CO_2 CAPTURE IN THE CEMENT INDUSTRY

Technical Study

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IEA GHG OVERVIEW

Background

The cement industry is one of the world's largest industrial sources of CO_2 emissions, accounting for 1.8 Gt/y in 2005, i.e. more than 6% of global emissions from the use of fossil fuels. Over the years the cement industry has substantially reduced emissions of CO_2 per tonne of cement by improved energy efficiency, replacing fossil fuels with wastes which can sometimes be regarded as 'carbon neutral' and by increasing the use of additives in the cement product. The scope for further reductions by these means is becoming limited but there is an increasing need to reduce CO_2 emissions to avoid climate change. CO_2 capture and storage (CCS) presents one of the few opportunities to make further major reductions in emissions and the industry is currently considering the feasibility of applying this technique in order to plan for the future. In many ways the cement industry represents a good opportunity for CCS, because cement plants are relatively large point sources of CO_2, the CO_2 concentration in cement plant flue gas is relatively high (about 25mol%, dry basis) and over 60% of total CO_2 emissions from a modern cement plant are from mineral decomposition and this CO_2 cannot be avoided by use of alternative energy sources.

IEA GHG has undertaken a study to assess the technologies that could be used to capture CO_2 in cement plants and their performances and costs. The study was undertaken for IEA GHG by Mott MacDonald. The British Cement Association collaborated and helped to obtain input from the cement manufacture and plant supply industries.

Study Description

Scope of the study
The scope of the study was to:

- Provide descriptions of cement plants and the global cement industry
- Review CO_2 capture processes that would be suitable for cement plants
- Evaluate the performance and economics of cement plants with and without CO_2 capture
- Discuss retrofitting CO_2 capture and CO_2 capture ready plants
- Identify information gaps and R&D needs

Study basis
The technical and economic assessments were based on a new cement plant in the UK producing 1 million tonnes/year of cement (910,000 t/y of clinker), using typical current technology: a dry feed plant with 5 stages of pre-heating. A 3 million t/y plant in Asia using the same technology was assessed as a sensitivity case.

The study was based on existing CO_2 capture technologies, or technologies that could be developed for use in cement plants in the near future with moderate risk. It should be recognised that further development of technologies may significantly reduce the cost per tonne of CO_2 and increase the fraction of CO_2 that can be captured.

The costs of cement production and CO_2 capture were calculated assuming a 10% annual discount rate in constant money values, a 25 year plant life, 90% load factor, a coal price of €2.51/GJ (LHV basis) and a petroleum coke price of €2.34/GJ (3.76 and 3.51 US$/GJ respectively[1]). A full list of the economic criteria used in the study is given in the main report.

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[1] Based on an exchange rate of 1.5 $/€

Results and Discussion

Cement production

Cement is composed of calcium silicates, calcium aluminates and calcium aluminoferrite. It is produced from a mixture of raw materials, principally calcium carbonate. The raw materials are ground and delivered to a kiln where they react at high temperature to form an intermediate ‘clinker’ which is ground together with some gypsum to form cement, the finished product. The various clinker production technologies are described in the main report. Historically clinker production has evolved from wet processes in which raw materials are ground in water, through semi-wet and semi-dry processes to dry processes, in which the raw materials are dried and ground before feeding to preheaters and a kiln. The dry process required less energy than the wet process and thus is generally favoured where its application is feasible. In industrialised countries rotary kilns, which can have capacities of 10,000t/d, are used. Vertical shaft kilns are also used in plants of less than 300t/d, now mainly in India and China, but their use is being phased out because they produce lower quality cement.

Global production of cement has grown steadily over many years, reaching about 2,500 million tonnes in 2006. The average growth rate was 5.5% per year during the period 1990 to 2006 and the average rate accelerated to 8.7% per year since 2002. The main growth has been in Asia, and China in particular. China now accounts for almost half of all global cement production.

CO_2 capture technologies

The three main CO_2 capture technologies are:

- Post combustion capture, in which CO_2 is separated from flue gas.
- Pre-combustion capture, in which fuel is reacted with oxygen and steam to produce a mixture of CO_2 and H_2, the CO_2 is separated and the H_2 is used as fuel.
- Oxy-combustion, in which fuel is burnt in oxygen instead of air, to produce a flue gas consisting mainly of CO_2.

Pre-combustion capture was not evaluated in detail in this study, mainly because it would only be able to capture the fuel-derived CO_2, not the larger quantity of CO_2 from decomposition of carbonate minerals.

Post combustion capture

Post combustion capture is a downstream process which would not affect the core of the cement production process. The main additions to the plant would be:

- A CO_2 capture plant which includes a solvent scrubber and regenerator
- A compressor to increase the pressure of the CO_2 product for transport by pipeline
- High efficiency FGD and De-NOx to satisfy the flue gas purity requirements of the CO_2 capture process
- A plant to provide the steam required for regeneration of the CO_2 capture solvent.

Cement plant flue gas has a relatively high CO_2 concentration; typically about 25 mol% compared to about 14% for a coal fired power plant. The post combustion solvent scrubbing processes that are being developed for CO_2 capture in coal fired power plants would in principle be suitable for use in cement plants. This study is based on the use of monoethanolamine (MEA) solvent scrubbing, which is a conservative choice. Alternative proprietary solvents with lower steam consumptions are being developed and used, so the sensitivity of costs to a lower energy consumption solvent was assessed. Post-combustion amine scrubbing is widely used in small plants producing up to 400t/d of CO_2 for the food, drinks and chemicals industries but scale up to around 3,000t/d would be needed for a 1 Mt/y cement plant which included capture of the CO_2 produced from an on-site steam generation plant.

Post combustion capture produces a high purity CO_2 stream, typically 99.9% dry basis. In this study the CO_2 is compressed on-site to 11 MPa to enable it to be transported by pipeline to an underground storage site. It was assumed that 85% of the CO_2 would be captured, for consistency with IEA GHG’s other studies but information obtained during IEA GHG's study on post combustion capture at power plants and other published work indicates that 95% capture should be feasible without significantly affecting the cost per tonne of CO_2 captured. The CO_2 concentration in cement plant flue gas is higher than in power plant flue gas, so similar or greater percentage CO_2 capture is expected to be feasible.

A major issue for post combustion capture is the supply of low pressure steam for CO_2 capture solvent regeneration. In this study the steam is provided by a coal-fired combined heat and power (CHP) plant, which includes high pressure steam generation and a back-pressure steam turbine. The electricity produced by the CHP plant can satisfy all of the electricity demand of the cement and CO_2 capture plants and there is a small surplus which is exported. Coal was selected as the CHP plant fuel because most, although not all, cement plants already use coal. The flue gases from the CHP and cement plants are combined and fed to the CO_2 capture plant.

Oxy-combustion

In an oxy-combustion plant fuel is combusted in oxygen, produced in a cryogenic air separation unit, and some CO_2-rich flue gas is recycled to control the flame temperature. The flue gas with a CO_2 concentration of about 80 mol% (dry basis) is purified to 95% CO_2 in a relatively simple cryogenic separation unit during compression. Higher purities can be achieved if necessary by employing cryogenic distillation.

Fuel is fed to two places in a modern cement plant: the precalciner, which helps to preheat the feedstock and calcine the raw material, and the high temperature kiln where cement clinker is produced. Most of the CO_2 is normally released from the limestone raw material in the preheaters and precalciner. Four oxy-combustion flow schemes were considered in the first phase of this study. The scheme that was selected for more detailed study involves oxy-combustion of the precalciner but air combustion of the kiln. This scheme minimises the possible impact of a high-CO_2 atmosphere on the clinker production processes which occur in the kiln and minimises the impact of air in-leakage, which is substantial in some parts of a cement plant and which would have a large impact on the CO_2 concentration and the losses of CO_2 during purification. Further R&D may show that cement kilns could be successfully operated with a high CO_2 atmosphere and in-leakage could be greatly reduced. If so, oxy-combustion of the kiln as well as the precalciner could be feasible.

There is some experience of oxygen enrichment in cement plants to improve plant throughput but not for CO_2 abatement and not at high levels of enrichment. Oxy-combustion power generation pilot plants are being built but oxy-combustion is at an earlier stage of development than post combustion capture.

Performance and costs of cement plants with CO_2 capture

The estimated performances and costs of cement plants with and without CO_2 capture are summarised in Tables 1 and 2.

Table 1 Plant performance

	Unit	Base case (no capture)	Post combustion capture	Oxy-combustion
Fuel and power				
Coal feed	kt/y	63.3	291.6	72.1
Petroleum coke feed	kt/y	32.9	32.9	27.1
Total fuel consumption (LHV basis)	MW	96.8	304.0	97.8
Average power consumption	MW	10.2	42.1	22.7
Average on-site power generation	MW	-	45.0	0.7
Average net power consumption	MW	10.2	-2.9	22.0
CO_2 emitted and captured				
CO_2 captured	kt/y	-	1067.7	465.0
CO_2 emitted on-site	kt/y	728.4	188.4	282.9
CO_2 emissions avoided at the cement plant[2]	kt/y	-	540.0	445.6
	%	-	74	61
CO_2 associated with power import/export	kt/y	42.0	-11.8	90.8
Overall net CO_2 emissions	kt/y	770.4	176.6	373.7
CO_2 emissions avoided, including power import and export	kt/y	-	593.8	396.8
	%	-	77	52

Table 2 Costs

	Unit	Base case (no capture)	Post combustion capture	Oxy-combustion
Capital cost[3]	€M	263	558	327
Operating costs				
Fuel	€M/y	6.7	21.5	6.9
Power	€M/y	4.0	-1.1	8.7
Other variable operating costs	€M/y	6.1	10.6	6.4
Fixed operating costs	€M/y	19.1	35.3	22.8
Capital charges	€M/y	29.7	63.1	36.9
Total costs	€M/y	65.6	129.4	81.6
Cement production cost	€/t	65.6	129.4	81.6
CO_2 abatement costs				
Cost per tonne of cement product	€/t	-	63.8	16.0
Cost per tonne of CO_2 captured	€/t	-	59.6	34.3
Cost per tonne of CO_2 emissions avoided[4]	€/t	-	107.4	40.2

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[2] The CO_2 emissions avoided are the emissions of the base case plant without capture minus the emissions of the plant with CO_2 capture.

[3] The capital costs include miscellaneous owners' costs but exclude interest during construction, although this is taken into account in the calculation of overall production costs.

[4] The costs per tonne of CO_2 emissions avoided take into account the emissions associated with imported and exported power.

Post combustion capture
The CO_2 emissions avoidance at the cement plant site is 74% but taking into account CO_2 avoided because of electricity exports the emissions avoidance increases to 77%. This could be increased to 93% if the percentage capture was increased to 95%.

The cost of CO_2 emissions avoidance is high at €107/t (US$161/t)[1]. The high concentration of CO_2 in the flue gas of a cement plant compared to that of a coal or gas fired power plant reduces the size of the absorber tower and associated ducts, fans etc. However this is more than offset by lower economies of scale, the need to include FGD, which is normally already included in most power plants without CO_2 capture but not in most cement plants, and the relatively high costs of providing steam for solvent regeneration from a modest-sized CHP plant.

Oxy-combustion
Oxy-combustion in just the pre-calciner avoids 61% of the CO_2 produced at the cement plant. If required, this could be increased to 66% relatively easily by re-capturing the CO_2 which is emitted in a vent stream from CO_2 purification, using membranes or a small scrubber. Oxy-combustion involves a significant increase in the on-site power consumption, mainly for oxygen production and CO_2 compression and purification. Taking into account CO_2 emitted during generation of this power, the overall reduction in CO_2 emissions is 52%. If the imported power was generated in power plants with low CO_2 emissions, such as plants with CCS, the overall avoidance of CO_2 emissions would be close to the on-site emissions avoidance. Oxy-combustion of the kiln as well as the pre-calciner, combined with processing of the vent stream, could increase the on-site CO_2 avoidance to close to 100% but this would involve greater technical uncertainties, as described earlier.

The cost of CO_2 emissions avoidance is €40/t (US$60/t)[1], substantially lower than the cost of post combustion capture. Oxy-combustion is particularly suitable for cement plants because oxygen only needs to be provided for the CO_2 that originates from fuel combustion. No oxygen needs to be provided for the CO_2 from mineral decomposition, which accounts for about two thirds of the on-site emissions of CO_2 from a modern cement plant.

Comparison with CO_2 capture in power generation
The costs of CO_2 capture in cement plants need to be viewed in the context of the costs of CO_2 abatement in other energy sectors. IEA GHG published studies on post-combustion and oxy-combustion capture of CO_2 in power generation in 2004-2005[5]. The costs of CO_2 avoidance were €30/t CO_2 for post combustion capture and €37/t CO_2 for oxy-combustion. The differences are within the limits of accuracy of the assessments. These costs are within the range of costs quoted in the IPCC Special Report on CCS, published in 2005[6], bearing in mind that the €/$ exchange rate was about 1:1 at the time when the source data for the IPCC report were produced. Since then the costs of all types of power and process plants have increased substantially due to higher raw materials prices, tight labour supply and market conditions in the equipment supply industries. Fuel costs have also increased substantially and the value of the US$ has declined against the Euro. After adjusting for inflation[7] the cost of post combustion capture in a coal fired power plant is estimated by IEA GHG to be €39/t (US$58/t)[1] of CO_2 avoided, which is similar to the cost of oxy-combustion at a European cement plant but less than the cost of post combustion capture.

Sensitivities
Sensitivities to technical and economic parameters are shown in Figures 1 and 2.

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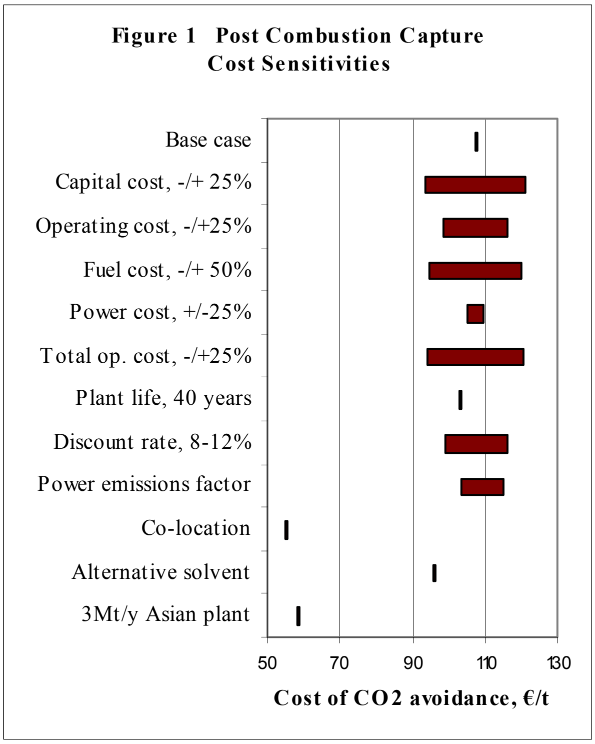
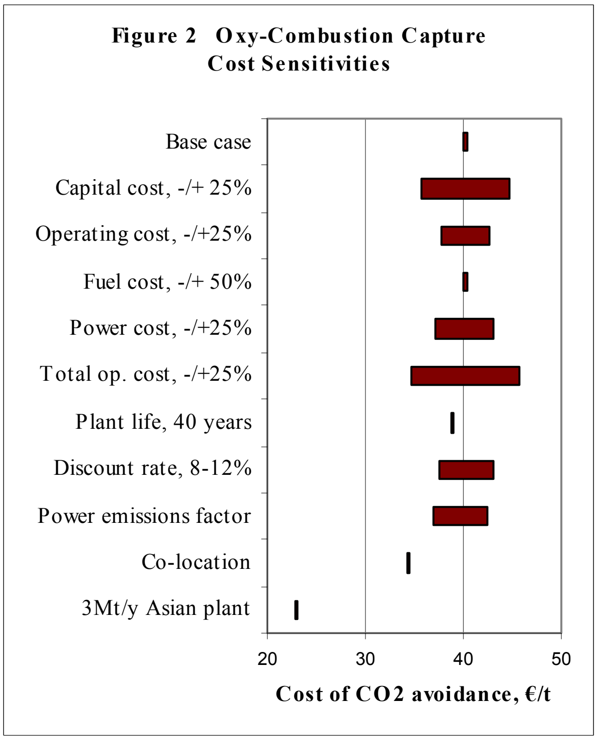
[5] Improvement in power generation with post combustion capture of carbon dioxide, IEA GHG report PH4/33, Nov. 2004. Oxy Combustion processes for CO_2 capture from power plants, IEA GHG report 2005/9, July 2005

[6] Intergovernmental Panel on Climate Change, Special Report on Carbon dioxide Capture and Storage, 2005, available at www.ipcc.ch

[7] Capital costs of power plants in Euros are assumed to have increased by 25% between the beginning of 2004 and the end of 2007 and the price of coal is assumed to be €2.51/GJ, in line with this study.





Power emissions factor

The emissions associated with the power import/export affect the quantity of net CO_2 emissions avoided. For the base case it was assumed that the emissions were the average of electricity generated in the UK (0.52 kg CO_2/kWh). The sensitivity cases are based on emissions from a new coal fired power plant without CO_2 capture (0.745 kg CO_2/kWh) and a coal fired plant with 85% CO_2 capture (0.14 kg CO_2/kWh).

Alternative post combustion capture solvent

Some alternative proprietary amine solvents, for example MHI's KS-1 solvent, have lower energy consumptions than the MEA solvent used as the basis for this study. The sensitivity to a 25% reduction in steam consumption is shown in Figure 1. Some developers are targeting even lower steam consumptions. Post-combustion capture processes which are not based on amines, for example aqueous ammonia scrubbing, also have the potential to significantly reduce the cost of capture in cement plants but insufficient data are currently available from process developers to enable costs to be estimated.

Co-location with a power plant and use of low sulphur raw meal

Another way of reducing the net cost of steam for post-combustion capture would be to obtain it from an adjacent power plant or another type of large industrial plant, thereby avoiding the need for an on-site CHP plant and its associated FGD. If in addition a raw meal (i.e. raw material) with a low sulphur content was used in the cement plant it may be possible to avoid the need for an FGD plant entirely. The sulphur compounds in cement plant flue gas originate mainly from decomposition of minerals during drying and preheating as most of the fuel-derived sulphur is retained in the cement product. A sensitivity case involving supply of steam from an adjacent large power plant at a cost of €10/t (US$15/t)[1] and use of a low sulphur raw meal was assessed. The cost of CO_2 avoidance is reduced by almost half to €55/t (US$83/t)[1].

Oxy-combustion cement plants could benefit from co-location with an oxy-combustion or IGCC power plant, because there would be economies of scale in larger oxygen and CO_2 compression and purification plants. The reduction in the cost of CO_2 avoidance is estimated to be approximately €6/t (US$9/t) at an assumed oxygen cost of €30/t (US$45/t).

Co-location would also help to reduce the cost of transporting the CO_2 to an underground store, because pipelines have large economies of scale. The European cement plants in this study capture 0.5-1.1 Mt/y of CO_2 and the bigger Asian plants capture 1.4-3.3 Mt/y of CO_2. For comparison a modern 1000 MW_e coal fired power plant operating at base load with 85% CO_2 capture would capture about 6 Mt/y of CO_2.

Large Asian plants
Substantially larger cement plants are being built and operated in some developing countries, particularly in Asia. The cost of CO_2 capture at a 3Mt/y cement plant in Asia was estimated to be significantly lower because of economies of scale and the lower costs of plant construction and operation in developing countries. The estimated cost of CO_2 avoidance by oxy-combustion is €23/t (US$34/t)[1] and the cost of post combustion capture is €59/t (US$88/t)[1].

Retrofit of CO_2 capture
Retrofit of post combustion capture to a cement plant would be relatively straight forward, provided sufficient space is available, as it would be a downstream addition to the plant. Oxy-combustion would require most of the core units in the cement plant to be rebuilt, although some of the ancillary solids handling equipment which accounts for a major proportion of the plant cost could be retained. Because oxy-combustion is expected to have significantly lower costs of CO_2 avoidance, it may be worthwhile retrofitting oxy-combustion at the time of a major plant re-build but this was not assessed in this study.

Capture ready plants
A CO_2 capture ready plant is a plant which can include CO_2 capture when the necessary economic and regulatory drivers are in place. Cement plants can be made capture ready by undertaking a design study on capture retrofit, including sufficient space and access for the capture equipment and by identifying reasonable routes to CO_2 storage. Some additional investment could be made to make retrofit easier and cheaper but such investment would need to take into account the effects of the time lag between investment and financial return and the uncertainties regarding future developments in capture technologies and whether or when capture would be retrofitted.

Expert Reviewers' Comments

The draft study report was reviewed by various external experts, including representatives from a cement manufacturer, a cement plant supplier, a cement industry research institute and experts on CO_2 capture. The report was generally well received by the reviewers and a substantial number of helpful comments were received. IEA GHG is very grateful to those who contributed to the review. The issues raised by the reviewers were discussed with the study contractor and representatives of the British Cement Association and modifications were made to the report to address most of the reviewers' comments.

Major Conclusions

The cement industry has made considerable strides in reducing CO_2 emissions but to make further major emission reductions CO_2 capture is required. The addition of CO_2 capture to new build cement plants to significantly reduce CO_2 emissions is technically feasible. Both post combustion and oxy-combustion options can be considered although the oxy-combustion option is not technically mature enough for deployment yet. Pre-combustion capture would be less suitable for cement plants because it would not capture the CO_2 from carbonate mineral decomposition, which account for about two thirds of the CO_2 from a modern cement plant.

The estimated costs of post combustion capture at new build cements plants are €107/tonne of CO_2 emissions avoided (US$161/t) for a 1 Mt/y European cement plant and €59/t (US$88/t) for a 3 Mt/y Asian plant. Use of alternative solvents and integration with an adjacent power plant could more than halve the costs. Oxy-combustion offers the lowest cost solution for CO_2 capture at new-build cement plants but further research and development is needed to address a number of technical issues to enable this technique to be deployed. Costs are estimated to be €40/tonne (US$60/t) of CO_2 avoided for a 1 Mt/y European cement plant and €23/t (US$34/t) for a 3 Mt/y plant in Asia.

The cost of CO_2 capture at a cement plant using oxy-combustion is expected to be similar to the cost of capture at a typical coal-fired power plant. The quantity of oxygen required per tonne of CO_2 captured is about three times lower at a cement plant but the economies of scale are less favourable. The cost of post combustion capture at a cement plant is expected to be substantially higher than at a power plant, mainly because of lower economies of scale and the need to install FGD, DeNOx and a boiler to provide steam for the regeneration of CO_2 capture solvent.

Post combustion capture could be readily retrofitted to existing cement plants provided sufficient space is available. Substantial rebuilding would be necessary to accommodate an oxy-combustion retrofit but this may nevertheless be the least cost option.

Recommendations

The following studies by IEA GHG should be considered for CO_2 capture at cement plants:

1. A study on assessing the capture readiness of cements plants.
2. An assessment of cement plants with novel post-combustion CO_2 capture processes including ammonia scrubbing should be undertaken when process performance and economic data become available from licensors.

Practical R&D on oxy-combustion at cement plants is outside the scope of IEA GHG but members should consider such work to address uncertainties identified in this study.

This study focussed on oxy-combustion in only the pre-calciner. Oxy-combustion for the whole plant including the kiln involves more technical uncertainties but it could be pursued as a study in the future, particularly if higher percentage capture of CO_2 is required.

Further work should be undertaken to assess the benefits of integration between cement plants with CO_2 capture and other large industrial plants, especially power plants. Potential barriers to co-location should also be assessed. Some of this work could be included in the proposed study on capture readiness.

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Glossary of Terms

ASU	Air Separation Unit
BAT	Best Available Techniques
BCA	British Cement Association
BET	Brunauer Emmett Teller
CAP	Chilled Ammonia Process
CCS	Carbon Capture and Storage
CHP	Combined Heat and Power
CLC	Chemical Looping Combustion
DCC	Direct Contact Cooling
EC	European Commission
EOR	Enhanced Oil Recovery
ESA	Electric Swing Adsorption
ESP	Electrostatic Precipitator
ETS	Emissions Trading Scheme
EU	European Union
FGD	Flue Gas Desulphurisation
HAZOP	Hazard and Operability Study
HHV	Higher Heating Value
IEA GHG	International Energy Agency Greenhouse Gas R&D Programme
IGCC	Integrated Gasification Combined Cycle
ILC	In Line Calciner
IPPC	Integrated Pollution Prevention and Control
ITM	Ion Transport Membrane
LHV	Lower Heating Value
LP	Low Pressure
MEA	Monoethanolamine
NOx	Oxides of Nitrogen
OEM	Original Equipment Manufacturer
OPC	Ordinary Portland Cement
ORC	Organic Rankine Cycle
PCA	Portland Cement Association
PCDD	Polychlorinated Dibenzodioxins
PCDF	Polychlorinated Dibenzofurans
PGNAA	Prompt Gamma Neutron Activation Analysis
PFD	Process Flow Diagram
PSA	Pressure Swing Adsorption
R&D	Research and Development
RSP	Reinforced Suspension Preheater
SCR	Selective Catalytic Reduction
SLC	Separate Line Calciner
SNCR	Selective Non-Catalytic Reduction
SOx	Oxides of sulphur
TOC	Total Organic Carbon
TSA	Temperature Swing Adsorption
UK	United Kingdom
US	Untied States
VOC	Volatile Organic Compound

Executive Summary

Background to Study

Cement production is an energy intensive production process, accounting for 2% of the global primary energy consumption (Worrell *et al.*, 2001) and between 6-7% of global industrial energy use (IEA, 2007). Its reliance on carbon intensive fuels such as coal and petroleum coke in clinker making is responsible for substantial CO_2 emissions coming directly from the combustion of fossil fuels. CO_2 is also emitted from the calcination process and from the consumption of electricity generated from fossil fuels. Cement production accounted for 1.8 Gt/y of CO_2 emissions worldwide in 2005 (IEA, 2007).

Carbon capture and storage (CCS) has been studied extensively for power plants by IEA GHG and others and there are plans to build several commercial scale plants. However, few studies have been undertaken on CCS in the cement industry. The IEA GHG commissioned Mott MacDonald to undertake this study to assess the technologies that could be used for CO_2 capture in cement plants, their costs, environmental impacts and barriers to their use.

Scope of Study

The study covered both new-build cement plants with post-combustion CO_2 capture and oxy-combustion CO_2 capture. The basis of the study was a 5-stage preheater with precalciner dry process cement plant with an output of 1 Mt/y of cement located in NE Scotland, UK. Process Flow Diagrams (PFDs) and heat and mass balance calculations for both options were developed. The plant costs and performance were estimated and the suitability for retrofit and the option of capture-ready cement plants was examined.

Post-Combustion CO_2 Capture Cement Plant

The major technical issues identified for applying post-combustion capture at a cement plant were as follows:

- concentration of sulphur dioxide (SO_x) in flue gas
- concentration of nitrogen dioxide (NO_2) in flue gas
- concentration of dust in flue gas
- air in-leaks in the raw mill, preheater and kiln
- the additional power required for the compression of CO_2
- the additional steam required for capture of CO_2 with amine
- the flue gas cooling required for efficient capture of CO_2 with amine.

All these issues can be addressed using existing technology. The main barrier to the application of post-combustion CO_2 capture at a cement plant is therefore the cost of deploying these technologies.

Oxy-Combustion CO_2 Capture Cement Plant

The major technical issues identified for applying oxy-combustion capture at a cement plant were as follows:

- The elevated flame temperatures achieved by oxygen combustion which are adjusted by recycle of the flue gas.
- The changing heat transfer characteristics in the combustion chamber due to the change of atmosphere.
- The influence of changing the atmosphere on feed lifting in the preheater.
- The balance between achieving high temperatures for the cement process and having to replace the kiln wall lining.
- The effect of changing the atmosphere on the thermodynamics, kinetics and equilibrium chemistry of the clinker production process.
- Air in-leaks in the raw mill, preheater and kiln.
- The additional power required for the compression of CO_2.

Many of these issues require fundamental research to prove the viability of oxy-combustion CO_2 capture at a cement plant although none was identified as a "show-stopper" at the time of writing this Report. However, at the present time, oxy-combustion capture is considered a higher technical risk solution than post-combustion capture.

Performance and Costs

Due to the fact that the majority of CO_2 emissions result within the precalciner, and due to significant air in-leaks in other stages of the cement plant, the option of undertaking oxy-combustion in the precalciner only was investigated in detail. In the case of post-combustion capture, CO_2 capture from all the flue gas was studied. This resulted in differing levels of CO_2 capture between the two options under consideration.

The results of the analysis are shown in Table S-1.

Table S-1: Summary of Oxy-combustion and Post-combustion Cement Plants with CO_2 Capture

	Parameter	Unit	Base Case	Post-Combustion	Oxy-Combustion
Inputs	Coal	t/y	63,317	291,633	72,061
	Petroleum Coke	t/y	32,876	32,876	27,091
Outputs	Clinker produced	t/y	910,000	910,000	910,000
	Cement produced	t/y	1,000,000	1,000,000	1,000,000
	CO_2 captured	t/y	0	1,067,734	465,014
	CO_2 emitted on-site	t/y	728,422	188,424	282,853
	Net Power Demand	MWh/y	80,809	-22,734	174,562
Performance	CO_2 associated with net power imports	t/y	42,021	-11,822	90,772
	Net quantity of CO_2 avoided (including net power imports)	t/y	N/A	593,841	396,818
	CO_2 avoided, net of power imports	%	-	77	52
Costs	Total Investment Cost	€M	263	558	327
	Net Variable Operating Costs	€M/y	17	31	22
	Fixed Operating Costs	€M/y	19	35	23
	Cost per tonne of CO_2 emissions avoided [i]	€/t	N/A	107.4 [ii]	40.2 [iii]
	Costs per tonne of cement product	€/t	65.6	129.4 (Δ = +63.7)	81.6 (Δ = +16.0)

[i] Cost per tonne of CO_2 emissions avoided is calculated by first determining the difference between the CO_2 emissions associated with the design under consideration and the base case. The additional costs incurred (compared to the base case) to achieve this reduction are then evaluated and the result expressed as €/tCO_2.

[ii] Previous studies (IEA GHG, 2004) have calculated the cost of avoidance of CO_2 emissions at a USCPF post-combustion power plant using coal at 29.5 $/tCO2 (25 €/t$CO_2$ at exchange rate of day). An updated estimate of this cost based on 25% inflation of capital costs and capital-related operating costs is 39 €/tCO_2 (Davison, 2008).

[iii] Previous studies (IEA GHG, 2005) have calculated the cost of avoidance of CO_2 emissions at an oxy-combustion power plant using pulverised coal at 37 $/t$CO_2$ (31 €/tCO_2 at exchange rate of the day).

The economic impact of implementing CO_2 capture can be summarised through the increase in cement production cost, or equivalently the same impact is shown through the cost per unit of avoided CO_2 emissions. The results show that the increase in the cost of producing cement is lower for the oxy-combustion process. Correspondingly, the capture cost per tonne of CO_2 emissions avoided for the oxy-combustion plant is also considerably lower than for the post-combustion process. The main reasons for this are the large capital costs required for the flue gas desulphurisation plant (FGD) and the CHP plant (the chosen method of generating LP steam for the amine scrubber) required in the post-combustion capture option. It should be noted that the costs associated with the FGD system would be reduced if the raw material used in the cement process was low in sulphur and the fuel source used in the CHP plant also contained low levels of sulphur.

The cost of CO_2 capture at cement plants is more expensive than recent reports for the power industry. One of the principle reasons for this is the economies of scale provided at power plants but it should be recognised that this study has adopted a conservative approach and that costs could be reduced with more refined design strategies. Costs will also reduce as knowledge is gained from operation of carbon capture systems at power plants. Co-location of the cement plant with a power plant offers further scope for savings. The co-location of a post-combustion cement plant with a post-combustion power plant would offer significant cost-saving opportunities as equipment (e.g. FGD, CO_2 absorbers, CO_2 compressors) could be shared. The limit of reduction in the cost of CO_2 avoidance at the cement plant would be the cost of abatement at the power plant.

From an environmental emissions perspective it should be noted that an added advantage of applying either post-combustion or oxy-combustion CO_2 capture at cement plants is that the cement plant will achieve emissions for dust, SO_x, NO_x and most other contaminants well below current cement plants.

Sensitivities

A number of sensitivities to the cases presented in Table S-1 were examined. This included an examination of undertaking CO_2 capture at a cement plant in an Asian developing country where cement plants are generally larger and cheaper to construct. The results from this analysis are shown in Table S-2.

Table S-2: Summary of Sensitivities Examined

Parameter	Unit	Asian Developing Country Scenario			Low S Raw Meal and External Steam Supply	Alternative Solvent for CO_2 Absorption	Co-location with Oxy-Combustion Power Plant
		No CO_2 Capture	Post-Combustion	Oxy-Combustion	Post-Combustion	Post-Combustion	Oxy-Combustion
Cement produced	t/y	3,000,000	3,000,000	3,000,000	1,000,000	1,000,000	1,000,000
Cost per tonne of CO_2 avoided (power emissions excluded)	€/t	n/a	64.6	20.4	51.2	105.0	33.9
Cost per tonne of CO_2 avoided (power emissions from grid included)	€/t	n/a	58.8	22.9	55.2	95.9	34.4
Costs per tonne of cement product	€/t	37.3	72.2	46.4	97.3	121.6	80.8

The cost of cement production is significantly lower in the Asian scenario. This is principally due to the economy of scale of operating a larger plant, the lower equipment costs and the cheaper labour. If carbon capture is applied then the cost of production increases significantly (more for post-combustion than oxy-combustion). However, it should be noted that the cost of cement production is only comparable to the European base case when post-combustion capture is applied. This indicates that it is more cost effective to undertake carbon capture at a cement plant in Asia than at a European plant.

The other sensitivities investigated have shown that the cost of CO_2 avoidance at cement plants can be reduced through co-location with power generation plant (including external steam supply), if a low sulphur raw meal is used, or if more efficient solvents than MEA are used in the absorption process.

Retrofitting

With regard to retrofitting CO_2 capture at an existing cement plant the following points need to be considered:

- It is technically feasible to retrofit post-combustion capture using amine absorption to any existing cement plant. The key constraints influencing the economic feasibility are:
    - the concentration of NO_2 and SO_x in the flue gases
    - the space requirements for the capture and compression equipment
    - the availability of sufficient steam for the amine absorption and regeneration system.
- The technical solution considered in this Report is to apply oxy-combustion in the precalciner only. The most feasible scenario for retro-fitting this solution to an existing cement plant would be if the existing precalciner operated as a separate line calciner (SLC). SLC operation is not currently the preferred technology within the cement industry.
- Given the significant cost differential between employing oxy-combustion and post-combustion capture it may be more economic to undertake a major rebuild of the kiln (i.e. from single to twin preheater towers) and undertake the oxy-combustion retrofit rather than implementing the post-combustion option. The costs of undertaking the retrofit would need to be assessed on a case by case basis.

Capture Ready

With regard to providing capture ready cement plant the following points need to be considered:

- A cement plant that was designed to be capture ready for post-combustion capture using amine would need to make provision for the following:
    - Leaving sufficient space for the flue gas treatment, CO_2 absorption, CO_2 purification and CO_2 compression equipment.
    - Fuel supply capacity, grid supply capacity and land for a power plant capable of generating the additional steam and electricity required by the capture equipment, or else potential for over-the-fence steam purchase from neighbouring cogeneration facilities.

  - Including SO_x and NO_2 abatement technologies within the plant or providing the space and capability to introduce these later.
  - Ability to transport (via pipe) the CO_2 to a suitable storage location.

- The total land requirement estimated by this study for post-combustion capture is in the range 3.6-4.3 ha, depending on site conditions.

- It is clear that it would be beneficial if the cement plant was co-located near an existing readily available steam supply (e.g. a power plant) that could be used when the post-combustion capture plant was installed. There is also important potential for capturing economies of scale through co-location with power plants, or other large emitters, also implementing CCS.

- Installation of a cement plant that is designed to be capture-ready for oxy-combustion in the precalciner only would require twin preheater towers with separate induced draft fans dedicated to each tower. Ideally the SLC configuration would be installed. However, switching from a parallel precalciner combustion chamber to fully separate venting would only require minor preheater ductwork modifications.

- Sufficient space would need to be provided for the air separation unit and CO_2 purification and compression plant and would ideally be installed adjacent to the preheater tower and kiln on the side closest to the precalciner. This study has estimated that the space required for oxy-combustion equipment at a 1 Mt/y cement plant would be approximately 0.5 ha.

Conclusions

Mott MacDonald have evaluated CO_2 capture at cement plants and have estimated the cost and performance of applying post-combustion and oxy-combustion capture at a 1 Mt/y plant. Post-combustion capture has been found to be the most expensive solution but could feasibly be applied at any cement plant as a retrofit solution. Oxy-combustion capture is a cheaper technology but more research and development of this option is required if this option is to become viable. The cost differential between oxy-combustion and post-combustion is such that, even if a major rebuild is required, retrofitting oxy-combustion may still be cost-effective.

1 Introduction

1.1 Background

The International Energy Agency Greenhouse Gas R&D Programme (IEA GHG) was established in 1991 to evaluate technologies that could be used to avoid emissions of greenhouse gases, particularly from the use of fossil fuels. IEA GHG is an international organisation, supported by eighteen countries worldwide, the European Commission, OPEC and twenty multi-national industrial sponsors[1].

The cement industry is one of the largest industrial sources of CO_2, accounting for 1.8 Gt/y of emissions worldwide in 2005 (IEA, 2007). It is expected that cement production, and hence CO_2 emissions, will grow significantly in the future, especially in developing countries. Cement plants emit CO_2 from fuel combustion (about 40% of the total in modern plants) and decomposition of carbonate minerals.

Modern cement plants have high efficiencies and the scope to reduce CO_2 emissions by further efficiency improvements is small. One of the few ways of greatly reducing CO_2 production from cement production is carbon capture and storage (CCS). CCS has been studied extensively for power plants by IEA GHG and others and there are plans to build several commercial scale plants. However, few studies have been undertaken on CCS in the cement industry.

Mott MacDonald is one of the world's largest multi-disciplinary consultancies and has won recognition for technical excellence in power engineering, water, transportation, building and infrastructure. Dr Michael Clark from cement industry consultants Whitehopleman assisted Mott MacDonald and provided expert advice on aspects associated with the cement plant.

The British Cement Association (BCA), which includes representatives of many major global cement companies, provided data from typical modern cement plants to assist with the study. The study also benefited from contributions from Lafarge and Polysius. Mott MacDonald encouraged this participation in order to ensure that, as far as possible, the results were accepted by the cement industry and their major concerns were addressed.

1.2 Scope of Work

The study assessed technologies that could be used to capture CO_2 in new cement plants and retrofits to existing modern plants. Capture in older, smaller plants was not considered. The major tasks covered in the study are listed below.

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[1] The members of IEA GHG are currently Australia, Austria, Canada, Denmark, Finland, France, Germany, India, Japan, Korea, The Netherlands, New Zealand, Norway, Spain, Sweden, Switzerland, UK, USA, OPEC, The European Commission, Alstom Power Technology, Babcock & Wilcox, BG Group, BP, CEZ Group, Chevron Corporation, ConocoPhililps, E.On, EniTechnologie SpA, EPRI, ExxonMobil, Repsol YPF, RWE AG, Schlumberger, Shell, Statkraft, Statoil, Total and Vattenfall.

1.2.1 Cement Plant Descriptions

Outline descriptions of the technologies used for cement production, typical plant sizes and variations between regions worldwide are provided. Descriptions of flue gas clean-up methods currently used in cement plants are included.

1.2.2 CO_2 Capture Process Descriptions

The leading technologies that could be used to capture CO_2 in cement plants are described. This includes post-combustion, oxy-combustion and pre-combustion. The reasons for not assessing pre-combustion capture are also explained.

1.2.3 Major Issues Affecting Application of CO_2 Capture Processes in Cement Plants

The main technical issues which would affect the application of post-combustion CO_2 capture technology and oxy-combustion CO_2 capture technology in cement plants are assessed.

