system or to corrosion in the gas line and blower feeding flue gases to the carbonator).

Table 2 summarizes the main range of operating conditions used during the experimental campaigns carried out. Different carbonator operation conditions have been tested, with different inventories (up to 1000 kg/m²) and different sorbent activity (from fresh calcined limestone to highly cycled solids with CO₂ carrying capacities close to the residual activity). The gas velocity through the carbonator was varied up to values of 5 m/s and the average temperature allowed to change between 600 and 715 °C.

Carbonator temperature (°C)	T _{carb}	600-715
Carbonator superficial gas velocity inlet (m/s)	u _{gas carb in}	2.0-5.0
Inlet CO ₂ volume fraction to the carbonator	V _{CO2 carb in}	0.12-0.14
Inlet SO ₂ concentration to the carbonator (ppm _v)	C _{SO2}	100-250
Inventory of solids in the carbonator (kg m ⁻²)	W _s	100-1000
Maximum CO ₂ carrying capacity of the solids	X _{ave}	0.10-0.70
Calciner temperature (°C)	T _{calc}	820-950 °C
Inlet O ₂ volume fraction to the calciner	V _{O2 calc in}	0.21-0.35
Inlet CO ₂ volume fraction to the calciner	V _{CO2 calc in}	0-0.75
CO ₂ capture efficiency	E _{carb}	0.4-0.95
SO ₂ capture efficiency	E _{sulf}	0.95-1.00

Table 2. Range of operating conditions and the main variables involved during CO₂ capture tests in “la Pereda” 1.7MWth pilot plant.

Concerning the performance of the carbonator reactor of the “La Pereda” pilot plant, the previous experience from small test facilities in continuous mode obtained in WP 1 have been used to define the experimental methodology and to characterize the main variables that affect the performance of the carbonator as a chemical reactor (see Arias et al. International Journal on Greenhouse Gas

Control 2013 for more details). The inventory of solids in the carbonator is determined continuously through the measurement of the pressure difference between the plane above the distributor and the exit of the reactor. The average composition of the solids arriving to the carbonator and their activity towards CO_2 and SO_2 capture is determined frequently from solid sampling using suitable entry and exit ports. Chemical analysis is carried out of the solids samples taken during each experiment and CO_2 carrying capacity and carbonation reaction rates are measured in a TG equipment. The molar flow of CO_2 entering the carbonator with the flue gas as the concentration of CO_2 is measured continuously, as well as the mass flow of flue gas entering the carbonator reactor. The molar circulation rate between carbonator and calciner reactors can be estimated by two methods in parallel: the closure of the carbon balance and the closure of the heat balance in the carbonator. Also, the total solid circulation rate through the risers can also be measured using a suction probe in isokinetic conditions at the exit of the riser.

The analysis of the experimental information obtained has been done by solving the CO_2 mass balances in the carbonator reactor. As in the small facilities, the CO_2 removed from the gas phase is the most reliable term as this can be calculated directly from the continuous measurements of flue gas fed into the carbonator and the gas composition entering and leaving the reactor. To determine CaCO_3 formed in the circulating stream of solids, it is necessary to know the molar flow between reactors and the increment of carbonate conversion of the solid between the inlet and the outlet of the carbonator. Assuming the calciner and carbonator as perfect mixed reactors, the increment of carbonate conversion can be estimated from the composition of the solid taken from the carbonator and calciner.