1.2.4 Detailed Assessment of Plant Cost and Performance

The impacts on the performance, capital costs and operating costs of adding CO_2 capture to a new-build cement plant is assessed in detail for both the post-combustion and oxy-combustion CO_2 capture technologies.

1.2.5 Suitability for Retrofit

The suitability of each CO_2 capture technology for retrofit to existing modern cement plants has been assessed. This has taken into account estimates of the additional land area required and the need to make modifications to the existing cement plant equipment.

1.2.6 Capture-ready Plants

During the timeframe of the study the IEA GHG published another report (IEA GHG, 2007a) which assessed ways of making power plants ‘capture-ready’ i.e. designing them to be easily retrofitted in future if necessary. Drawing on the results from this report, ways of making cement plants ‘capture-ready’ have been assessed.

1.2.7 Information Gaps and R&D needs

Important information gaps have been identified and recommendations for research, development and demonstration are made.

1.3 Approach

This report is based on the following sources of information:

- A review of published literature.

- Meetings and discussions with IEA GHG and BCA.
- A site visit to Cauldon Cement Works.
- Process modelling undertaken by Mott MacDonald and Whitehopleman.
- Cost estimating undertaken by Mott MacDonald and Whitehopleman.
- Economic modelling undertaken by Mott MacDonald.
- Consultations with a number of cement plant equipment suppliers and CO_2 capture technology providers.
- Consultations with researchers working on related topics.

1.4 Structure of Report

The report is structured in a way that largely mirrors the Terms of Reference. The report consists of the following sections:

- Section 2 provides background to the cement industry including details of flue gas cleaning methods currently used at cement plants.
- Section 3 provides background on CO_2 capture processes.
- Section 4 contains the technical evaluation of CO_2 capture at a new-build cement plant for both post-combustion and oxy-combustion solutions.
- Section 5 provides an analysis of the costs of providing CO_2 capture.
- Section 6 describes the issues pertaining to the retro-fitting of post-combustion and oxy-combustion CO_2 capture at cement plant.
- Section 7 examines the issues associated with making cement plants ‘capture ready’ for either post- or oxy-combustion CO_2 capture.
- Section 8 identifies information gaps and further research required for the further development of post- and oxy-combustion at cement plants.
- Section 9 summarises and compares the post-combustion and oxy-combustion solutions.
- Section 10 presents the conclusions from this study.

Appendices are attached to the Report which include various technical data used and generated as part of the study.

2 Cement Plants

2.1 Introduction

This section begins with a brief overview of the global cement industry including discussion on variations between different worldwide regions. This is followed by outline descriptions of the technologies used for cement production including descriptions of flue gas clean-up methods currently used in cement plants.

2.2 The Global Cement Industry

2.2.1 Cement Nature and Properties

Cement is one of the most abundantly produced materials in the world due to its widespread use in construction activities (concreting), low price and the geographic abundance of its main raw materials (EC, 2001).

Cement is composed of calcium silicates, calcium aluminates and calcium aluminoferrite minerals. It is produced from a mixture of raw materials, with calcium carbonate being the primary raw material and provider of the lime necessary to combine with the silica, alumina and iron oxide in these cement minerals. The silica, alumina and iron oxide are provided from a variety of secondary raw materials, including alumino-silicate clays and shales, silica sand, bauxite and iron ore.

These raw materials are ground together into a fine, intimately mixed powder and delivered to a cement kiln. In the kiln the raw materials react together during the heating process to form an intermediate 'clinker', composed of the cement minerals, that is subsequently ground together with some gypsum to form the cement finished product.

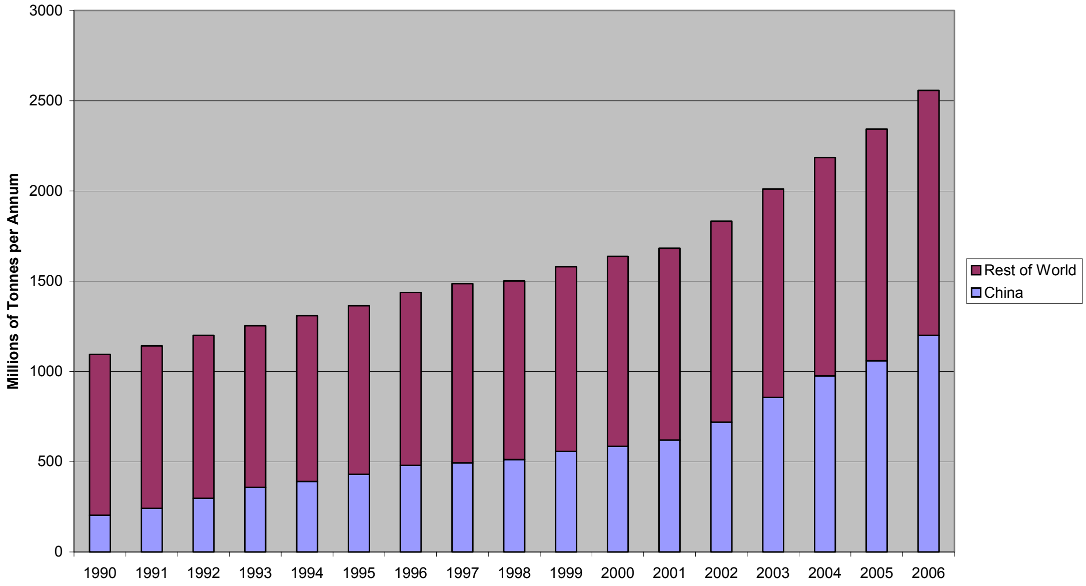
The cement product is a finely ground, non-metallic, inorganic substance with hydraulic binding properties. Cement forms a strong hard paste which binds together sands and aggregates in mortar or concrete when mixed with water due to the formation of calcium silicate and aluminate hydrates. 'Portland Cement' is the most common cement type and is composed of 95% clinker and 5% gypsum. The gypsum is ground together with the cement clinker to control the setting of cement and allow the mortars and concretes produced with the cement to be worked into moulds and placed around reinforcement. International standards for cement now allow the mixture of up to 5% limestone with this Portland Cement in Ordinary Portland Cements (OPCs).

The exact composition of cement determines its properties (sulphate resistance, alkalinity, heat of hydration etc.) whereas its fineness is related to the development of strength and rate of setting.

2.2.2 Production Trends

World cement production has grown steadily since the 1950s, with increased production in Asia and China in particular, accounting for the bulk of growth from the 1990s. Figure 2-1 shows that global cement production reached an estimated 2,557 million tonnes in 2006. Cement consumption has grown at an average rate of 5.48% per year during the period 1990 to 2006 with this average growth rate having accelerated to 8.73% per year since 2002.

Figure 2-1: World and Chinese Cement Consumption Trend 1990-2006 (Global Cement Report, 2007)



The largest cement producer, by far, is China, whose annual consumption had risen to 1,200 million tonnes in 2006. Chinese cement consumption has risen at an average rate of 11.90% per year during the period 1990 to 2006 with this average growth rate having accelerated to 14.17% per year in the years 2002 to 2006.

In 2002 cement consumption in India (of 110.86 million tonnes) overtook that in the USA for the first time to become the second largest consumer of cement in the world. Since that date Indian consumption has grown at an average rate of 11.31% per year, compared with 1.91% in the USA, consolidating India's position as the second largest cement consuming country in the world. Indian cement consumption had risen to 153 million tonnes per year in 2006 compared with 127.58 million tonnes in the USA.

In Western Europe cement consumption has risen at an average rate of 1.45% per year during the period 1990 to 2006, from 220 million tonnes in 1990 to 276 million tonnes in 2006. The rate of increase in cement consumption in Europe has accelerated to 3.13% in the years 2002 to 2006.

2.3 Global Variations in the Cement Industry

2.3.1 Introduction

The continuing world cement boom, as described in the preceding section, is marked by regional variations due to economic, technological and policy drivers coming about in the main cement producing countries/regions of China, India, the USA, Europe and the Middle East.

2.3.2 The Cement Industry in Europe

As discussed in section 2.2.2 growth in cement demand and consumption has been lower in Europe than in Asia or the USA. However, four of the five top multinational cement companies are based and have their origins in Europe. This means that the European cement industry has a truly global presence enjoying a market share of 95% in Europe and 70% in North America.

In 2006, the European cement industry produced 314 Mt, accounting for 12.4% of global output (2.54 billion tons). Table 2-1 shows the main cement consumers in Europe in 2006. Cement production followed construction trends closely, with new residential (26.7%) and renovation (45.2%) works accounting for the rise in production (CEMBUREAU, 2006).

Table 2-1: Main Cement Consumers in Europe in 2006

Country	Cement Consumption (Mt)
Spain	55.74
Italy	46.88
Turkey	41.65
Germany	28.57
France	23.87

In 1995 there were 252 installations producing cement clinker and finished cement in the European Union and a total of 437 kilns but not all of them in operation (EC, 2001). In addition there were a further 68 grinding plants (mills) without kilns. In recent years typical kiln size has come to be around 3,000 tonnes clinker/day (1 Mt/yr).

In 2001, about 78% of Europe's cement production was from dry process kilns, a further 16% of production was accounted for by semi-dry and semi-wet process kilns, with the remainder of European production, about 6%, coming from wet process kilns (EC, 2001). The wet process kilns operating in Europe are generally expected to be converted to dry process kiln systems when renewed, as are semi-dry and semi-wet processes kiln systems. These technologies are more fully described in section 2.4.

Growth in cement production is forecast to increase due to prospective construction projects and infrastructure development. Imports to the EU are also growing (6.8 million tonnes from China alone in 2006 (CEMBUREAU, 2006)) and new sea terminals for cement are being constructed.

2.3.3 The Cement Industry in the USA

The cement industry is a relatively small but significant component of the US economy; in 2003 it had 39 companies operating 118 cement plants in 38 states. Worldwide, the United States ranked third in cement production, behind China and India for that same year (PCA, 2003).

US cement production is widely distributed with the largest company producing over 13% of the industry total. The top five multinational cement companies collectively produce around 53% of the total. Foreign companies now own approximately 81% of US cement capacity, up from about 22% in 1980 (PCA, 2003). Investments during the 1980s by European companies, as well as Asian entities, were spurred by the favourable position of the US dollar against foreign currencies.

In 2006, US cement consumption was 127.6 million tonnes (Global Cement Report, 2007). Cement consumption was spurred by strong performance in the construction industry as a whole; however, individual sector growth, such as highway construction, still affects cement consumption more heavily. Strong construction markets in the 1990s and since 2003 have maintained the average growth rate of 2.90% in cement consumption during the period 1990 to 2006. However, in 2007 and 2008 actual cement consumption is forecast to decline due to difficulties in the US residential construction industry arising from sub-prime lending problems and a tightening credit environment.

The cement industry has boosted efficiency by concentrating new capital investment in plants that use the dry process of cement manufacture and by phasing out operations that rely on the more energy-intensive wet process. In the period 1974 to 2003, the number of wet process kilns dropped from 234 to 54 (77% decline) while the number of dry process kilns was reduced from 198 to 136 as older, low capacity kilns were replaced by higher capacity, modern, dry-process kilns. In 2003 about 81% of the cement produced in the United States was manufactured using dry process technology (PCA, 2003). This process of modernisation of the US cement manufacturing base has continued during the period 2003 to 2006.

The cement industry is regional in nature. As the cost of shipping cement quickly overtakes its value, customers traditionally purchase cement from local sources. Nearly 96% of US cement is shipped to consumers by truck (PCA, 2003). Barge and rail modes account for the remaining distribution modes.

The USA has been the largest market for imported cement throughout the period 1990 to 2006. In 2006 these are estimated to have reached a peak of 35.51 million tonnes or 28% of the total cement consumption of 127.58 million tonnes. The four largest sources of these imports, in order of scale were China, Canada, Thailand and Taiwan. Other large suppliers were South Korea, Mexico, Greece, Colombia, Venezuela, Sweden and Peru. These import volumes are expected to be significantly lower in 2007 and 2008 as cement demand contracts and domestic cement producers increase their market share at the expense of imports.

Exports of cement seldom exceed 1% of total US production (PCA, 2003; USGS 2007a; USGS 2007b; USGS 2007c). Like imports, exports are cyclical reaching marginally higher levels during economic recessions when domestic markets are slack. In 2006, the United States is estimated to have exported 1.55 million tonnes of cement, mainly to neighbouring Canada and Mexico.

2.3.4 The Cement Industry in Asia

(i) China

Much of China's development over the past few decades has depended on cement, as the country has built numerous power plants, millions of houses and buildings and paved thousands of kilometres of highways, all demanding enormous quantities of cement.

China's cement industry has grown remarkably since economic reforms began in the 1970s. By 1985, China had become the world's leading cement producer and by 1998, the cement output was twice as much as the next three largest producing countries combined. In 2000, per capita cement output reached 448 kg, which is about 200 kg higher than the world average (WBCSD, 2002a). Per capita annual cement consumption reached 812 kg in 2005 and 915 kg in 2006.

The cement industry in China remains highly fragmented and suffers from significant overcapacity and supply/demand imbalances. In 2005 the number of integrated cement plants had grown to 4,777 as a result of speculative expansion of production capacity. Excess supply eroded prices at a time when energy and transportation costs were rising. As a result Chinese cement company margins were seriously eroded and 36% of companies reported losses in 2005. Margins are reported to have improved in 2006.

Overcapacity problems are exacerbated by the pressing need to modernise the Chinese cement production base. 60% of capacity remains reliant on wet process or shaft kiln technology. The estimated total number of Chinese cement plants ranges from 8,000 to 9,300, although the actual number is uncertain due to the fragmented nature of the industry, the small size of many plants, the fact that some plants exist illegally and data reliability issues. About 50% of these facilities are rural township enterprises with average annual output of less than 30,000 tonnes (WBCSD, 2002a).

Historically, Chinese cement has not been of particularly high quality due to the widespread use of vertical kilns. However, with the gradual increase in use of the advanced dry rotary kiln process it is to be expected that cement production will become more efficient and the cement will be of better quality.

China's central government is aware of the challenges and has formulated several proposals to restructure the cement industry in the 11th five year development plan. This includes the following targets to be met by 2010:

- Total annual cement production of 1.25 billion tonnes.
- Consolidation with top 10 cement groups producing 30% of total.
- Modernisation with 250 million tonnes of shaft kiln annual cement production capacity being eliminated and the proportion of production from modern, dry-process cement plants rising to 70%.

The government has also taken steps to cool the economy and control the growth of fixed-asset investment and the cement manufacturing capacity. There is evidence that the rate of capacity expansion is slowing. However, delays in the closure of obsolete capacity continues to be a major threat to addressing the supply/demand imbalances (Global Cement Report, 2007).

(ii) India

India, with a total consumption of 153 million tonnes in 2006, has emerged as the second largest market for cement after China. In 2007 and 2008 India is forecast to be the fastest growing market in the world (Global Cement Report, 2007). Spending on infrastructure, particularly on roads, ports and airports, a surge in housing construction and expansion in corporate production facilities are responsible for the spurt in cement consumption growth in India. In particular, the shift in cement use patterns is accounted for by the housing sector, which is responsible for more than 50% of the total cement demand (IBEF, 2004).

The production base of the Indian cement industry includes more than 100 integrated cement plants and more than 25 dedicated grinding units (Global Cement Report, 2007). Installed capacity is 160 million tonnes per year and more than 300 mini-cement plants have an additional estimated capacity of 11.10 Mt per year (IBEF, 2004) although these mini-cement plants are only estimated to have produced 5.5 million tonnes in 2006 (Global Cement Report, 2007). 86% of production is from modern, dry-process cement plants with capacity of more than 1 million tonnes per year.

In 2007 the Indian cement industry will approach full capacity utilisation and multiple capacity expansion projects have been announced. Table 2-2 shows independent estimates of the projected capacity expansions until 2011. This will lift Indian cement production capacity to 280 million tonnes per year. All these capacity expansions will be based on modern, dry-process cement manufacturing technology.

Table 2-2: Independent Estimates of Projected Cement Capacity Expansions in India

Year	Projected Capacity Expansions (Mt)
2007	7
2008	16
2009	46
2010	30
2011	20

India produces various types of cement, including the most widely used Portland type. Pozzolan Portland blended cement constitutes 52% of total cement sales, with Ordinary Portland Cement (OPC) making up 39% of sales.

Major consolidation of the Indian cement industry has taken place in recent years. The Holcim acquisition of stakes in Gujarat Ambuja Cement and Associated Cement Companies has resulted in a group with 33 million tonnes per year capacity. The merger of Grasim and Ultratech in 2004 has resulted in a local cement company with 28.8 million tonnes per year capacity. However, the Indian cement industry still largely remains unconsolidated with more than 40 other players and some new entrants. The southern region accounts for the largest share in overall production (32%) due to the availability of limestone, followed by the north with 21% (Global Cement Report, 2007).

Over the years, demand in the east has been driven by the housing sector, whereas infrastructure and investments in industrial projects have propelled demand in other regions. Cement and clinker exports grew at a rate of 18.1% from 1995 to 2004 when they peaked with total exports of 10 million tonnes. However, export volumes have declined in 2005 and 2006 as domestic cement demand has increased and this downward trend of export is predicted to continue in 2007 and 2008.

2.4 Cement Plant Descriptions

2.4.1 Cement Manufacture Processes

(i) General Description

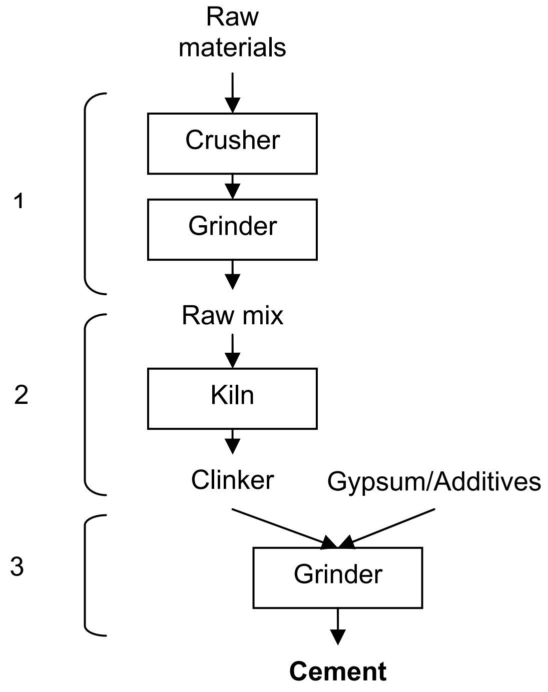
The process of manufacturing cement (shown in Figure 2-2) requires the following three stages:

- **Raw material preparation:** grinding a mixture of raw materials to make a 'raw mix'.

- **Clinker burning (Pyro-processing):** heating the raw mix to a sintering temperature of >1400ºC in a cement kiln to produce 'clinker'.

- **Cement preparation:** grinding the resulting clinker with gypsum and other additives to make cement.

Figure 2-2: Simplified Cement Making Process Schematic



The most common raw materials used for cement manufacture are limestone, marls and clay (although others can be used). The selected materials are crushed, ground and mixed to produce a raw mix of suitable composition and form for pyro-processing. Approximately 1.50-1.60 t of (dry) raw mix are required to produce 1 t of clinker, due to the calcination of the limestone component of the raw mix.

The second stage in the process involves feeding the raw mix into a kiln system where it is gradually heated. Successive chemical reactions take place as the temperature of the raw mix rises. These reaction stages are shown in Table 2-3.

Table 2-3: Kiln Chemical Reactions (CEMBUREAU, 1999)

Kiln temperature (ºC)	Chemical reaction
20 – 900	Drying and Preheating: the release of free and chemically bound water
850 – 950	Calcination: the release of CO_2 from calcium carbonate (limestone) and initial reactions with formation of clinker minerals and intermediate phases
1250 – 1450	Sintering or clinkerisation: the formation of calcium silicates and partial melting
1350 – 1200	Kiln internal cooling: crystallisation of calcium aluminate and calcium ferrite from the partial melt

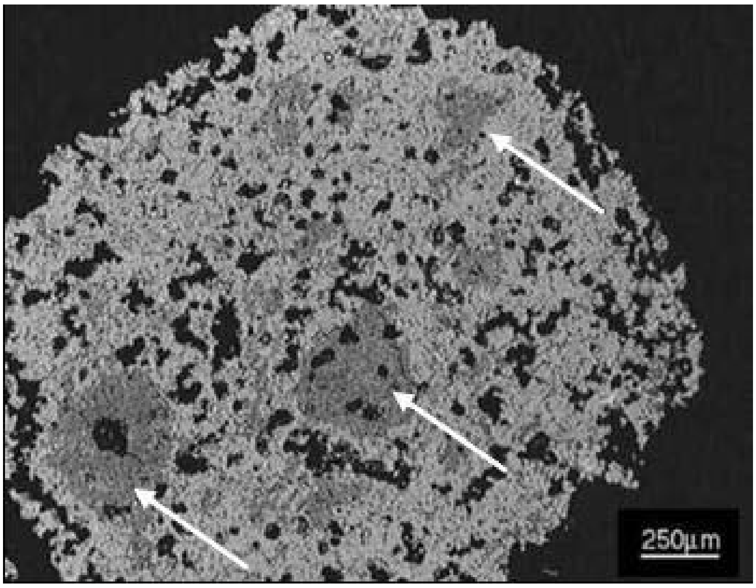
Typically, a peak temperature of 1400-1450 ºC is required to complete the reactions and produce alite, the characteristic constituent of Portland cement. Partial melting causes the material to aggregate into lumps or nodules known as clinker which is then cooled with air to 100-200ºC. Waste heat from the kiln system is usually used for drying the raw materials in the raw grinding section of the production process. A typical cement clinker consists of the compounds shown in Table 2-4.

Table 2-4: Typical Composition of Cement Clinker (IEA GHG, 1999)

Compound	Chemical formula	Content (%)		
		Min	Average	Max
Tricalcium silicate (Alite)	$3\ CaO.SiO_2$	45	62	75
Dicalcium silicate (Belite)	$2\ CaO.SiO_2$	5	15	35
Calcium aluminoferrite (Ferrite)	$4\ CaO.Al_2O_3.FeO_3$	4	8	15
Tricalcium aluminate (Aluminate)	$3\ CaO.Al_2O_3$	4	11	15
Free Calcium oxide	CaO	0.1	1	4
Free Magnesium oxide	MgO	0.5	1.5	4.5

Figure 2-3 shows an electron microscope image of a (polished) clinker nodule. Most of the nodule is comprised of alite (light grey) and some clusters of belite are visible (arrowed). Aluminate and ferrite are present but not visible at this relatively low magnification.

Figure 2-3: Magnification of a Clinker Nodule (Understanding Cement, 2007)



The final stage in the manufacturing process involves grinding clinker with additives, such as gypsum, limestone, granulated blast furnace slag and fly ash, to control the properties of the cement.

(ii) The Four 'Integrated Plant' Process Routes

The three manufacturing stages previously mentioned are common to four main 'integrated plant' process routes by which cement can be produced. These process routes are distinguished by the moisture content of the feed going into the kiln and are known as follows:

- **Wet process:** raw materials are ground in water to form a slurry which is either fed directly into the kiln or first to a slurry drier.
- **Semi-wet process:** raw slurry is first dewatered in filter presses and the filter cake extruded into pellets and fed either to a grate preheater or directly to a filter cake drier for raw mix production.
- **Semi-dry:** dry raw mix is pelletised with water and fed into a grate preheater before the kiln or to a long kiln equipped with chains and crosses to facilitate the heat exchange between the kiln feed and the combustion gases.
- **Dry process:** raw materials are ground and dried to form a dry raw mix in the form of a flowable powder. The dry raw mix is fed into the preheater or precalciner kiln, or more rarely to a long dry kiln.

Each process requires a slightly modified approach and use of equipment in the first two of the three manufacturing stages, i.e. the raw grinding and pyroprocessing stages. These differences are reviewed in the sections below.

Historically, the clinker manufacturing process has evolved from 'wet' to 'dry' systems with the intermediate steps of the 'semi-wet' and 'semi-dry' process routes. The first rotary kilns, introduced around 1895, were long wet kilns (CEMBUREAU, 1999). 'Wet' kilns allowed for easier blending and homogenisation of raw materials. However, as technology developed it became possible to produce a homogenous raw mix using the dry process. The 'dry' production process requires less energy and so has lower fuel costs than the wet process and thus is generally favoured where its application is feasible.

Wet Process

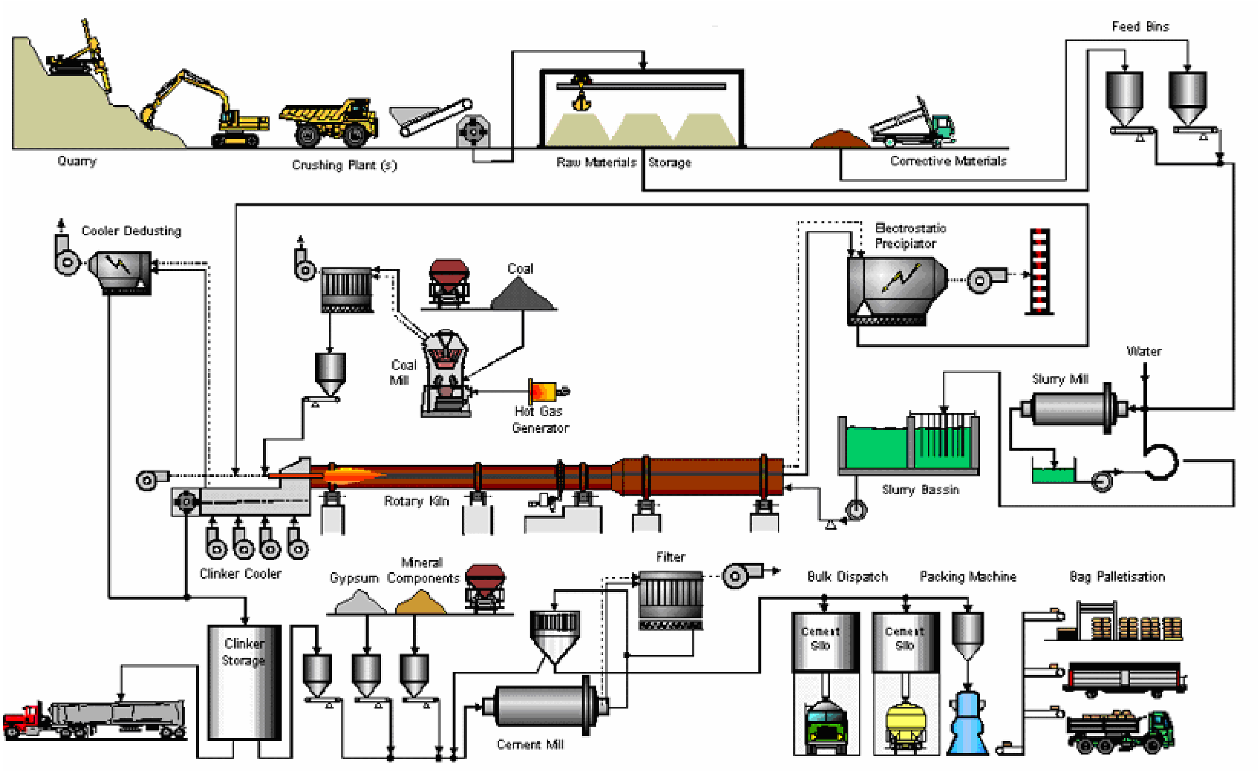
In the wet process the preparation of the raw mix is achieved by grinding the dry raw materials with water in wash mills, or ball mills with rubber linings. The resultant slurry is mixed in slurry basins where compressed air is introduced and the slurry is continuously stirred to produce a slurry of consistent composition.

In the conventional wet process this thick liquid slurry, which may have water content of 30-40%, is fed directly into the kiln where the water is evaporated in the drying zone at the kiln inlet. This drying zone is fitted with curtains of chains which promote heat exchange between the slurry and the exhaust gases from the combustion of fuel in the kiln passing counter-current with the slurry feed up the kiln. After passing through the drying zone the raw material moves down the kiln through the preheating zone to be calcined and finally burnt to clinker in the sintering zone. Conventional wet kiln technology has a high heat consumption and produces large volumes of exhaust gases. In more modern wet kiln systems water is driven off the raw slurry prior to entering the kiln in a slurry drier. Slurry “thinning” chemicals may be added to allow the slurry to be pumped to the kiln with lower water content. These more modern wet kiln systems have lower specific heat consumption compared to conventional wet kilns.

In this process the fuel consumption is the highest (in the region of 5000 to 7000 kJ/kg of clinker) but the power consumption is typically lower at 110-115 kWh/tonne of cement (Chemical Universe, 2007).

A simplified flow diagram of a typical wet process is shown in Figure 2-4.

Figure 2-4: Production of Cement by the Wet Process (CEMBUREAU, 1999)



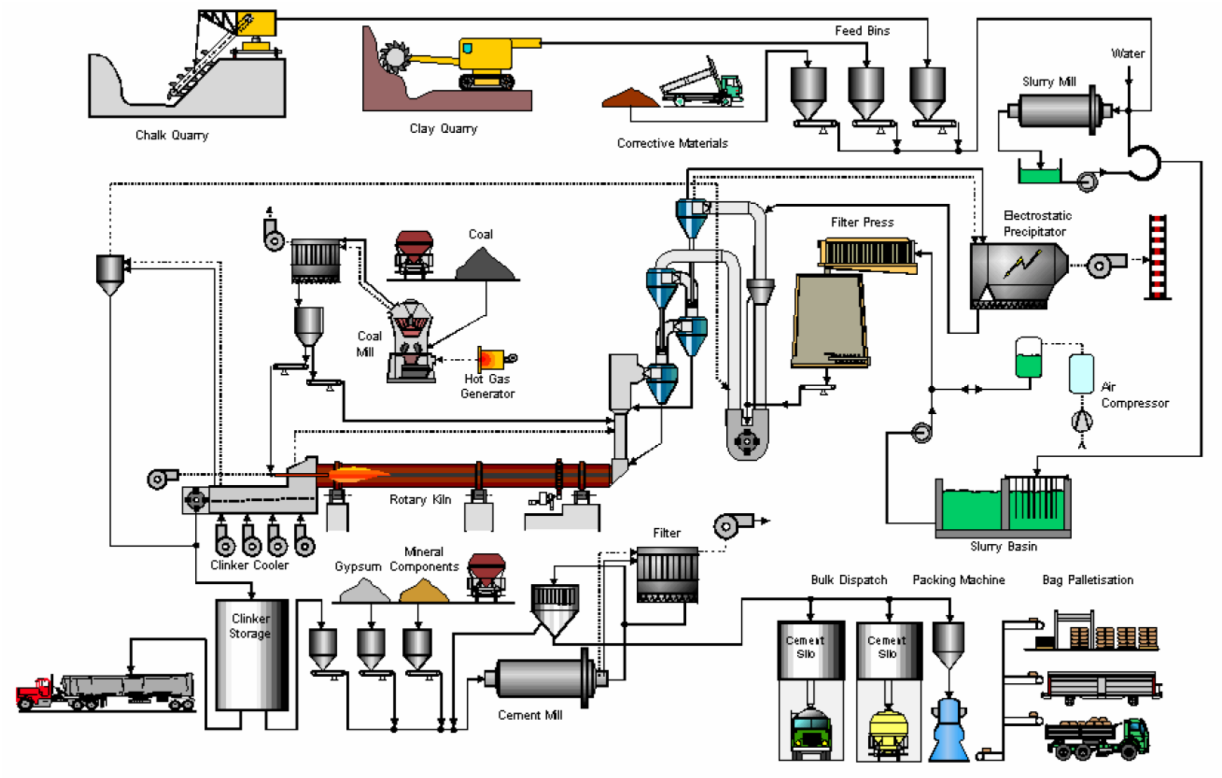
Semi-wet Process

As in the wet process the raw mix used in the semi-wet process is prepared as a slurry. Unlike the wet process the slurry is then dried in crusher-dryer before being delivered to a preheater for preheating, calcination and clinker formation in the rotary section of the kiln.

The fuel consumption of the semi-dry and semi-wet process is about 3500-5000 kJ/kg of clinker. The power consumption is typically in the range 115-120 kWh/tonne of cement (Chemical Universe, 2007).

A simplified flow diagram of a typical semi-wet process is shown in Figure 2-5.

Figure 2-5: Production of Cement by the Semi-wet Process (CEMBUREAU, 1999)



Semi-dry Process

In the semi-dry process a dry-powder raw mix is made into pellets in a nodulizing pan with the addition of 10-13% water. The pellets are loaded onto a *grate preheater* which dries, preheats and partially calcines the raw mix. The pellets then drop into the kiln for completion of the calcination and combination into clinker.

If the semi-dry process is used kiln exhaust gases cannot be utilised in the raw mix drying and grinding process due to their lower temperature following their prior use. This combined with high grate preheater maintenance costs mean that the semi-dry process is no longer installed in modern cement plants.

The fuel consumption of the semi-dry and semi-wet process is about 3500-5000 kJ/kg of clinker. The power consumption is typically in the range 115-120 kWh/tonne of cement (Chemical Universe, 2007).

Dry process

In the dry process, the raw materials are dried in a combined drying and grinding plant to reduce the moisture content below 1%. The materials are ground in either a ball or a roller mill with drying of the materials being achieved by sweeping the mill with the exhaust gases from the kiln. The drying capacity of the raw grinding system may be supplemented by auxiliary hot air generating furnaces.

The dried, finely ground raw mix powder is then blended and homogenised in silos with the mixing being provided by compressed air.

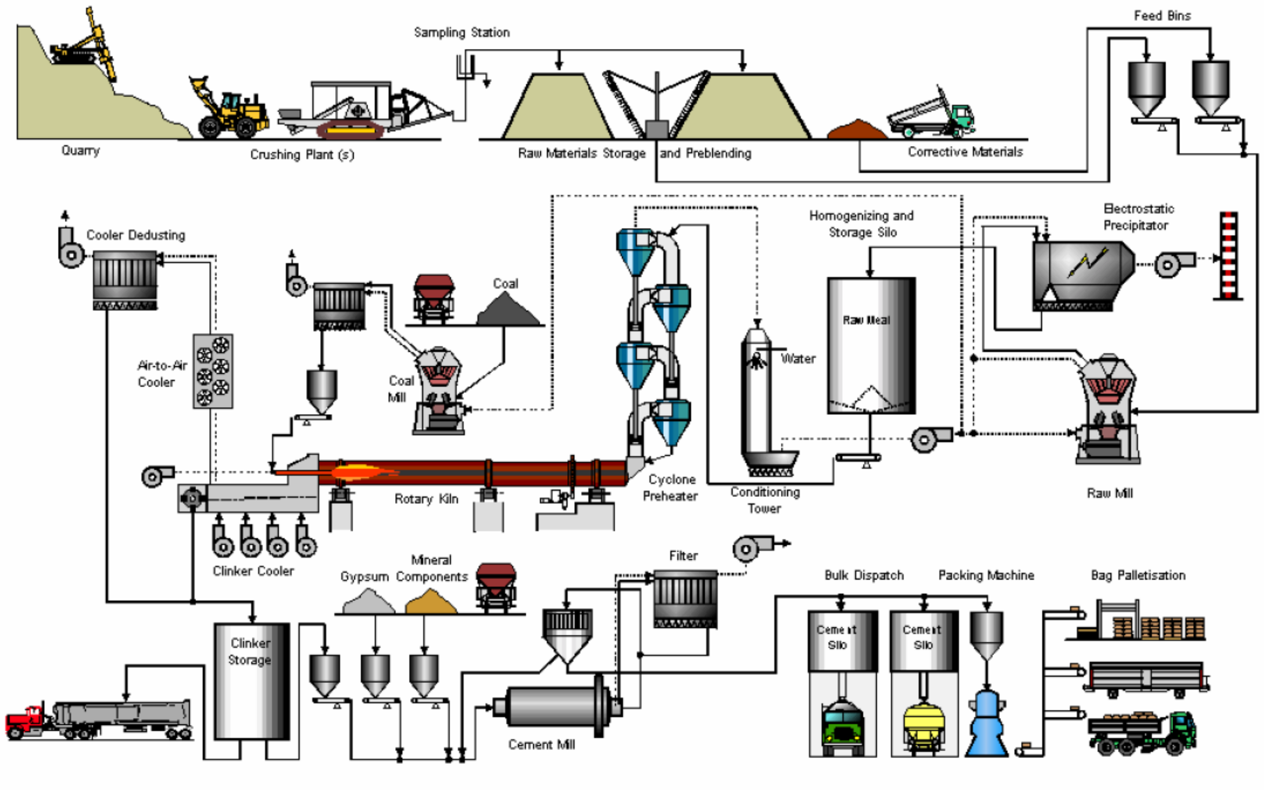
The blended and homogenised raw mix is then usually fed to a *gas-suspension preheater ahead of a short rotary kiln.* These preheaters consist of successive heat exchanger and collection cyclone stages arranged above one-another in series in a tower. Partial calcination of the raw mix takes place prior to the mix entering the rotary section of the kiln.

Additionally a *precalciner* can be integrated between the kiln and the suspension preheater which ensures complete calcining of the raw mix before it enters the kiln. Precalciners increase kiln capacity and reduce energy consumption.

Fuel consumption is lowest in this process and is in the range of 2750-4000 kJ/kg of clinker. The power consumption is in the range of 120-125 kWh/tonne of cement (Chemical Universe, 2007). However, the most modern, dry-process cement plants can have power consumption in the range 80-100 kWh/tonne of cement by deployment of compression grinding equipment for raw and finish milling rather than ball mills. Production of blended cements is also widely used to reduce unit electrical energy consumption.

A simplified flow diagram of a typical dry process cement kiln is shown in Figure 2-6.

Figure 2-6: Production of Cement by the Dry Process (CEMBUREAU, 1999)



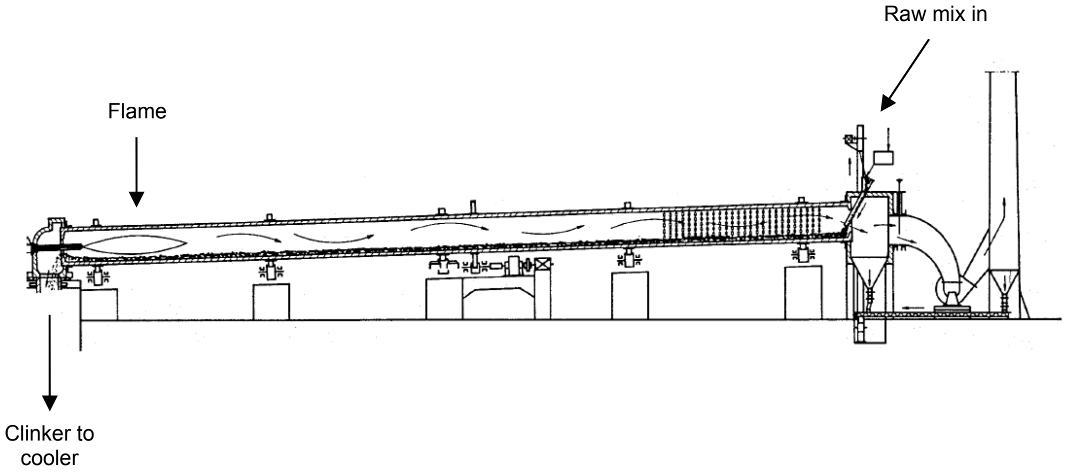
(iii) Key Kiln System Technologies

Rotary Kilns

In industrialised countries the ground mixture of raw materials for cement production are processed in rotary kilns. These rotary kilns form the core element of industrial cement production. They are refractory lined, slightly inclined rotating steel cylinders that are internally fired with fossil fuels and designed to produce clinker through the intense heating of the raw materials. They can be fed with slurry, crushed filter cakes, nodules or dry raw mix and are thus suitable for all process types. Raw mix is fed into the upper, cool end of the kiln while fuels are normally fed into the lower, hot end of the kiln. The inclination and rotation of the kiln causes the ground mixture of raw materials to move down the kiln from inlet to outlet in counter-current with the combustion gases generated by the firing of the fossil fuels in the lower, hot end of the kiln. In this way the ground mixture of raw materials are progressively heated as they travel down the length of the kiln being successively dried, preheated, calcined and combined into the cement clinker.

The first rotary kilns were long wet kilns (Figure 2-7) and all the heat consuming thermal processes required for the manufacture of cement occurred in the kiln. To obtain the necessary heat transfer for water evaporation, wet kilns must be of a sufficient length. The length of a long wet kiln is generally 32-35 times its diameter. The longest are 230 m long and these large units produce around 3500 t/d.

Figure 2-7: Long Wet Rotary Kiln (Adapted from CEMBUREAU,1999)

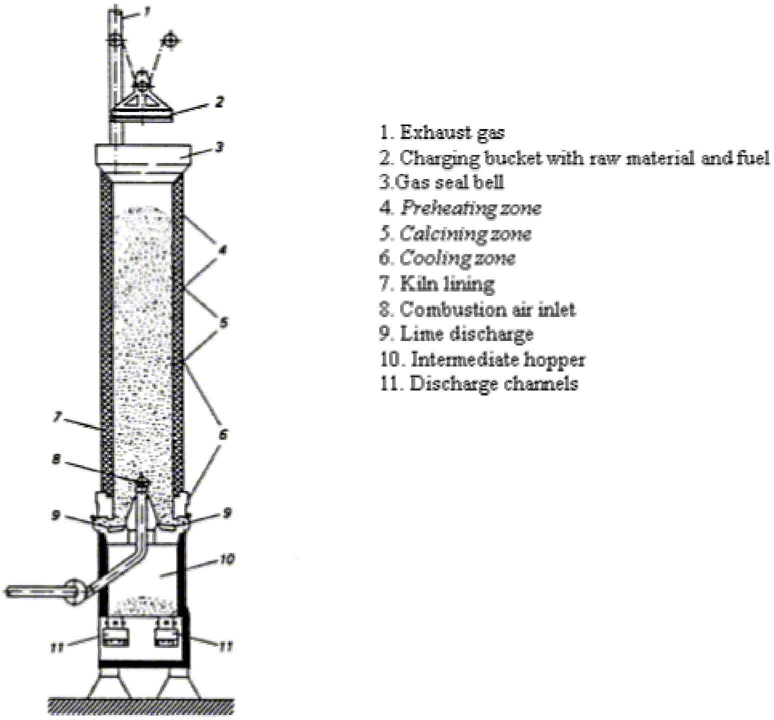


The introduction of the dry-process led to an optimisation of the cement clinker manufacturing process that allowed drying, preheating and calcining to take place in installations outside of the rotary kiln. Consequently dry process kilns are shorter than otherwise comparable wet process kilns. Their length is generally 11-20 times their diameter. Today the highest capacity dry-process cement kilns can have capacities of 10,000 t/d.

Shaft Kilns

Now predominantly found in India and China, shaft kilns (shown in Figure 2-8) consist of a refractory-lined, vertical cylinder 2-3 m in diameter and 8-10 m high. They are fed from the top and the raw mix passes through a short preheating, calcining and sintering zone at the top of the kiln where it is heated by hot combustion gasses. It is then cooled by air blown in from the base of the kiln and leaves as clinker at the bottom. The product quality from the vertical shaft kilns is generally lower than that from rotary kilns and capacity is less than 300 t/d. Shaft kilns are being phased out in China as they are incapable of producing cement with high enough strength development characteristics for high rise construction projects.

Figure 2-8: Shaft Kiln (VDI, 1984)



Preheating and Precalcination

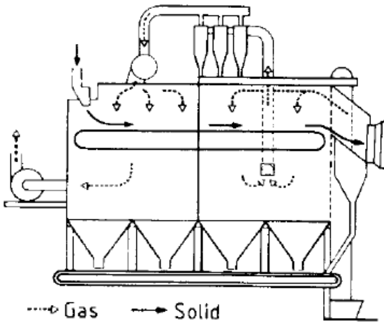
The introduction of the dry-process led to optimisation that allowed drying, preheating and calcining to take place in stationary installations outside of the rotary kiln. These 'additions' to the standard rotary kiln are discussed in more detail below.

Grate Preheater Kilns

The introduction of grate preheaters, also known as a Lepol kilns, in 1928 allowed part of the clinkering process to proceed outside of the rotary kiln for the first time. This allowed the rotary kilns to become shorter, reducing heat loses and increasing energy efficiency.

In a grate preheater (shown in Figure 2-9) nodules made by mixing dry raw material powder with water in a nodulising pan (semi-dry process) or pellets made from wet slurry filter cakes in an extruder (semi-wet process) are fed onto a moving grate that passes through two chambers: a drying chamber and a preheating chamber. A fan draws hot exhaust gases from the rear of the kiln into the top of the preheating chamber and through the bed of nodulised or pelletised kiln feed as it travels along the grate through the chamber to the kiln. This preheats the material before it enters the rotary kiln. A second fan then draws the exhaust gas from beneath the preheating chamber into the drying chamber above the grate and through the bed of wet nodules or pellets as they travel along the grate. This dries the nodules or pellets before they pass to the preheating chamber.

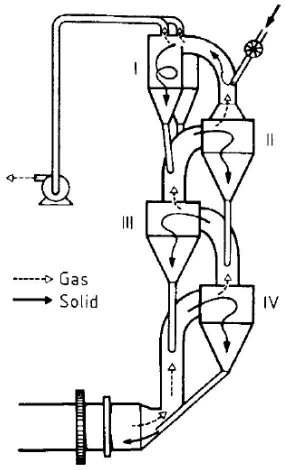
Figure 2-9: Grate Preheater (EC, 2001)



Suspension Preheater Kilns

Suspension preheaters (Figure 2-10) were developed as an alternative to travelling grate preheaters in the early 1950s. Preheating and partial calcination of the raw mix powder occurs in these preheaters by heat exchange between the hot kiln exhaust gases rising up the preheater from the rotary kiln and the raw mix powder descending the preheater. The process is thermally efficient since the suspension of the raw mix powder in the hot gases produces a large contact surface area that allows for efficient heat exchange between the gas and the powder.

Figure 2-10: Four Stage Suspension Preheater (EC, 2001)



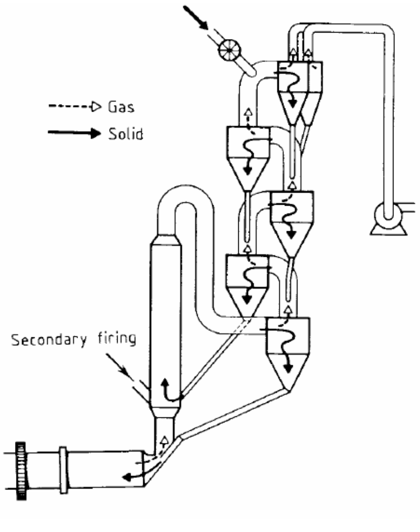
The key components of these gas-suspension preheaters are the gas risers between the successive preheater stages and the cyclones for collection of the preheated raw mix powder and passing the preheated raw mix powder down to the next stage of the preheater. Heat exchange between the hot exhaust gases and the raw mix powder takes place in gas risers as the raw mix is lifted by the exhaust gases into the collection cyclones. The raw mix powder is added to the uppermost gas riser of the preheater tower and lifted into the uppermost cyclone stage. The preheated raw mix powder and gases are separated in the cyclone by the centrifugal motion induced by the geometry of the cyclone, with the exhaust gases passing out of the top of the cyclone and the preheated raw mix passing out of the bottom of the cyclone to the next lower stage of the preheater. This alternate preheating and separation improves the heat exchange efficiency of the process. The number of these preheating stages in a preheater tower system varies between 2 and 6, with heat exchange and overall thermal efficiency increasing with each additional preheater stage.

The four stage cyclone preheater was standard technology in the 1970s. At that time many plants were built with capacities in the 1000-3000 t/d range. After passing down the preheater calcination of the kiln feed is already about 30% complete.

Precalciners

The precalciner is an evolution of the suspension preheater developed in the 1970s. It comprises a combustion chamber (precalciner) that is located between the suspension preheater and the kiln in which the preheated raw mix can be almost fully calcined prior to entering the kiln (Figure 2-11). The use of such a precalciner allows for a shorter kiln length and can reduce energy consumption by 5-10% compared with a kiln equipped only with a suspension preheater. Approximately 60-70% of the total fuel supply to the cement kiln is fired in the precalciner with the remainder fired in the main burner of the rotary kiln itself.

Figure 2-11: Five Stage Suspension Preheater with Precalciner (EC, 2001)



Kiln systems with five cyclone preheater stages and precalciner are considered ‘standard’ technology for today’s modern, dry-process plants. Typical capacities are between 3,000–5,000 t/d. In Asia there is an increasing trend towards the installation of higher capacity cement kilns with an increasing number of 10,000 t/d plants operating. There are plans for the installation of a 12,000 t/d cement kiln in the USA.

2.4.2 Emissions Associated with Cement Manufacturing

(i) Carbon Dioxide Emissions

Cement production is an energy intensive production process, accounting for 2% of the global primary energy consumption (Worrell *et al.*, 2001) and between 6-7% of global industrial energy use (IEA, 2007). Its reliance on carbon intensive fuels such as coal and petroleum coke in clinker making is responsible for substantial CO_2 emissions coming directly from the combustion of fossil fuels. CO_2 is also emitted from the calcination process and from the consumption of electricity generated from fossil fuels. Cement production accounted for 1.8 Gt/y of CO_2 emissions worldwide in 2005 (IEA, 2007).

It is estimated that about 5-6% of global CO_2 emissions from stationary sources originate from cement production (ECRA, 2007). Approximately half of these emissions are from the calcination process and the other half from the combustion process (this ignores emissions attributable to mobile equipment used in raw material mining, material and cement transportation, and plant site activities) (Hendriks *et al.*, 1998).

Process CO_2 Emissions from Calcination

As shown in the following chemical reaction, the conversion of limestone to CaO, the primary precursor to cement, unavoidably generates CO_2 as a by-product:

$$CaCO_3 \rightarrow CaO + CO_2$$

CO_2 emissions from cement production depend on the ratio of clinker/cement, which varies from 0.5 to 0.95 but are estimated at about 0.5 kgCO_2/kg clinker for calcination only (Worrell *et al.*, 2001).

CO_2 Emissions from Fossil Fuels

The majority of fossil fuel is used during the pyro-processing step of cement production. The amount of CO_2 emitted during this step is influenced by the following factors:

- the type of production process and its efficiency
- the nature of fuel used (e.g. coal, fuel oil, natural gas etc.)
- the ratio of clinker/cement and percentage of additives.

Table 2-5 shows CO_2 emissions from dry and wet-process cement production in relation to the clinker/cement ratio and nature of fuel used.

Table 2-5: Dry and Wet-process CO_2 Emissions from Fossil Fuels (Hendriks *et al.*, 1998)

Process emissions		**Process and fuel-related emissions (in kg/kg of cement produced)**[1]							
		Dry process				**Wet process**			
Clinker /cement ratio	*Clinker*	*Coal*	*Fuel Oil*	*Natural gas*	*Waste*[2]	*Coal*	*Fuel Oil*	*Natural gas*	*Waste*
55%	0.28	0.55	0.50	0.47	0.36	0.67	0.59	0.53	0.36
75%	0.38	0.72	0.66	0.61	0.47	0.88	0.77	0.69	0.47
95%	0.49	0.89	0.81	0.75	0.57	1.09	0.95	0.90	0.57

[1] Emissions from electricity consumption are included.

[2] Assumed to be a zero emission fuel.

CO_2 Emissions from Electricity Use

Grinding is the largest electricity demand in the cement industry with around 100 kWh/t of cement consumed in rotary kilns for grinding of raw materials, at the kiln and for grinding cement (IEA, 2007). The current best practice is reported to be around 80-90 kWh/t of clinker with most countries for which data is available falling within the range of 90-120 kWh/t of cement (IEA, 2007).

Global CO_2 Emissions from the Cement Production Process

In 1994 the average energy intensity associated with global cement production was about 4.8 MJ/kg cement, with the most energy intensive regions being in Eastern Europe and the Former Soviet Union (5.5 MJ/kg), followed by North America (5.4 MJ/kg) and the Middle East (5.1 MJ/kg) (Hendriks *et al.*, 1998). The resulting CO_2 emissions for that year were 1,126 Mt CO_2, 587 Mt CO_2 from process emissions and 539 Mt CO_2 from energy use (Hendriks *et al.*, 1998).

The carbon intensity of cement manufacture is subject to some variation in the literature. Hendriks *et al.* (1998) reported a carbon intensity of cement making of 0.81 kg CO_2/kg cement with the carbon intensity being about 10% higher than the average in India, North America and China. However, Mahasenan *et al.* (2005) report the average gross unit-based emissions for the industry to be 0.87 kg CO_2/kg with regional variation from 0.73 CO_2/kg in Japan to 0.99 kg CO_2/kg in the United States.

IEA (2007) reported that total CO_2 emissions per tonne of cement from energy and calcination in 2003-2004 ranged from about 0.65 kg CO_2/kg of cement in Spain, Brazil and Italy to 0.93 kg CO_2/kg of cement in the United States.

(ii) Other Emissions

Other emissions that are potentially associated with the cement industry and are therefore controlled include nitrogen oxides, sulphur dioxide and dust. Carbon monoxide, volatile organic compounds (VOCs), polychlorinated dibenzodioxins (PCDD), dibenzofurans (PCDF) and heavy metals are possible trace emissions.

Typical emissions from European cement kilns in 1997 are shown in Table 2-6.

Table 2-6: Emissions from European Cement Kilns (EC, 2001)

Emissions	mg/Nm^3	Kg/tonne clinker	Tonne/year
NO_x (as NO_2)	<200-3,000	<0.4-6	400-6000
SO_2	<10-3,500	<0.02-7	20-7000
Dust	5-200	0.01-0.4	10-400
HF	<0.4-5	<0.8-10 g/t	<0.8-10
HCl	<1-25	<2-50 g/t	<2-50
PCDD/F	<0.1-0.5 ng/Nm^3	<200-1000 ng/t	< 0.2-1 g/year
Metals			
∑ (Hg, Cd, Tl)	0.01-0.3 (mainly Hg)	200-600 mg/t	20-600 kg/year
∑ (As, Co, Ni, Se, Te)	0.001-0.1	2-200 mg/t	2-200 kg/year
∑ (Sb, Pb, Cr, Cu, Mn, V, Sn, Zn)	0.005-0.3	10-600 mg/t	10-600 kg/year

Note: Mass figures are based on 2000 m^3/tonne clinker and 1 million tonnes clinker/year. Emission ranges are one –year averages and are indicative values based on various measurement techniques. O_2-content is normally 10%.

2.4.3 Kiln Flue Gas Clean-up Methods

The exhaust gases from a kiln amount to between 1,350 – 3,000 Nm^3/tonne of clinker depending on the type of kiln (CEMBUREAU, 1999). Depending on the local discharge requirements these exhaust gases require treatment before discharge. A summary of flue gas clean-up methods that are employed (to varying degrees) within the cement industry is shown inTable 2-7. These methods are discussed in further detail in the sections below.

Table 2-7: Common Flue Gas Clean-up Methods (Adapted from EC, 2001)

Technique	Kiln system applicability	Reduction efficiency (%)	Reported emissions		Reported costs (10^6€/tonne of clinker)[3]	
			mg/m^3 [1]	kg/tonne [2]	Investment	Operating
Dust						
Electrostatic precipitators	All	-	5-50	0.01-0.1	2.1-4.6	0.1-0.2
Fabric filters	All	-	5-50	0.01-0.1	2.1-4.6	0.15-0.35
NO_x						
SNCR	Preheater and precalciner	10-85	200-800	0.4-1.6	0.5-1.5	0.3-0.5
SCR	Possibly all	65-95 [6]	100-500	0.2-0.4	ca.2.5 [4] 3.5-4.5 [5]	0.2-0.4 [4] uncertain [5]
SO_x						
Absorbent addition	All	60-80	400	0.8	0.2-0.3	0.1-0.4
Dry scrubber	Dry	Up to 90	<400	<0.8	11	1.4-1.6
Wet scrubber	All	>90	<200	<0.4	6-10	0.5-0.1
Activated carbon	Dry	Up to 95	<50	<0.1	15 [7]	uncertain

[2] kg/tonne clinker: based on 2000 m^3/tonne clinker

[3] Investment costs 10^6 euros and operating cost in euros/tonne of clinker. Dust: to reduce emissions to 10-50 mg/m^3 from initial emissions of up to 500 g dust/m^3. NO_x: to reduce emissions from initial levels of up to 2000 mg NO_x /m^3.

[4] Costs estimated by Ökopol for a full scale installation (kiln capacities from 1,000 – 5,000 tonne clinker/day and initial emissions from 1,300 – 2,000 mg NO_x /m^3), operating costs ca. 25% lower than for SNCR

[5] Costs estimated by CEMBUREAU

[6] There has been limited study of the application of SCR to the cement manufacturing process and the literature that exists provides conflicting data on SCR performance. The wide efficiency range shown reflects this uncertainty.

[7] This cost also includes an SNCR process, referring to a kiln capacity of 2000 tonnes clinker/day and initial emissions of 50-600 mg SO_2 /m^3

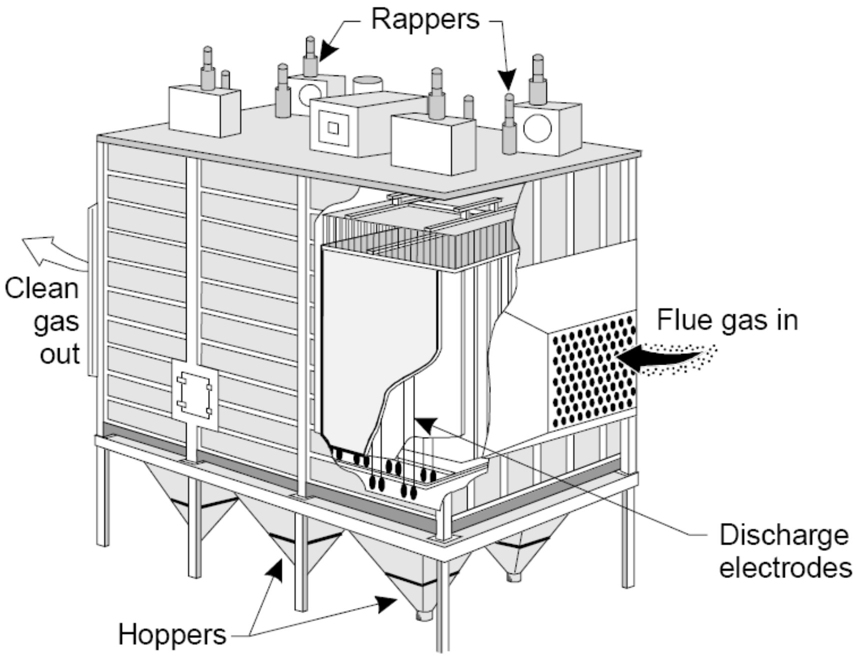
(i) Techniques for Controlling Dust Emissions

Cement kiln flue gases carry a large amount of dust - typically 3,000 mg/m^3. Whilst various de-dusting techniques have historically been used in cement manufacturing, the modern plants now almost exclusively use electrostatic precipitators and bag filters. These are recognised as the best available techniques (BAT) for reducing dust emissions (EC, 2001).

Electrostatic Precipitators

Electrostatic Precipitators (ESPs) use electrostatic forces to separate dust from flue gases. An electrostatic field is generated across the path of particulate matter in the air stream. Particles become negatively charged and migrate towards positively charged collection plates (Figure 2-12). Dust accumulated on the collection plates is dislodged and collected by electrode rapping.

Figure 2-12: An Electrostatic Precipitator (EPA, 2007)



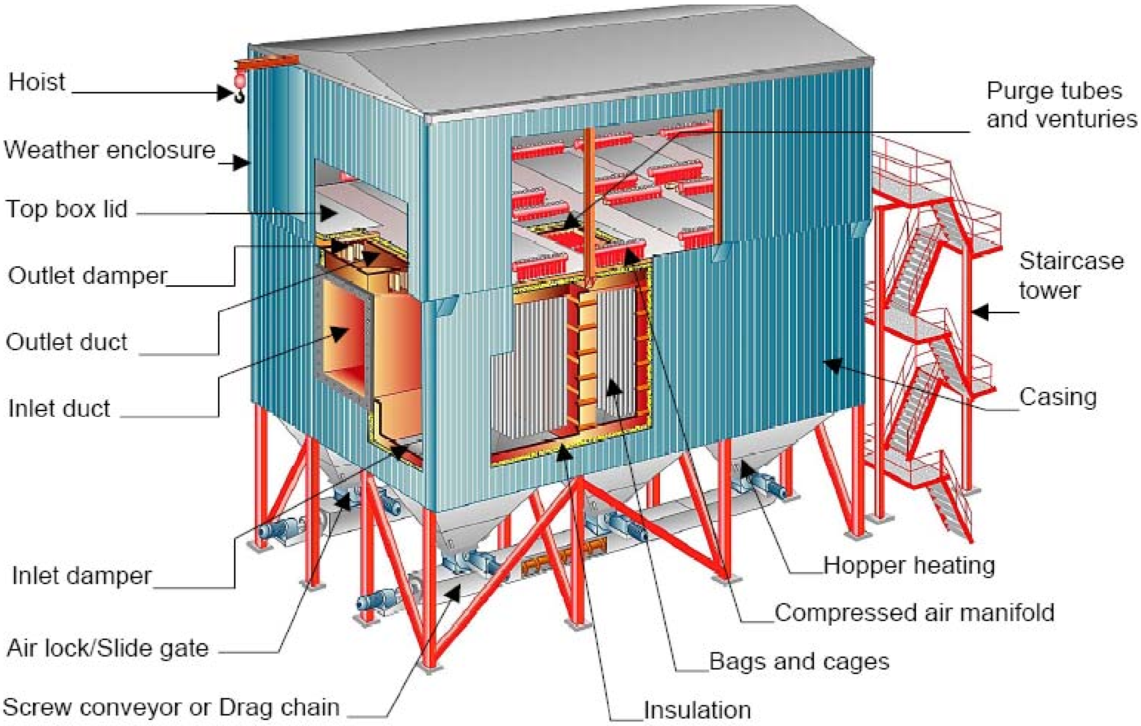
A properly designed ESP combined with gas conditioning and an optimised ESP cleaning regime can reduce dust levels down to 5-15 mg/m^3 (Federal Environment Agency, Austria 1997).

EC (2001) reports that ESPs are characterised by their ability to operate under conditions of high temperatures (up to approximately 400 ºC) and high humidity. However, Mott MacDonald understands that the performance of ESPs at high temperatures with dry exhaust gases is problematic and that they operate most efficiently with cool (~100ºC) moist gas.

Bag Filters

Bag filters (Figure 2-13) for de-dusting gaseous waste streams comprise a fabric membrane which is permeable to gas flowing towards and through it but which will retain particulates. As dust accumulates on the fabric membranes (bags) of these 'bag filters' the resistance to gas flow increases and thus regular cleaning is required to maintain performance. The most common cleaning methods include reverse airflow, mechanical shaking, vibration and compressed air pulsing.

Figure 2-13: A Pulse Jet Fabric Filter (FLSmidth)



Modern fabric filters can reduce dust emissions to below 5 mg/m^3 (Federal Environment Agency, Austria 1997).

High temperature applications of bag filters require the use of more exotic fabric types than are ‘normally’ supplied (EC, 2001).

(ii) Techniques for Controlling NO_x Emissions

NO_x is a generic term used to refer, collectively, to all the gaseous oxides of nitrogen, nitrogen monoxide (NO), nitrogen dioxide (NO_2), nitrous oxide (N_2O) and nitrogen tetroxide (N_2O_4). There are other less common oxides of nitrogen with variable stoichiometry represented by the formula, N_xO_y.

NO_x are formed during the combustion of fuel by oxidisation of the molecular nitrogen of the combustion air (thermal NO_x) as well as the nitrogen compounds in the fuels (fuel NO_x) and raw materials. Thermal NO_x forms at temperatures above 1200ºC and is produced mainly in the kiln burning zone where it is hot enough to achieve this reaction (EC, 2001). The amount of thermal NO_x produced in the burning zone is related to both burning zone temperature and oxygen content (air excess factor) (EC, 2001). Fuel NO_x is generated by the combustion of the nitrogen present in the fuel. In a precalciner the prevailing temperature is in the range 850-900ºC, which is not high enough to form thermal NO_x, but fuel NO_x will occur (EC, 2001).

The temperatures in the burning zone of a cement kiln and the long residence time of the exhaust gases in the process favour the formation of NO. Less than 10% of the NO_x emissions from a kiln system are composed of NO_2 (EC, 2001). Rushworth (2008) reported the NO_2 fraction as less than 5% and Leibacher *et al.* (2006) reported that in the order of 98% of the NO_x present in the flue gas from cement kiln is in the form of NO. However, these NO emissions are oxidised to NO_2 in the atmosphere after emission from the kiln. Ultimately it is the NO_2 emissions which are responsible for acid rain, smog and ozone formation and this necessitates limits on the emissions of NO_x from cement kilns. NO_x emissions from cement kilns are therefore converted into their NO_2 emission equivalents.

The NO_x emissions (expressed as NO_2) associated with cement kilns vary generally between <200 and 3000 mg NO_2/Nm^3 (dry gas, 0ºC, 101.3 kPa, 10% O_2) (EC, 2001).

NO_x emissions can be reduced via primary process optimisation measures and some of these techniques are listed below (CEMBUREAU, 1999):

- optimisation of the clinker burning process
- computer based expert system for kiln operation
- optimisation of main burner (low NO_x burner)
- multi-stage combustion for in-line precalciners.

NO_x emissions in the kiln flue gas can also be abated using the clean up options detailed below.

Selective Non-catalytic Reduction (SNCR)

NO_x in flue gases can be converted into N_2, and water by reaction with NH_4–X compounds injected into the flue gas. The ammonia-containing solution may be supplied in the form of anhydrous ammonia, aqueous ammonia or urea. To work effectively SNCR requires:

- That NH_4–X be introduced into the flue gas at temperatures of between 800 – 1000 ^oC.
- That exhaust streams have a relatively high concentration of NO_x as opposed to other potential reactants.
- An oxidising or fuel lean atmosphere.
- The accurate injection of the optimum quantity of reagent NH_3.

This optimal temperature window and other conditions can be obtained in suspension preheater kilns, precalciner kilns and potentially in some Lepol kilns. It is important for the reaction to occur within the optimal temperature range, as if the temperature falls then unconverted ammonia may be emitted (known as 'ammonia slip') and react in the atmosphere to form visible plumes above the exhaust gas stack. However, if the temperature is too high then the injected ammonia can itself react to form additional NO_x thus increasing the emission problem.

SNCR is routinely used at some European cement plants and there are currently at least two cement kilns in the USA using SNCR to control NO_x emissions. As such it is generally considered to be 'commercially available' (Schreiber *et al.*, 2006) for modern preheater/precalciner kiln systems. Most SNCR installations operating today are designed for NO_x reduction rates of 10-50% and emission levels of 500-800 mg/m^3 of NO_x. However, two plants in Sweden have been utilising SNCR to comply with tight Swedish emission limits since 1997 and achieving control efficiencies of 80-85% which corresponds to emissions of less than 200 mg/m^3 of NO_x – reaching flue gas purity levels, at least in these first demonstrations, consistent with the current requirements of post-combustion capture, discussed in section 3.3.4.

Selective Catalytic Reduction (SCR)

SCR has been successfully employed in the power industry and involves the injection of ammonia-based reagents into the flue gas stream in the presence of a catalyst. Common reagents include anhydrous ammonia, aqueous ammonia and urea. NO_x compounds are converted to N_2 and water and the catalyst allows these reactions to occur at lower temperatures (300^oC – 400^oC) than those required for SNCR.

The application of SCR technology to cement plants is relatively new for the industry. There have been several small pilot studies and two large scale installations; one at the Solnhofen cement works in Germany and the other at the Cementeria di Monselice in Italy.

Solnhofen was the first plant to adopt the full scale technology and actually has both SCR and SNCR on-site. A recent study into the application of SCR at cement plants (Schreiber *et al.*, 2006) concluded that SCR technology does not, on currently available experience, appear to be a more effective NO_x treatment method than SNCR, that SCR suppliers could not provide performance guarantees and that considerable pilot testing would be required at any given plant to optimise SCR performance at an unknown cost. The report concluded that SCR systems could not yet be considered to be 'commercially available' technologies in the cement industry.

At the time of the report the plant had not reached the 200 mg/Nm^3 imposed target for NO_x emissions. It was also not operating the SCR unit suggesting that the SNCR unit was satisfactory to meet the operating permit emission rate of 500 mg/Nm^3. However, many modifications and improvements were made to the SCR system to cater for the high dust environment experienced in cement plants. The main new item was equipment to remove the dust clogging the honeycomb catalyst. Many lessons were learnt for introducing the technology to other cement plants.

Using the modified design from Solnhofen, Swiss company ELEX installed another full scale SCR system at the Monselice Cement Plant in Italy. A range of results were achieved from the Monselice plant ranging from 95% NO_x removal efficiency down to 43% removal (Leibacher *et al.*, 2006). A 25% ammonia solution was used as the reductant injected in one of the early stages of the preheater. The ammonia slip through the system was negligible due to the long residence time of the vapourised ammonia.

(iii) Techniques for Controlling SO_x Emissions

Sulphur dioxide, SO_2, is formed:

i. during the combustion of fuel, released in the burning zone of the kiln and the precalciner (if present); and

ii. due to the oxidisation of pyrite/marcasite and organic sulphur in raw feed materials in the preheater or the kiln inlet of long wet or dry kilns.

However, the SO_2 formed by the combustion of fuel is absorbed by the intimate contact between the combustion exhaust gases and the basic feed material in the preheaters of modern cement kilns. In these modern kilns any SO_2 emissions from the cement plant are predominantly due to oxidation of pyrite/marcasite in the feed materials in the upper stages of the preheater. Modern cement plants which do not have these sulphide minerals in their raw materials have very low emissions of SO_2.

The SO_2 emissions from kilns vary between <10 and 3,500 mg/m^3 (daily average basis and standard conditions of 0ºC, 101.3 kPa, 10% O_2 and dry gas) with the emissions of below 10 mg/m^3 from cement kilns which do not have sulphide minerals in their raw materials (EC, 2001).

SO_2 emissions can be reduced via primary process optimisation measures and some of these techniques are listed below (CEMBUREAU, 1999):

- reduction in sulphur content in fuels and kiln feed
- optimisation of the clinker burning process
- addition of slaked lime to the kiln feed with the use of a dry scrubber if necessary.

SO_2 emissions in kiln flue gas can also be abated using the clean up options detailed below.

Addition of Absorbents

Adding absorbents, in dry or wet form, such as slaked lime, quicklime or activated fly ash with high CaO content to the exhaust gas of a kiln can absorb SO_2. Where emissions are below 1,200 mg/m^3 the most efficient way of utilising slaked lime is to add it to the kiln feed prior to clinker burning and let the SO_2 absorption occur in the kiln, rather than inject it into the flue gases after sintering. However, where SO_2 concentrations are higher this is not economically feasible.

SO_2 reductions of 60-80% can be achieved by absorption injection in suspension preheater kiln systems and for emission levels of up to 1,200 mg/m^3 it is possible to reduce eventual flue gas emissions to around 400 mg/m^3 (EC, 2001). The technique is in principle applicable to all kiln systems and is currently in use at several plants.

Dry Scrubbers

Where emission levels exceed 1,500 mg/m^3 a separate scrubber may be required to reduce emission to below 500 mg/m^3. A 'dry' scrubber can be used prior to the raw meal entering the kiln.

One type of scrubber uses a venturi reactor column to produce a fluidised bed of slaked lime and raw meal. The intensive contact between gas and absorbent, the long residence time and the low temperature allow for efficient absorption of SO_2 and absorbent rates of 90% can be achieved (EC 2001).

As of 2001 there was only one SO_2 dry scrubber in operation in Europe at the HCB-Untervaz plant in Switzerland. It has a 4 stage suspension preheater kiln system and a maximum capacity of 2,000 tonnes of clinker per day. Unabated SO_2 emissions would be around 2,500 mg/m^3 in direct operation and 2,000 mg/m^3 in compound operation. Using dry scrubbers the abated average emission level was reported to be 385 mg/m^3 in 1998 (EC, 2001).

Wet scrubbers

Wet scrubbers are the most commonly used technique for flue gas desulphurisation in coal fired power plants. Calcium carbonate, hydroxide or oxide in a liquid or slurry is used to absorb SO_2 from the flue gases. The slurry is sprayed in a counter current direction to the exhaust gas and collected in a recycle tank at the bottom of the scrubber where the formed sulphite is oxidised with air to sulphate and forms gypsum, $CaSO_4.2H_2O$. The gypsum is separated and used as set controlling agent in cement milling, while the water is returned to the scrubber.

Wet scrubbers have been installed at various cement plants throughout Europe and flue gases with peak SO_2 concentrations of more than 2,000 mg/m^3 have been treated to produce exhaust gases with SO_2 concentrations below 200 mg/m^3 (CEMBUEAU, 1999).

The wet scrubber also significantly reduces the HCl, residual dust, metal and NH_3 emissions (EC 2001).

Activated Carbon

Various pollutants such as SO_2, organic compounds, metals, NH_3 (and NO_x if NH_3 is present), NH_4 compounds, HCl, HF and residual dust (after an ESP or fabric filter) can be absorbed from flue gases using activated carbon (EC, 2001). The flue gases pass through an activated carbon filter constructed as a packed bed and the used activated coke is periodically replaced with fresh absorbent.

As of 2001 there was only one activated carbon filter installed in Europe at Siggenthal in Switzerland. This 4 stage cyclone preheater kiln has a capacity of 2,000 tonnes of clinker per day and prior to installation of an activated carbon filter was emitting SO_2 in concentrations of between 50 and 600 mg/m^3. Following installation emissions at the outlet were significantly below 50 mg/m^3 (EC, 2001). Activated carbon filters can be fitted to all dry kiln systems.

(iv) UK National Emission Limits

Emission limits for discharges from cement production vary across the globe with the measurement criteria (averaging time, reference conditions measurement techniques, compliance criteria etc.) different from one country to another. Tables 2-8 and 2-9 indicate the emission limits for dust, SO2 and NOx in the UK for production of cement (EC, 2001).

Table 2-8: UK[1] Emission Limits for Dust for the Production of Cement

Plant	Unit	Kiln Stack	Clinker Cooling	Cement Grinding	Other Point Sources
New/modified	mg/Nm^3	40 [2]	50 [2]	40 [2]	50 [2]
Existing [3]	mg/Nm^3	-	-	-	-

[1] IPC Guidance Note S2.3.01
[2] 'Benchmark releases'
[3] Benchmark releases are, in particular, not applicable to existing plant but are a factor in considering appropriate limits.

Table 2-9: UK[1] Emission Limits for SO_2 and NO_x for the Production of Cement

Plant	**Unit**	**SO_2 Normal situation**	**SO_2 S-rich raw materials**	**NO_x**
New/modified	mg/Nm^3	200 [2]		900 [2]
Existing [3]	mg/Nm^3	Note[3]	600-2500 [4]	500-1200 [3, 5]

[1] IPC Guidance Note S2.3.01
[2] 'Benchmark releases'
[3] Benchmark releases are, in particular, not applicable to existing plant but are a factor in considering appropriate limits.
[4] Limit values reflect the actual levels of releases. Daily averages and reference condition of dry gas and actual O_2 content.
[5] Actual releases, daily averages, not all plants currently have limits.

(v) European Discharge Limits

The Integrated Pollution Prevention Control (IPPC) Directive (EC/96/61) employs a permitting system to achieve an integrated approach to controlling the environmental impacts of industrial activities. In order to achieve permitting the permitting authority will consider whether the plant meets the BAT emission limits. Permits may be issued for a plant that does not meet the BAT emission limits but a good justification is required. Table 2-10 summarises the BAT emission limits for cement plants (EC, 2001).

Table 2-10: BAT Emission Levels (EC, 2001)

Emission	**Best Available Techniques (BAT)**	**Emission Level[1]**
NO_x	Combination of general primary measures, primary measures to control NO_x emissions, staged combustion and selective non-catalytic reduction (SNCR)	200-500 mg NO_x/m^3 (as NO_2)
	Selective catalytic reduction (SCR)	100-200 mg NO_x/m^3 (as NO_2)
SO_2	Combination of general primary measures and absorbent addition for initial emission levels not higher than about 1200 mg SO_2/m^3 and a wet or dry scrubber for initial emission levels higher than about 1200 mg SO_2/m^3	200-400 mg SO_2/m^3
Dust	Combination of general primary measures and efficient removal of particulate matter from point sources by application of electrostatic precipitators and/or fabric filters	20-30 mg dust/m^3

[1] Emission levels are expressed on daily average basis and standard conditions of 273 K, 101.3 kPa, 10% oxygen and dry gas

3 CO_2 Capture Processes

3.1 Introduction

In this section the leading technologies that could be used to capture CO_2 in cement plants are described. These technologies include post-combustion, oxy-combustion and pre-combustion CO_2 capture. The reasons why pre-combustion capture is not suitable for capture of CO_2 from cement plants are explained.

3.2 Overview

Hendriks *et al.* (1998) summarise the methods by which CO_2 emissions can be reduced in the production of traditional cement as:

1. Improvement of the energy efficiency of the process.
2. Shifting to a more energy efficient process (e.g. from a (semi) wet to (semi) dry process).
3. Replacing high carbon fuels by low carbon fuels.
4. Replacing fossil fuels with alternative fuels (e.g. waste-derived fuels).
5. Applying lower clinker/cement ratio (i.e. increasing the ratio additives/cement).
6. Removal of CO_2 from the flue gases (CO_2 capture).

The cement industry has, for various reasons and to varying degrees, implemented or begun to implement the first five strategies. This has resulted in a reduction over time in the CO_2 emitted per tonne of clinker produced. In Japan and Europe, for example, energy efficiency improvements led to a reduction in the energy intensity of clinker production of roughly 30% between the 1970s and the early 1990s (Japan Cement Association, 2006; CEMBUREAU, 1998). However, these efficiency improvements appear to have plateaued as incremental improvements became less cost effective and global emissions of CO_2 per tonne of clinker have remained essentially stable since 1990 (WBCSD, 2002b). In the face of rising production this has led to an inevitable increase in absolute CO_2 emissions associated with cement manufacture.

Large predicted growth in demand for cement when coupled with realistic assumptions on a hypothetical global CO_2 emissions reduction regime imply that the global cement industry may face significant emission reduction requirements in the future (Mahasenan *et al.*, 2005). Although there remains undoubted scope to improve CO_2 efficiency through the application of existing strategies (see IEA GHG, 1999), the industry may need to seriously consider the application of CO_2 capture to meet significant reduction targets particularly as this technology offers a way to secure deep emission cuts. The application of CO_2 capture technologies in the cement industry is the focus of this Report.