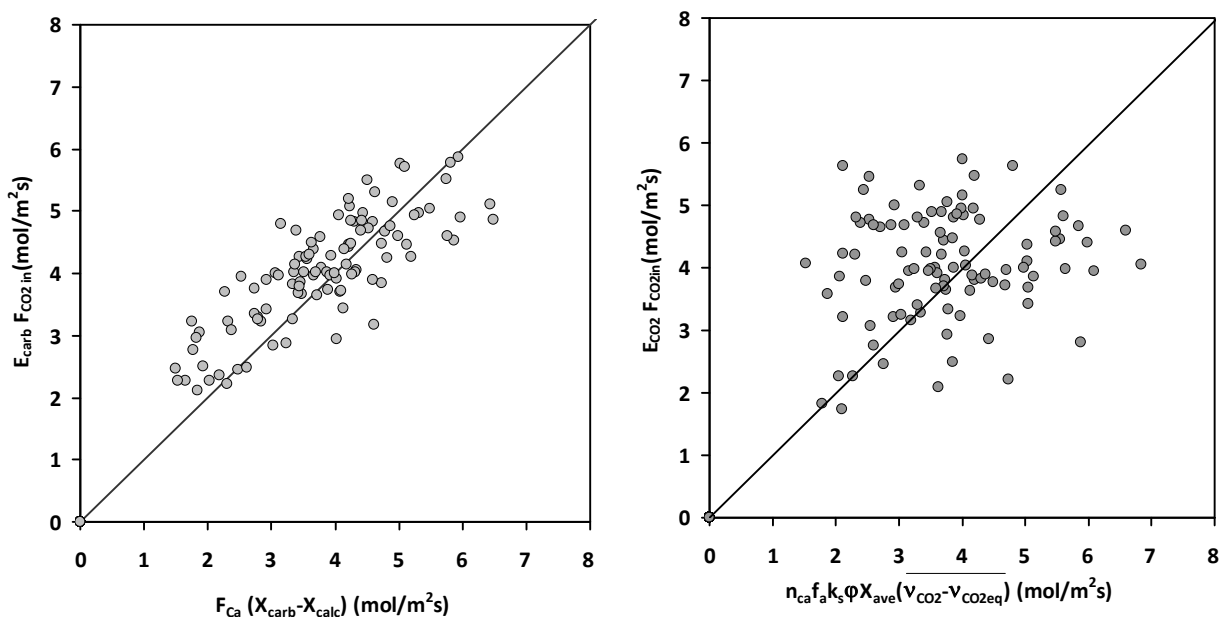


Figure 13. (a) Comparison between the CO_2 removed from the gas phase in the carbonator and the increment in CaCO_3 flow between reactors. (b) Comparison between the CO_2 removed from the gas phase and the CO_2 reacting with CaO in the carbonator bed

The mass flow of solids between reactors can be calculated by solving an energy balance to the carbonator. Since the pilot is refractory lined, heat losses are modest and can be calibrated. All the terms in the heat balance can be estimated continuously from instrumentation available in the pilot plant and the measurements from the heat balance calibrated to correct systematic deviations. Therefore, it is possible to determine during steady state conditions the CaO molar circulation rate of solids arriving to the carbonator from the calciner from the energy balance and the composition of

the samples taken from the reactors. Figure 13a compares the CO₂ removed from the gas phase and CaCO₃ formed in the circulating stream of solids for all stationary states achieved in the operation of the plant in CO₂ capture mode. As can be seen from this figure, there is a good agreement between both values.

The other relevant CO₂ mass balance closure particularly useful for reactor design concerns the comparison of the flow of CO₂ captured from the gas phase and the flow of CO₂ reacting with the CaO particles present in the carbonator bed inventory at any particular time. To close this mass balance, we have applied the same methodology as in WP1 considering the carbonator of the pilot plant as a carbonator as a perfect mixed reactor for the solid phase and a plug flow reactor for the gas phase. For the average reaction rate of the solids, we have also assumed that the particles react at a constant rate until they reach their maximum carbonate conversion (X_{ave}) and after that point the reaction rate is zero. The constant rate (k_s) measured in TGA of samples taken during these tests gives an average value of 0.45 s⁻¹ which is consistent with those found in WP1 and WP2.

Figure 13b shows the final comparison between the CO₂ removed from the gas phase and the CO₂ reacting with CaO in the carbonator bed. As can be seen, there is only a rough closure of this mass balance, which can still be considered reasonable when taking into account the inherent uncertainties in the determination of the parameters of the involved in closure of the CO₂ mass balance. The qualitative similarities in the closure of the CO₂ mass balances represented in Figures 13a and 13b with those reported from a smaller pilot provides confidence about the scalability of these results in what refers to CO₂ capture in the carbonator reactor.

Other aspects have been studied in the pilot plant as the evolution of the sorbent activity and conversion with time and the impact of the irreversible reaction of CaO with SO₂. As example, Figure 14a show the results obtained during an steady state experiment conducted with this purpose during 12 hours without feeding limestone, allowing the sorbent activity to fall towards the residual value. Attrition losses were negligible during this test because the bed inventory was made up with what was left after calcination (in air combustion mode) of the large batch of fresh limestone. Figure 14a shows the evolution with time of the CO₂ carrying capacity of the solid samples extracted from the system during this test against the effective number of cycles. This variable is only an approximation that accounts for the number of times that the moles of CO₂ captured could carbonate the total inventory of Ca up to its average CO₂ carrying capacity. As can be seen in Figure 14a, the drop in the CO₂ carrying capacity observed during this experiment is consistent with the deactivation curve when it is tested in a TGA in absence of SO₂ (shown as a solid black line in Figure 14a). Sulfation conversion of the solids increases with the effective number of cycles, as the total inventory of material in the reactor system (including loop seals and standpipes) progressively reacts with SO₂.

Furthermore, by knowing the amount of Ca in the system and the SO₂ that is capture in the carbonator and calciner, the evolution of the sulphate conversion can be calculated (shown as solid grey line in Figure 14a). Excellent closure of the SO₂ mass balance was achieved, as can be seen from the good agreement between experimental and calculated values. Figure 14a also shows the total sorbent utilization which is the sum of the CO₂ carrying capacity and CaSO₄ molar conversion. As can be seen in this figure, the total sorbent is higher than the expected and after 35 cycles the sorbent. This qualitative trend (effective sorbent utilizations higher than expected) has been confirmed in other similar test and has economic implications for the CaL systems as it will make more feasible to operate the system with low make up flows of fresh limestone.