3.3 CO_2 Capture Processes Descriptions

3.3.1 Overview

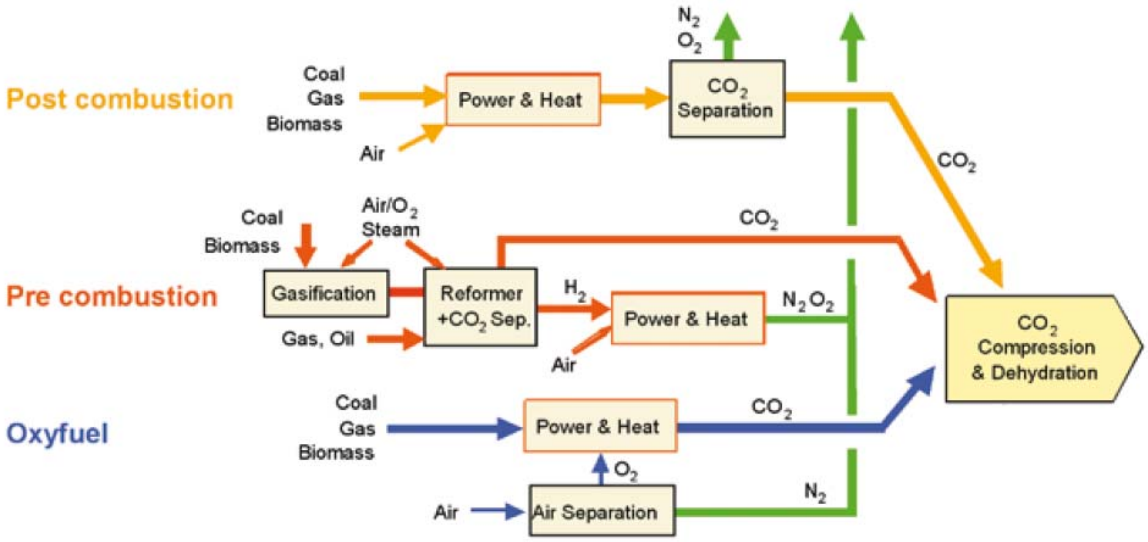
CO_2 capture produces a concentrated stream of CO_2 that can then be transported to a CO_2 storage site where it can remain in isolation from the atmosphere. The technology is most suited to large single point sources of CO_2 and its application is often discussed in relation to the capture of CO_2 emissions from power plants which are the biggest (stationary) sources of CO_2 emissions. Cement plants are the next biggest source of emissions and thus, there is increasing interest in the potential application of CO_2 capture technology to the cement industry. Cement kiln flue gases have a relatively high CO_2 concentration (approximately 15-30%) that should provide some advantages over CO_2 capture from the power generation industry.

The approach to carbon capture can be split into three distinct categories:

- **Pre-combustion** – conversion of fuel to H_2 and CO_2 prior to combustion
- **Oxy-combustion** - burning fuel in pure oxygen rather than air with recycle of flue gas to control combustion temperature
- **Post-combustion** – capture of CO_2 from kiln flue gases.

A simplified process schematic for the three routes is shown in Figure 3-1. Although CO_2 is removed from the system at opposite ends of the process in pre- and post-combustion, the technologies available to achieve this CO_2 separation from other process streams are similar and are dealt with individually in section 3.4. Ideally oxy-combustion should produce a flue gas with a very high concentration of CO_2 and thus negate the need for a further CO_2 separation step although some purification of the flue gas stream is likely to be required.

Figure 3-1: CO_2 Capture Processes (IPCC, 2005)



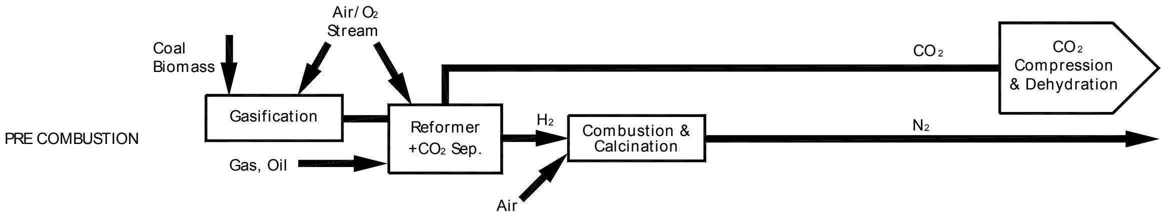
Post combustion capture has been applied in industrial manufacturing processes since the 1920s (Kohl and Neilsen, 1997) and the same solvent technology is widely used in refining and gas processing although predominantly with gases under reducing conditions rather than with gases containing significant quantities of oxygen. However, given such experience this approach to CO_2 capture is considered to be 'commercially available' although it has yet to be applied at a large power or cement plant on a commercial basis. In contrast pre- and oxy-combustion technologies are less well developed and are considered to be 'emerging technologies' which will require further development before finding commercial acceptance.

Each of these approaches to the process of CO_2 capture and its application at cement plants is discussed in further detail in the sections below and is followed by a review of commercially available and emerging CO_2 separation technologies.

3.3.2 Pre-combustion

Pre-combustion CO_2 capture generally involves three stages. Firstly, a hydrocarbon-containing primary fuel such as coal, natural gas or biomass is reacted with either steam or oxygen to produce a mixture of hydrogen and carbon monoxide known as syngas. The process of reacting the fuel with steam is known as 'reforming' while reaction with oxygen is known as 'partial oxidation' when applied to gaseous or liquid fuels and 'gasification' when the primary fuel is a solid. The second step in the process is to catalytically react the carbon monoxide with further steam to produce CO_2 and more hydrogen. This is known as the 'water gas shift reaction' and the result is a stream of hydrogen with a high concentration of CO_2 – typically 15 to 60% (dry basis). Finally, the CO_2 is removed from the CO_2/H_2 mixture using one of the capture methods discussed in section 3.4 leaving hydrogen to be cleaned for use as a carbon free fuel. The process is shown in Figure 3-2.

Figure 3-2: Pre-combustion CO_2 Capture Process for Cement Manufacture (Adapted from IPCC, 2005)



Although pre-combustion capture could be used to 'de-carbonise' the fuel source used in the cement manufacturing process it is considered to have two major drawbacks when applied to the cement industry that make its application unsuitable.

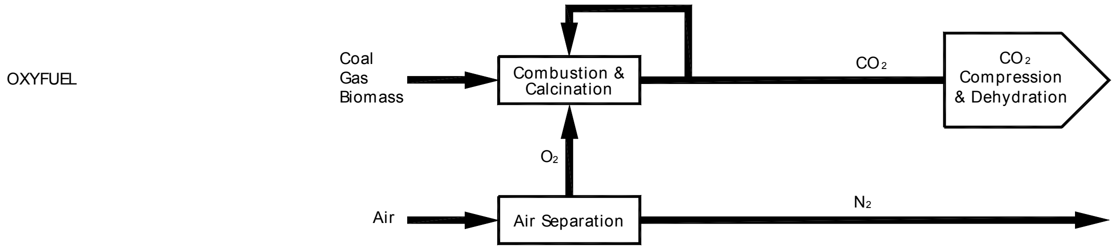
Firstly, since a significant proportion of the CO_2 generated through cement manufacture is due to the calcination process, which takes place after the CO_2 associated with the fuel has been separated, the exit gases from the kiln would still contain large quantities of CO_2 despite the application of pre-combustion capture. This would require additional and expensive CO_2 removal (given the lower CO_2 concentrations) by application of one of the other CO_2 removal technologies. Secondly, there is little opportunity to integrate the fuel conversion process into the existing production process - the cement manufacturing process does not, unlike power generation facilities, have an existing steam cycle it can tap into and this will increase costs associated with the process.

For these reasons pre-combustion capture is considered not to be an attractive method by which to reduce CO_2 emissions associated with the cement industry (IEA GHG, 1999; IPCC, 2005; and Hegerland *et al.*, 2005) and is not considered further in this report.

3.3.3 Oxy-combustion

In the process of oxy-combustion (Figure 3-3) fuel is burnt in almost pure oxygen, rather than air. The oxygen required for combustion is obtained by the removal of nitrogen from air, which is typically 78% nitrogen, in an Air Separation Unit (ASU).

Figure 3-3: Oxy-combustion CO_2 Capture Process for Cement Manufacture (Adapted from IPCC, 2005)



After combustion the flue gas is cooled to condense water vapour and the resulting flue gas has a high CO_2 concentration (80-98% depending on fuel and particular oxy-combustion process). The flue gas will also contain other products associated with fuel combustion such as NO_x, SO_x, inert gases from the fuel or leakage into the system and diluents used in the oxygen feed. Fuel burnt in a pure oxygen environment has a very high stoichiometric combustion temperature (3,500^oC). To control this a proportion of the flue gas can be recycled back to the burner inlet, as a diluent, to moderate the temperature.

The high concentration of CO_2 in the flue gas eliminates or reduces the need for costly post combustion CO_2 capture from the flue gases (see section 3.4 for a discussion on CO_2 separation technologies). If the flue gas is simply captured and stored without further processing then other trace pollutants (such as NO_x and SO_x) can also be compressed and stored without the need for additional costly steps to remove these components (Marin *et al.*, 2003). However, recent work by Air Products (White, 2007) indicates that most of the SO_x and NO_x will be removed as acid during the CO_2 compression.

The production of oxygen is a key step in an oxy-combustion system. The standard technology capable of delivering the high quantities of oxygen required for oxy-combustion is the distillation of oxygen from air at cryogenic temperatures. In a typical cryogenic air separation plant, air is compressed and purified before being cooled through a collection of heat exchangers and finally separated into its fractions in a double distillation column. The process has been in use for over 100 years and can reliably deliver high quantities of oxygen for industrial processes. Typical power consumption for the delivery of 95% O_2 at low pressure is 200-240 kWh/tO_2 (IPCC, 2005). Commentators have argued that this energy cost makes CO_2 capture via oxy-combustion uncompetitive compared with post-combustion technologies (Canadian Clean Power Association, 2004). There are a number of processes being explored that could reduce the cost of oxygen production and these are discussed briefly under emerging technologies in section 3.4.3.

Although individual parts of the oxy-combustion process are in use in various industries, including the key step of separating O_2 from air, the use of the technology specifically for the purposes of CO_2 capture in an integrated process has yet to be deployed on a commercial scale.

The application of the process to power generation cycles has attracted significant attention and there have been a number of studies (Yantovskii *et al.*, 1992; Mathieu, 2003; Jericha *et al.*, 2003; Nsakala *et al.*, 2003; Simbeck and McDonald, 2001; Kvamsdal *et al.*, 2004; Buhre *et al.*, 2005) and pilot tests (Babcock Energy Ltd *et al.*, 1995; Croiset and Thambimuthu, 2000; Anderson *et al.*, 2004) of various approaches to applying oxy-combustion in the process of electrical generation.

Among other examples, a 6 MW demonstration plant incorporating an oxy-combustion steam turbine cycle is now online (IPCC, 2005) and a 68 MW 'zero emissions' oxyfuel power plant developed by Dutch company SEQ Nederland B.V. is planned at Drachten in the Netherlands. A 30MWth oxyfuel pilot plant is also planned for operation in 2008 at Schwarze Pumpe, Germany and another 30 MW demonstration project is planned for operation in 2009 at Callide-A in Queensland, Australia .

Although in principle the process of oxy-combustion could be applied to the cement production process (Zeman and Lackner, 2006) with a mixture of oxygen and recycled CO_2 being fed to the burner in the kiln, existing research into oxy-combustion has focused predominantly on the power generation industry. Literature on the application of the technology to the cement industry is limited and no pilot projects exist at present.

The available literature has highlighted the following issues that may affect the use of oxy-combustion in the cement manufacturing process (Hendriks *et al.*, 1998; IEA, 2004):

- A process redesign may be required in order to prevent excessive equipment wear. This may make the application of this technology as a retrofit solution difficult.

- Where a precalciner is used a second combustion point using recycled CO_2 may be required.

- Effects on the process chemistry need to be fully assessed. In particular, the impact of a change of the CO_2 atmosphere on the calcination process must be fully understood. If, for example, a higher partial CO_2 pressure required a higher temperature to complete reactions this may shift more of the process into the kiln rather than the precalciner, since heat transfer is less efficient in the kin this may reduce production capacities at a plant.

- Whether or not the process can be sufficiently free of in-leaks to prevent air dilution of the concentrated CO_2 stream.

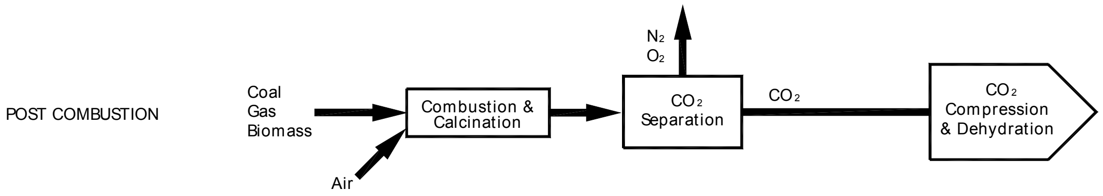
- During periods of start-up there may not be enough CO_2 from exhaust gases to recycle for flame cooling which may result in onsite CO_2 storage being required. It should be noted that recent developments for oxy-combustion power plants indicate that start up would normally be done using air with the flue gas being vented before switching over to recycled flue gas and oxygen. Hence, this would only be an issue if it was not environmentally acceptable to vent the flue gas during start-up.

All these factors will be discussed further in section 4.5.

3.3.4 Post-combustion

Post-combustion CO_2 capture adds a CO_2 removal stage onto the process of flue gas clean-up. Rather than being discharged directly to the atmosphere the flue gas is passed through equipment that separates out CO_2 which is then stored while the remaining flue gas is discharged to the atmosphere (Figure 3-4). CO_2 separation technologies are already widely applied in industrial manufacturing processes, refining and gas processing and it is expected that they could be applied to post-combustion capture of CO_2 from cement manufacturing flue gases.

Figure 3-4: Post-combustion CO_2 Capture Process for Cement Manufacture (Adapted from IPCC, 2005)



The post combustion capture of CO_2 from a flue gas can be achieved by a variety of different methods. The leading commercial technologies utilise a chemical process, that offers a high capture efficiency, selectivity and the lowest energy use and costs when compared with other existing and emerging capture processes. These capture technologies are reviewed in section 3.4.2.

Given the high CO_2 concentrations in cement plant flue gas (15-30%) post-combustion capture appears, on the one hand, to be well suited to use at a cement plant. However, some commentators (Thambimuthu, 2006; IEA GHG, 1999) have argued that chemical solvent scrubbing is not suited for use at a cement plant since the large quantities of low grade heat required for solvent regeneration are not generally available at a cement works. They suggest that given these constraints oxy- rather than post-combustion capture is a more attractive CO_2 capture route. In contrast, a recent concept study by Hegerland *et al.* (2006) which looked at retrofitting capture technologies to an existing cement works in Norway concluded that, given the stage of development of alternatives, post-combustion capture using amine solvents is the only viable option.

3.4 Types of CO_2 Capture Technology

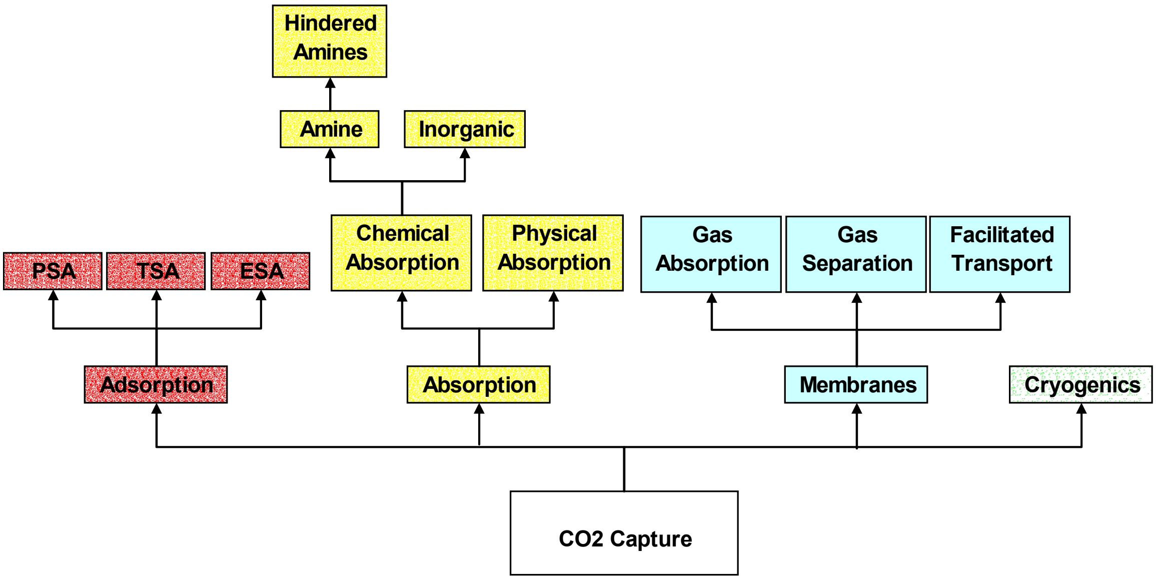
3.4.1 Introduction

There are four basic approaches to capturing CO_2 from process streams (Figure 3-5). These are as follows:

- separation with chemical/physical solvents (absorption)
- adsorption
- separation with membranes
- cryogenic separation.

The use of chemical or physical solvents to separate CO_2 is considered to be a commercially available technology and is widely used while adsorption, membranes and cryogenic separation are considered developing technologies and although used in niche applications are not considered commercially available for carbon capture at large scale power or cement plants. These commercial capture technologies are reviewed below and the developing technologies are briefly discussed.

Figure 3-5: Types of CO_2 Capture Technology (Adapted from Gupta *et al.*, 2003)



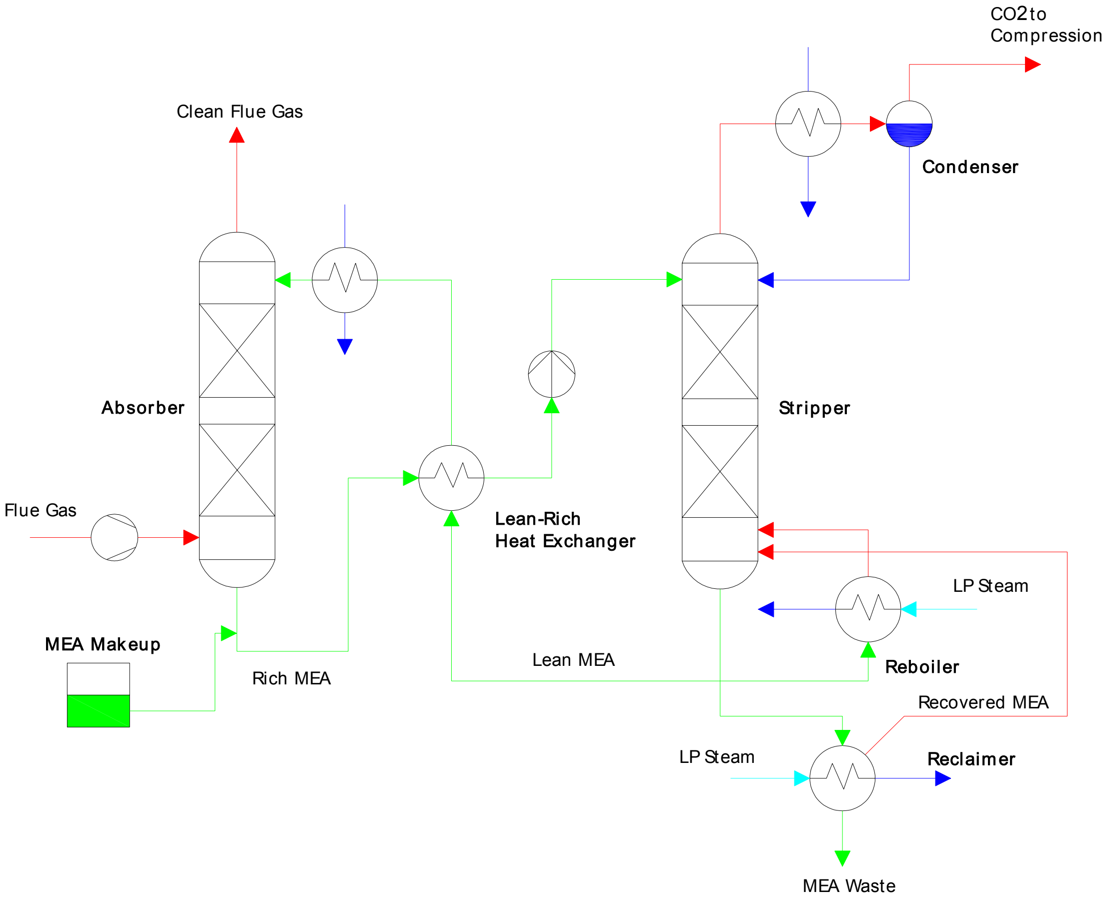
3.4.2 Leading CO_2 Separation Technologies

(i) Chemical Solvent Scrubbing

Chemical solvent scrubbing is currently the most widely used method for capturing CO_2 from low pressure and low concentration flue gases. The majority of chemical solvents are amine based and the most widely used is Monoethanolamine (MEA). Since CO_2 is an acid gas, alkaline solvents like MEA will form chemical bonds with it. This property can be used to absorb CO_2 from a flue gas stream. Once the CO_2 has been absorbed from the flue gas, heat can be applied to the absorbent to release the CO_2 for storage while simultaneously regenerating the solvent for reuse in the process.

A process flow diagram for a commercial MEA absorption system is shown in Figure 3-6 The flue gas is first cleaned of particulates and other impurities. It then moves into the absorption vessel where it is brought into contact with the absorber at temperatures of between 40 and 60°C allowing CO_2 to chemically react with the solvent to form a loosely bound intermediate compound. The CO_2 'rich' solvent is then fed into the top of the stripper column, via a heat exchanger, and heat is applied (100-140°C) to remove the chemically bound CO_2 from the solvent. This produces a CO_2 stream and regenerates the original solvent which can be recycled back to the absorption vessel via a heat exchanger and a cooler. In most industrial applications heat is supplied via steam taken from a steam cycle. Typical CO_2 recovery is between 80-95% with a product purity in excess of 99% (Rao and Rubin, 2002).

Figure 3-6: CO_2 Recovery from Flue Gas by Chemical Absorption (Wolf *et al.*, 2006)



Chemical solvent scrubbing has been used for decades in the chemical and oil industries and has begun to be applied tentatively in power generation facilities. Although the technology has yet to be applied to large scale power facilities or have had any application in the cement industry the process itself is considered to be commercially available and could be used to treat cement flue gases should the right incentives exist.

The major disadvantages with the use of MEA and other amine solvents are:

- A low carbon dioxide loading capacity - 0.4 $kgCO_2/kgMEA$ (Yeh and Pennline, 2004).
- High energy consumption during absorbent regeneration (Herzog *et al*, 1997; Turkenburg and Hendriks, 1999; David and Herzog, 2000).
- A high equipment corrosion rate unless additives are used.
- High solvent degradation rates in the presence of a range flue gas constituents including SO_2, NO_2 and O_2.
- Removal (by filtration) and disposal or regeneration of solid salts formed when the MEA combines with SO_x or NO_2.

- Consumption of activated carbon used for removal of degradation products to prevent the acceleration of the solvent degradation processes – typically around 0.075 kg activated carbon per tonne CO_2 (Chapel *et al.*, 1999).

To improve the performance of amine solvents, sterically hindered amines have been developed which can require less energy for absorption and regeneration and have higher CO_2 loading capacity than MEA. KS-1, a proprietary sterically hindered amine, has been used in a commercial gas scrubbing operation in Malaysia since 1999 to produce a pure CO_2 stream for urea production (Mimura *et al.*, 2000). The KS-1 solvent scrubbing process uses 2.7 MJ/kg CO_2 of steam compared to approximately 3.7 MJ/kg CO_2 (IPCC, 2005). Other chemicals can be added to amines to improve performance.

Inorganic alternatives to amine based scrubbers are also commercially available including the use of potassium carbonate combined with a promoter.

In search of even greater efficiency novel process designs and alternatives to chemical scrubbers are being investigated. These emerging technologies are briefly discussed in section 3.4.3.

(ii) Physical Solvent Scrubbing

During physical solvent scrubbing, CO_2 from a gas stream is weakly absorbed by the solvent under conditions of high pressure and released when the pressure is reduced to allow solvent regeneration. The most commonly used physical solvent is Selexol consisting of dimethylethers of polyethylene glycol. Alternatives include cold methanol (Rectisol), N-methyl-2-pyrolidone (Purisol) and propylene carbonate (Fluor solvent). These physical solvents have a relatively low affinity for NO_x and O_2. However, SO_2 is readily absorbed and must be removed from a gas stream prior to physical solvent use. Physical solvent scrubbing is best applied to pressurised gas streams containing high concentrations of CO_2 and is generally not considered suitable for post-combustion capture. They are better suited for application at an Integrated Gasification Combined Cycle (IGCC) pre-combustion capture plant where the fuel gases contain CO_2 concentrations of 35-40% and are at high pressures.

Although cement kiln flue gases have a relatively high CO_2 concentration (15-30%) they emerge at atmospheric pressure and thus would require considerable energy input for pressurisation before physical solvents could be usefully applied. As such physical solvent scrubbing is not considered suitable to clean up the flue gas from a conventional cement plant. An exception may be where an oxy-combustion process (as described in Section 3.3.3) was first applied which would result in very high CO_2 flue gas concentrations (Croiset and Thambimuthu, 1999; McDonald and Palkes, 1999; Wolsky *et al.*, 1994).

(iii) Hybrid Chemical/Physical Solvents

Under certain conditions a combination of chemical and physical solvents may be efficiently used. The Sufinol and the Amisol processes are examples of this hybrid solvent approach and are currently used to remove CO_2 and sulphur from coal syngas.

A summary of the commercially available CO_2 solvents is given in Table 3-1.

Absorption process	Solvent	Process conditions
Physical solvents		
Rectisol	Methanol	-10/-70˚C, >2 MPa
Purisol	n-2-methyl-2-pyrolidone	-20/+40˚C, >2 MPa
Selexol	Dimethyl ethters of polyethyleneglycol	-40 ˚C, 2-3 MPa
Fluor solvent	Propylene carbonate	Below ambient temperatures, 3.1-6.9 MPa
Chemical solvents		
Organic (amine based)		
MEA	2,5 n monoethanolamine and inhibitors	40 ˚C, ambient-intermediate pressures
Amine guard	5n monoethanolamine and inhibitors	40 ˚C, ambient-intermediate pressures
Econamine	6n diglycolamine	80-120˚C, 6.3 MPa
ADIP	2-4n diisopropanolamine 2n methyldiethanolamine	35-40˚C, >0.1 MPa
MDEA	2n methyldiethanolamine	
Flexsorb, KS-1, KS-2, KS-3	Hindered amine	
Inorganic		
Benfield and versions	Potassium carbonate and catalysts. Lurgi and Catacarb processes with arsenic trioxide	70-120˚C, 2.2-7 MPa
Physical/chemical solvents		
Sulfinol-D, Sulfinol M	Mixture of DIPA or MDEA, water and tertahydrothiopene (DIPAM) or diethylamine	>0.5 MPa
Amisol	Mixture of methanol and MEA, DEA, diisopropylamine (DIPAM) or diethylamine	5/40˚C, >1 MPa

Table 3-1: Commercially Available CO_2 Scrubbing Solvents (Gupta *et al.*, 2003)

3.4.3 Developing CO_2 Separation and Capture Technologies

As well as investigating novel physical/chemical solvents and potential improvements to the absorption process a number of speculative capture technologies are currently being explored. It is hoped that these developing technologies may offer cheaper and more efficient ways of capturing CO_2. These developing approaches to CO_2 capture are described below.

(i) Absorption

Novel technologies are being investigated in order to improve, complement or replace the chemical and physical absorption techniques previously described. Innovative designs are being investigated with regards to the packing process (Aroonwilas *et al.*, 2003), the increase of aqueous MEA concentrations (Aboudheir *et al.*, 2003) and the prevention of oxidative degradation of MEA by deoxygenation of the solvent solutions (Chakravarti *et al.*, 2001).

Besides novel process designs, various novel solvents are also being tested with the aim of reducing energy consumption for solvent regeneration (Chakma, 1995; Chakma and Tontiwachwuthikul, 1999; Cullinane and Rochelle, 2003; and IEA, 2004). One of the most promising of these new solvents is aqueous ammonia.

Although research on its use for CO_2 capture is at an early stage there have been a number of papers and articles published highlighting several important perceived advantages that aqueous ammonia has over amine scrubbers and MEA in particular. These include:

- The CO_2 removal efficiency of aqueous ammonia is higher than MEA (99% compared to 94%) (Yeh *et al.*, 2002)
- The CO_2 loading capacity of aqueos ammonia is 3 times that of MEA (1.2kgCO_2 /kg NH_3 compared to 0.40 kgCO_2 /kg MEA) (Yeh and Pennline, 2004)
- NH_3 solution is also significantly cheaper by weight than MEA which, when combined with its better efficiency, make it over 12.5 times cheaper, in terms of chemical make-up costs, than MEA (Ciferno *et al.*, 2005)
- The thermal energy requirement is approximately 50% less in a duel alkali system using ammonia to absorb CO_2 and anion exchange resins to regenerate ammonia for reuse than using amine to absorb CO_2 and steam stripping to dissociate the resulting carbamates (Huang and Chang, 2002)
- Aqueous ammonia could be used to capture all three major acid gases (SO_2, NO_x, and CO_2) simultaneously since, unlike MEA, it does not suffer from absorbent degradation problems in the presence of SO_2 or O_2. This is expected to reduce the total cost of and complexity of emission control systems (Yeh and Pennline, 2004).

Though a promising future replacement for amine based scrubbing, challenges remain to the use of ammonia in this manner. As a result of ammonia's high volatility it will tend to vaporise in the absorption tower and escape with the flue gas (commonly referred to as ‘ammonia slip’). To mitigate this the absorption process must take place at either elevated pressure or very low temperature and additional tail gas control measures will be necessary. The regeneration of the ammonia solvent will take place at an elevated pressure (Wolf *et al.*, 2006). Given these uncertainties laboratory-scale testing and more rigorous process analysis and modelling will be required before ammonia based solvents can be utilised on a commercial basis.

(ii) Adsorption

In the adsorption process, CO_2 from a gas stream fix to the surface of a solid sorbent by either chemical or physical processes and are thereby removed from the gas stream. For applications to CO_2 capture, adsorbing beds of alumina, zeolite and activated carbon are currently the most effective sorbents. Though promising, this technology incurs the costs of removing the adsorbed gas, which is energy intensive (Anderson and Newell, 2003).

The main types of adsorption processes are as follows:

- Temperature swing adsorption (TSA)
- Pressure swing adsorption (PSA)
- Electric swing adsorption (ESA).

In PSA, the gas mixture flows through packed beds of spherical materials at elevated pressures and low temperatures until the adsorption of the desired component approaches equilibrium conditions at the bed exit and the adsorbent is regenerated by reducing the pressure. In TSA, heat is used to regenerate adsorbents by raising their temperature (Hassan, 2005). PSA and TSA are commercially used for gas separation and in the removal of CO_2 from natural gas, though TSA requires more energy, time for regeneration and larger adsorbent beds (Hassan, 2005; IPCC, 2005). It is understood that PSA requires high feed partial pressure of CO_2 or a vacuum regeranration step to desorb CO_2. PSA is used for treating 5 bar pressure process off-gas from the COREX iron making facility at Saldana Bay, South Africa. Though not yet commercially available, ESA is said to offer the prospect of lower energy consumption than PSA and TSA (Hassan, 2005). In ESA, regeneration takes place by passing a low voltage electric current through the adsorbing material (Hassan, 2005). TSA, PSA and ESA generally require the gases to be cooled and dried, thereby limiting the attractiveness of this technology (IPCC, 2005), although it is understood that the PSA process referred to at Saldana Bay, South Africa does not utilise gas drying. Adsorption is not yet considered attractive for large-scale separation of CO_2 from flue gases because the capacity and CO_2 selectivity of available absorbents is low (Gupta *et al.*, 2003), new sorbents with improved performance characteristics are needed to make this approach to CO_2 capture commercially viable.

One such novel sorbent is CaO. The use of CaO as a sorbent to separate CO_2 from gas streams entails the carbonation reaction of CaO to separate CO_2 from hot gases and the regeneration of the sorbent by calcining the $CaCO_3$ into CaO and pure CO_2. This carbonation-calcination cycle was successfully tested in experimental settings (Abanades *et al.*, 2004; Grasa and Abanades, 2006), at a pilot plant (Curran *et al.*, 1967) and in a post-combustion system (Shimizu *et al.*, 1999). This process is not suitable for CO_2 capture at a cement plant as the CaO produced at the cement plant is produced by calcining $CaCO_3$. Therefore, there is no CO_2 emission reduction available by this route.

(iii) Membrane Separation

Membranes are specially manufactured materials that allow the selective passage of a gas through them; a process that depends on the nature of the materials and the pressure difference across the membrane itself which favours high-pressure streams. Although this technology has not yet been applied in large-scale and demanding industrial settings, a considerable research effort is in progress to make it more applicable (IEA, 2004; IPCC, 2005). The main membrane technologies are described below.

Gas separation membranes

Gas separation membranes have the potential of capturing CO_2, depending on the permeability and selectivity of the membrane. The applicability of this technology to CO_2 capture is thus ideal for a concentrated CO_2 source stream with few contaminant gases, flowing through a permeable membrane that is highly selective with respect to CO_2. These conditions would only apply to the oxy-combustion capture model for the cement industry. Despite the high concentration of CO_2 in cement plant flue gas, applications of this technology for post-combustion capture would require a multistage process requiring a large number of membranes that would substantially increase capital costs (Anderson and Newell, 2003).

Though most have been used only in research laboratory settings, common types of gas separation membranes include polymer, palladium, and molecular sieve membranes (Riemer *et al.*, 1993). Gas separation membranes are likely to play a key role in CO_2 capture systems in the future as their energy efficiency can be higher than for absorption separation systems (IEA, 2004).

Gas absorption membranes

In gas absorption, a microporous membrane keeps the gas and liquid flows separate by acting as a contacting device between gas mixtures and liquid absorbents (such as MEA solvents), thereby increasing the efficiency of physical or chemical absorption.

In contrast to gas separation, it is the absorption liquid and not the membrane that determines the process' selectivity. This technology minimises entrapment and foaming, in addition to capital costs due to its compacted nature (Miesen and Shuai, 1997; Hassan, 2005) but is limited to gas streams and absorption liquids of similar pressure levels (Anderson and Newell, 2003).

Facilitated transport membranes

These membranes rely on the formation of complexes or reversible chemical reactions of components present in a gas stream with compounds present in the membrane which are then transported through the membrane. Carbozymes (Trachtenberg *et al.*, 2007), amines and molten salt hydrates have been suggested for CO_2 separations with porous membranes used for support. These membranes have only been tested on a laboratory scale and are fraught with limitations related to membrane selectivity and saturation but could be used in a pre-concentration step as part of other capture processes (IPCC, 2005).

(iv) Cryogenic Separation

Cryogenic separation involves the compression and cooling of gas mixtures in multiple stages to induce phase changes in CO_2 and other gases which allow them to be separated. Studies by Herzog *et al.* (1997) have shown this process to be most effective with gases containing components with very different boiling points. This is not typical of gas streams in the cement industry. Furthermore, the behaviour of CO_2 itself is complicated and can lead to the formation of solids that damage equipment and reduce heat transfer rates. Cryogenic separation could be effective for large and highly concentrated source streams of CO_2, which compensate for the high energy demand of such a process (Anderson and Newell, 2003).

In the future, the most promising applications of cryogenics are predicted for the separation of CO_2 in oxyfuel combustion in which the input gas has high concentrations of CO_2 or from high-pressure gases such as in pre-combustion capture processes (Hassan, 2005).

Clathrate Hydrate

While clathrate hydrates have the potential to impede the cryogenic processes described above, their formation could be used to separate CO_2 from gas mixtures. With more research, this could be achieved by combining CO_2 and water at various combinations of high pressure and low temperature, leading to the formation of clathrate hydrate crystals that can be separated from other gases, transported as slurry and stored in a dedicated location (Chargin and Socolow, 1997).

(v) Emerging Air Separation Technologies

The production of a large quantity of concentrated O_2 is a vital and currently expensive step in the process of oxy-combustion, in particular. There are a number of emerging air separation technologies that may, in the future, lead to more efficient O_2 production and thus make oxy-combustion more attractive. These emerging technologies are discussed briefly below.

Ion Transport Membranes

The use of membrane technologies for the production of oxygen is an emerging alternative to cryogenic air separation which may, if proven, make oxy-combustion a more attractive process. Ion Transport Membranes (ITMs) are semi permeable ceramic mixed metal oxide membranes that can be used to separate oxygen from a stream of heated air. A voltage or pressure differential across the membranes causes oxygen ions from the airflow to migrate through the membrane where, on the other side, they recombine to form oxygen gas and electrons (Bredesden *et al.*, 2004). These membranes operate at temperatures similar to those of the tertiary air drawn from the cooler for combustion in the precalciner of a modern cement kiln (Clark, 2006) and as such offer potential for integration into cement manufacturing process. If membrane technologies are effective at larger scales then they may offer a more efficient alternative to cryogenic air separation (Stein and Foster, 2001).

Chemical Looping Combustion

An alternative to having to produce a separate stream of oxygen to be used in an oxy-combustion process is Chemical Looping Combustion (CLC) first proposed by Richter and Knoche (1983). The concept is based on the use of a metal/metal oxide system to provide a reversible chemical reaction for oxygen supply. Fuel combustion is split into separate oxidation and reduction reactions in two reactors. In one reactor a suitable metal reacts with air to produce a metal oxide, which is then circulated to the other reactor where it reacts with the fuel to produce syngas and a metal (that can be recycled back into the process). This so called 'flameless combustion' avoids the need for energy intensive air separation to provide a pure oxygen supply. Although the concept of CLC has been around for over 25 years it has only been applied in a laboratory and has never been tested on a commercial scale – as such it should be considered a speculative technology (IPPC, 2005).

4 Evaluation of CO_2 Capture at a New-Build Cement Plant

4.1 Introduction

This section of the Report describes the detailed technical evaluation of applying CO_2 capture to a new-build cement plant. Both post-combustion and oxy-combustion capture have been studied. Outline designs including mass and energy balances have been prepared for both options. The performance of the cement plants with CO_2 capture is described.

4.2 Design Basis

The design basis for the study was as follows:

- The capacity of the cement plant was 1 Mt/y of cement. This is typical of a modern European cement plant.
- The cement plant is a 5-stage preheater with precalciner dry process cement plant. This was chosen as the best available technique (BAT) for the production of cement clinker for new plants and major upgrades is considered to be a dry process kiln with multi-stage preheating and precalcination (EC, 2001).
- The plant is located in NE Scotland, UK. It should be noted that the IEA GHG's standard site location of the NE coast of The Netherlands was not used due to the absence of cement plants in this region of Europe. The location of NE Scotland was considered more realistic particularly due to its proximity to the North Sea oilfields which have been proposed as potential CO_2 storage locations. The ambient conditions of the site in Scotland were assumed to be the same as those for IEA GHG's standard site in The Netherlands.
- The fuel to the plant is 60% coal and 40% petroleum coke. This is typical although due to rising costs the use of petroleum coke is reducing in the UK. Due to the heat balance a fuel ratio of 65% coal and 35% petroleum coke was accepted in the oxy-combustion case as described later.
- The cement plant is located adjacent to a limestone quarry.

4.3 Assumptions

The main technical assumptions used for this study were based on the IEA GHG R&D programme Technical & Financial Assessment Criteria (IEA GHG, 2003). These have been developed for the technical and financial assessment of power plants with CO_2 capture. Discussions were held with the BCA to modify some of the assumptions to be relevant to the cement industry.

The key points are summarised below:

Site Conditions

- Ambient air temperature: 9 ºC

- Ambient air relative humidity: 60%
- Ambient air pressure: 1.013 bar

Operational hours

- Days of operation per year: 330
- Hours of operation per day: 24

Raw meal

- Moisture content: 8% (w/w)
- Sulphur content: 0.33% (w/w – dry basis)

Raw mill

- Moisture content of raw meal exiting raw mill: 1% (w/w)
- Air in-leaks in raw mill: 0.5 kg/kg clinker

Preheater

- Air in-leaks in preheater: 0.13 kg/kg clinker

Precalciner

- It is assumed that the precalciner is fed with coal only.

Kiln

- It is assumed that the kiln is fed with petroleum coke only.
- Air in-leaks in kiln inlet: 0.15 kg/kg clinker
- Air in-leaks in kiln outlet: 0.05 kg/kg clinker

Coal

- The coal specification is presented in Appendix A.

Petroleum coke

- The petroleum coke specification is presented in Appendix B.

Fuel drying

- Moisture content of dried coal: 1% (w/w)
- Moisture content of dried petroleum coke: 1% (w/w)

Cement product

- Clinker composition in cement: 91% (w/w)
- Gypsum composition in cement: 4% (w/w)
- Limestone composition in cement: 5% (w/w)

Flue gas composition

- Fraction of NO_2 in NO_x: 10% (w/w)

CO_2 processing

- For consistency with other IEA GHG studies the minimum CO_2 capture level is 80% with the preferred level 85%. However, a lower percentage capture was accepted in the oxy-combustion case as described later.
- CO_2 is to be compressed to 110 bara before injection into the transfer pipeline.

O_2 purity

- The O_2 purity supplied by the Air Separation Unit (ASU) was specified at 95% (v/v), this figure being derived from previous studies (IEA GHG, 2005) as the optimum purity associated with oxy-combustion power plant given tramp air ingress.
- N_2 content in O_2 supplied by Air Separation Unit (ASU): 2% (v/v)
- Argon content in O_2 supplied by Air Separation Unit (ASU): 3% (v/v)

Seawater

- It is assumed that seawater is available for cooling. Direct cooling is used for large compressor intercoolers and an indirect cooling system is used for other process coolers.
- Average inlet temperature of seawater cooling water: 12 ºC
- Maximum temperature rise: 7 ºC
- Salinity: 22 g/l

4.4 Evaluation of Post-Combustion CO_2 Capture

4.4.1 Overview

For this study, CO_2 absorption using MEA was selected as the preferred technology for post-combustion CO_2 capture. As such, the design, technical issues and process descriptions are based around this technology. The selection of MEA is made on the basis of commercial acceptance for this proven chemical solvent. It should be recognised that substantial scope may exist for future designs using more economic post-combustion capture technologies being developed in the power sector.

4.4.2 Major Technical Issues

The addition of post-combustion CO_2 capture to a cement plant involves overcoming some major technical issues if an efficient and cost effective solution is to be delivered. A typical cement plant operates in such a way that will allow the incorporation of the capture equipment with limited modifications to the existing plant. However, the following issues must be considered:

(i) Sulphur Dioxide (SO_2)

Sulphur oxides (SO_x) can be produced in the cement production process from two different sources; either from the raw meal itself or from the fuel. Any sulphides (e.g. pyrite or marcasite) present in the raw meal will be oxidised when exposed to heat and oxygen in the precalciner or the kiln. SO_x are also produced from the combustion of coal or pet-coke. The sulphur in the fuel reacts in the same way as the sulphides in the raw meal to produce SO_x. Both these processes mainly occur in the precalciner and the kiln. However, ordinarily in the cement process the majority of the SO_x and other acidic gases are re-absorbed into the system in the raw mill and preheater. SO_x emissions are not often of concern at modern cement plants, therefore, as such emissions result almost exclusively from the oxidation of the sulphides in the raw mill occurring in the upper stages of the pre-heater. Thus, cement plants using low sulphide minerals as the raw material have very low emissions of SO_2, even as low as 10 mg/m^3. For plants using raw meal with significant sulphide minerals, however, SO_2 emissions can be as high as 3,500 mg/m^3.

The level of sulphides in the raw meal depends on the location of the source. For example, raw meal from Northern UK contains relatively higher levels of sulphides compared to the South of the country. As the location of the plant for this study is in the North East of Scotland and it is assumed that the raw material is sourced locally then Mott MacDonald has assumed a sulphide concentration of 0.33% (w/w) in the raw material for the cement plant.

When calcination occurs in the precalciner and kiln Mott MacDonald has assumed that all sulphur oxides are absorbed in the clinker. Additionally, the exit gas from the preheater contains some SO_x which are partially absorbed in the raw mill.

The concentration of SO_2 in the flue gas from the cement process is important for post-combustion capture due to the nature of amines. Amines react with acidic compounds to form amine salts that will not dissociate in the amine stripping system. These amine salts must be removed from the amine solution and must be replaced with fresh make-up amine solution. Hence, the consumption (and cost) of amine within the amine absorption system is directly related to the concentration of SO_2 in the inlet gas. It is important, therefore, that the flue gas exposed to the amine has a very low concentration of SO_2 as amine solution is currently priced at around 1100 €/t. It has been reported that the SO_2 concentration in the inlet to the amine absorption process should be restricted to approximately 10 ppmv [@ 6% O_2] if CO_2 absorption using amine is to economic (Chapel *et al.*, 1999). This study has assumed that the flue gas should be treated to this level prior to the amine absorption system.

In the process flow scheme considered for this study SO_2 can also originate from another source: the auxiliary boiler required to produce steam for the stripper. The coal used in the CHP plant has an assumed sulphur content of 1.1% (w/w). The SO_x that escapes through the raw mill and is produced by the auxiliary boiler is removed by a wet limestone flue gas desulphurisation (FGD) system. This system was chosen due to its demonstrated ability to achieve high levels of flue gas desulphurisation, as required for post-combustion capture, and its current application in cement plant in Eastern Scotland, similar to the design basis for this study. Wet limestone FGD is also currently the most common way to mitigate SO_x emissions in power plants, and benefits from current power sector trials of post-combustion capture.

Seawater FGD was also considered but it was considered that a higher level of SO_2 reduction could be achieved using the wet limestone technique. Further study of hybrid combinations of limestone FGD with seawater FGD or absorbent injection within the cement process is suggested to establish the most economic means of achieving these high levels of purity in cement plant applications.

Importantly, a by-product of the wet limestone process is gypsum which can be used later in the process to mix with the clinker to form the final cement product. The level of SO_x reduction required will determine the amount of gypsum produced. For the process studied here approximately 50% of the gypsum required by the cement plant will be produced from reducing SO_2 emissions to the required level for economic amine absorption.

(ii) Nitrogen Dioxide (NO_2)

NO_x are formed as a result of the combustion of the fuel. It can be formed in two ways during combustion: thermal NO_x is formed by the oxidisation of the molecular nitrogen in the combustion air; and fuel NO_x is formed from the oxidisation of the nitrogen compounds in the fuel. The fuel nitrogen content assumed in this study is 1.5% (w/w) for pet-coke and 1.8% (w/w) for coal. In modern precalciner cement plants combustion occurs in both the precalciner and the kiln. The high temperatures involved in the rotary kiln combustion process cause a large amount of thermal NO_x to be formed.

NO_x are problematic for MEA absorption systems as they result in solvent degradation. The chief culprit in NO_x is NO_2 which reacts to form nitric acid in the amine solvent and ultimately heat stable salts (Chapel *et al.*, 1999). However, typically less than 10% of the NO_x emissions from the cement process are NO_2 with the large majority being NO (Hegerland 2006; EC, 2001; Rushworth, 2008; and Leibacher, 2006) and only a fraction of the NO_2 is absorbed in the solvent (Knudsen *et al.*, 2006 and Iijima *et al.*, 2007). NO_x emissions associated with cement kilns vary generally between <200 and 3000 mg/Nm^3 (EC, 2001) containing NO_2 emissions of 20-300 mg/Nm^3. It is reported that the concentration of NO_2 in the flue gas should be restricted to approximately 41 mg/Nm^3 (20 ppmv at 6% O_2) for economic post-combustion capture using amine (IEA GHG, 2004).

There are two main ways to reduce the NO_x in a flue gas. These have been described in section 0. The most common way to mitigate NO_x emissions from power plants is to use selective catalytic reduction (SCR). SCR has been applied in this study because SCR is generally capable of achieving greater levels of NO_x reduction as compared to SNCR. This is a conservative approach and it is recognised that SNCR may be able to achieve the required levels of NO_x reduction.

To apply SCR to the cement process requires some modification of existing cement plant equipment (described later in section 4.4.5).

(iii) Dust

The cement process produces significant quantities of dust which are carried in the flue gas and are in the region of 3000 mg/m^3. For a typical cement plant the main issues with dust are meeting the emission standards for dust and loss of end product (i.e. clinker). Both of these issues are mitigated through the use of electrostatic precipitators (ESPs) or bag filters. Generally, the dust is removed from the exit flue gas and recycled with the raw meal in one of the storage silos.

The presence of dust can effect the operation of the SCR technology and the amine CO_2 absorber.

High dust conditions for SCR can cause problems with the dust sticking to the catalyst and clogging the system. The SCR unit will be located downstream (in the direction of the gas) of the preheater, which does not allow the reduction of dust levels prior to it. Therefore, the SCR unit must be designed to operate in a high dust environment. A pilot plant for high dust SCR is currently being tested in Monselice, Italy with reportedly positive results (Leibacher *et al.*, 2006). For this application a standard SCR unit for a power plant application has been modified to prevent clogging; the catalyst is manufactured with slightly larger holes in the honeycomb design and additional high pressure air blowers are fitted to remove dust sticking to the catalyst.

Dust also has a negative effect on the amine solution and causes wastage once it comes into contact with the amine. To operate post-combustion capture economically the dust level must be below 15 mg/Nm^3 (Chapel *et al.*, 1999). Both ESPs and bag filters are capable of removing 99.9% of the dust loading in the flue gas and are therefore appropriate for this application.

(iv) Air Dilution

Air in-leaks occur in the raw mill, preheater and kiln due to the pressure within the system operating at slightly below atmospheric. Excessive air in-leaks will result in contamination of the CO_2-rich exhaust gas with nitrogen, oxygen and argon. This will increase the volume flow rate of the flue gas and reduce the CO_2 concentration. Both these factors will reduce the efficiency of the capture process and increase the operating costs. The air in-leaks have been included in the process model as they cannot be avoided.

(v) Additional Power Requirements

There will be additional power required for the compression of the CO_2. Typically, this additional power requirement for compression of the CO_2 will be around 0.146 kWh/kgCO_2 (IEA GHG, 2005) In this study the total power requirement for the cement plant and the post-combustion capture plant will be supplied by an onsite CHP plant. The CHP plant will also meet the steam demand to strip the CO_2 from the MEA.

(vi) Additional Steam Requirements

One of the major technical issues with using MEA CO_2 capture is the large steam requirement. After the CO_2 has been absorbed by the MEA it then has to be regenerated by using heat in the stripper column. The heat is supplied from low pressure steam that can be generated by an auxiliary boiler or a power plant (either co-located or on-site). The steam conditions required are approximately 3.5 bara and 140 - 150°C at approximately 200 – 250 t/hr (for the cement application).

Another alternative could be to use a different post-combustion capture technology that does not have such a large steam demand. There are other techniques currently being developed but are not as commercially and technically advanced as MEA CO_2 capture. These alternatives are discussed in section 4.4.11.

(vii) Reducing Conditions

It is recognised that the clinker must not be generated in reducing conditions. This necessitates that an excess of oxygen is maintained in the process. The oxygen concentration is required to be >2% (w/w) in the preheater, precalciner and the kiln.

(viii) Heat Reduction for MEA Absorption

The flue gas leaves the cement process (at the raw mill) at approximately 110°C. This must be cooled to approximately 50°C to meet the ideal temperature required for CO_2 absorption in the MEA. This can be achieved by passing the flue gas through the FGD system. Due to the contact with the wet limestone slurry and the long residence time the temperature of the flue gas will drop to 45-50°C.

(ix) Other Gases

The oxygen content required for CO_2 absorption in the MEA is >1.5% (v/v) (Chapel *et al.*, 1999). This should not pose any issues for the cement plant considered in this study as the oxygen content in the flue gas after the raw mill is estimated to be 7.8% (v/v). Combined with the other exit gases from the fuel drying and CHP, the oxygen content drops down to 4.3% (v/v) but this is still significantly above the target value.

Hydrochloric acid (HCl) can be present in small quantities within cement flue gases; typical concentrations are around 8 mg/Nm^3 (Hegerland *et al.*, 2006). The presence of any acidic components such as HCl will reduce the efficiency of the MEA absorption process. HCl concentrations will be reduced by SCR and by FGD so concentrations should not pose a significant problem.

Total Organic Carbon (TOC) can also appear in similar quantities to HCl (~8 mg/Nm^3) but should not affect CO_2 absorption in the MEA (Hegerland *et al.*, 2006).

(x) Heat Integration

Although the cement process has been highly optimised already, post-combustion capture offers a potential opportunity to maximise performance through heat recovery from the CO_2 compression system.

Although heat integration between the CO_2 compression plant and the cement plant could lead to energy savings it is important to understand that this may reduce the operational flexibility of the plant. There may be valid strategic reasons for having the capability to operate the cement plant without the CO_2 compression plant. Full heat integration may make this difficult to achieve.

4.4.3 Configurations Considered

During the development of the conceptual design of a post-combustion cement plant several different configurations for CO_2 capture were considered and are illustrated in Figure 4-1 to Figure 4-4. They can be summarised as:

- **Configuration 1:** Capture of all flue gases from cement plant in existing orientation.
- **Configuration 2:** Diversion of preheater exit gases for fuel drying and subsequent capture.

- **Configuration 3:** Diversion of cooler exhaust for fuel drying with capture of all other flue gases.

- **Configuration 4:** Dual preheaters with capture of gases only from the precalciner.

Figure 4-1: Post-Combustion Cement Plant Configuration 1 - Capture of all Flue Gases from Cement Plant in Existing Orientation

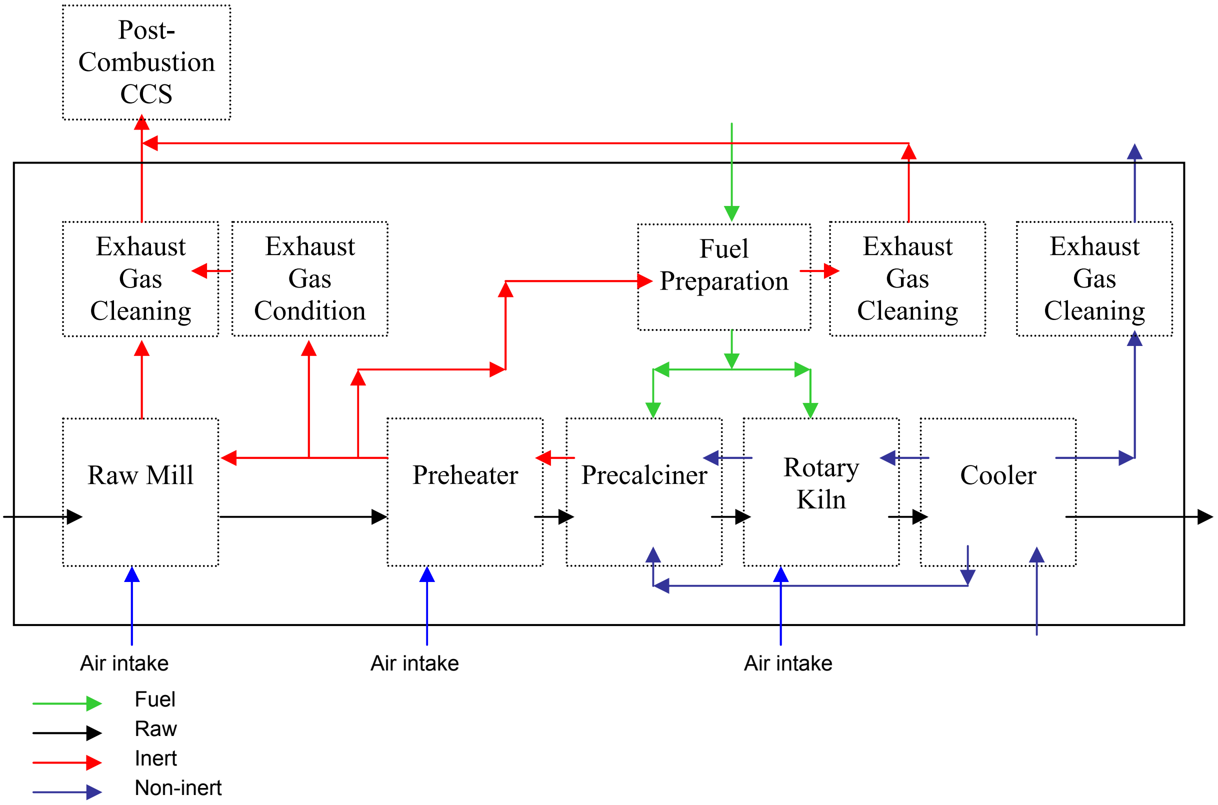


Figure 4-2: Post-Combustion Cement Plant Configuration 2 - Diversion of Preheater Exit Gases for Fuel Drying and Subsequent Capture

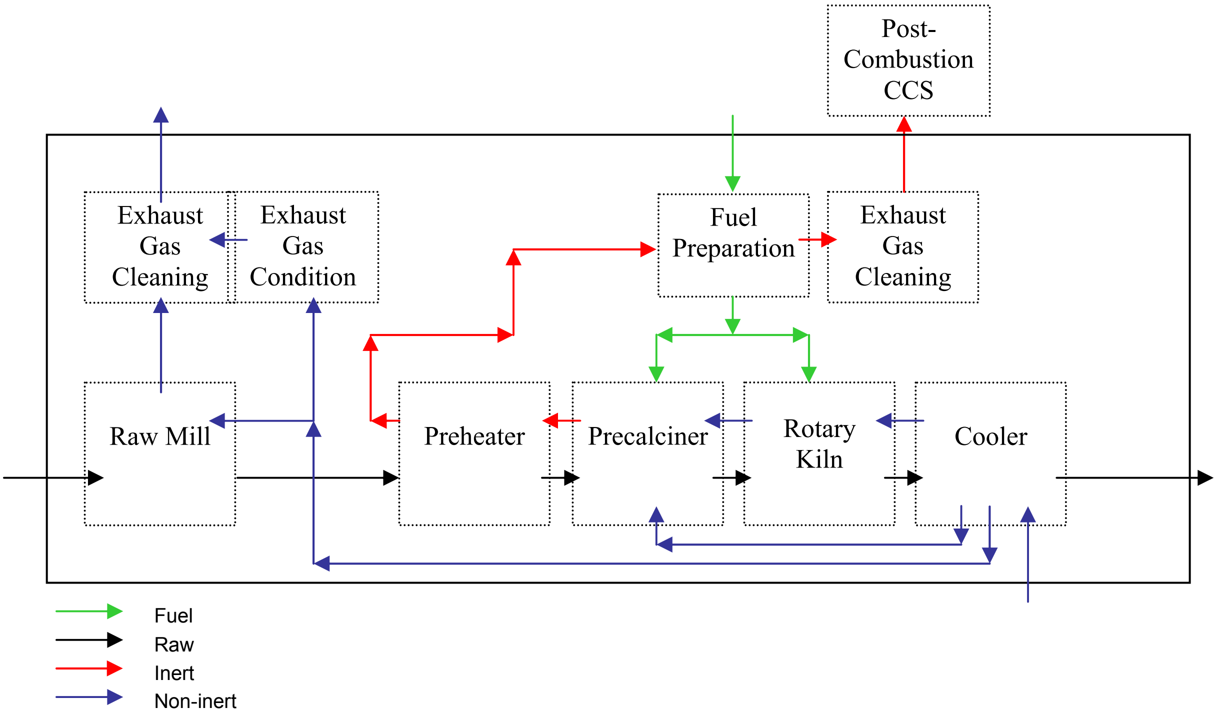


Figure 4-3: Post-Combustion Cement Plant Configuration 3 - Diversion of Cooler Exhaust for Fuel Drying with Capture of all other Flue Gases

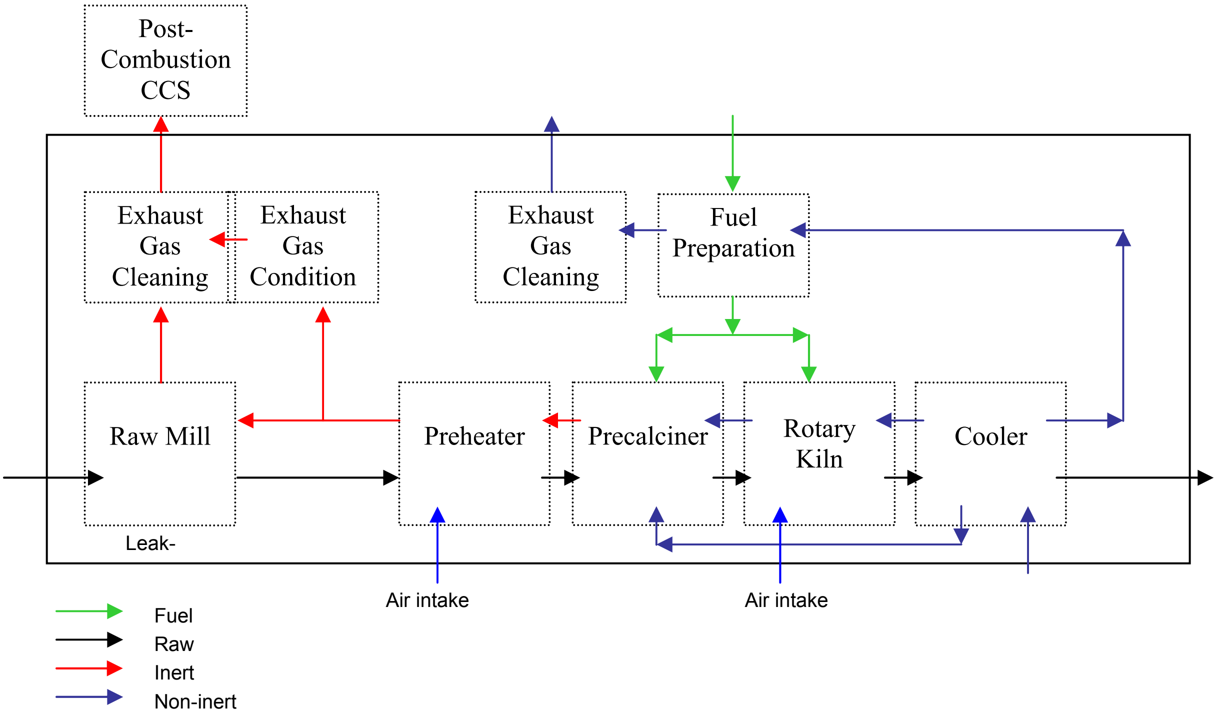
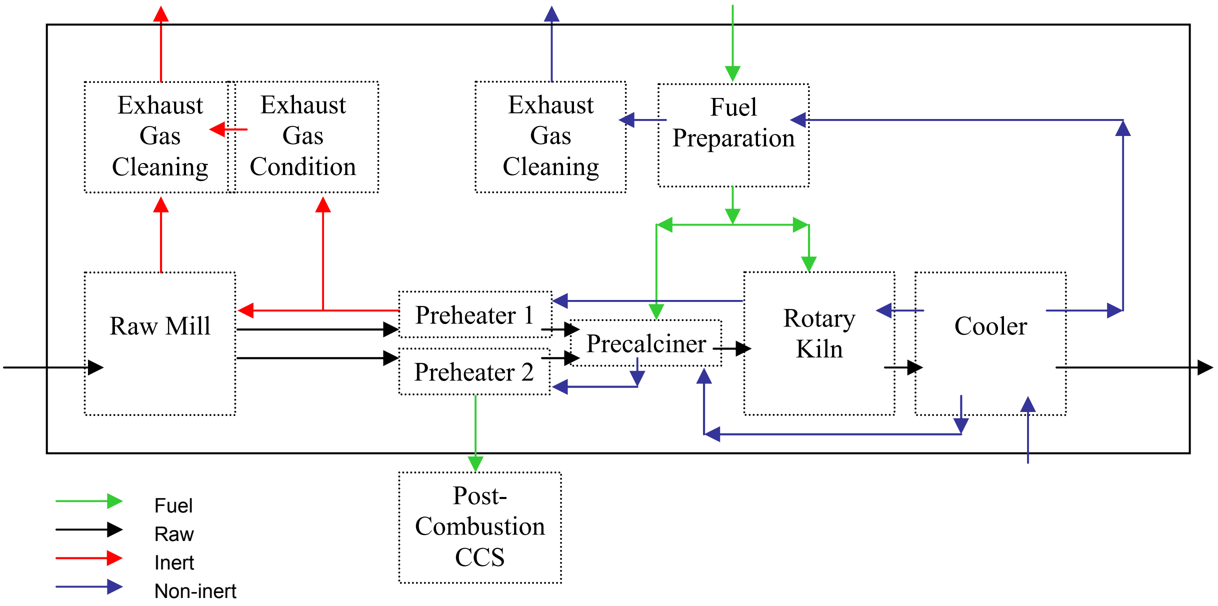


Figure 4-4: Post-Combustion Cement Plant Configuration 4 - Dual Preheaters with Capture of Gases only from the Precalciner



The configurations are compared with a summary of their respective advantages and disadvantages shown in Table 4-1.

Table 4-1: Comparison of Different Post-Combustion Configurations

Configuration	**1**	**2**	**3**	**4**
Advantages	Can potentially capture all the CO_2 the cement plant. Does not affect the cement process.	Less gas will need to be separated from the CO_2 (possibly a higher CO_2 concentration), therefore lower Capex and Opex. Air in-leaks from the raw mill are avoided	Using cooler excess air to dry fuel is already proven.	Air in-leaks from the raw mill are avoided.
Disadvantages	Air in-leaks reduce the efficiency of the process by diluting the flue gas.	Only applicable to plants where the raw mill is remote. Not all the CO_2 has the potential to be captured.	Non-inert gas is used for drying the fuel. This can cause operational difficulties as it is a less safe mode of operation. Air in-leaks will dilute the flue gas.	Dual pre-heaters required which makes retrofitting difficult. Not all CO_2 is captured.
Technical risk	Low	Low	Low	Low

4.4.4 Chosen Configuration

Configuration 1 was chosen for further evaluation because there is the possibility to capture all of the CO_2 emitted from the cement plant, it has minimal impact on the standard cement process (only the addition of SCR will affect the process directly) and it could easily be retrofitted to an existing plant. None of the other three options can offer all these advantages.

4.4.5 Process Description

A detailed evaluation of configuration 1 was undertaken.

PFD 234996/PC/FS02 contained in Appendix D illustrates the major process units and process streams.

(i) Process Modelling

Mott MacDonald and Whitehopleman developed a steady state mass and energy balance of the process using Excel. Each individual process unit was modelled. The mass flows into and out of each process unit were calculated based on simple performance equations derived from typical industry data or parameters. The energy flows were calculated using the basic thermodynamic properties of the components. Where necessary recycle loops were solved using an iterative approach.

The outputs of this exercise were the compositions of the individual streams contained in Appendix D.

(ii) General Description

PFD 234996/PC/FS02 contained in Appendix D shows the arrangement selected for the basis of this particular study. There are five key features that distinguish the process from a conventional cement plant:

- An SCR unit is fitted between the preheater and the raw mill for NO_x abatement. The ammonia will be injected into the first stage of the preheater to increase the residence time of the reaction.

- An auxiliary boiler or CHP plant is required for the low pressure steam for MEA stripping. The CO_2 from this process will also be captured and will mix with the cement flue gas before the wet limestone FGD unit.

- An FGD unit is fitted to remove SO_x from the flue gas stream.

- CO_2 capture equipment based on amine solvent separation is installed.

- The net CO_2 product is compressed, dried, the inerts separated and the pure CO_2 is further compressed to pipeline pressure of 110 bara.

A more detailed process description is described below:

(iii) Raw Material Preparation

Limestone is extracted from the adjacent limestone quarry by drilling and blasting. The blasted limestone is loaded onto haul trucks by wheel loaders and hauled to the crusher. The crusher is a double rotator hammer crusher rated at 1000 tph and capable of producing limestone of the required size of less than 90mm in a single pass through the crusher. Crushed limestone is conveyed to a longitudinal prehomogenisation stockpile where it is stacked together with the shale secondary raw material using the windrow technique.

The shale secondary raw material is imported to a primary stockpile from which it is loaded to a hammer crusher rated at 250 tph. Crushed shale is delivered onto the crushed limestone conveyor belt and the two raw materials are delivered together to the prehomogenisation stockpile. The mixture of raw materials passes through a prompt gamma neutron activation analysis (PGNAA) cross-belt analyser. This PGNAA analyser determines the composition of the mixture of raw materials allowing the proportions of limestone and shale in the mixture to be adjusted to required chemistry targets by varying the feed rates to the limestone and shale crushers.

Raw material crushing operations are carried out on a day work basis, working eight hours per day for five days per week. The prehomogenisation stockpile comprises two longitudinal beds of 35,000 tonnes capacity each, providing sufficient crushed raw materials for more than 7 days raw milling operations. One stockpile is built while the other is reclaimed to the raw mill on successive weeks. Power consumption of the raw material crushing, stacking and reclaiming operations totals 2 kWh/t of clinker produced from the kiln.

(iv) Raw Mill

The premixed raw materials are extracted to a premix bin which feeds the vertical roller mill for raw grinding. Additional corrective raw material feed bins are provided for sand and iron oxide. These materials are imported to primary stockpiles and loaded into the raw mill feed bins via apron feeders and conveyor belts, as required by the process. The premix and corrective raw materials are extracted from their feed bins at a measured rate and conveyed to the raw mill in the proportions required to meet the kiln feed chemistry targets. 52.42 kg/s of raw meal is required, with a raw material consumption of 1.51 t/t clinker. This value compares favourably to the 1.57 tonnes of raw meal required to produce 1 tonne of clinker for the typical average consumption of raw materials in the EU (EC 2001).

The raw mill is a vertical roller mill. Feed is delivered onto the centre of the rotating table of the mill and is drawn under the rollers of the mill which are pressed onto the material bed on the table by hydraulic cylinders. Ground raw material falls over the dam ring round the edge of the table and is lifted by the hot gases (tempered tertiary air) sweeping the mill through the mill body to the integral separator. Coarse material is rejected by the separator and falls back to the mill table for regrinding while fine material is drawn out of the separator and mill with the hot gases by the mill induced draft fan.

Air from the SCR tower unit is used to sweep the raw mill and dry the raw materials from their initial moisture content of 8% (w/w) to 1% (w/w). This requires approximately 292 kJ/kg of energy. The dried raw mix and sweeping air exit the raw mill at 100ºC. The temperature of the SCR unit exit air entering the mill is tempered by recirculation of mill exhaust air in order to protect the bearings of the rollers and separator.

The raw milling system has a rated capacity of 255 tph in order to provide 33% overtaking capacity above the requirements of the kiln for raw mix kiln feed. This enables planned stops of the raw milling system for maintenance to take place for up to 6 hours per day. During these periods when the raw milling system is stopped the SCR exhaust air bypasses the raw milling circuit and the gas conditioning tower is brought into operation, introducing moisture via atomising sprays into the flue gas to reduce the temperature.

The fine, raw mix product is collected from the hot gases after the raw mill induced draft fan by collection cyclones and the bag filter for dedusting. The bag filter recovers approximately 0.96 kg/s of dust that is recycled into either the storage silo or directly into the preheater entry. The de-dusted gases are drawn through a fan and exhausted to the gas mixer. With the addition of air in-leaks, (which are significant in the raw mill at approximately 15.96 kg/s) the total exit gas from the raw mill transported to the carbon capture equipment is 75.59 kg/s, of which 29.6% is CO_2. This is considerably higher than the typical concentration in flue gases from power plant applications and should provide some technical and economic advantages to CO_2 capture at a cement plant. The collected raw mix product is transported mechanically to the combined raw mix homogenisation and storage silo. This homogenisation silo has a capacity of 10,000 tonnes in order to provide over 2 days buffer storage of ground raw mix kiln feed. Homogenisation of the raw mix in the silo is achieved by sequential activation of air slides in the segmented base of the silo by compressed air and extraction of material from a pressure relieved central chamber with a conical roof.

Power consumption of the raw milling and homogenisation system totals 25 kWh/t of clinker produced from the kiln.

(v) Selective Catalytic Reduction

The SCR tower is situated between the preheater and the raw mill. The reductant used in this case is a 25% ammonia solution. The vapourised ammonia is injected into the first stage of the modified preheater to allow a long residence time with the gas. This reduces the ammonia slip (the amount of ammonia that passes through the system) to less than 5 mg/Nm^3 (Schreiber and Russell, 2006). Ammonia slip can be a problem with SNCR applications. The temperature of the gas at this stage will be sufficiently high to allow the reaction to take place. The reaction can only occur on the surface of the catalyst and requires a temperature between 300°C and 450°C for optimal operation. The ammonia rich flue gas is blown through the SCR system which consists of several layers of catalyst. For the cement application the catalysts are honeycomb shaped (as opposed to a flat plate catalyst) to allow the dust to pass through. The basic equations for the reactions that will take place are as follows:

$$4NO + 4NH_3 + O_2 \rightarrow 4N_2 + 6H_2O$$

$$4NH_3 + 2NO_2 + O_2 \rightarrow 3N_2 + 6H_2O$$

It should be noted that the MEA is unaffected by the NO but the SCR system is still capable of removing it from the flue gas. The catalyst is usually manufactured from active metals and substrates (a combination of vanadium oxide and titanium dioxide is considered best for cement kiln applications) (Leibacher *et al.*, 2006) and is arranged in several honeycomb layers. The holes in the catalyst are slightly larger than usual for SCR used in power stations due to the high dust environment created by the cement process. The holes can be approximately 10 mm in diameter to avoid clogging of the dust. The tower is designed so that each catalyst layer is periodically cleaned using pressurised air.

Based on the reported BAT emission level for SCR of 100-200 $mgNO_x/Nm^3$ (as NO_2) (EC, 2001) then this system should comfortably be able to achieve a maximum outlet NO_x concentration of 410 mg/Nm^3 (at 6% O_2) in order to achieve the target NO_2 concentration of 41 mg/Nm^3 (at 6% O_2) required for economic post-combustion capture with amine.

(vi) Preheater

A 5-stage preheater would be slightly modified to integrate with the SCR system required for CO_2 capture using MEA. Vapourised ammonia injection equipment must be integrated into the first cyclone stage and the process is described below.

The milled raw meal product is transported from the silos via conveyors and elevators to the preheater feed bin. This is situated at the top of the preheater. Here it enters the first of the 5 cyclone stages, vaporised ammonia is also added at this stage for NO_x abatement. It is added to the preheater because it gives the NO_x a longer residence time to mix with the ammonia (although the reaction will only take place on the surface of the catalyst), it gives a higher NO_x removal efficiency and decreases the level of ammonia slip (ammonia that passes through the SCR system to the exit gas stream).

The preheater uses exhaust air from the precalciner (which operates at 900°C) to dry the raw meal and drive off any remaining moisture; this requires approximately 167 kJ/kg of heat. It is estimated that 18% of the calcination also occurs in the preheater. Calcination is an endothermic reaction and absorbs 353 kJ/kg of heat. As the preheater is not perfectly insulated some heat is lost as radiation; this is estimated at 75 kJ/kg. The preheated (and partially calcined) raw meal enters the precalciner at 780°C at 61.76 kg/s with no moisture content.

(vii) Precalciner

The bulk of the calcination process is carried out in the precalciner. Dried pulverised coal is used as the fuel and is delivered at the required rate (2.03 kg/s at 1% (w/w) moisture) to ensure constant temperature operation of the precalciner at 900ºC. This amounts to 60% of the total calorific input to the kiln system and is in line with the normal practice in the operation of precalciner cement kilns.

In a conventional, in-line precalciner the oxygen for the combustion of the precalciner fuel is part drawn from the excess air in the hot kiln exhaust gases rising from the rotary kiln and part from the hot tertiary air drawn from the hot end of the clinker cooler which have bypassed the kiln via the tertiary air duct. The tertiary air has been heated to 908°C to maintain the high precalciner temperature.

The Mott MacDonald mass and energy balance model assumes 77% of the calcination occurs in the precalciner and consumes 1511 kJ/kg of heat. There are also radiation losses in the precalciner that account for 25 kJ/kg of energy. The release of CO_2 in the calcination reaction significantly reduces the amount of solid feed passing forward to the rotary kiln. The preheated raw meal entering the precalciner is 61.76 kg/s but the calcined raw meal (95% calcined) only has a mass flow rate of 37.46 kg/s. The calcined raw meal enters the rotary kiln at 870°C.

Air in-leaks are also present in the precalciner and account for 4.81 kg/s of the total precalciner exit gas of 70.51 kg/s. The temperature of the exit gas is 900°C.

(viii) Kiln

The material from the precalciner is collected in the lower stage cyclone of the preheater tower and enters the kiln via the kiln inlet hearth. A burner is situated both centrally and co-axially with the kiln at the front end of the kiln. Petroleum coke is fired into the front end of the kiln along with the 'primary air'. The kiln burning conditions are constantly monitored in terms of clinker temperature, burning zone temperature and nitrous oxides, oxygen and carbon monoxide emissions (at the back end of the kiln). The pet-coke is heated to approximately 65ºC in the fuel drying process and is transported to the kiln using a fan at 1.5 kg/s. The secondary combustion air is drawn into the kiln from the cooler and enters the kiln at approximately 1025°C. The hot clinker leaving the kiln falls into the clinker cooler. The kiln rotates at between 3 and 4 revolutions per minute driven by the main drive of the kiln. By the end of its time in the kiln the raw meal has been completely calcined and combined to form clinker and exits the kiln at approximately 32.57 kg/s at 1350°C.

(ix) Fuel Preparation

Raw coal is imported to site and stored under cover. It is transported from the stockpile by a belt conveyor transport system that feeds an intermediate raw coal storage hopper. The raw coal is extracted from this storage hopper at a controlled rate and fed to the vertical roller coal mill. This mill has a rated capacity of 10 tph in order to provide 25% overtaking capacity above the requirements of the precalciner for fine coal. This enables planned stops of the coal milling system for maintenance to take place for up to 5 hours per day.

This vertical roller mill is swept by inert (<10% w/w oxygen) exhaust gases from the preheater. These exhaust gases have a temperature of approximately 330ºC and are used for drying the coal while it is being ground in the mill. The temperature at the mill exit is controlled to a temperature of approximately 70ºC dependent on the dewatering curve of the coal being used to fire the precalciner.

Raw coal is delivered onto the centre of the rotating table of the vertical roller mill and drawn under the rollers by centrifugal force induced by the rotation of the table. The rollers are pressed onto the bed of coal on the rotating table by hydraulic cylinders. After passing under the rollers the ground coal falls over the dam ring around the edge of the rotating table and is picked up by the hot, inert gases sweeping the mill and carried to the internal classifier of the mill. Coal which is not of sufficient fineness for firing in the precalciner is returned to the centre of the rotating table for regrinding. Fine coal passes out of the mill with the exhaust gases and is collected in a bag filter. The collected pulverised and dried coal is delivered into a fine coal storage bin for onward delivery to the precalciner at the rate demanded by the process. The exhaust gases from the filter are vented to atmosphere through their own dedicated stack. The metered coal from the precalciner fine coal bin is delivered pneumatically to the precalciner coal burners using tertiary air as the transport gas.

Raw petroleum coke is handled in a similar way to the coal. It is imported to site, stored under cover and transported from the stockpile by a belt conveyor transport system that feeds an intermediate raw petroleum coke storage hopper. The raw petroleum coke is extracted from the storage hopper at a controlled rate and fed to the petroleum coke mill. This mill has a rated capacity of 5 tph in order to provide 25% overtaking capacity above the requirements of the kiln for fine petroleum coke. This enables planned stops of the petroleum coke milling system for maintenance to take place for up to 5 hours per day.

This vertical roller mill is swept by inert (<10% w/w oxygen) exhaust gases from the preheater. These exhaust gases have a temperature of approximately 330ºC and are used for drying the petroleum coke while it is being ground in the mill. The temperature at the mill exit is controlled to a temperature of approximately 70ºC.