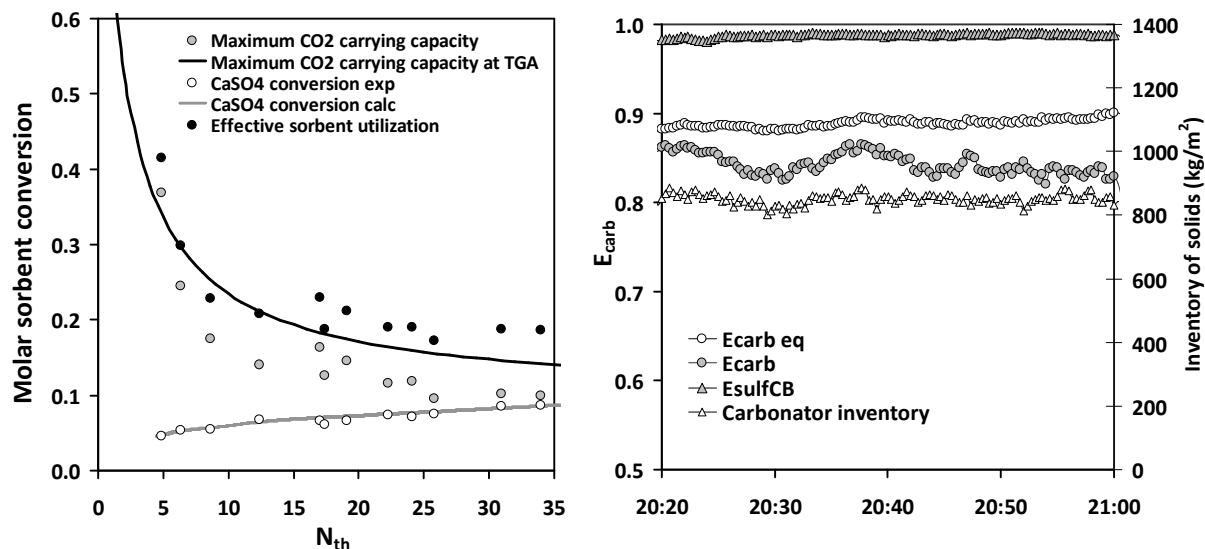


Figure 14.. (a) Evolution of sorbent utilization with the average number of carbonation-calcination cycles of particles in the system and (b) CO₂ capture and SO₂ capture efficiencies when operating with solids with residual CO₂ carrying capacity of Figure 14a.

It is interesting to note that the “deactivated” material circulating between calciner and carbonator is still able to capture CO₂ in the carbonator with an efficiency over 80%, as long as there is sufficient bed inventory in the reactor. This is shown in Figure 14b, that corresponds to an experimental period of 50 minutes at the end of the experimental run of 12 hours without addition of limestone when the CO₂ carrying capacity of the solids present in the system was around 0.10. Figure 14b also represents SO₂ capture efficiency in the carbonator at the same time period. In this figure we can see that SO₂ capture efficiency in the carbonator was kept well above 0.95.

Main conclusions

Several tested campaigns have been carried out in “La Pereda” pilot plant under operating conditions that resemble those expected in large CaL postcombustion CO₂ capture systems in terms of reactor temperatures, gas velocities, solid compositions and circulation rates and reaction atmospheres. The results obtained in a 1.7MW_{th} pilot confirm that postcombustion calcium looping is a promising technology for CO₂ capture that can strongly benefit for scaling up purposed from existing knowledge on mature Circulating Fluidized Bed Combustion Technologies. CO₂ capture efficiencies over 90% have been achieved in a wide range of experimental conditions in the CFB carbonator, including continuous operation using solids with modest CO₂ carrying capacities. Closure of carbon and sulphur balances has been satisfactory during steady state periods lasting for up to 380 hours of accumulated experimental time with CO₂ capture, including 170 hours with the calciner operating under oxy-fuel combustion mode.

A valuable data base of results have been acquired for model validation and scale up purposes, including test conducted in full oxy-combustion mode in the calciner. A basic reactor model has been used to interpret the results obtained. The apparent reaction rates and CO₂ carrying capacities of the materials in the system are in agreement with those found from smaller facilities. The positive experience in the 1.7 MW_{th} La Pereda pilot plant should facilitate the scale up of this new technology and provide the necessary confidence for the demonstration of Ca-looping technology making use of the available expertise in CFB combustion and parallel developments in related oxy-fuel combustion.

WP6: Scaling parameters for industrial plant demo

The aim of this workpackage was on one hand to generate a conceptual design for a 20 MWth calcium looping demonstration plant that prepares the next development step of the technology and on the other to complete a techno-economical analysis of the commercial application of this technology.