Raw petroleum coke is delivered onto the centre of the rotating table of the vertical roller mill and drawn under the rollers by centrifugal force induced by the rotation of the table. The rollers are pressed onto the bed of petroleum coke on the rotating table by hydraulic cylinders. After passing under the rollers the ground petroleum coke falls over the dam ring around the edge of the rotating table and is picked up by the hot, inert gases sweeping the mill and carried to the internal classifier of the mill. Petroleum coke which is not of sufficient fineness for firing in the main burner of the rotary kiln is returned to the centre of the rotating table for regrinding. Fine petroleum coke passes out of the mill with the exhaust gases and is collected in a bag filter. The collected pulverised and dried petroleum coke is delivered into a fine petroleum coke storage bin for onward delivery to the main burner of the rotary kiln at the rate demanded by the process. The exhaust gases from the filter are vented to atmosphere through their own dedicated stack. The metered, fine petroleum coke from the main burner fine petroleum coke bin is delivered pneumatically to the main burner of the rotary kiln where it is blown into the kiln with additional primary air.

(x) Clinker Cooler

The hot clinker falls from the kiln onto the inclined, static grate of the clinker cooler at a temperature of 1300ºC. A layer of cooled clinker permanently sits on this grate protecting the plates against abrasion and overheating. Sliding of fresh, hot clinker over this layer of cooled clinker is induced by the transport of clinker away from the base of the inclined, static grate over the subsequent, horizontal grate of the cooler. Clinker is progressively transported over this horizontal grate by the reciprocating action of the tracks or bars moving above the grate within the bed of clinker using the principal of a "walking floor". Cooling air is blown into air beams beneath and through the grates along the entire length of the cooler. The air beams and grates are stationary along the entire length of the cooler with the movement of the clinker induced by the moving elements above the grate.

The clinker is cooled from 1350ºC to 50ºC above ambient temperature at the discharge of the cooler. Heat is transferred from the clinker to the cooling air stream for subsequent recycling in the process. The hottest secondary air at a temperature of 1125ºC is drawn into the rotary kiln to provide the secondary combustion air for the firing of the petroleum coke in the main burner of the kiln. Hot tertiary air at 908ºC is drawn into the tertiary air duct from above the cooler horizontal grate. The lower temperature excess air resulting from the final clinker cooling is exhausted through a bag filter system before passing to the stack.

The cooled clinker falls into the hammer crusher and then one of two metal clinker conveyors. From the main clinker conveyor the clinker can be fed to either the main clinker storage silo or to the cement mill clinker feed bins. The second, shorter clinker conveyor feeds to a clinker export hopper for lorry loading. This arrangement provides redundancy of the clinker transport system away from the kiln and cooler.

The main clinker storage silo has 60,000 tonnes capacity in order to provide sufficient storage for 20 days operation of the kiln and allow for continued supply of the market during kiln refractory lining repairs of this duration. Clinker can be recovered from the main clinker store to the cement mill clinker feed bins.

(xi) Cement Milling

Two vertical roller mills are used for cement finish grinding. Each mill has a rated capacity of 90 tph at the average Blaine 3500 cm^2/g of the cement products produced on the cement plant. This provides for the 1,000,000 tonnes per year production of the cement plant to be ground in 310 days with the mills operating for 18 hours of those days. This allows ample overtaking capacity to facilitate maintenance of the mills and for seasonality in the demand for cement from the plant.

Ambient air is used for ventilation of the mills and dedusting is provided by bag filter and fan units.

Power consumption of the finish milling system totals 35 kWh/t of cement produced.

(xii) Packing and Loading

Two cement storage silos, each of 10,000 tonnes capacity are provided for finished cement storage. This is sufficient for almost 7 days dispatches of cement from the plant. These silos are each divided into six compartments to allow for storage of up to 12 different cement products.

Cement can also be transferred to a bag packing plant from where the bagged product is conveyed to a palletiser for warehousing and storage. The warehouse is capable of storing 2,000 tonnes of palletised, bagged cement, sufficient for more than 3 days dispatches of cement in this form of packaging.

(xiii) Gas Mixer

A gas mixer is required to mix the gas from three sources prior to the wet limestone FGD system. The gases come from the kiln/calciner system, the fuel drying and the auxiliary boiler/CHP.

(xiv) Flue Gas Desulphurisation

Wet limestone flue gas desulphurisation (FGD) is situated after the flue gases have been mixed from the auxiliary boiler/CHP and the clinker production process. The flue gases pass to the absorber column where they are mixed in the reaction tank with the sprayed limestone slurry. The slurry has a residence time of approximately ten hours. The inlet temperature of the flue gas coming from the raw mill will be approximately 110°C. The limestone slurry will cool the flue gas to 45-50°C during the process. This is the ideal temperature for the CO_2 capture in the MEA and reduces the need for additional flue gas cooling. The absorber column, inlet and outlet pipes must all be manufactured using corrosion resistant materials to avoid corrosion caused by the wet portion of the flue gas. The slurry is typically a 15% concentration of calcium sulphate, calcium sulphite, unreacted alkali, inert materials, fly ash and various dissolved materials (IEA GHG, 2004). The slurry is produced using several layers of spray banks. The slurry required depends on the level of SO_2 removal required. For absorption of CO_2 with MEA very low levels of SO_2 are required (<10 ppmv) and the amount of limestone required is approximately 13,000 t/y. These should be considered as preliminary figures as substantial testing and optimisation will be required before the FGD system is installed.

The basic equation for the reaction with SO_2 is shown below:

$$CaCO_3 + SO_2 + 2H_2O + 0.5O_2 \rightarrow CaSO_4.(2H_2O) + CO_2$$

The resulting product from the reaction is wet gypsum and carbon dioxide. The gypsum can be dried and added to the clinker later in the process to create cement. The gypsum formed from the FGD process can account for around half of the gypsum required in the cement product. The carbon dioxide formed will be captured along with the rest of the CO_2 in the flue gas and the water can be recycled to form the wet limestone slurry.

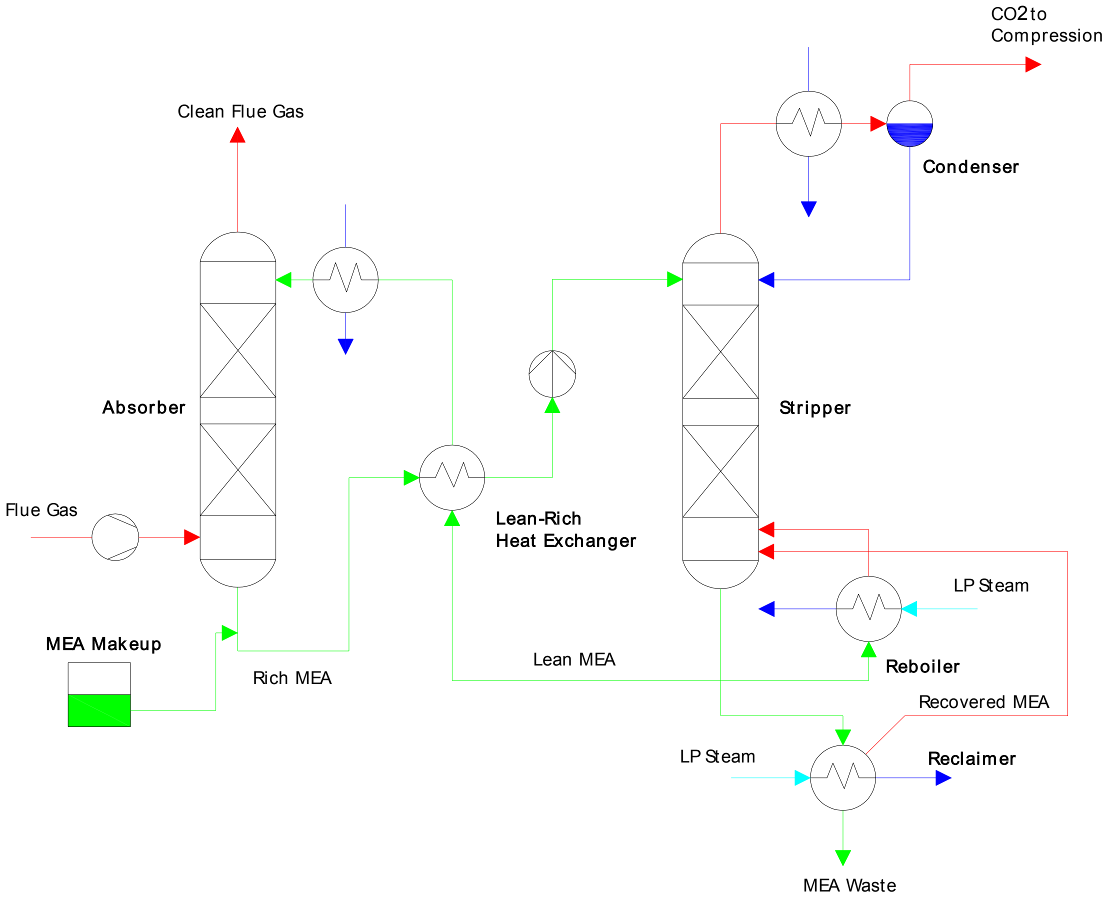
Limestone processing and storage may also be required. Limestone must be ground with the recycled FGD water to form the slurry and added to the reaction tank when required by control valves. Alternatively, pre-ground limestone could be used. Since large quantities of limestone will be delivered to site for the cement process, it is most likely that it will be more cost-effective to process the limestone on site.

(xv) Amine Capture Process

The de-sulphurised gas should leave the FGD system at approximately 50°C and will be suitable for the CO_2 absorption process. The clean flue gas is blown from the FGD unit to the packed absorber column; it enters from the bottom of the column and is forced in counter current flow with the lean MEA. The lean MEA is fed in from the top of the absorber column along with any makeup MEA required. Ideally, the CO_2 absorption and stripping cycle should be a closed loop system where all the MEA is regenerated in the stripper column and returned to the absorber. However, certain compounds in the flue gas combine with MEA to form amine salts and must be removed from the system. The amount of make-up MEA required will influenced by the levels of NO_2, SO_2, dust and other degenerative components present in the flue gas. The MEA make-up required has been estimated at 2,400 t/yr.

The CO_2 first comes into contact with rich MEA and as it travels up the packed column the MEA contains less CO_2 (leaner) and therefore will more readily absorb. The clean flue gas leaves at the top of the column and goes through a water wash (to remove any MEA slip) before exiting to the atmosphere. Depending on the amount of CO_2 left in the clean flue gas, stack gas preheating may also be required to prevent slumping. A basic process flow diagram of the CO_2 capture process is shown in Figure 4-5.

Figure 4-5: MEA CO_2 Capture System



The rich MEA flows through a heat exchanger with the lean MEA from the stripper column to increase its temperature. The rich MEA is fed into the top of the stripper column; low pressure (LP) steam increases the temperature of the rich MEA to release the CO_2 which flows out from the top of the stripper column. The LP steam demand has been estimated at 200-250 t/hr with a temperature of 140-150°C and a pressure of approximately 2.2 barg. The steam can be provided by either an auxiliary boiler or a CHP plant and these options are discussed in Section 4.4.5(xvii).

Any excess steam residing in the CO_2 is recovered using a condenser from where the CO_2 goes onto compression. A slip steram of MEA is fed into a reclaimer where it is heated to a higher temperature in the presence of Sodium Hydroxide. This recovers some of the degraded MEA and is returned to the cycle. The Waste MEA leaves the reclaimer and must be disposed. There is the possibility of using this as a fuel in the kiln but this would require further investigation (see Section 4.4.9). The hot lean MEA goes through the heat exchanger with the rich MEA and reduces in temperature before entering the top of the absorber.

Compared to power plant applications, the gas stream from the cement process is more favourable for CO_2 capture. There is less gas and the concentration of CO_2 is greater. This results in smaller absorber and stripper columns being required. The absorber column will be approximately 11.87 m in height and 10.3 m in diameter. The stripper column will have the same diameter but a height of only 8.23 m.

(xvi) CO_2 Drying and Compression

The CO_2 leaving the stripper column is already more than 95% (w/w) pure CO_2 and does not require further purification. The CO_2 leaves the condenser at atmospheric pressure and temperature. This needs to be dried and compressed to 110 bara before transportation.

(xvii) Electrical Supply

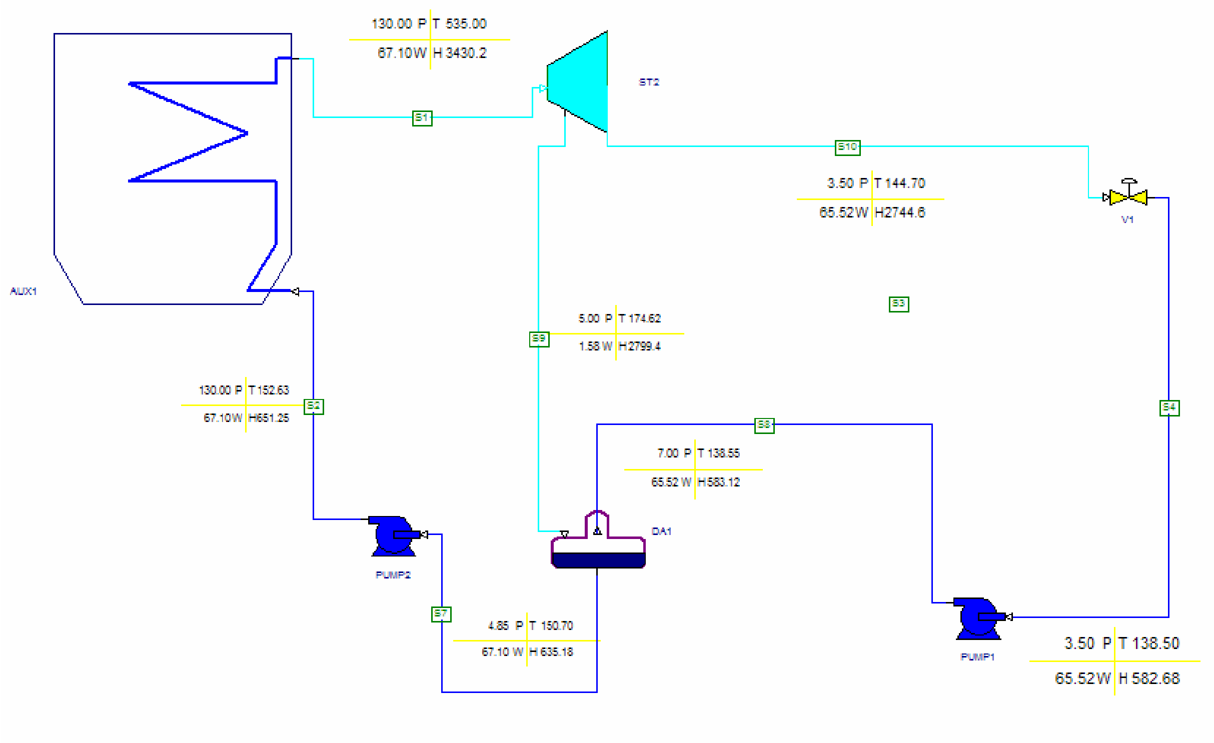
The total installed power required by the cement plant is estimated at 16.3 MWe with an additional 26.9 MWe required for the CO_2 capture plant (see Table 4-2) giving a total power demand of 43.2 MWe. The operating power, operating hours and the power consumed of all the major process units are also shown in Table 4-2. The estimation of overall CO_2 emissions of a post-combustion capture cement plant includes a coal-fired CHP plant to supply heat and power to the entire plant and it is assumed that a suitable grid connection is available.

A significant addition of power is required for the CO_2 capture plant. It can be seen from Table 4-2 that nearly 50% of the additional power required to capture the CO_2 is in the compression to 110 bara.

The steam demand for the MEA stripping can be met with a 45 MWe CHP plant. If a CHP plant is constructed onsite then there may also be some commercial benefits from exporting energy to the grid, especially if biomass is used to raise the steam as the plant would be entitled to ROCs, but this is not considered further here. The CHP could also qualify for enhanced capital allowances if it was deemed a good quality CHP. From preliminary calculations, it appears good quality CHP could be achieved, although no capital allowances are assumed for this study.

The CHP plant was sized on a 45 MWe back pressure steam turbine. A simple model using the GateCycle software assessed the sizing of the plant required. The output from the model can be seen in Figure 4-6.

Figure 4-6: GateCycle Output for CHP Plant



The model shows that the turbine requires approximately 242 t/h of steam at 130 bara and 530°C. The steam requirement calculated for the stripper is 202 t/h at 3.5 bara and 140°C. The exhaust from the turbine would be at 3.5 bara and would more than meet the steam demand of the CO$_2$ capture system. This is simplified model of the system and the steam cycle has been slightly oversized, a condenser may be required to optimise the steam flow to the stripper column. The quantity of coal required to raise this steam has been estimated at approximately 228,000 t/y.

An alternative to installing a CHP plant would be to import the electricity and provide the steam using an auxiliary boiler. This would reduce the amount of coal required and reduce the capital cost of the plant as a steam turbine would not be required. However, the benefits of producing power would not be realised.

Another option for a new-build cement plant would be to co-locate with an existing power station that could provide the LP steam to the stripper column. This would reduce the need for either a CHP plant or an auxiliary boiler.

Table 4-2: Electrical Demand of Post-Combustion Cement Plant with CO_2 Capture

Process Area	Main Equipment Items	Installed Power (MWe)	Operating Power (MWe)	Operating hours per year	Power consumed (MWh/y)
Raw Material Preparation	Limestone crusher	0.60	0.51	2080	1061
	Shale crusher	0.20	0.17	2080	354
	Stacker-reclaimer system	0.28	0.24	2080	500
Raw Mill	Raw mill drive	1.20	1.02	5730	5845
	Raw mill fan	1.50	1.28	5730	7306
	Raw mill ancillaries	-	-	-	-
	Raw mill bag filter	-	-	-	-
	Bag filter fan	1.20	1.02	5730	5845
	Conditioning Tower	1.20	1.02	5730	5845
	Homogenising chamber silo	0.12	0.10	5730	560
Preheater	Preheater feed bin	-	-	-	-
	5 stage preheater	-	-	-	-
	Preheater fan	0.90	0.77	7920	6059
Precalciner	Precalciner	-	-	-	-
Rotary Kiln	Kiln	0.65	0.55	7920	4349
Clinker Cooler	Grate cooler	1.00	0.85	7920	6732
	Cooler bag filter				
	Cooler bag filter				
	Clinker crusher				
	Clinker conveyor				
	Clinker storage silo				
	Export clinker hopper				
Coal preparation	Coal milling system	0.30	0.26	6095	1554
Petroleum coke preparation	Petroleum coke mill	0.30	0.26	6095	1554
Cement milling	Mill 1	3.15	2.68	5580	14940
	Mill 2	3.15	2.68	5580	14940
Packing and loading	Silo 1	-	-	-	-
	Silo 2	-	-	-	-
	Bag packing plant	0.50	0.43	7920	3366
Total for cement plant		**16.24**	**13.81**	**-**	**80809**
Selective Catalytic Reduction (SCR)	Reactor Casing	0.35	0.30	7920	2356
	Bypass System				
	Catalyst				
	Ammonia Injection Equipment				
	Handling Equipment				

Process Area	**Main Equipment Items**	**Installed Power (MWe)**	**Operating Power (MWe)**	**Operating hours per year**	**Power consumed (MWh/y)**
	Control System				
Flue Gas Desulphurisation (FGD)	Spray Tower Absorption Reaction Tank Mist Eliminators Limestone Storage Limestone Preparation Dewatering System	2.10	1.79	7920	14137
Other	Gas Mixer	-	-	-	-
CO_2 Capture	Pumps (MEA, Reflux, Stripper etc) Amine Filter Package Water Tankage Cross Exchangers Stripper Coolers (DCC and water wash) DCC Towers Overhead Stripper Condenser Stripper Reboiler Lean Solvent Cooler Packing Absorption Tower Flash Drums	3.90	3.32	7920	26255
CO_2 compression and purification	Dryer Compressors CO_2 Pumps	23.71	20.15	7920	159618
CHP Plant	Coal and Ash Handling (incl. milling) Boiler Island Steam Turbine plant and Generator Selective Catalytic Reduction Balance of Plant	7.50	6.38	7920	50490
Total for CO_2 Capture Plant		**37.56**	**31.93**	**-**	**252857**
Total for whole plant		**53.80**	**45.73**	**-**	**333666**

4.4.6 Plant Performance

The performance of the cement plant with post-combustion CO_2 capture will only be affected by the addition of an SCR unit between the preheater and the raw mill. This will reduce the temperature of the flue gas used for drying in the raw mill. The MEA CO_2 capture process will not affect the cement process.

It should be noted that the flue gas leaves the FGD unit with a CO_2 concentration of 26.4% (w/w). Research and preliminary testing indicate that the CO_2 capture rate using amine scrubbing could be between 85% and 95%. For consistency with other IEA GHG studies the CO_2 capture rate has been conservatively assumed at 85%.

Table 4-3 summarises the performance of the post-combustion CO_2 capture cement plant.

Table 4-3: Performance of Cement Plant with Post-combustion CO_2 capture

	Parameter	Unit	Value
Inputs	Raw meal consumed	t/y	1,494,526
	Coal required	t/y	291,633
	Petroleum coke required	t/y	32,876
	Limestone for FGD	t/y	12,795
	Water for FGD	t/y	115,152
	Ammonia	t/y	1,855
	MEA Make Up	t/y	2,400
	Low Pressure Steam (@ 3.5 bara and 141°C)	t/y	1,662,000
	Electricity consumption	MWh/y	333,666
Outputs	Clinker produced	t/y	910,000
	Cement produced	t/y	1,000,000
	Gypsum produced (FGD)	t/y	16,374
	Net Electricity Exported	MWh/y	22,734
	CO_2 captured	t/y	1,067,734
	CO_2 emitted	t/y	188,424

(i) Evaluation of MEA Absorption

Although it has been assumed in this study that the CO_2 capture rate using MEA is 85%, the University of Regina, Canada has modelled CO_2 capture from the predicted cement flue gas with a 95% capture rate. The modelling was based on early data provided by Mott MacDonald and cannot be compared directly with other outputs in this Report. However, it gives a further indication of the possibilities of CO_2 capture from this gas composition. The results from this study can be seen in Table 4-4. Two options to achieve the 95% capture rate by using different CO_2 loading and liquid circulating rates were considered.

Table 4-4: Outputs from MEA Modelling with 95% Capture

Parameter	Option 1	Option 2
CO_2 Feed (t/day)	4680	4680
Absorption liquid		
MEA concentration (wt%)	30	30
CO_2 lean loading (mol/mol)	0.22	0.18
CO_2 rich loading (mol/mol)	0.45	0.43
Liquid circulating rate (t/h)	3,823	3,485
Solvent regeneration		
Energy requirement (kJ/kg CO_2)	3,384	4,832
Energy requirement (kJ/h)	6.246×10^8	8.918×10^8
Steam pressure (MPa)	0.35	0.35
CO_2 Captured (t/day)	4,429	4,429

4.4.7 Plant Layout

Cement plants are typically linear in layout for an economical flow of material. A 1 Mt/y cement plant without CO_2 would expect to occupy approximately 15 ha (500 m x 300 m) of land but could accommodate less space if necessary. Post combustion capture would require a considerable amount of additional land. Although it is difficult to estimate the area required, Table 4-5 gives some indication of the footprint required for the CO_2 capture plant.

Table 4-5: Estimated Footprint for CO_2 Capture in a Cement Plant

Equipment	Footprint (ha)
Cement Plant	Approx. 15
SCR	Included above
FGD and limestone processing	1.0 – 1.4
CHP Plant	Approx. 2
MEA Capture System	0.5 - 0.7
CO_2 Drying and Compression	0.1 - 0.16
TOTAL	**18.6 – 19.3**

There may also be a limestone quarry that would occupy a much larger area, possibly as much as 200 ha (2000 m x 1000 m). A detailed layout of post-combustion cement plant is beyond the scope of this study but it is important to note that the following will need to be considered:

- Hazardous materials (e.g. MEA, ammonia) should be stored at a safe distance from other buildings.
- The ancillary buildings should be located so as to minimise the time spent by personnel in travelling between buildings.
- Administration offices should be located well away from potentially hazardous processes.
- Utility buildings should be sited to give the most economical run of pipes to and from the process units.

- Equipment should be located so that it can be conveniently tied in with any future expansion of the site.

4.4.8 Operation Considerations

(i) Operating with Raw Mill Off

It has been assumed that the raw mill will be inactive for 15% of the time due to the high maintenance requirements. The raw milling has been oversized and has an overtaking capacity above the kiln which enables planned maintenance on the mill to take place. During the periods of maintenance the exhaust gas from the SCR will bypass the raw mill circuit. A gas conditioning tower is situated in parallel with the raw mill for this purpose. Moisture is introduced to the gas via atomising sprays to reduce the temperature. It should still be possible to capture all of the CO_2 if the raw mill is not operating although the SO_2 concentration in the flue gas will increase (due to the lack of absorption within the raw mill) and will increase the SO_2 removal required from the FGD system. Hence, it is necessary to design the FGD system for the maximum loading rate when the raw mill is off.

(ii) SCR or FGD not Operating

If the SCR or FGD is not operating this will allow increase concentrations of SO_2 and NO_2 to pass through to the MEA CO_2 absorption system. Amines react with acidic compounds to form amine salts that will not dissociate in the amine stripping system. Thus, under these conditions, there will be more waste MEA that must be removed from the capture cycle. A decision must be made at this point as to whether it is more economic to pay for the extra cost of MEA required or to stop the CO_2 capture altogether whilst the SCR and FGD units are non-operational.

Maintenance of the units should be planned for periods when the cement plant is not operating to minimise the amount of downtime of the CO_2 capture system.

4.4.9 Emissions and Environmental Considerations

A post-combustion capture cement plant will achieve emission standards well below current cement plants. This is an added advantage of CO_2 capture. Some emissions will occur and these are discussed in the following sections.

(i) Emission Sources

During normal operation there will only be two major gaseous discharges to the atmosphere. These will be the exhaust gases from the MEA absorber and excess cooling air from the clinker cooler. The flue gases will be discharged via a single stack. Due to the requirements of the MEA CO_2 capture system, flue gas clean up systems must be employed to reduce the levels of contaminants down to acceptably low values. Any contaminants passing through to the MEA system (e.g. ammonia, NO_2, SO_2) will combine with the MEA and be removed as part of the MEA waste. This results in the discharged flue gas being virtually free of any contaminants.

(ii) Dust

A large amount of dust is carried through in the flue gases during the cement process. From the raw mill itself approximately 1 kg/s of dust must be eliminated from the flue gas before it reaches the MEA. This is achieved with the addition of bag filters that can capture 99.9% of the dust in the flue gas. The requirement for the CO_2 capture system is 15 mg/Nm^3 and the bag filters can achieve as low as 5 mg/Nm^3. Any dust that does enter the absorber will combine with the amine and will be removed with the waste. This results in negligible amounts of dust reaching the stack. Any dust caught in the bag filters will be recycled into the clinker at a later stage.

Dust also forms in the coal and pet coke drying system, as well as the clinker cooler. Discharges from these processes will pass through bag filters prior to discharge.

(iii) NO_x Emissions

NO_2 emissions must be reduced to 20 ppmv for the MEA CO_2 capture system. This is achieved using SCR. Therefore, the overall NO_x emissions to the atmosphere will be very low. This level of NO_2 is well below the usual emission regulations that apply at cement plants.

(iv) SO_x Emissions

As discussed in Section 0, modern cement plants which do not have sulphide minerals in their raw materials have very low emissions of SO_x and often achieve emission levels below 10 mg/m^3 of SO_2 without using abatement techniques (EC, 2001). In this case it is assumed that the raw meal originates from NE Scotland and contains 0.33% (w/w – dry basis) sulphur.

It is expected that 30% or more of the sulphur in the raw material will be evaporated and be emitted from the first stage of the preheater and that SO_2 will be the main (99%) sulphur compound to be released (EC, 2001). The gases from the preheater are used for drying within the raw mill (when in operation) and for fuel drying. It is reported that between 20-70% of the SO_2 that is passed to the raw mill will be captured by the finely ground raw materials (EC, 2001).

SO_2 will also be formed in the CHP plant and will combine with the flue gas from the cement process in the gas mixer before the wet limestone FGD system. The FGD system will reduce the SO_2 content to approximately 10 ppmv to meet the requirements for MEA CO_2 capture. Thus, the flue gas emitted to the atmosphere will contain minimal amounts of SO_2.

(v) Effluent

During normal operation there will be three main effluent discharges. These are:

- chloride purge in the FGD system
- waste MEA
- water obtained from temperature swing dual bed desiccant dryer.

These discharges will either be recycled or disposed of in an appropriate manner.

The condensed water obtained from the temperature swing bed desiccant dryer will contain dissolved acid gas components which may require neutralisation prior to discharge.

Periodically some of the recycle effluent from the FGD system must be removed to avoid build up of chloride in the limestone slurry. The waste from this effluent can be dried and disposed in a landfill site.

MEA waste is caused by contaminants from the flue gas and has been estimated at 2,400 t/y for this cement plant. MEA is an odorous and corrosive substance and should be handled carefully when being disposed. It also has a calorific value of approximately 22 MJ/kg which offers the possibility of burning it in the cement kiln. Disposing of the waste MEA in this manner requires further detailed evaluation as there will be environmental regulation issues associated with disposing of a waste material in this manner (e.g. Waste Incineration Directive). It is likely that combustion of the MEA would produce more NO_x which would increase the capital and operating costs for the SCR system. No costs associated with any effluent treatment have been included in the cost estimate for the post-combustion process option.

4.4.10 Design Risk Assessment

It is of paramount importance that a post-combustion cement plant is implemented safely and without significant impact upon operational flexibility. Mott MacDonald have undertaken a preliminary design risk assessment to identify the major additional plant operation risks that need to be considered beyond those that apply for a cement plant without carbon capture. The results of this exercise are shown in Table 4-6. It is important to note that this exercise would have to be repeated in greater detail if a detailed design of an post-combustion cement plant was undertaken.

Table 4-6: Preliminary Design Risk Assessment for Post-Combustion CO_2 Capture Cement Plant

Hazard Ref	Hazard	Stage of work	Risk Control Measures	Owner	Comment
1	Risk of operator suffocation as CO_2 is an asphyxiant	Operation	1. Standards, codes of practice etc. to be adopted during design. 2. Appropriate material selection during design. 3. HAZOP study to be undertaken.	Contractor	Hazards associated with handling CO_2 are well understood from other industries.
2	Risk of dust explosion in fuel milling and drying stages.	Operation	1. Standards, codes of practice etc. to be adopted during design. 2. HAZOP study to be undertaken. 3. ATEX assessment to be	Contractor	Hazards associated with explosive dusts are well understood within cement

Hazard Ref	Hazard	Stage of work	Risk Control Measures	Owner	Comment
			performed.		industry.
3	Risk of corrosion damage due to presence of acidic components (e.g. SO_2, SO_3 and HCl) in gas streams.	Operation	1. Standards, codes of practice etc. to be adopted during design. 2. Appropriate material selection during design.	Contractor	Hazards associated with acidic components in gas streams are well understood from other industries.
4	Risk of injury to operator due to handling of ammonia solution.	Operation	1. Standards, codes of practice etc. to be adopted during design. 2. Appropriate material selection during design. 3. HAZOP study to be undertaken.	Contractor	Hazards associated with handling of ammonia are well understood from other industries.
5	Risk of operator suffocation as N_2 is an asphyxiant.	Operation	1. Standards, codes of practice etc. to be adopted during design.	Contractor	Hazards associated with handling N_2 are well understood from other industries.
6	Risk of operator injury as MEA is a corrosive material.	Operation	1. Standards, codes of practice etc. to be adopted during design. 2. Appropriate material selection during design. 3. HAZOP study to be undertaken	Contractor	Hazards associated with handling MEA are well understood from other industries.
7	Risk of damage caused by release of high pressure as high pressures present in CO_2 compression unit.	Operation	1. Standards, codes of practice etc. to be adopted during design. 2. HAZOP study to be undertaken	Contractor	Hazards associated with handling high pressures are well understood from other industries.

4.4.11 Possibilities for Alternative CO_2 Absorption Technologies

As discussed in section 3.4.2 there are several other alternatives to amine absorption for the capture of CO_2 that are being researched in the power generation industry. Although they are less commercially and technically advanced as amine absorption, there may be some differences in the technologies that are more advantageous for the cement industry. In the section below one alternative technology which has received significant interest from the power generation industry is examined as a possible alternative CO_2 absorption technology that could be applied to the cement process. There are plans to build larger chilled ammonia plants. AEP is planning on building a 200 MWe plant by 2011.

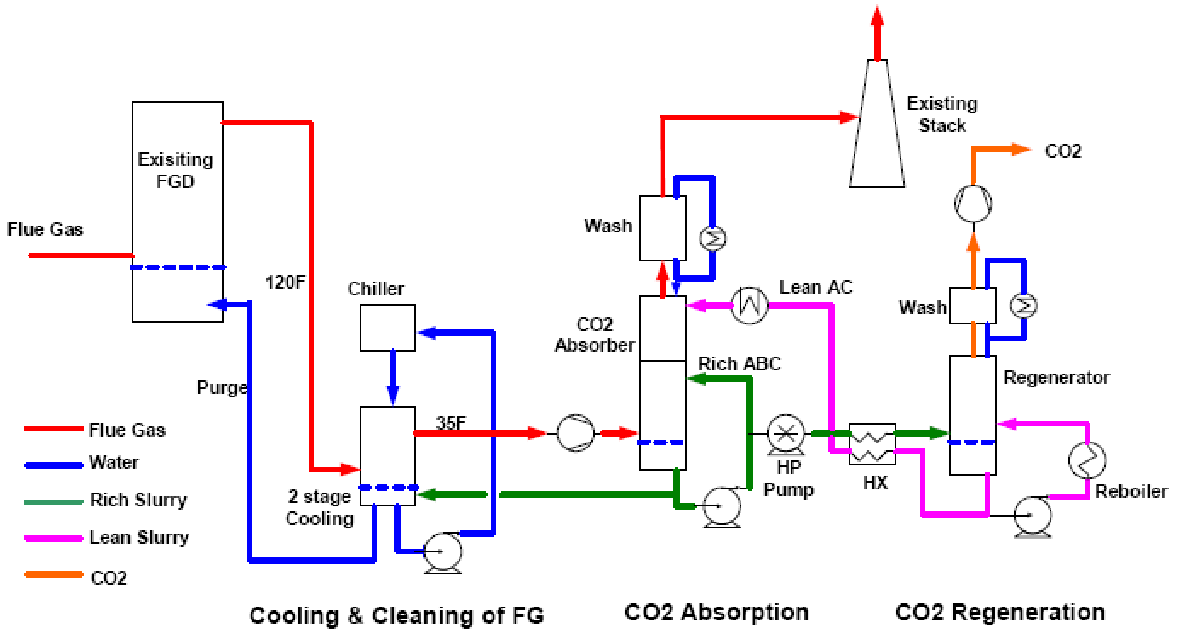
(i) Chilled Ammonia

The Chilled Ammonia Capture (CAP) process is currently being developed by Alstom. Small and large scale bench tests have been completed with a full scale (14,000 tons of CO_2 per year) field test to begin in the near future at We Energies Pleasant Prairie Power Plant located in Pleasant Prairie, Wisconsin, USA (Black *et al.*, 2007). EPRI will be conducting the techno-economic analysis to scale up system for commercial operation.

Process Description

The process description is based on the CAP process applied to power plants but there may be several differences when adapted to the cement industry. The CAP process has three main stages: flue gas cooling, CO_2 absorption and solvent stripping. Figure 4-7 shows a process flow diagram.

Figure 4-7: Chilled Ammonia Process Diagram



The flue gas from the cement plant will already have been treated to remove the NOx, SOx and dust and will be at approximately 50-60°C. The flue gas is cooled to 0-20°C in two stages using direct contact cooling (DCC). During the first stage, the flue gas is cooled to approximately 37°C and is fully saturated with water on exit. The condensate and cooling water are sent to a cooling tower after processing. Here, most of the condensate is evaporated and the water is cooled back down to 27°C.

During condensation in the DCC units, some residual contaminants will be entrained in the condensate making it acidic and corrosive; this must be taken into account when design the cooling units.

Before the second stage, the flue gas must be pressurised to approximately 1.25 bara to continue through the CO_2 absorption process. During the second stage, the flue gas is contacted with chilled water (approximately 3°C) and is cooled to 5°C. Some of the cooling water is then used to cool the water in the first stage and the rest goes back to the water chiller.

The CO_2 absorber is in the form of a spray tower designed to operate with slurry. The flue gas flows up the tower counter current to both an aqueous ammonium carbonate solution and a slurry containing aqueous ammonium bicarbonate, solid ammonium bicarbonate and aqueous ammonium carbonate. The clean flue gas is released at the top of the absorber to a water wash, removing any ammonia still entrained. The flue gas is then heated to 60°C to remove any visible plume formation and prevent ammonia odours reaching the ground before releasing it to the atmosphere.

Before going to the stripper all the solids in the CO_2 rich slurry must be dissolved. The lean CO_2 solution is used to heat the CO_2 rich slurry in a heat exchanger to approximately 75°C, at which point all the solids are dissolved.

The stripper is a packed column that operates at high pressure and temperature (31 bara and 127°C). The heat causes the absorption reaction to reverse leaving aqueous ammonium carbonate and the released CO_2. The CO_2 may still contain trace components of ammonia that can be removed by cooling to 20-30°C and contacted with a water wash unit. This reduces ammonia costs and helps prevent corrosion in pipes downstream.

The CO_2 can then be compressed to 110 bara but this requires less stages than the amine process because the CO_2 is already compressed to 30 bara.

Potential Advantages

Notwithstanding the energy demands of water cooling, the CAP process may be more energy efficient than the amine capture process for the following reasons:

- It re-uses greater low-grade waste heat generated in the process, reducing the useful heat consumption of the system.
- The slurry is pressurised to 30 bara, thus reducing the compression ratio required for the CO_2 at the end of the process.
- The power required to compress the CO_2 rich liquid is small in comparison to the power required to compress CO_2 gas. Although the total pressure increase required is ultimately the same, an energy saving is made by compressing the liquid rather than the gas in the CAP.
- The ammonia solution has a high capacity to dissolve CO_2, which can reduce the volume of solution with obvious advantages of reduced pump sizes, heat exchangers etc.
- Ammonia is a readily available chemical and will be cheaper and in direct comparison to amine it will be easier and cheaper to source. Ammonium carbonate and ammonium bicarbonate are very stable in solution and are not degradable, thus the make-up requirements are negligible.

- The stripping of the CO_2 from the ammonia solution also requires less heat input than that for the equivalent amine process.

Technical Barriers

There are some technical barriers that must be overcome before CAP becomes commercially viable for both the power and cement industries. These are:

- Additional heat is required to dissolve the ammonium bicarbonate solids before transferring to the stripper column that may cancel out the energy benefits of the low energy absorption.
- The precipitation of the ammonium bicarbonate solids is an exothermic reaction requiring additional cooling.
- Ammonia slip into both the clean flue gas and the CO_2 captured requires additional water scrubbing that may not be economical.
- A large amount of refrigeration is required to keep the cooling water at 3°C, which will consume electricity.
- Full scale tests have still not been conducted and may highlight other performance issues that have not been mentioned in this report.

Capital and Operating Costs

CAP is an emerging technology and so far has only been considered for the power generation industry. There is no cost data currently available for the process as applied to the cement industry but Mott MacDonald understands that the approximate EPC price for CAP is around $2,200/kW and that preliminary indications are that the price of CAP is higher than the cost of MEA absorption. Mott MacDonald recommends that these figures are regarded with a degree of caution as the cost for the technology is not fully established. The equipment list for the CAP process is listed below:

Flue Gas Cooling

DCC units

Cooling tower

Pumps

CO_2 Capture and Regeneration

Spray Tower Absorber

Packed Bed Stripping Unit

High Pressure Slurry Pumps

Slurry recycle pumps

Rich/Lean Heat Exchanger

Flue Gas/CO_2 Heat Exchanger

Water Wash Units and Accessories

Water Chillers

CO_2 Compression and Drying

The operational costs will likely be lower than for the MEA system because CAP requires less steam for regeneration of the CO_2. The cooling water, ammonia and power for the pumps will be the main costs linked with the operation of the plant.

4.5 Evaluation of Oxy-Combustion CO_2 Capture

4.5.1 Overview

As described in section 3.3.3 oxy-combustion technology as a means to achieve near-zero CO_2 emission power generation has been the subject of a significant number of research papers and reports (e.g. IEA GHG, 2005). Although Mott MacDonald does not anticipate full-scale commercialisation of oxy-combustion power plants to occur until around 2015 a small number of Original Equipment Manufacturers (OEMs) are developing the technology. Doosan Babcock, for example, is focussing its main efforts within CO_2 capture technologies on oxy-combustion for pulverised fuel power generation both as a retrofit technology and for applications to a new-build advanced power generation plant (Panesar and Sekkappan, 2007).

As described in section 3.3.3 the principle of oxy-combustion for power plants is to combust the fuel in an oxygen-rich atmosphere to create a CO_2-rich exhaust that can be captured. This principle can also be applied to the cement production process. In an oxy-combustion cement production process the calcination and fuel combustion are undertaken in an oxygen-rich atmosphere to produce a flue gas comprising mainly CO_2 and water vapour. Although the concept is not new for the cement industry it is not being widely researched. Frank Zeman and Klaus Lackner at Columbia University, USA appear to be the only researchers to have published papers on the subject. Dr Zeman was consulted as part of this study.

4.5.2 Major Technical Issues

The major technical issues associated with the oxy-combustion cement production process are discussed below. Most of these issues are topics for fundamental research and are beyond the scope of this study. Although the feasibility of an oxy-combustion cement process remains to be proven, none of the issues identified was considered a “show-stopper” at the time of writing the report.

(i) Flame Temperatures and Dilution

The oxy-combustion process is based on excluding the inert components of air from the combustion process. These components, mainly composed of nitrogen gas, pass through the combustion chamber during conventional air firing without chemical change. The presence of the inerts does, however, play an important role as it dilutes the combustion process gases to yield acceptable flame temperatures. Flame temperatures in excess of 3500 ºC can be achieved by oxygen combustion. This is too hot for normal operation so it is therefore essential that within an oxy-combustion process a proportion of the CO_2-rich flue gases are recycled back to the combustion zone to provide the necessary dilution. The flame temperature can be adjusted by varying the flue gas recycle ratio.

However, due to the requirement for high temperatures to achieve calcination and clinker formation the fact that it is far easier to reach high temperatures in oxygen combustion as compared to air combustion could be seen as an advantage for operating the cement process in this manner as it may lead to shorter residence times and permit reductions in the size.

(ii) Heat Transfer Characteristics

The radiant heat fluxes and convective heat transfer performance of the combustion chamber are a function of the gas composition in the chamber. Changing the atmosphere within the combustion chamber from one dominated by nitrogen to one dominated by CO_2 will have a significant effect on the heat transfer characteristics. It should be noted that CO_2 has a higher molar heat capacity [C_p = 37.84 J/(mol K) at 20 ºC, 1 bar(a)] than nitrogen [C_p = 28.86 J/(mol K) at 20 ºC, 1 bar(a)] which may actually improve the heat transfer characteristics of the combustion chamber.

(iii) Feed Lifting

Oxygen enrichment of cement kiln combustion has been practiced for a number of years on some kilns in the USA. Clark (2006) reports that the benefits in terms of increased clinker output are limited to ~15 per cent additional output for a ~2 per cent oxygen enrichment. Clark (2006) states that the reason for this is thought to be that nitrogen ballast in the exhaust gases plays an important role in lifting the feed between cyclone stages in the suspension preheater of a modern cement kiln. CO_2 is a denser gas than nitrogen and should be more effective in this feed lifting role within the suspension preheater. Hence, the recycled CO_2 is important not only for reducing the flame temperatures but also for a suspension preheater to work effectively.

(iv) Wear and Tear

Due to the higher flame temperatures it is generally understood that cement kiln wall deterioration will increase at higher oxygen concentrations. Hence, there is a balance between achieving high temperatures for the cement production process and having to replace the kiln wall lining. Literature (e.g. Wrampe and Rolseth, 1976) suggests that 22-23% (w/w) “contained oxygen level” (or equivalent oxygen content of combustion air) is the optimum concentration in conventional kilns. Zeman (2007) indicated that a higher oxygen concentration of approx. 28.5% may be more optimal for an oxy-combustion system.

(v) Process Chemistry

Research is on-going to determine whether clinker formation in a different atmosphere will still generate a useful product. Zeman and Lackner (2006) have indicated that they intend to undertake studies to compare conventional clinker with that produced in a CO_2 environment but no results have yet been published. Zeman (2008) reported the following to Mott MacDonald regarding his findings to date with changing the dominant gas in the kiln from nitrogen to carbon dioxide using a raw meal with a composition of 80.5% lime, 14.4% silica, 3.7% alumina and 1.4% hematite:

- In a CO_2 atmosphere the onset of calcination is delayed until the reaction temperature reaches 900ºC.

- In a CO_2 atmosphere re-carbonation occurs immediately after the temperature drops below 900ºC thus placing an emphasis on a stable temperature profile between the precalciner and the kiln as any re-carbonation will necessitate re-calcination.

- Despite the high partial pressure of CO_2 the evolution of CO_2 nears completion at temperatures above 1100ºC indicating that temperatures in the kiln are high enough to fully calcine the raw meal.

- Brunauer Emmett Teller (BET) measurements showed a marked reduction in specific surface area for raw meal heated in a CO_2 atmosphere compared to raw meal heated in a N_2 atmosphere. The reduced surface area may affect low temperature belite formation. However, once temperatures reach 1200ºC the available surface area is similar for both environments.

- Once the calcination reaction is complete there is no difference between samples prepared in either atmosphere prior to the onset of clinker formation.

- The degree of sintering and chemical reactions induced by clinker formation at high temperatures prevents re-carbonation during cooling with CO_2.

Further fundamental research into the thermodynamics, kinetics and equilibrium chemistry of driving off the CO_2 in a CO_2-rich atmosphere is required to further develop the oxy-combustion cement process. This research was beyond the scope of this report.

(vi) Air Dilution

Air in-leaks occur in the raw mill, preheater and kiln. Excessive air in-leaks will result in contamination of the CO_2-rich exhaust gas with nitrogen, oxygen and argon. These contaminants will require removal which will increase the costs of operating an oxy-combustion cement process. Any design for an oxy-combustion cement process must minimise air in-leaks.

(vii) Flue Gas Clean-up

Depending on the final storage location of the CO_2 the gas will require some clean-up. It may be necessary to remove water vapour, nitrogen, argon, NO_x and SO_x from the flue gas stream. However, as the majority of the nitrogen has been removed from the combustion chamber levels of thermally generated NO_x should be lower than in a conventional air fired system.

(viii) Additional Power Requirements

There will be additional power required for the compression of the CO_2. Typically, this additional power requirement for compression of the CO_2 will be around 0.146 kWh/kgCO_2 (IEA GHG, 2005).

(ix) Air Separation Unit (ASU)

An on-site air separation unit will be required to deliver oxygen to the oxy-combustion process. This will require energy and increase the electricity demand of the plant compared with a conventional plant. The power required by a cryogenic oxygen plant is typically in the range 200-240 kWh/tO_2 (IPCC, 2005).

IEA GHG have recently published a study (IEA GHG, 2007b) which includes descriptions of cryogenic and membrane oxygen production processes, their components and their integration into power generation and synthesis gas production processes with CO_2 capture. Readers are referred to this study for further information on recent and potential future improvements in these oxygen production processes.

(x) Reducing Conditions

It is recognised that the clinker must not be generated in reducing conditions. This necessitates that an excess of oxygen is maintained in the process. It is generally considered that the oxygen concentration should be maintained >2% (w/w).

(xi) Heat Integration

Oxy-combustion offers potential opportunities to maximise performance through integration of the additional oxy-combustion process units with the cement process. Options available are as follows:

- heat recovery from additional flue gas cooling
- heat recovery from CO_2 compression and ASU compressors
- direct mechanical drives for compressors.

Although heat integration between the CO_2 compression plant, ASU compressors and the cement plant could lead to energy savings it is important to understand that this may reduce the operational flexibility of the plant. There may be valid strategic reasons for having the capability to operate the cement plant without the CO_2 compression plant or ASU. Full heat integration may make this difficult to achieve.

4.5.3 Configurations Considered

During the development of the conceptual design of an oxy-combustion cement plant a number of different configurations were considered. These are illustrated in Figure 4-8 to Figure 4-11. They can be summarised as:

- **Configuration 1:** Diversion of preheater exit gases for fuel drying and subsequent capture or recycle.
- **Configuration 2:** Capture and recycle of all flue gases from cement plant with diversion of cooler exhaust for fuel drying.

- **Configuration 3:** Capture and recycle of all flue gases from cement plant with diversion of cooler exhaust for fuel drying and a 'leak-free' raw mill.

- **Configuration 4:** Dual preheaters with oxy-combustion in the precalciner only.

Figure 4-8: Oxy-Combustion Cement Plant Configuration 1 - Diversion of Preheater Exit Gases for Fuel Drying and Subsequent Capture or Recycle

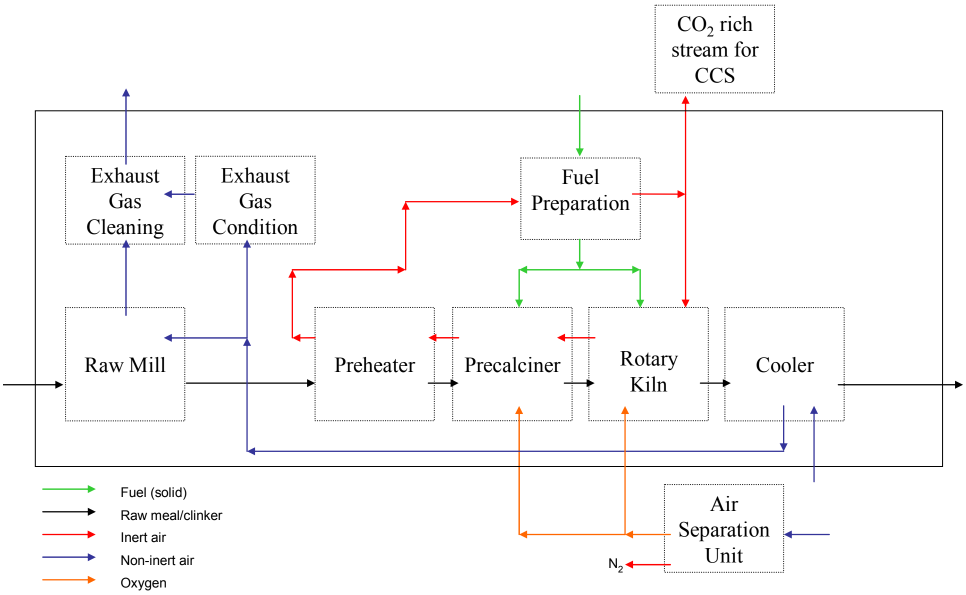


Figure 4-9: Oxy-Combustion Cement Plant Configuration 2 - Capture and Recycle of all Flue Gases from Cement Plant with Diversion of Cooler Exhaust for Fuel Drying

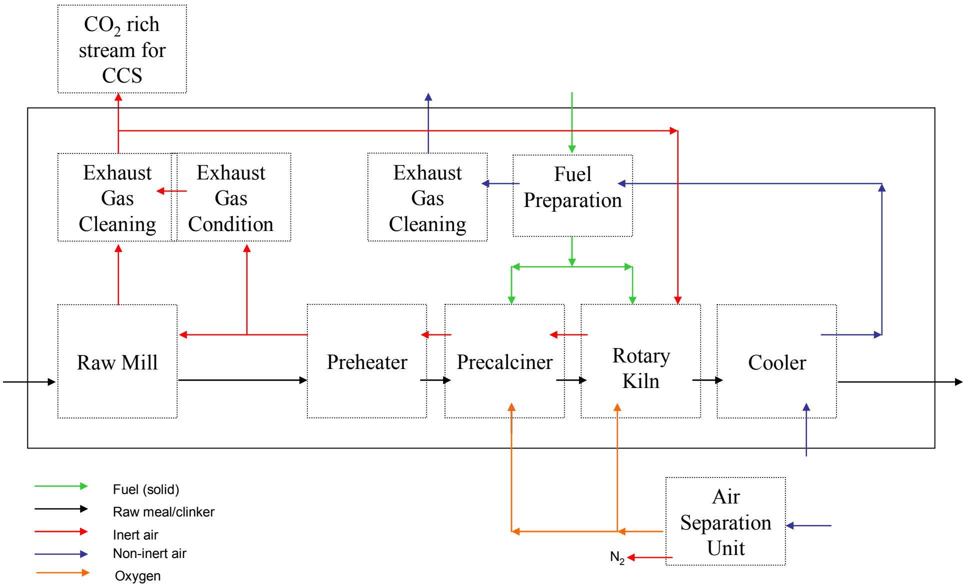


Figure 4-10: Oxy-Combustion Cement Plant Configuration 3 - Capture and Recycle of all Flue Gases from Cement Plant with Diversion of Cooler Exhaust for Fuel Drying and a 'Leak-free' Raw Mill

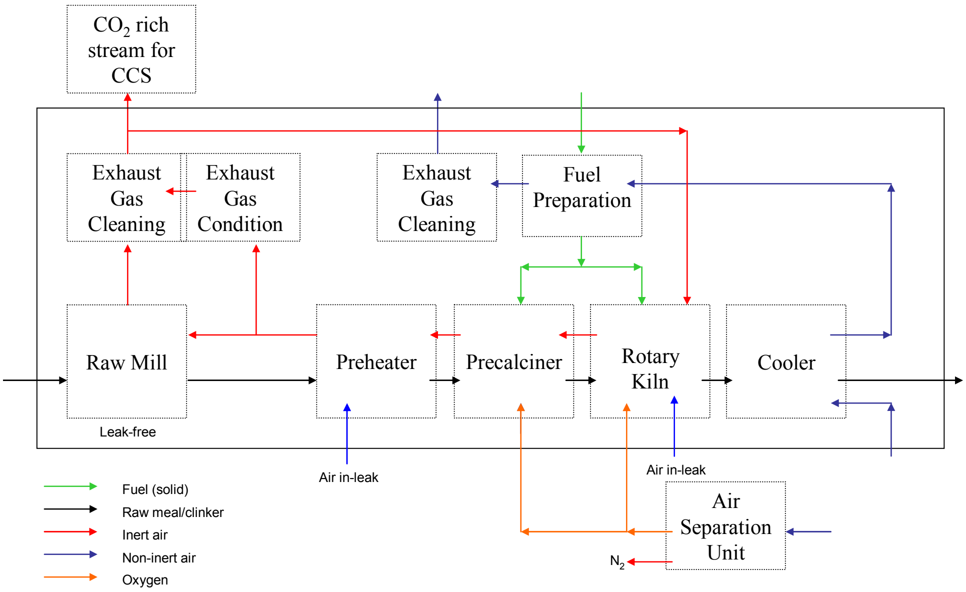
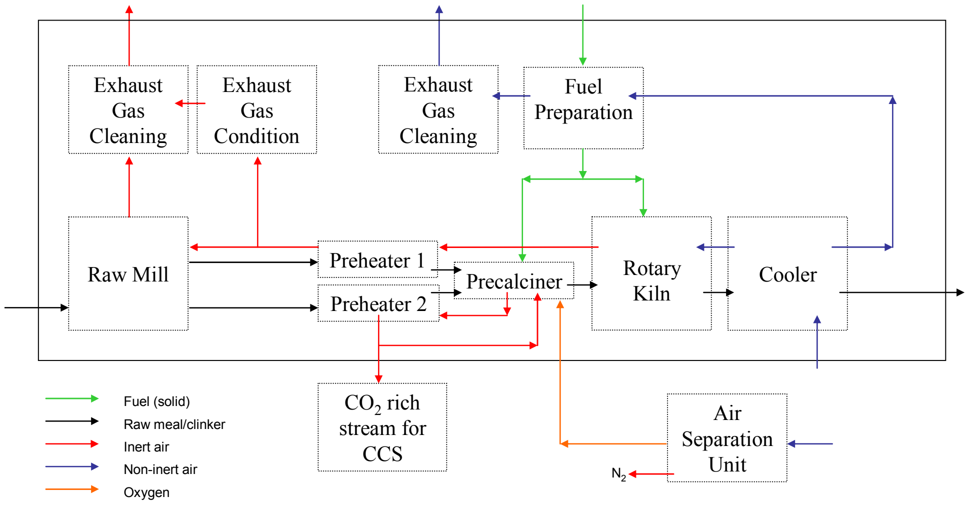


Figure 4-11: Oxy-Combustion Cement Plant Configuration 4 - Dual Preheaters with Oxy-Combustion in the Precalciner only



The advantages and disadvantages of the different configurations are compared in Table 4-7.

Table 4-7: Comparison of Different Oxy-Combustion Configurations

Configuration	**1**	**2**	**3**	**4**
Advantages	The CO_2 generated in both the precalciner and the kiln is captured. Air in-leaks from raw mill are avoided. Inert gas is used for drying the fuel.	The CO_2 generated in both the precalciner and the kiln is captured. Heat demands are better matched than 1.	The CO_2 generated in both the precalciner and the kiln is captured. Heat demands are better matched than 1. Air in-leaks from raw mill are avoided.	Heat demands are better matched than 1. Kiln operates conventionally. Air in-leaks from kiln and raw mill are avoided.
Disadvantages	Heat demands are not matched. The drying requirement is greater in the raw mill than the fuel preparation. The heat available from	Kiln operation with CO_2-rich atmosphere unproven. Non-inert gas is used for drying the	Kiln operation with CO_2-rich atmosphere unproven. Non-inert gas is used for drying the	Only the CO_2 generated in the precalciner is captured. Non-inert gas is used for drying the fuel.

	the preheater exhaust gases is not used effectively. Kiln operation with CO_2-rich atmosphere unproven. The layout of cement plants is such that the raw mill is often remote from the air cooler. Air in-leaks in kiln and preheater are not avoided.	fuel. This can cause operational difficulties as it is a less safe mode of operation. Air in-leaks in kiln, preheater and raw mill are not avoided.	fuel. This can cause operational difficulties as it is a less safe mode of operation. Requires design of a leak-free raw mill resulting in significant capital cost increase. Air in-leaks in kiln and preheater are not avoided.	This can cause operational difficulties as it is a less safe mode of operation.
Technical risk	High	High	High	Medium

4.5.4 Chosen Configuration

The operation of the precalciner was critical in determining the configuration which was taken forward for detailed evaluation. Precalciners in a modern cement plant consume ~60% of the fuel demand and are operated to undertake up to 95% of the calcination. Hence, the majority of the CO_2 emissions within a modern cement plant originate within the precalciner. The option to capture the majority of CO_2 emissions from the cement plant by focussing on a single unit without affecting the rest of the cement plant was considered to be the most cost effective and lowest risk configuration for reducing the CO_2 emissions from a cement plant using oxy-combustion. Furthermore, by focussing only on the precalciner the difficulties of undertaking oxy-combustion in the kiln and the inherent problem of air in-leakage in the kiln and raw mill were avoided. Hence, configuration 4 was taken forward for detailed evaluation.

4.5.5 Process Description

A detailed evaluation of configuration 4 was undertaken. It should be noted that the final configuration analysed in detail was a refinement of the preliminary configuration.

PFD 234996/OC/FS03 contained in Appendix E illustrates the major process units and process streams.

(i) Process Modelling

Mott MacDonald and Whitehopleman developed a steady state mass and energy balance of the process using Excel. Each individual process unit was modelled. The mass flows into and out of each process unit were calculated based on simple performance equations derived from typical industry data or parameters. The energy flows were calculated using the basic thermodynamic properties of the components. Where necessary recycle loops were solved using an iterative approach.

The outputs of this exercise were the compositions of the individual streams contained in Appendix E.

(ii) General Description

PFD 234996/OC/FS03 in Appendix E shows the arrangement selected for the basis of this particular study. There are three key features that distinguish the process from a conventional cement plant:

- The combustion air for the precalciner is separated prior to the precalciner in an air separation unit (ASU). Only the oxygen (95 mol% oxygen with 2 mol% nitrogen and 3 mol% argon) is supplied to the precalciner, the nitrogen and argon are vented to atmosphere.
- Part of the precalciner exhaust gases, around 50% of the total, is recirculated back to the precalciner burners to provide the part of the ballasting effect which is carried out by the atmospheric nitrogen in a conventional system.
- The net flue gas from the precalciner is cooled and passed to the CO_2 compression and inerts separation plant. The net product is compressed, dried, the inerts separated and the pure CO_2 is further compressed to pipeline pressure of 110 bara.

A more detailed process description is described below.

(iii) Raw Material Preparation

The raw material preparation is identical to the post-combustion capture cement plant described in section 4.4.5.

(iv) Raw Mill

The premixed raw materials are extracted to a premix bin which feeds the vertical roller mill for raw grinding. Additional corrective raw material feed bins are provided for sand and iron oxide. These materials are imported to primary stockpiles and loaded into the raw mill feed bins via apron feeders and conveyor belts, as required by the process. The premix and corrective raw materials are extracted from their feed bins at a measured rate and conveyed to the raw mill in the proportions required to meet the kiln feed chemistry targets.

The raw mill is a vertical roller mill. Feed is delivered onto the centre of the rotating table of the mill and is drawn under the rollers of the mill which are pressed onto the material bed on the table by hydraulic cylinders. Ground raw material falls over the dam ring round the edge of the table and is lifted by the hot gases sweeping the mill through the mill body to the integral separator. Coarse material is rejected by the separator and falls back to the mill table for regrinding while fine material is drawn out of the separator and mill with the hot gases by the mill induced draft fan.

A combination of gases from elsewhere in the process are used to sweep the raw mill and dry the raw materials from their initial moisture content of 8% (w/w) to 1% (w/w). The gases comprise cooled ‘tertiary air’ (~411ºC), excess exhaust gas from the kiln preheater (~330ºC) and exhaust air from the clinker cooler (~270ºC). The dried raw mix and sweeping air exit the raw mill at 100ºC. The temperature of the hot gases entering the mill is tempered by recirculation of mill exhaust air in order to protect the bearings of the rollers and separator.

The raw milling system has a rated capacity of 240 tph in order to provide 33% overtaking capacity above the requirements of the kiln for raw mix kiln feed. This enables planned stops of the raw milling system for maintenance to take place for up to 6 hours per day. During these periods when the raw milling system is stopped the gases that are used for drying are diverted to an Organic Rankine Cycle (ORC) heat recovery system.

The fine, raw mix product is collected from the hot gases after the raw mill induced draft fan by collection cyclones and a bag filter for dedusting. The de-dusted gases are drawn through a fan and exhausted to atmosphere through the main stack. The collected raw mix product is transported mechanically to the combined raw mix homogenisation and storage silo. This homogenisation silo has a capacity of 10,000 tonnes in order to provide over 2 days buffer storage of ground raw mix kiln feed. Homogenisation of the raw mix in the silo is achieved by sequential activation of air slides in the segmented base of the silo by compressed air and extraction of material from a pressure relieved central chamber with a conical roof.

Power consumption of the raw milling and homogenisation system totals 25 kWh/t of clinker produced from the kiln.

(v) Kiln Preheater

Kiln feed raw mix is extracted from the homogenisation silo on airslides and delivered to the kiln preheater feed bin. From the kiln preheater feed bin the raw mix feed is extracted at a controlled rate (as specified by the operator) into an elevator which transports the meal to the top of the preheater where it is delivered into the gas riser duct between stage 2 and 1 cyclones (where the preheater cyclones are numbered in the order in which the material flows through them, i.e. stage 1 is the top stage and stage 4 is the bottom stage of the preheater). The kiln feed raw mix enters the kiln preheater at approximately 18.5 kg/s and 1% (w/w) moisture content when the kiln is operating at its design capacity.

The kiln preheater tower includes 4 preheating and collection stages with heat exchange taking place in the gas risers between cyclones and collection of the preheated feed taking place in the cyclones. Exhaust gases from the rotary kiln are drawn through the kiln preheater by the kiln preheater induced draft fan. Successive heat exchange between the rising exhaust gas and the feed in each successive stage of the preheater lifts the temperature of the kiln feed raw mix from 100ºC to around 870ºC at the discharge from the lower, fourth stage cyclone into the precalciner.

The temperature of the exhaust gases exiting the rotary section of the kiln into the kiln preheater is at 1100ºC. Some of the partially preheated feed from the third stage cyclone of the kiln preheater therefore has to be delivered into the gas riser from the kiln to the lower, fourth stage of the kiln preheater tower in order to cool these exhaust gases. In this way melt formation and clogging in the kiln preheater tower is avoided. The feed used to cool the kiln exhaust gases in the gas riser from the kiln is therefore calcined in the kiln preheater tower and 32% of the calcination of the total feed takes place in the kiln and kiln preheater tower. The CO_2 derived from calcination of this portion of the total feed to the kiln is therefore not captured, in addition to the CO_2 derived from the combustion of the fuel in the main burner of the kiln.

The temperature of the exhaust gases passing through the kiln preheater fan is 330ºC and these are ducted to: (i) the fuel grinding systems to provide inert gases for drying of the coal and petroleum coke in the fuel grinding systems; and (ii) the raw mill to provide additional drying capacity.

(vi) Precalciner Preheater

Kiln feed raw mix is extracted from the homogenisation silo on airslides and delivered to the precalciner preheater feed bin. From the precalciner preheater feed bin the raw mix feed is extracted at a controlled rate (as specified by the operator) into an elevator which transports the meal to top of the preheater where it is delivered into the gas riser duct between stage 2 and 1 cyclones (where the preheater cyclones are numbered in the order in which the material flows through them, i.e. stage 1 is the top stage and stage 5 is the bottom stage of the preheater). The kiln feed raw mix enters the precalciner preheater at approximately 33 kg/s and 1% (w/w) moisture content.

The precalciner preheater tower includes 5 preheating and collection stages with heat exchange taking place in the gas risers between cyclones and collection of the preheated feed taking place in the cyclones. Exhaust gases from the precalciner are drawn through the precalciner preheater by the precalciner preheater induced draft fan. Successive heat exchange between the rising exhaust gas and the feed in each successive stage of the preheater lifts the temperature of the kiln feed raw mix from 100ºC to around 870ºC at the discharge from the fourth stage cyclone into the precalciner.

This preheated feed from the fourth stage cyclone is joined by the feed from the lower, fourth stage cyclone of the kiln preheater in the precalciner. The total calcined feed exiting the precalciner is then separated from the precalciner exhaust gases in the lower, fifth stage of the precalciner preheater and delivered to the inlet hearth of the rotary kiln.

The temperature of the exhaust gases passing through the precalciner preheater fan is 330ºC and these are split into a recycle stream (approximately 50% of total flow) and a stream for CO_2 capture (approximately 50% of total flow). The stream for CO_2 capture is passed through a gas-to-gas heat exchanger which heats up the oxygen generated from the ASU and then to an electrostatic precipitator (ESP) for dust collection prior to further purification.

(vii) Precalciner

The bulk of the calcination process is carried out in the precalciner. Dried pulverised coal is used as the fuel and is delivered at the required rate (2.4 kg/s at 1% (w/w) moisture) to ensure constant temperature operation of the precalciner at 900ºC. This amounts to 65% of the total calorific input to the kiln system and is in line with the normal practice in the operation of precalciner cement kilns. For this particular configuration it is advantageous to maximise the proportion of the total fuel fired in the precalciner for CO_2 capture purposes as this increases the amount of CO_2 generated from both the combustion of fuel and the calcination of the feed which can be subsequently captured. However, the ability to raise the proportion of fuel fired in the precalciner is limited by the requirement to fire a safe minimum proportion of the fuel on the main burner of the kiln to ensure stable operation. It is estimated that 68% of the calcination takes place in the precalciner and precalciner preheater tower.

In a conventional in-line precalciner the oxygen for the combustion of the precalciner fuel is part drawn from the excess air in the hot kiln exhaust gases rising from the rotary kiln and part from the hot tertiary air drawn from the hot end of the clinker cooler which have bypassed the kiln via the tertiary air duct. In this case the oxygen for the combustion of the precalciner fuel is provided by oxygen from the air separation unit (ASU). The oxygen stream has had its temperature raised by heat exchange with the stream diverted for CO_2 capture. This heated oxygen stream is mixed with the recycle stream which has a high concentration of CO_2 before the combined stream is heated by the 'tertiary' air from the clinker cooler in a gas-to-gas heat exchanger. The cooled tertiary air then passes to the raw mill for drying.

The quantity of recycle gas is set to ensure:

- complete combustion
- that reducing conditions do not occur within the precalciner (an excess of oxygen of 1.2% w/w at exit of precalciner is provided)
- the weight of ballast gas is maintained at the amount required for conventional combustion.

(viii) Rotary Kiln

The precalcined material from the precalciner is collected in the lower stage cyclone of the precalciner preheater tower and enters the kiln via the kiln inlet hearth. From this point forward the rotary kiln is operating in the conventional manner of a precalciner cement kiln. A burner is situated both centrally and co-axially with the kiln at the front end of the kiln. Petroleum coke is fired into the front end of the kiln whilst monitoring the kiln burning conditions. The hot clinker leaving the kiln falls into the clinker cooler. The kiln rotates at between 3 and 4 revolutions per minute driven by the main drive of the kiln.

(ix) Fuel Preparation

The raw coal preparation is identical to the post-combustion capture cement plant described in section 4.4.5 except that: (i) the vertical roller mill is swept by exhaust gases from the kiln preheater (which have a temperature of approximately 330ºC); and (ii) the metered coal from the precalciner fine coal bin is delivered pneumatically to the precalciner coal burners using precalciner recycle gas as the transport gas.

The raw petroleum coke preparation is identical to the post-combustion capture cement plant described in section 4.4.5 except that the vertical roller mill is swept by exhaust gases from the kiln preheater (which have a temperature of approximately 330ºC).

(x) Clinker Cooler

The clinker cooler is identical to the post-combustion capture cement plant described in section 4.4.5 except that the hot tertiary air at 795ºC is drawn into the tertiary air duct from above the cooler horizontal grate and used to: (i) heat up the recycled CO_2 and oxygen for combustion of the coal in the precalciner; and (ii) to dry the raw materials in the raw mill. The lower temperature excess air resulting from the final clinker cooling is also passed to the raw mill for drying.

(xi) Cement Milling

The cement milling is identical to the post-combustion capture cement plant described in section 4.4.5.

(xii) Packing and Loading

The cement packing and loading is identical to the post-combustion capture cement plant described in section 4.4.5.

(xiii) Air Separation Unit (ASU)

The oxygen flow required in the precalciner (5.51 kg/s) is supplied at 95% mol purity from the Air Separation Unit (ASU).

Oxygen will be produced in an ASU with a maximum capacity of 600 t/d. This is well within the range of installed plant output. The process is well known and produces gaseous oxygen by boiling liquid oxygen and is ideally suited to this application as the delivery pressure is low. There is no requirement for either pumping the liquid oxygen or compressing the gaseous product.

A low purity cycle is used which produces 95% mol oxygen purity. Studies (e.g. IEA GHG, 2005) have been carried out to show that for oxy-combustion power plants this is the optimum purity. This is principally due to two factors:

- There will always be some air in-leakage and there will always be some inerts that must be removed in the CO_2 purification plant.
- The increase in power required for the ASU to produce 99.5% mol purity oxygen is greater than the increase in CO_2 compression power required to remove inerts introduced due to lower purity oxygen.

These factors also apply for a cement plant.

The oxygen is delivered from the plant at around 16ºC. This is heated up to 310ºC in a gas-to-gas heat exchanger with the stream for CO_2 purification and compression which requires cooling.

The separated air stream containing predominantly nitrogen and argon is vented to atmosphere. It should be noted that there are potential revenues available from using this stream to produce nitrogen and argon for sale. Further uses of this gas stream have not been considered as part of this study.

(xiv) Gas Mixer

A proportion (approximately 50%) of the precalciner preheater exhaust gases is drawn by fan to a gas mixer which mixes this stream (at 330ºC) with the heated oxygen (at 310ºC) to form the precalciner recycle (at 325 ºC). The combined stream is mostly CO_2 (63.7% w/w) and oxygen (23.1% w/w). The mixing is undertaken by introducing turbulence in a length of pipe by the use of orifices or baffles.

(xv) Gas-to-Gas Heat Exchanger (Recycle/'Tertiary Air')

The recycle stream (at 325ºC) passes to a gas-to-gas heat exchanger where the hot 'tertiary' air (at 795ºC) is used to heat up the recycle stream to temperatures approaching the operation temperature of the precalciner. The cooled tertiary air is used for raw mill drying.

It is recognised that the operation of such a high temperature gas-to-gas heat exchanger with a high dust load is a significant technical challenge. However, during the course of their investigations for this study, Mott MacDonald became aware that there is some experience within the cement industry of operating such high temperature and high dust load gas-to-gas heat exchangers and that currently an inlet temperature of around 800ºC is considered to be the acceptable maximum.

(xvi) ESP

The proportion of precalciner preheater exhaust gases that have been sent for purification and CO_2 capture exit the gas-to-gas heat exchanger at around 248ºC. This gas contains dust from the preheater which must be removed prior to further purification steps. An electrostatic precipitator (ESP) is used for this purpose rather than a bag filter as the literature (e.g. EC, 2001) suggests that they operate better at higher temperatures compared to bag filters.

If the CO_2 is not to be captured (e.g. during start-up) then the exit stream from the ESP can be vented directly to atmosphere.

(xvii) Product Gas Purification

The dedusted gas is cooled further, dried and then passed through a compression and inerts (nitrogen and argon) and oxygen removal stage that delivers a final CO_2 product (16.9 kg/s) of 95% (mol) at 110 bara.

The CO_2-rich flue gas leaves the ESP at approximately 248ºC. The first part of the CO_2 treatment system cools the flue gas, thus removing the moisture by condensation, and compresses it to 30 bara.

A venturi mixer is used to quench the gas with water to a temperature where a conventional indirect seawater contact cooler can be used. A two column system is used. The first column cools the flue gas to 35ºC by direct contact with condensate that has been cooled against seawater in titanium plate-frame heat exchangers. The exit stream from this column is then further cooled against seawater to a temperature of 13ºC in a second direct contact cooler.

The net flue gas is now around 79% (mol) CO_2 and at atmospheric pressure. It is now compressed to 30 bara for further drying before purification. Compression to 30 bara is carried out in two stages. The first compressor compresses the CO_2 adiabatically to 15 bara. Cooling water is used to further cool the raw CO_2 to 20ºC before compression to 30 bara with intercooling using cooling water.

The raw CO_2 gas passes through a temperature swing dual bed desiccant dryer.

The inerts removal process uses the principle of phase separation between condensed liquid CO_2 and insoluble inerts gas. The actual CO_2 pressure levels used for the separation are fixed by the specification of > 95% (mol) CO_2 product purity and the need to reduce the CO_2 vented with the inerts to an economic minimum.

It has been assumed that any excess O_2, SO_2 or NO_x present in the CO_2 need not be removed. For specific applications requiring higher purity CO_2 (e.g. > 98%) the process would need some modification.

It should be noted that the condensed water obtained from the indirect seawater cooler and temperature swing bed desiccant dryer will contain dissolved acid gas components. Recent work by Air Products (White, 2007) indicates that most of the SO_x and NO_x will be removed as acids during the CO_2 compression. This acidic waste stream will require neutralisation prior to discharge.

(xviii) CO_2 Compression

The final compression stage takes the gas up to 110 bara for pipeline transmission.

(xix) Organic Rankine Cycle (ORC) unit

During periods when the main process is still operating but the raw mill is not running then the gases that are normally used for the drying (i.e. cooled 'tertiary' air, kiln preheater exhaust and clinker cooler exhaust gas) will be diverted to an Organic Rankine Cycle (ORC) heat recovery system. This unit will be used to generate electrical power which could be used on the site or exported to the grid.

The basic operating principle of the system is as follows:

- The heat contained in the waste gases is transferred to a closed loop circulating low pressure thermal oil.
- The thermal oil heat content is transferred to the energy converter motive fluid (typically a hydrocarbon such as pentane) in a vaporiser.
- The motive fluid is used to generate electrical power through a specially designed impulse type turbine.
- The motive fluid is condensed and returned to the vaporiser.

Mott MacDonald estimates that it would be possible to generate a maximum electrical power of 4.7 MW if all the hot gases were sent to the ORC unit.

(xx) Electrical Supply

The total installed power required by the plant is estimated at 31.3 MW (see Table 4-8). The operating power, operating hours and the power consumed of all the major process units are also shown in Table 4-8. It has been assumed that a suitable grid connection is available.

It should be noted that the ORC will be capable of generating a maximum electrical power of 4.67 MW when the hot gases are not being used for raw mill drying. It has been assumed that this will only occur for 15% of operational hours.

Table 4-8: Electrical Demand of Oxy-combustion Cement Plant with CO_2 capture

Process Area	**Main Equipment Items**	**Installed Power (MWe)**	**Operating Power (MWe)**	**Operating hours per year**	**Power consumed (MWh/y)**
Raw Material Preparation	Limestone crusher	0.60	0.51	2080	1061
	Shale crusher	0.20	0.17	2080	354
	Stacker-reclaimer system	0.28	0.24	2080	500
Raw Mill	Raw mill drive	1.20	1.02	5730	5845
	Raw mill fan	1.50	1.28	5730	7306
	Raw mill ancillaries	-	-	-	-
	Raw mill bag filter	-	-	-	-
	Bag filter fan	1.20	1.02	5730	5845
	Homogenising chamber silo	0.12	0.10	5730	560
Kiln preheater	Preheater (kiln) feed bin	-	-	-	-
	4 stage preheater	-	-	-	-
	Preheater (kiln) fan	0.35	0.30	7920	2356
Precalciner preheater	Preheater (precalciner) feed bin	-	-	-	-
	5 stage preheater	-	-	-	-
	Preheater (precalciner) fan	0.55	0.47	7920	3703
Precalciner	Precalciner	-	-	-	-
Rotary Kiln	Kiln	0.65	0.55	7920	4349
Clinker Cooler	Grate cooler	1.00	0.85	7920	6732
	Cooler bag filter				
	Cooler bag filter				
	Clinker crusher				
	Clinker conveyor				
	Clinker storage silo				
	Export clinker hopper				
Coal preparation	Coal milling system	0.30	0.26	6095	1554
Petroleum coke preparation	Petroleum coke mill	0.30	0.26	6095	1554
Cement milling	Mill 1	3.15	2.68	5580	14940
	Mill 2	3.15	2.68	5580	14940
Packing and loading	Silo 1	-	-	-	-
	Silo 2	-	-	-	-
	Bag packing plant	0.50	0.43	7920	3366
ORC heat recovery unit	Electrostatic precipitator	0.08	0.06	1188	76
	ORC heat recovery unit	-	-	-	-
	Exhaust gas fan	0.70	0.60	1188	707

Process Area	Main Equipment Items	Installed Power (MWe)	Operating Power (MWe)	Operating hours per year	Power consumed (MWh/y)
Air Separation Unit	Main air compressors Air purification system Main heat exchanger ASU compander ASU column system Pumps Gas-to-gas heat exchanger for oxygen/product Backup storage vessel	6.00	5.10	7920	40392
Recycle	Recycle stream fan	0.33	0.28	7920	2188
	Gas mixer	-	-	-	-
	Gas-to-gas heat exchanger for 'tertiary air'/recycle	-	-	-	-
Electrostatic precipitator	Electrostatic precipitator	0.08	0.06	7920	505
	Electrostatic precipitator fan	0.23	0.19	7920	1515
CO_2 compression and purification	Venturi scrubber Indirect contact cooler Direct contact seawater cooler Compressors Heat exchangers Flue gas expander	8.88	7.55	7920	59763
Total		**31.32**	**26.62**	**-**	**180109**

4.5.6 Plant Performance

The performance of the oxy-combustion cement plant with CO_2 capture is shown in Table 4-9. Note that for the purposes of this analysis the CO_2 emitted from the electricity delivered to site is not included in the CO_2 emitted figure.

The CO_2 capture rate of the plant is 52%.