The 20 MWth demo plant has been designed for oxygen firing mode and for air firing mode. This last mode shall be used mainly for plant start-up. Regarding the model used to simulate the plant, in the main oxygen firing operating mode for the calciner, the temperature profile of the calciner is kept within the required temperature range. The profile, as expected, is similar to those ones found in a CFB boiler without secondary air injection. The density profile shown, as expected, it is clearly divided in two zones. The first one is the dense phase of the bed and the upper expanded bed zone is the so called freeboard zone. It is in the lower dense bed zone where most of the calcination reaction takes place.

The temperature profile of the carbonator remains rather flat with an average temperature being in agreement to the baseline target of 650 C. The wall layer effects and the high turbulent mixing in the reactor tend to flatten this temperature. On the subject of the pressure/density bed, as in the calciner, the bed is divided in two distinct zones, lower dense zone and expanded zone. The highly turbulent characteristic of this dense bed causes the carbonation reactor to be mostly held within this zone. The superficial velocity in the reactor is within the common range of the fast beds for these applications and solid particle size.

Anyway the height of the reactors will be driven for the need to fit the solids separator, loop seal and INTREX heat exchanger to cool down the solids stream returning from the calciner. The 20MWth plant shall be designed for testing certain features needed for recovering heat from the calciner flue gas and from the solid stream travelling from the calciner to the carbonator. The flue gas coming from the induced draft fan outlet of the power plant is fed to the carbonator through a forced draft fan in order to overcome the pressure of the carbonator fluidized bed.

On the other hand, the calcination of the limestone taking place on the calciner will be done with concentrated oxygen and diluted up to 25 to 40% with flue gas recirculation (FGR). The oxygen will be supplied from a liquid storage tank. The O₂/FGR mixture is preheated up to 250 °C.

In order to control the carbonator temperature, the solid stream coming from the calciner will be cooled and heat will be recovered from a gas-solid heat exchanger (INTREX) in order to superheat steam. This way the carbonator bed temperature can be controlled, together with in-bed heat exchanger tubes inserted in the bed. This heat exchanger is one of the main differences between this demo plant and the pilot plant. Another aspect to explore in this demo plant, it is the controllability of the steam temperature against the control of the carbonator bed temperature. The calciner bed temperature shall be controlled through the coal feed, increasing or decreasing the heat input to the reactor.

The flue gas coming out from the calciner shall be sent to a heat recovery area, made of a convective evaporator and an economizer, where it will be cooled down, and then delivered to a bag filter for particulate removal before being sent to the power plant stack.

Since the main target of the 20 MWth demo plant would be for testing and scaling-up the calcium looping, no heat will be recovered from the carbonator flue gas outlet. The cost of manufacturing and erecting heat exchanging surfaces would not give any compensation since it is not a new technology to explore. The carbonator, almost free of CO₂, at 650 °C shall be cooled down through cold air injection and blended through a mixer. Flue gas is then sent to a filter for particle removal and finally delivered to the stack.

A 3-D model (Figure 15) for the reactors general arrangement has been performed for laying out the different components of the reactors.

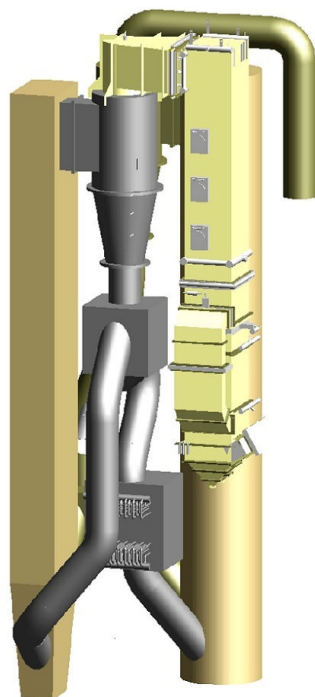


Figure 15. Reactors 3-D model. Isometric view.

The diameter has been selected in order to keep an adequate superficial gas velocity. The height has been selected with the condition of having a residence time of the gas in the reactor very similar or even greater to the one used in the pilot plant. Anyway the height of both reactors is driven mainly by the need of having enough room available to fit the arrangement of the cyclones, the downlegs and the loopseal in the reactors arrangement. The carbonator will be made of carbon steel plated inner insulated with castable refractory.

Regarding the calciner, the cross sectional area of this reactor will be rectangular and the reactor height will be selected to achieve almost full calcination. This reactor will have a dual purpose, as calciner and as steam generator. In the backend of the reactor, after the cyclone, it will be laid out the heat recovery area (HRA) to cool the gas above its acid dew point. The HRA is made of a convective evaporator and an economizer.

An INTREX superheater will be located at the discharge of one of the double loop seal outlets. Solids at high temperature returning from the calciner loop seal will exchange heat with the saturated steam coming from the drum and superheat the steam. The solids discharge of the INTREX after losing heat will return to the carbonator, in such a way that is possible to control the carbonator bed

temperature. INTREX shall have installed spray attenuators for steam temperature control. The INTREX selected for this application is of the upstream type.

The oxidant system will consist in primary oxidant, secondary oxidant and high –pressure oxidant. The primary oxidant will serve for fluidizing the calciner bed through the grid nozzles. Secondary oxidant will be supplied to different nozzles elevations situated in the furnace wall of the calciner. The flow between the nozzle elevations can be regulated through dedicated dampers. A high pressure oxidant system will supply the flow needed for the loop seal of the calciner. The calciner shall get its oxidant either from the primary, secondary and high-pressure oxidant fans in case of air firing mode or from the mixture of O₂ with flue gas recirculated through these same fans from the bag filter outlet. The O₂ will come from an appropriate storage tank given than for this plant size and, that the plant is a test facility, installing an Air Separation Unit (ASU) could not be the most economical choice.

The primary oxidant will delivered to the grid through a primary oxidant centrifugal fan. This fan will be dual purpose. It will be able to operate with air during plant start-up and in air mode, while it will operate with flue gas recirculated from the calciner bag filter outlet. The flue gas will be mixed with the O₂ by means of a primary oxidant static mixer coming from the O₂ storage plant and heated up to 250°C through a steam coil station. The O₂ shall be supplied at the adequate pressure for this purpose. This steam coil station will used saturated steam from the steam drum as heat exchanging fluid.

Like the primary oxidant, the secondary oxidant will conveyed to the secondary oxidant nozzles by the secondary oxidant centrifugal fan. This flue gas shall be mixed with the O₂ delivered from the storage plant using another static mixer for this purpose and heated up to 250 °C through the steam coil station.

The high pressure oxidant needed for the loop seal fluidization shall be forced through the high pressure centrifugal fan. Through a third static mixed high pressure O₂ delivered from a dedicated header of the O₂ storage tank and heated up to 250 °C through the steam coil station.

Air for starting the carbonator and fluidizing the bed made solids circulating between both reactors shall be supplied through a dedicated forced draft flue gas centrifugal fan. This centrifugal fan shall be able to handle air and the flue gas coming from La Pereda Power Plant Once the carbonator reaches a stable operation with temperatures above 600 °C, it can be changed to accept flue gas from the existing power plant.

For fluidizing the carbonator loop seal and the INTREX chamber, a high pressure flue gas fan shall be used. This fan shall be dual purpose, like the other ones, being able to operate with air during start-up and with flue gas once the carbonator reaches 600°C and flue gas can be admitted to the system. In case of air firing, the primary oxidant fan will supply air also to the sight ports, flame detectors and to the sealing system of the coal and limestone feeding system. There will be a flowmeter through a Venturi and individual measurement to each consumer (calciner sight ports, calciner flame detectors and sealing system). In case of oxy firing, the oxidant will be supplied from the oxidant header after O₂ is mixed with the recirculated flue gas in the static mixer. Then it shall be heated in the steam coil station.

In this workpackage a techno-economical study for the commercial application of the calcium looping has been carried out. This analysis has included all the experience gained in the project pilot plant, the biggest pilot in operation with this technology. In order to establish a better comparison

with other technologies and to ensure the consistency of the results, the project has selected the criteria developed in the European Benchmarking Task Force (EBTF), which has been used to evaluate the cost of other CCS technologies in other FP7 projects. In order to establish a realistic base case, the boundary conditions considered for the Calcium Looping unit (CaL) have been quite conservative. The existing plant where the CaL system has been integrated in an advanced ultrasupercritical plant of circa 800 MWe, which is described in the EBTF guidelines. The following figure shows the integration between both units

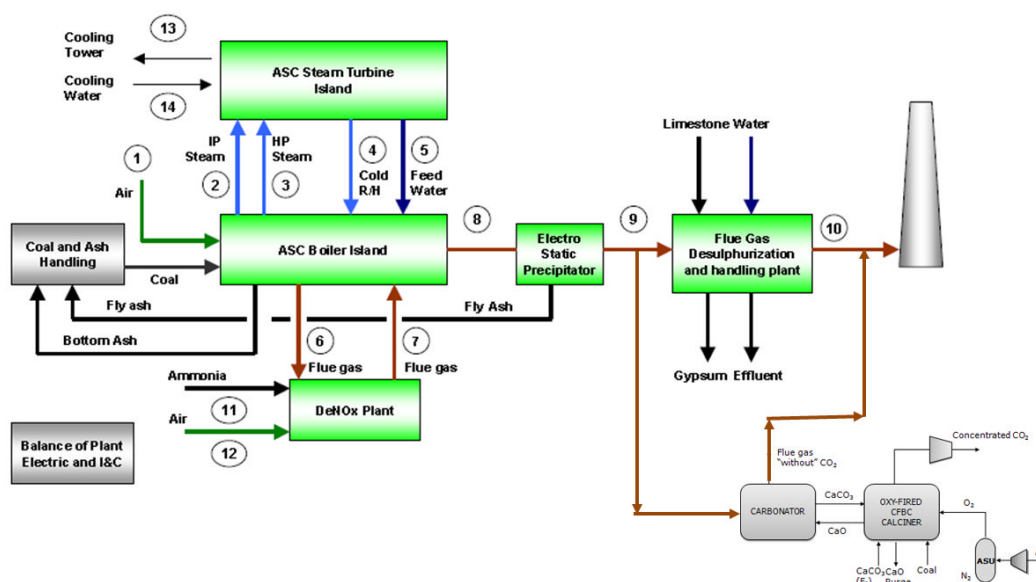


Figure 16. Integration between existing plant and Calcium looping unit

To recover the heat released in the CaL, a supercritical water steam cycle has been considered. Temperatures have been kept to 600°C both main steam and reheated steam, and 300°C for the feedwater temperature. Main Steam pressure was fixed to 300 bar. In order to propose a conservative and simplified approach for the heat recovery, a very low integrated concept for the steam cycle has been considered, trying to wean the operation of the two reactors. Therefore there are options to improve the figures obtained in this study

Results of the study show that calcium looping is a very competitive technology compared. For the selected existing power plant case taken from the EBTF document, an additional extra 416 net MWe can be produced with the heat available in the Calcium looping. This additional power is the main reason of the reduction of the overall costs of the electricity and leads the technology to a very competitive cost compared against other technologies considered in the EBTF document. The cost of the electricity is reduced a 25% compared with the other technologies for the base case and the cost of the CO₂ capture can range between 20-25 €/tonCO₂. A set of improvements have been identified in the project which could lead to even better figures that obtained for the base case.

To complete the techno-economical analysis, a study of possible applications of the purge material from the Calcium looping system has been also carried out. Cement industry, flue gas desulphurization, soil treatment, road construction, agricultural soil amendment, mining and building and construction have been identified as possible fields of application.

1.4 Impact

CaOling project has given a significant step in the development of one of the most promising CO₂ capture technologies, demonstrating the feasibility of the concept at the 1,7MWt scale. This is clearly in line with the expected level of development for this technology outlined in the European Technology Platform for Zero Emission Fossil Fuel Power Plants and also with wider targets to accelerate the development of breakthrough technologies for CO₂ capture under the EU Strategic Energy Technology Plan. The challenging objectives of the European SET-PLAN require a fast development of new technologies that could reduce the high costs of the CO₂ capture technologies. Development of calcium looping, and other second generation capture technologies, could help to fulfil the objectives of the European Union and a broader roll-out of CCS.

One of the major drawbacks of the CCS technologies is the associated impact on the efficiency of the plant where it is implemented. Results obtained in CaOling project support the initial idea that Postcombustion Calcium Looping technology has the potential to achieve a substantial reduction in energy penalties. Comparison of the techno-economical analysis shows that this technology could reduce significantly the energy penalty of other first generation CCS technologies. Additionally there is still interesting opportunities for further developments that could improve these figures even more.

Between the second generation CCS technologies, this is the concept that has experienced the fastest developing pace in the last years, thanks to the results obtained in CaOling and other projects. The most aggressive scale up of the technology has been pursued under CaOling project, that has include the design, commissioning and operation of a 1.7 MWt pilot in la Pereda (Spain) to test the postcombustion Ca-looping concept under a real industrial environment.

The development of this technology is being leaded by European entities. So, besides CaOling project, the biggest initiatives on this technology are also placed in Europe. Therefore, the successful of the implementation of this process at commercial scale could have undoubtedly a notable impact in the European industrial sector.

The innovative contributions of CaOling Proposal are:

- It has demonstrated the feasibility at 1,7 MWt size of a CO₂ technology with low efficiency penalty, suitable for the retrofitting of any type of coal combustion power plant.
- The project has researched new techniques for improving sorbent performance. These techniques are able to reduce the total amount of sorbent consumed in the process. Two of the processes are currently being patented.
- The project has created new simulation tools, and models, that could be used for scale-up of the process towards a large demonstration.
- CaOling project has consolidated the operation procedures of a calcium looping plant under real conditions, with the interconnection with an existing power plant, and the use of real combustion gases.
- The project has contributed to the detailed design and operation procedures of interconnected fluidised bed reactors that could be used also by other technologies such as chemical looping.
- It has contributed to the development of a technology feasible for other industrial sectors with high CO₂ emissions.

The main targets in the design of the 1.7 MWt pilot plant was to be flexible to operate under full continuous mode in terms of solid materials, temperature conditions, solid circulation rates and inventories, superficial gas velocities, real flue gases and oxy-combustion conditions in the calciner. This flexibility gives a particular interest to this pilot, allowing to obtain data of great value for simulating different operating conditions.

To ensure consistency in the benchmarking assessments among other European CCS projects, CaOling has used the criteria developed on the European Benchmarking Task Force (EBTF), for comparing the CO₂ capture costs of this technology. Results show a significant reduction on the Cost of Electricity (COE) and in the cost of the ton of CO₂ captured. Estimation for a conservative case show a reduction of 13% in the COE and a reduction of 35% on the cost of the Ton of CO₂ captured. More integrated concepts with full recovery of the energy available at low temperature and the implementation of improvements already identified in the CaOLing project will improve these figures significantly. Therefore there is a considerable potential for improving the technology from the conservative scheme considered in the analysis. Further R&D efforts in the following years could bring the process to a very competitive situation, being able to be competitive also in cases with low CO₂ credit prices.

There are several doubts concerning technical aspects of the use of some of CO₂ capture technology postcombustion alternatives for retrofitting pulverised coal power plants. Limits imposed by sulphur oxides content in the flue gas translate into extra costs for these technologies. In this context, results obtained in CaOling demonstrate that Calcium looping is one of the best placed CO₂ capture technologies for retrofitting existing coal power plants due to its capability for the capture of both SO_x and CO₂ in the same reactor, with a very low penalty compared with more expensive sorbents. In that sense, the technology is very flexible in respect to fuel considered in the retrofitted plant and process makes this technology suitable for a wide range of coals, including local European coals. Additionally the use of the fluidised bed technology in the calciner opens the option of using different kinds of fuels including biomass in the process. Therefore the development of this technology would contribute to maintaining jobs in coal mining and rural areas, permitting the retrofit of existing power plants. Additionally the development of this concept enables to decrease dependency of external energy supply in EU as various local fuels including co-combustion of biomass can be utilized.

Another advantage of the calcium looping technology is that this technology re-powers the existing site with new clean energy. Besides capturing the CO₂ from the existing power plant, new power generation with CO₂ capture integrated is added to the grid by recovering the heat produced in the calcium looping in a new steam cycle. By contrast, other postcombustion alternatives will reduce the power available in the site considerably. So, the increase of the energy demand of the European countries in the future years could be covered by this extra contribution jointly with the renewable contribution.

A major concern in the calcium looping technology is the generation of large amount of solid waste from the purged sorbent. An initial analysis of industrial reutilization of the purge material has been made in CaOling project, showing a wide range of potential industrial uses that can suit with the composition of the purge. This reutilization will also suppose a decarbonisation of these industries, and a reduction of emissions and auxiliary energy consumption. This will also suppose a direct benefit avoiding the costs of the disposal of the material. Additionally, some of the methods researched in CaOling for improving the sorbent reactivity, can be used to reduce the production of waste material, limiting the amount of residues in case where synergies with other industries cannot be applied.

1.5 Dissemination activities

Dissemination activities

CaOling Consortium has understood the importance of the dissemination activities in the CCS field, and significant efforts have been dedicated to dissemination and exploitation activities during all the project. The results can be considered as very successful. Most of the active entities in this CCS field has visit the pilot and know in detail the results reached during the project. The following figures resume the main dissemination activities carried out and show how the initial objectives regarding dissemination have been widely exceeded:

- 27 publications in high quality scientific journals (5 promised in the proposal).
- 28 presentations in international congresses. (8 promised in the proposal).
- Dissemination in other European forums (EUROELECTRIC, ZEP...)
- More than 10 open visits to the pilot plant, covering institutions, companies and public authorities. (1 promised in the proposal).
- Dedicated workshop for presenting results of the project.
- More than 20 appearances in General Media.
- Website

Deliverable D7.6 ‘Plan for Dissemination of Knowledge and Exploitation of Project Results’ was prepared in the project with the aim to organise the dissemination and exploitation activities. This document compiles the main activities that integrate the plan on dissemination of knowledge and exploitation of results fields. The main objective of this Dissemination Plan is to identify and organise the activities to be performed in order to promote the commercial exploitation of the project’s results and to generate the widest dissemination of knowledge from the project. The document also presents the main communication channels, internal and external, that would be used for the development of these activities and all the possible interested stakeholders.

With the aim to improve the dissemination results, the roles of each of the partner in the CaOling Consortium were distributed. CaOling is formed by 8 partners from 5 different countries with an important presence of the Spanish partners, the country where the calcium looping pilot plant is located. The Consortium includes the presence of the industry, a manufacture company and academy which brings to different roles in the dissemination activities and increases the opportunities for disseminating the results in different forums.

CaOling project has its own website www.caoling.eu which has been used as a basic tool for promoting the dissemination activities and as a repository of information and public results generated in the project. The high number of contacts made via the website confirms the usefulness of the website and demonstrates the interest that the project has created. Endesa as coordinator has been in charge of keeping the information updated and has been the contact point of the project. The main information contained in the website includes

- Information about CaOling project and its activities including contact details, participants, main objectives, brochure, background information, a plant description, events (workshops, conferences) etc.
- Results obtained during the project, including all the scientific papers and presentations of conferences.
- Photos and videos of the pilot installation as the reference of the project.
- Frequent news and updates on the public material to keep the community informed.

CaOling dissemination activities have not only covered the scientific and industry field but also have focused on other targeted audiences such as politics and general public, both groups with a crucial importance in the success of the future deployment of the technology. In that sense different activities have been carried out in order to cover all the interested groups. Communications channels used include:

Communication channel	Targeted audience
Website	All groups
Conferences	Scientific groups and industry
Printed publications	Scientific groups, industry, public authorities
Scientific publications	Scientific groups and industry
Workshop	Scientific groups industry, public authorities
Visit to the plant	Scientific groups, industry, public authorities
General Media	General public

Table 3. Communications channels used in the development of the dissemination actions of CaOling project.

A Workshop to show the main results of CaOling project was held on April 19th, 2012 at INCAR-CSIC headquarters in Oviedo with great success. During the afternoon a visit to La Pereda calcium looping CO₂ capture plant was organised. Seventy-eight people of 37 different entities participated in the event. 14 countries were represented. The participants were equally distributed between research and industrial entities. Industry participants include members of the sectors of electricity, cement production and lime production, covering a wide spectrum of interest.



Figure 17. CaOling workshop in La Pereda.

The main objective of the workshop was to review and present the most important results obtained during the first 28 months of the CaOling project. Deliverable 7.4 “CaOling workshop and open day event – Proceedings” was edited. This deliverable includes more information regarding the workshop main topics and results. This public deliverable is available in the website of the project (www.caoling.eu).

The high number of scientific publications and international conferences confirm the relevance of the results obtained in the project. A complete list of all the publications edited in the project is included in chapter 2 of this document.

Regarding dissemination of results to public authorities, several events have been made in order to increase the awareness of the technology and in order to get the future support of the authorities for the deployment of CCS technologies. As a representative item, the pilot plant of La Pereda was presented to a group composed by 11 of the Spanish Energy Regional Directors who visit the plant on November 16th 2012.



Figure 18. Visit of Regional Energy Directors to La Pereda Pilot Plant

The project presentation to the regional and national press in Asturias, Spain, took place on the 28th of December 2009 at INCAR-CSIC (Oviedo). The President of CSIC, the President of HUNOSA, the Director General of ENDESA General and the Regional Vicepresident of Foster Wheeler attended to the meeting. Articles covering the event were published in the main regional newspapers, and radio and TV interviews also took place related to the project (including a short presentation in TVE, the public Spanish television station).

On October 28th 2010 an event was organized in La Pereda power plant by HUNOSA coinciding with the start of the construction of the pilot plant. The President of the Asturias region, the main unions, and other local authorities jointly with the general managers and directors of the companies involved in the construction attended to the event. It had an extensive coverage in newspapers and local radio and television, contributing to the public awareness and acceptance of the project.

A summary of the project was included in the EUROELECTRIC “Energy Wisdom Programme 2012-2013” where some of the most significant projects related to carbon emissions reduction led by European utilities are presented. A summary document was edited and can be downloaded in the EUROELECTRIC Webpage: <http://www.eurelectric.org/EWP> . The Programme was also presented in a meeting of EUROELECTRIC with presence of Members of the European Parliament.

During all the project a continuous follow-up of the status of the testing has been published in regional papers. This frequent presence in the press media has improved the public awareness on the initiative and the CCS technology.



Figure 19. Some references of news published in press media regarding CaOling project

Exploitation activities

The consortium has a structure that guarantees itself the exploitation of the results, as contains manufacturers and end users additionally to research centers. Therefore results could be directly exploited by the partners of the Consortium. The experience of the partners in energy production related with CFB technology makes the consortium powerful for bringing this concept to a pre-industrial state.

Two patents applications in the area of improvement of sorbent activity have been produced as result of the project execution. There is a project in development for the validation of one of these patents in a pilot scale size.

Additionally to the patents, a very valuable set of experimental information at lab and pilot scale have been obtained. Moreover different simulation tools have been created in the project and have been validated with the experimental data. All this information will be fundamental for the future scale-up and exploitation of the technology. Operational experience of the 1,7 Mwt pilot constructed in the project is also a very important output for the project, and will contribute to a clear definition of the following phases.

A result that will be directly exploited in a second development phase will be the conceptual design of the demo plant of 20MWt, developed in the WP6. This work will be the first step of the continuation of the scaling-up of the technology, and would suppose the basis for the engineering needed for the demo plant.

Besides what has been highlighted before, several meetings have been held with parties interested in the development of the technology, which could open additional future exploitation alternatives. During 2012 and the beginning of 2013 at least six interested entities including big industrial companies, establish contact with the project in order to know more about the project results, the facility and future continuation plans. Most of the meetings have included a visit to the pilot facility of La Pereda.

1.6 The address of the project public website and contact details

Webpage:

<http://www.caoling.eu>

Contact:

Mr. Andrés Sánchez-Biezma Sacristán, ENDESA Generación, Coordinator

andres.sanchez@endesa.es

Tel. +34 91 213 1145