Table 4-9: Performance of Oxy-combustion Cement Plant with CO_2 capture

	Parameter	Unit	Value
Inputs	Raw meal consumed	t/y	1,508,424
	Coal required	t/y	72,061
	Petroleum coke required	t/y	27,091
	Oxygen (at 95% mol) from ASU	t/y	166,298
	Electricity consumption	MWh/y	174,562
Outputs	Clinker produced	t/y	910,000
	Cement produced	t/y	1,000,000
	CO_2 captured	t/y	465,014
	CO_2 emitted	t/y	282,853

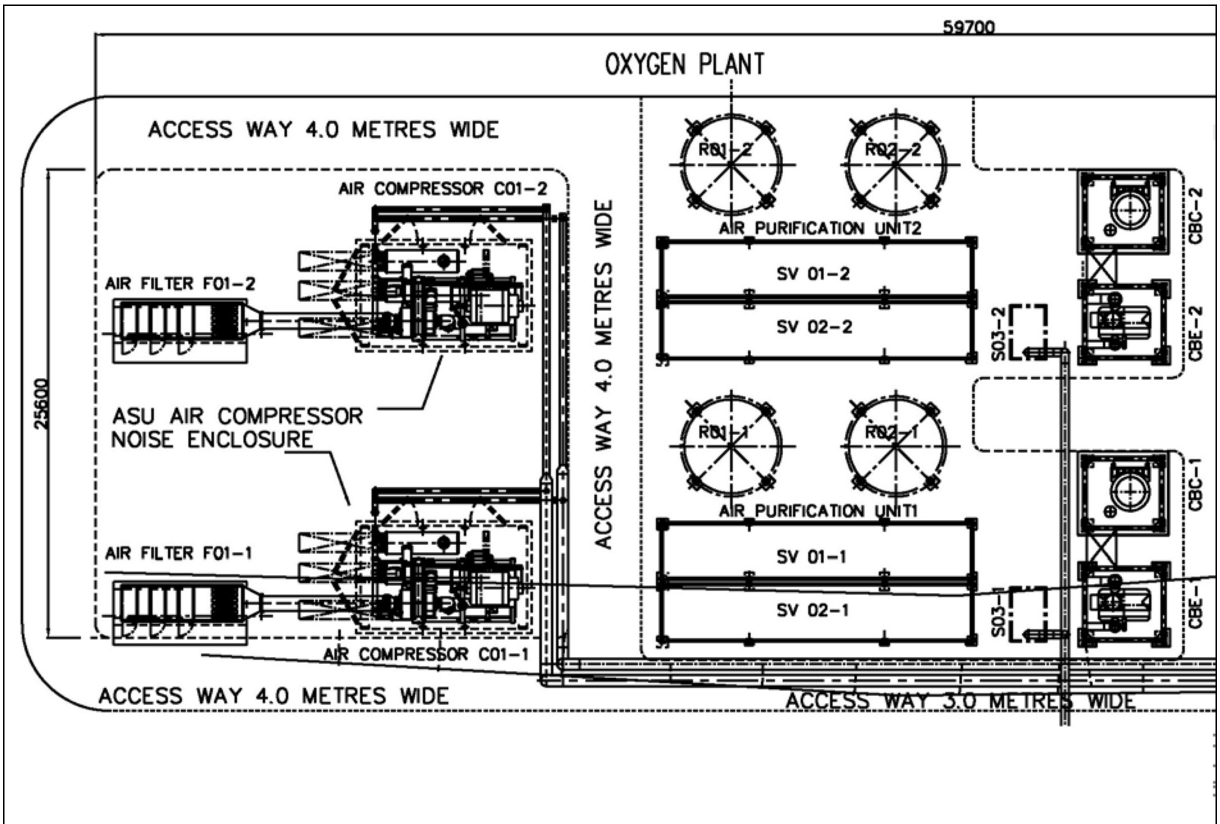
4.5.7 Plant Layout

The process units and ancillary buildings will need to be laid out to give the most economical flow of materials and personnel around the site. A detailed layout of an oxy-combustion cement plant is beyond the scope of this study but it is important to note that the following will need to be considered:

- Cement plants are typical laid out linearly.
- Hazardous materials (e.g. oxygen) should be stored at a safe distance from other buildings.
- The ancillary buildings should be located so as to minimise the time spent by personnel in travelling between buildings.
- Administration offices should be located well away from potentially hazardous processes.
- Utility buildings should be sited to give the most economical run of pipes to and from the process units.
- Equipment should be located so that it can be conveniently tied in with any future expansion of the site.

Mott MacDonald estimates that the footprint of an oxy-combustion cement plant would be similar to a cement plant without CO_2 capture. A typical 1 Mt/y cement plant without cement capture would occupy approximately 15 ha (500 m x 300 m) but could be accommodated in less space if necessary. Mott MacDonald estimates that the additional equipment required for an oxy-combustion cement plant (ASU, ORC, recycle and CO_2 compression equipment etc.) would require approximately 0.5 ha. The limestone quarry would occupy a much larger area, possibly as much as 200 ha (2000 m x 1000 m). For illustration purposes Figure 4-12 shows the typical layout for 2 x 330 tpd ASUs. The footprint of this plant is 25.6 m x 59.7 m.

Figure 4-12: Typical Layout for 2 x 330 tpd ASUs (Air Liquide)



LEGEND	
TAG NUMBER	EQUIPMENT
F01-1	- AIR FILTER ASU1
F01-2	- AIR FILTER ASU2
C01-1	- AIR COMPRESSOR ASU1
C01-2	- AIR COMPRESSOR ASU2
R01-1/02-1	- AIR PURIFICATION BOTTLE ASU1
R01-2/02-2	- AIR PURIFICATION BOTTLE ASU2
SV01-1	- APU VALVE SKID 1 ASU1
SV02-1	- APU VALVE SKID 2 ASU1
SV01-2	- APU VALVE SKID 1 ASU2
SV02-2	- APU VALVE SKID 2 ASU2
CBC-1	- COLD BOX COLUMN ASU1
CBC-2	- COLD BOX COLUMN ASU2
CBE-1	- COLD BOX EXCHNAGER ASU1
CBE-2	- COLD BOX EXCHNAGER ASU2
SO3-1	- O2 VALVE SKID ASU1
SO3-2	- O2 VALVE SKID ASU2
B80-1	- CRYOGENIC PURGE TANK ASU1
B80-2	- CRYOGENIC PURGE TANK ASU2

4.5.8 Operation Considerations

The following key operation considerations have been identified for the oxy-combustion CO$_2$ capture cement plant:

- The precalciner must be designed in such a way as to permit air-firing as a fall-back position should there be an interruption in supply from the ASU. Therefore, adequate backup for the ASU should be provided in order to allow a controlled change-over to air-firing. Backup should be in the form of liquid oxygen enough of which will be stored on site to allow controlled changeover to air-firing. A liquid oxygen capacity of 60 tonne capacity should be sufficient (3 hours of operation). It will be stored at 2.5 bara in a vacuum insulated storage tank which can be filled by gravity from the ASU. This tank would also be used during start-up of the ASU.

- As mentioned in section 4.5.5(xix), if the raw mill is not operating then the hot gases which are normally used for drying will be diverted to an Organic Rankine Cycle (ORC) unit for generating electrical power.

- As mentioned in section 4.5.5(xvi), if the CO_2 purification and compression plant is not available then it will be possible to discharge the purge stream following dust removal in the ESP. This route will also be used during start-up until a sufficiently concentrated CO_2 stream is achieved.

- For operational flexibility the provision to use tertiary air in the precalciner should be included. This will help during start-up.

- It has been noted within the cement industry that there are operational difficulties with using separate line calciners (SLCs). The principal issue relates to blocking that can occur if flow is stopped (e.g. during a power cut or a shut down). It is important that sufficient emergency generator power is available on site to ensure fans remain operational so that a controlled shut down can be undertaken. In addition, the design of the ductwork around the SLC must be such that the settling of calcined material is prevented.

4.5.9 Emissions and Environmental Considerations

An oxy-combustion capture cement plant will achieve emission standards well below current cement plants. This is an added advantage of CO_2 capture. Some emissions will occur and these are discussed in the following sections.

(i) Emission Sources

During normal operation there will be five main discharges to atmosphere. These are:

- exhaust gases from the raw mill
- exhaust gases from the coal dryer
- exhaust gases from the petroleum coke dryer
- waste gas from the ASU
- inerts and oxygen from the CO_2 purification plant.

The following discharges would also occur:

- purge stream gas during start-up and when CO_2 purification and compression plant is not available
- discharge gases from ORC unit when raw mill is not operating.

These gases could be discharged via a single stack or via individual stacks depending on the layout, cost and the environmental constraints imposed on the plant.

(ii) Dust

As discussed in section 0 the best available techniques for reducing dust emissions from point sources is the application of electrostatic precipitators (ESPs) and/or fabric filters. The BAT emission level[2] associated with these techniques is 20-30 mg dust/m^3 (EC, 2001). Discharges from the raw mill, the coal dryer and the petroleum coke dryer will pass through bag filters prior to discharge. Discharges from the ORC or the purge stream will pass through ESPs prior to discharge. The selection of ESPs over bag filters in these cases is due to the reported better performance of ESPs at higher temperatures. All other emissions will not contain dust and will not require treatment.

Through use of these abatement technologies it is expected that dust emissions from the plant will be within the BAT range i.e. 20-30 mg dust/m^3.

Any dust collected would be recycled back through the process.

(iii) NO_x Emissions

The level of NO_x present in the flue gases from oxy-combustion processes are considered to be lower than for conventionally operated processes due to the reduction in production of thermal NO_x. However, in the configuration under consideration, the kiln is operating in a conventional mode so the production of thermal NO_x and fuel NO_x from the kiln will be the same. Production of thermal NO_x in the precalciner will not be significant due to the reduced levels of nitrogen and the temperature of operation. Production of fuel NO_x will be generated as per a conventional process.

NO_x emissions will occur in:

- exhaust gases from the raw mill
- exhaust gases from the coal dryer
- exhaust gases from the petroleum coke dryer
- inerts and oxygen from the CO_2 purification plant
- purge stream gas during start-up and when CO_2 purification and compression plant is not available
- discharge gases from ORC unit when raw mill is not operating.

The BAT techniques for reducing NO_x emissions are a combination of general primary measures, primary measures to control NO_x emissions, staged combustion and selective non-catalytic reduction (SNCR). It is anticipated that a combination of these techniques would be used to ensure that the BAT emission level[3] of 200-500 mg NO_x/m^3 (as NO_2) was achieved from the discharges.

Recent work by Air Products (White, 2007) has indicated that most of the NO_x in the CO_2 stream will be removed as nitric acid during compression. Hence, the inerts and oxygen from the CO_2 purification plant should contain negligible levels of NO_x.

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[2] Emission levels are expressed on a daily average basis and standard conditions of 273 K, 101.3 kPa, 10% oxygen and dry gas.

[3] Emission levels are expressed on a daily average basis and standard conditions of 273 K, 101.3 kPa, 10% oxygen and dry gas.

The costs associated with any NO_x emissions reduction measures have not been included in the cost estimate for the oxy-combustion process option.

(iv) SO_x Emissions

As discussed in Section 0, modern cement plants which do not have sulphide minerals in their raw materials have very low emissions of SO_x and often achieve emission levels below 10 $mgSO_2/m^3$ without using abatement techniques (EC 2001). In this case it is assumed that the raw meal originates from NE Scotland and contains 0.33% (w/w – dry basis) sulphur.

It is expected that 30% or more of the sulphur in the raw material will be evaporated and be emitted from the first stage of the preheater and that SO_2 will be the main (99%) sulphur compound to be released (EC 2001). In this case the emissions will be emitted from both the kiln preheater and the precalciner preheater.

The gases from the kiln preheater are used for drying within the raw mill (when in operation) and for fuel drying. It is reported that between 20-70% of the SO_2 that is passed to the raw mill will be captured by the finely ground raw materials (EC 2001).

The gases from the precalciner preheater are either recycled back to the precalciner or are passed forward for CO_2 purification and compression. It is expected that SO_2 recycled back to the precalciner will be absorbed into the solid material. It is also expected that most of the SO_2 passed forward to the CO_2 purification and compression stages will be removed in the condensate during the contact cooling process prior to CO_2 compression. Recent work by Air Products (White, 2007) has indicated that most of the SO_2 in the CO_2 stream will be removed as sulphuric acid during compression. Hence, the inerts and oxygen from the CO_2 purification plant should contain negligible levels of SO_2.

It is expected that sulphur in the fuels fed to the precalciner or kiln will not lead to significant SO_2 emissions due to the strong alkaline nature in the sintering zone, the calcination zone and the lower stages of the preheaters. This sulphur will be captured in the clinker.

The BAT techniques for reducing SO_2 emissions are a combination of general primary measures and absorbent addition for initial emission levels not higher than about 1200 mg/m^3 of SO_2 and a wet or dry scrubber for initial emission levels higher than about 1200 mg/m^3 of SO_2. The BAT emission level[4] associated with these techniques is 200-400 mg/m^3 of SO_2.

The costs associated with any SO_x emissions reduction measures have not been included in the cost estimate for the oxy-combustion process option.

(v) Effluent

During normal operation there will be three main effluent discharges. These are:

- condensed water from the indirect seawater cooler
- seawater discharge from the direct seawater cooler

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[4] Emission levels are expressed on a daily average basis and standard conditions of 273 K, 101.3 kPa, 10% oxygen and dry gas.

- water obtained from temperature swing dual bed desiccant dryer.

The condensed water obtained from the indirect seawater cooler and temperature swing bed desiccant dryer will contain acidic components. Recent work by Air Products (White, 2007) indicates that most of the SO_x and NO_x will be removed as acids during the CO_2 compression. This acidic waste stream will require neutralisation prior to discharge. However, it may be possible to convert the sulphuric acid produced into gypsum by direct reaction with limestone in a stirred tank reactor. The gypsum could then be used in the cement production process. The effluent from this process would contain soluble calcium nitrate. Mott MacDonald has not investigated the feasibility of this option further.

The seawater obtained from direct seawater cooler will also contain some dissolved acid gas components and may require further treatment prior to discharge. It should be noted that previous studies examining oxy-combustion at power plants (IEA GHG, 2005) have shown this stream being discharged directly back to the sea.

The costs associated with any effluent treatment have not been included in the cost estimate for the oxy-combustion process option.

(vi) Others

As discussed in Section 2.4.2 in addition to dust, NO_x and SO_x there are a number of pollutants considered to be of concern for the production of cement. These include:

- Carbon monoxide
- Volatile organic compounds (VOCs)
- Polychlorinated dibenzodioxins and dibenzofurans (PCDDs and PCDFs)
- Metals and their compounds
- HF
- HCl.

The fate of these pollutants within the oxy-combustion process has not been examined in detail within this study. However, it is important to note that the recycling of exit gases from the precalciner preheater will result in material with a high calcium oxide content being in contact with gases containing acid species such as HF and HCl. It is therefore expected that these species will be absorbed within the clinker and will result in a reduction of emissions as compared to a standard cement plant.

4.5.10 Design Risk Assessment

It is of paramount importance that an oxy-combustion cement plant is implemented safely and without significant impact upon operational flexibility. Mott MacDonald have undertaken a preliminary design risk assessment to identify the major additional plant operation risks that need to be considered beyond those that apply for a cement plant without carbon capture. The results of this exercise are shown in Table 4-10. It is important to note that this exercise would have to be repeated in greater detail if a detailed design of an oxy-combustion cement plant was undertaken.

Table 4-10: Preliminary Design Risk Assessment for Oxy-Combustion CO_2 Capture Cement Plant

Hazard Ref	Hazard	Stage of work	Risk Control Measures	Owner	Comment
1	Risk of operator suffocation as CO_2 is an asphyxiant	Operation	1. Standards, codes of practice etc. to be adopted during design. 2. Appropriate material selection during design. 3. HAZOP study to be undertaken.	Contractor	Hazards associated with handling CO_2 are well understood from other industries.
2	Risk of fire or explosion as O_2 may ignite on contact with combustible materials.	Operation	1. Standards, codes of practice etc. to be adopted during design. 2. Appropriate material selection during design. 3. HAZOP study to be undertaken. 4. ATEX assessment to be performed.	Contractor	Hazards associated with handling O_2 are well understood from other industries.
3	Risk of dust explosion in fuel milling and drying stages.	Operation	1. Standards, codes of practice etc. to be adopted during design. 2. HAZOP study to be undertaken. 3. ATEX assessment to be performed.	Contractor	Hazards associated with explosive dusts are well understood within cement industry.
4	Risk of corrosion damage due to presence of acidic components (e.g. SO_2, SO_3 and HCl) in gas streams.	Operation	1. Standards, codes of practice etc. to be adopted during design. 2. Appropriate material selection during design.	Contractor	Hazards associated with acidic components in gas streams are well understood from other industries.
5	Risk of operator exposure to extreme temperatures as	Operation	1. Standards, codes of practice etc. to be adopted during design.	Contractor	Hazards associated with low temperatures

Hazard Ref	Hazard	Stage of work	Risk Control Measures	Owner	Comment
	very low temperatures present in cryogenic air separation unit.		2. Appropriate material selection during design. 3. HAZOP study to be undertaken.		are well understood from other industries.
6	Risk of operator suffocation as N_2 is an asphyxiant	Operation	1. Standards, codes of practice etc. to be adopted during design.	Contractor	Hazards associated with handling N_2 are well understood from other industries.
7	Risk of fire as pentane used in ORC which is flammable.	Operation	1. Standards, codes of practice etc. to be adopted during design. 2. Appropriate material selection during design. 3. HAZOP study to be undertaken	Contractor	Hazards associated with handling hydrocarbons are well understood from other industries.
8	Risk of damage caused by release of high pressure as high pressures present in CO_2 compression unit.	Operation	1. Standards, codes of practice etc. to be adopted during design. 2. HAZOP study to be undertaken	Contractor	Hazards associated with handling high pressures are well understood from other industries.

5 Economic Analysis of CO_2 Capture at a New-Build Cement Plant

5.1 Introduction

This section of the report describes the assessment of the capital and operating costs of adding CO_2 capture to a new-build cement plant. Both post-combustion and oxy-combustion capture options are reported.

Various sensitivities have also been investigated.

5.2 Assumptions

The main financial assumptions used for this study were based on the IEA GHG R&D programme Technical & Financial Assessment Criteria (IEA GHG, 2003). These have been developed for the technical and financial assessment of power plants with CO_2 capture. Discussions were held with the BCA to modify some of the assumptions to be relevant to the cement industry.

The key points are summarised below:

Location

- A green field site with no special civil works implications was assumed.
- Adequate plant and facilities to make the plant self-sufficient in site services were included in the investment costs.

Currency

- The results of the studies are expressed in Euro € applicable to a specific year.
- An exchange rate of 1 Euro = 0.694 UK £ has been applied where necessary.
- An exchange rate of 1 Euro = 1.389 US $ has been applied where necessary.

Design and Construction

- The ‘S’ curve of expenditure during construction of the cement plant was as follows:

Year 1	20% investment
Year 2	45% investment
Year 3	35% investment

Plant Life

- A design life of twenty five years is assumed as the basis for the economic appraisal.

Load Factor

- The load factor, defined as the achieved output as a percentage of rated nameplate capacity, for the cement plant is assumed as follows:

Year 1	60% of rated capacity
Subsequent years	90% of nameplate capacity

Cost of Debt

- All capital requirements are treated as debt at the same discount rate used to derive capital charges.
- There is no allowance for grants, cheap loans etc.
- Specific capital cost figures are presented without including an allowance for funds used during contingencies.

Capital charges

- Discounted cash flow calculations are expressed at a discount rate of 10%.
- Inflation assumptions have not been made. No allowance is made for escalation of fuel, labour or other costs relative to each other.

Contingencies

- A contingency has been added to the capital cost to give a 50% probability of a cost over-run or under-run. The value of the contingency is taken as 5% of the installed cement plant cost and 10% of the installed CO_2 capture technology plant cost.
- All plant is assumed to be built on a turnkey basis i.e. the cost of risk is built into the contractor's fees.

Fees

- The contractor's fees for design and build form part of the basic plant cost estimate.
- 5% of the installed plant cost (excluding contingency) has been included to cover owners costs such as land purchase, surveys and general site preparation etc.
- 2% of the installed plant cost (excluding contingency has been included to cover fees such as process/patent fees, fees for agents or consultants, and legal and planning costs.

Commissioning and Working Capital

- A 3 month commissioning period has been assumed for all plant.
- Sufficient storage for 7 days operation at rated capacity will be allowed for limestone.

- Sufficient storage for 15 days operation at rated capacity will be allowed for coal, petroleum coke, other raw materials, products and consumables.

- Allowance is made for receipts from product sales during the commissioning period.

Decommissioning

- The remediation cost of the cement plant quarries was not included in this analysis.

Taxation and Insurance

- 1% per year of the installed plant cost (excluding contingency and fees) has been allowed to cover specific services e.g. local rates.

- Taxation on profits has not been included in the assessments.

- An allowance of 1% per year of the installed plant cost (excluding contingency and fees) has been included to cover insurance.

Maintenance

- Routine and breakdown maintenance has been allowed for at 4% per year of installed plant cost (excluding contingency and fees) for solids handling plant and at 2% per year for plants handling gases and liquids and services plant.

- The cost of maintenance labour is assumed to be included within the labour costs.

Labour

- Operating labour only has been identified and assumed to work in a 5 shift pattern.

- The annual salary of an operator has been assumed to be €40,000/yr.

- An allowance of 20% of the operating labour direct costs has been included to cover supervision.

- A further 30% of direct labour costs has been included to cover administration and general overheads.

Fuels and raw materials

- The cost of limestone delivered to the site is assumed to be €3/tonne.

- The cost of shale delivered to the site is assumed to be €1.5/tonne.

- The cost of sand delivered to the site is assumed to be €50/tonne.

- The cost of iron oxide delivered to the site is assumed to be €50/tonne.

- The cost of gypsum delivered to the site is assumed to be €10/tonne.

- The cost of coal delivered to the site is assumed to be €65/t (2.51 €/GJ (LHV basis)).

- The cost of petroleum coke delivered to the site is assumed to be €80/t (2.34 €/GJ (LHV basis).
- The cost of power delivered to the site is assumed to be €0.0500/kWh. This figure represents an annual average wholesale price without carbon for the UK (i.e. the price does not include EU ETS compliance costs).
- The revenue for power generated on-site and exported to the grid is assumed to be €0.0500/kWh. This figure represents an annual average wholesale price without carbon for the UK (i.e. the price does not include EU ETS compliance costs).
- The cost of water and seawater delivered to the site is assumed to be €0.1/t.
- The cost of cooling water is assumed to be €0.02/t.
- The cost of ammonia is assumed to be €200/t.
- The cost of MEA is assumed to be €1100/t.
- The cost of additive inhibitor is assumed to be 20% of price of MEA make-up.
- It is assumed that the SCR catalyst must be changed every 3 years.

5.3 Base Case: Cement Plant with no CO_2 Capture

5.3.1 Base Case

For the purposes of calculating the additional costs for installing CO_2 capture at a cement plant the capital and operating costs of a base case cement plant with no CO_2 capture were estimated. The flowsheet for the base case (234996-BC-FS01) and the stream data are contained in Appendix C.

5.3.2 Capital Costs

A cement plant equipment supplier provided a budget cost for a 1 million tonnes per year cement plant in Northern Europe of €120 million for the mechanical and electrical equipment. For the purposes of this study this total budget has been allocated to the different process sections of the cement plant as set out in Table 5-1. The estimated accuracy of the figures quoted is ± 25%.

Table 5-1: Capital Cost Allocation for Cement Plant with no CO_2 Capture (Base Case)

Cement Plant Process Section	**CAPEX budget €M**
Raw material crushing and blending	
Limestone crushing	2.50
Shale crushing	1.25
Stacker-reclaimer	2.08
Raw milling and homogenisation	
Raw mill drive	16.50
Raw mill fan	0.50
Raw mill ancillaries	1.90
Raw mill bag filter	2.00
Bag filter fan	0.50
Conditioning Tower	0.50
Homogenisation chamber silo	5.28
Preheater	
Preheater feed bin	Included below
Five stage preheater	5.40
Preheater fan	0.80
Precalciner	
Precalciner	0.50
Rotary Kiln	
Kiln	12.00
Clinker Cooler	
Grate cooler	12.00
Cooler bag filter	2.00
Cooler bag filter fan	0.50
Clinker crusher	Included below
Clinker conveyor	1.50
Clinker storage silo	9.00
Export clinker hopper	0.30
Coal preparation	
Coal milling system	5.00
Petroleum coke preparation	
Petroleum coke milling system	5.00
Cement milling	
Mill 1	10.00
Mill 2	10.00
Cement packing and loading	
Compartmental silo 1	5.00
Compartmental silo 2	5.00
Bag packing plant	3.00
Total	**120**

The following must be added to the total mechanical and electrical cost:

- construction costs including mechanical erection, instrument and electrical installation, civil works, buildings and site preparation

- design and engineering costs
- other costs including temporary facilities, training, commissioning, start-up costs and spare parts
- EPC services including contractors home services and construction supervision and freight
- contingency
- fees
- owners costs.

The breakdown of these costs for the base case plant are shown in Table 5-2.

Table 5-2: Capital Costs for Cement Plant with no CO_2 Capture (Base Case)

Description	€M
Mechanical and Electrical Equipment Cost	120
Design and Engineering	42
Construction	48
Other Costs	8
EPC Services	17
Installed Costs	**235**
Contingency	12
Fees	5
Owners Costs	12
Total Investment Cost	**263**

5.3.3 Operating Costs

The operating costs are formed from variable costs and fixed costs.

The variable cost element in the total operating cost is derived from the consumables associated with the plant. The consumables are listed in Table 5-3 and are based on 330 working days per year.

Table 5-3: Variable Operating Costs for Cement Plant with no CO_2 Capture (Base Case)

Description	Unit Cost (€/unit)	Consumption	€M/year
Limestone	3/t	1,245,973 t/y	3.74
Shale	1.5/t	283,974 t/y	0.43
Sand	50/t	7,473 t/y	0.37
Iron Oxide	50/t	7,473 t/y	0.37
Gypsum	10/t	40,000 t/y	0.40
Coal	65/t	0.0633 Mt/y	4.11
Petroleum Coke	80/t	0.0319 Mt/y	2.55
Miscellaneous Materials			0.72
Power	0.05/kWh	80,809 MWh/y	4.04
Process Water	0.1/t	240,000 t/y	0.02
Total variable operating cost			**16.76**

The fixed cost contribution to the operating costs is shown in Table 5-4. These costs comprise the items listed below:

- maintenance
- operating labour
- supervision
- administration and general overheads
- local rates
- insurance.

Table 5-4: Fixed Operating Costs for Cement Plant with no CO_2 Capture (Base Case)

Description	€M/year
Maintenance	9.4
Operating labour	3.2
Supervision	0.6
Administration and general overheads	1.2
Local rates	2.4
Insurance	2.4
Total fixed operating costs	**19.1**

5.4 Economic Analysis of Post-Combustion CO_2 Capture

5.4.1 Capital Costs

The capital cost allocation of the post-combustion cement plant described in section 4.4.5 is set out as in Table 5-5. The estimated accuracy of the figures quoted is ± 25%. These figures were derived from budgetary prices provided by suppliers, costs scaled from published information and prices obtained from Mott MacDonald's database of equipment prices.

The CHP capital costs assume that there is currently a grid connection and that no reinforcement of the grid or additional capital costs are required.

Potential cost reductions could occur if the CHP was replaced with an auxiliary boiler to provide the low pressure steam. This would remove the need for the steam turbine, reduce the coal requirements and reduce the balance of plant costs. However, the operating costs would increase because the cement plant will not be a net exporter of electricity (as it would be if the CHP was operating).

Table 5-5: Capital Cost Allocation for Post-Combustion Cement Plant with CO_2 Capture

Cement Plant Process Section	CAPEX budget €M	Total €M
Raw material preparation		
Limestone crusher	2.50	
Shale crusher	1.25	
Stacker-reclaimer system	2.08	
Raw milling and homogenisation		
Raw mill drive	16.50	
Raw mill fan	0.50	
Raw mill ancillaries	1.90	
Raw mill bag filter	2.00	
Bag filter fan	0.50	
Conditioning Tower	0.50	
Homogenisation chamber silo	5.28	
Preheater		
Preheater feed bin	Included below	
Five stage preheater	5.40	
Preheater fan	0.80	
Precalciner		
Precalciner	0.50	
Rotary Kiln		
Kiln	12.00	
Clinker Cooler		
Grate cooler	12.00	
Cooler bag filter	2.00	
Cooler bag filter fan	0.50	
Clinker crusher	Included below	
Clinker conveyor	1.50	
Clinker storage silo	9.00	
Export clinker hopper	0.30	
Coal preparation		
Coal milling system	5.00	
Petroleum coke preparation		
Petroleum coke milling system	5.00	
Cement milling		
Mill 1	10.00	
Mill 2	10.00	
Cement packing and loading		
Compartmental silo 1	5.00	
Compartmental silo 2	5.00	
Bag packing plant	3.00	
Total mechanical and electrical cost of cement plant		**120.0**
Selective Catalytic Reduction (SCR)	4.6	
Reactor Casing		
Bypass System		
Catalyst		

Cement Plant Process Section	**CAPEX budget €M**	**Total €M**
Ammonia Injection System		
Handling Equipment		
Control System		
Flue Gas Desulphurisation	22.5	
Spray Tower Absorption		
Reaction Tank		
Mist Eliminators		
Limestone Storage		
Limestone Preparation		
Dewatering System		
Other		
Gas Mixer	0.1	
CO_2 Capture	31.8	
Pumps (MEA, reflux, stripper etc)		
Amine Filter Package		
Water Tankage		
Cross Exchangers		
Stripper		
Coolers (DCC and water wash)		
DCC Towers		
Overhead Stripper Condenser		
Stripper Reboiler		
Lean Solvent Cooler		
Packing		
Absorption Tower		
Flash Drums		
CO_2 compression and purification	7.8	
Dryer		
Compressors		
CO_2 Pumps		
CHP Plant	66.2	
Coal and Ash Handling		
Boiler Island		
Steam Turbine Plant and Generator		
Selective Catalytic Reduction (SCR)		
Balance of Plant		
Total mechanical and electrical cost of additional post-combustion plant		**132.9**
Total mechanical and electrical cost of post-combustion cement plant		**252.9**

The following must be added to the total mechanical and electrical cost:

- construction costs including mechanical erection, instrument and electrical installation, civil works, buildings and site preparation
- design and engineering costs

- other costs including temporary facilities, training, commissioning, start-up costs and spare parts
- EPC services including contractors home services and construction supervision and freight
- contingency
- fees
- owners costs.

The breakdown of these costs for the post-combustion cement plant are shown in Table 5-6.

Table 5-6: Capital Costs for Post-Combustion Cement Plant with CO_2 Capture

Description	Cement Plant €M	Post-Combustion Plant €M	Total €M
Mechanical and Electrical Equipment Cost	120	133	253
Design and Engineering	42	29	71
Construction	48	57	105
Other Costs	8	24	32
EPC Services	17	14	31
Installed Costs	**235**	**256**	**492**
Contingency	12	20	32
Fees	5	5	10
Owners Costs	12	13	25
Total Investment Cost	**263**	**294**	**558**

5.4.2 Operating Costs

The variable cost element in the total operating cost is derived from the consumables associated with the plant. The consumables are listed in Table 5-7 and are based on 330 working days per year.

Table 5-7: Variable Operating Costs for Post-Combustion Cement Plant with CO_2 Capture

Description	Unit Cost (€/unit)	Consumption	€M/year
Limestone for raw meal and blending	3/t	1,245,679 t/y	3.74
Limestone for FGD	3/t	12,830 t/y	0.04
Shale	1.5/t	283,974 t/y	0.43
Sand	50/t	7,473 t/y	0.37
Iron Oxide	50/t	7,473 t/y	0.37
Gypsum	10/t	23,634 t/y	0.24
Ammonia	200/t	1,853 t/y	0.37
MEA	1100/t	2,242 t/y	2.47
Additive Inhibitor	-	-	0.49
Catalyst for SCR	Change every 3 years	-	1.19
Coal for Precalciner	65/t	0.0633 Mt/y	4.11
Coal for CHP	65/t	0.2283 Mt/y	14.84
Petroleum Coke	80/t	0.0319 Mt/y	2.55
Miscellaneous Materials	-	-	0.72
Power – Cement Plant	0.05/kWh	80,809 MWh/y	4.04
Power – Post Combustion Plant	0.05/kWh	252,857 MWh/y	12.64
Process Water – Cement plant	0.1/t	0.24 Mt/y	0.02
Process Water – Post-Combustion plant	0.1/t	0.15 Mt/y	0.02
Cooling Water – Cement Plant	0.02/t	-	-
Cooling Water – Post-Combustion Plant	0.02/t	4.38 Mt/y	0.09
Total variable operating cost			**48.75**
Power – Generated in the CHP	0.05/kWh	356,400 MWh/y	(17.82)
Net variable operating cost			**30.93**

The fixed cost contribution to the operating costs is shown in Table 5-8.

Table 5-8: Fixed Operating Costs for Post-Combustion Cement Plant with CO_2 Capture

Description	Cement Plant €M	Post-Combustion Plant €M	Total €M
Maintenance	9.4	7.4	16.8
Operating labour	3.2	2.4	5.6
Supervision	0.6	0.5	1.1
Administration and general overheads	1.2	0.9	2.0
Local rates	2.4	2.6	4.9
Insurance	2.4	2.6	4.9
Total fixed operating costs	**19.1**	**16.2**	**35.3**

5.4.3 Cash Flow Calculations

The assumptions set out in section 5.2 have been applied to the IEA Greenhouse Gas R&D Programme model which has been adapted to reflect the operations of a cement manufacturing plant. The model projects the cash flows arising from the construction and operation of a plant for a 25 year period and calculates the net increase in costs as a result of employing CO_2 capture technology. Revenues are not considered, as these are assumed to be identical for all potential plants. Similarly, any implications for a cost of carbon are omitted as these are not relevant to the study. The cash flow spreadsheet for the base case and post-combustion capture are included with Appendix F.

The model assumes that electricity generated on site at the cement plant can be wholly used to offset the consumption of the plant and, where applicable, any excess can be sold to the grid.

A discount rate of 10% is used to calculate the levelised cost of cement manufacture, which represents the cost per tonne of cement produced in current monetary value (at 2008). The results of the model for the base case and for a post-combustion plant are shown in Table 5-9.

Table 5-9: Cost of Cement Manufacture

	€/t
No CO_2 capture (Base Case)	65.7
With capture (Post-Combustion)	129.4

As would be expected given the large increase in the capital cost of the plant, the cost per tonne of cement is increased considerably when post-combustion carbon capture is added.

5.4.4 Calculations of CO_2 Cost

The increase in cost per tonne of cement produced has been used to derive a cost per tonne of CO_2 emissions avoided when compared with the base case. Emissions from the base case plant, with no CO_2 capture, were calculated to be 728,422 t CO_2/yr or 0.728 tonnes CO_2 per tonne of cement produced. For the post-combustion plant, two scenarios have been considered:

- a conservative scenario, in which only the CO_2 captured from cement manufacture at the site is considered when calculating the cost of reduction;
- an holistic scenario, including both the CO_2 captured from cement manufacture, plus the CO_2 saved by on-site power generation and, as the post-combustion cement plant is a net exporter of electricity, displaced CO_2 generation from the grid.

For the second scenario, a CO_2 emissions level of 0.52 kg/kWh for external power has been assumed. Appendix G includes a note on the derivation of this emission factor.

The results for the model are shown in Table 5-10.

Table 5-10: Cost of CO_2 Avoidance

	€ / t CO_2
Power emissions excluded	118.1
Power emissions from grid included	107.4

As indicated by the high increase in cost per tonne of cement produced, the cost per tonne of CO_2 emissions avoided is also relatively high. Given the excess power produced by the plant, which is assumed to be sold to the grid, including the CO_2 not emitted due to displacing fossil fuel power generation results in a reduction in the cost per tonne emitted. However, this is a generous assumption and it may be prudent not to include this in an analysis of post-combustion capture.

The cost per tonne of CO_2 captured is calculated by dividing the incremental cost per tonne increase in producing cement in a post-combustion cement plant by the quantity of CO_2 captured per tonne of cement. For this system the cost per tonne of CO_2 captured is 59.6 €/tCO_2.

5.4.5 Sensitivities

In order to test the robustness of the model results, a number of sensitivity cases for key inputs have been modelled. The key inputs are capital costs, operating costs, plant life, fuel costs and discount rate, and the outcomes are discussed below.

(i) Capital Cost

The impact of a ± 25% variation in the capital cost of the whole plant has been considered. The results are shown in Table 5-11. It should be noted that some of the operating costs (e.g. local rates, insurance and maintenance) are calculated as percentages of the installed plant cost. For the purposes of this sensitivity analysis these costs were assumed to remain constant (i.e. the same as for the base case).

Table 5-11: Capital Cost Sensitivities

		-25%	**Base Case**	**+25%**
€/t cement	Base Case	58.2	65.6	73.1
	Post-combustion CO_2 capture plant	113.7	129.4	145.1
€/t CO_2 avoided	Power emissions excluded	102.7	118.1	133.4
	Power emissions from grid included	93.4	107.4	121.3

As would be expected given the relatively large impact of the capital cost on the cost of cement production and CO_2 capture, the results are very sensitive to variations in the capital cost assumptions.

(ii) Operating Costs

The impact of a ± 25% variation in the operating costs of all plant has been considered. Table 5-12 shows the results. It should be noted that for the purposes of the sensitivity analysis the cost of fuel and power are not included in the operating costs. In the analysis they are assumed to remain at the base case values.

Table 5-12: Operating Cost Sensitivities

		-25%	Base Case	+25%
€/t cement	Base Case	59.4	65.6	71.9
	Post-combustion CO_2 capture plant	117.9	129.4	140.9
€/t CO_2 avoided	Power emissions excluded	108.4	118.1	127.7
	Power emissions from grid included	98.6	107.4	116.2

These results show that the model is sensitive to changes in operating costs, within the range shown.

(iii) Fuel Costs

Fuel costs are difficult to forecast with any degree of accuracy over the longer-term and can fluctuate significantly. The base case costs of coal and petroleum coke were €65/t (2.51 €/GJ (LHV basis)) and €80/t (2.34 €/GJ (LHV basis)) respectively. The impact of a ± 50% variation in the cost of coal and pet coke has been considered. Table 5-13 shows the results.

Table 5-13: Fuel Costs Sensitivity

		-50%	Base Case	+50%
€/t cement	Base Case	62.3	65.6	69.0
	Post-combustion CO_2 capture plant	118.6	129.4	140.2
€/t CO_2 avoided	Power emissions excluded	104.3	118.1	131.8
	Power emissions from grid included	94.9	107.4	119.9

The table shows that the results are affected by variation in fuel costs although to a lesser degree than for capital costs. Post-combustion capture requires a large increase in coal consumption so any increase in cost is bound to impact on the cost per tonne of CO_2 avoided.

(iv) Power Costs

The impact of a ± 25% variation in the cost of power (purchase and sale) has been considered. Table 5-14 shows the results.

Table 5-14: Power Costs Sensitivity

		-25%	Base Case	+25%
€/t cement	Base Case	64.6	65.6	66.7
	Post-combustion CO_2 capture plant	129.7	129.4	129.1
€/t CO_2 avoided	Power emissions excluded	120.5	118.1	115.7
	Power emissions from grid included	109.5	107.4	105.2

The table shows that the results are not greatly affected by variation in power costs. However, as post-combustion capture results in the net export of electricity the cost per tonne of cement or avoidance decreases as the price of power increases due to the increases in revenue.

(v) Compound Operating Costs

The impact of a ± 25% variation in all the operating costs (including fuel and power purchase costs) has also been considered. Table 5-15 shows the results.

Table 5-15: Compound Operating Cost Sensitivities

		-25%	Base Case	+25%
€/t cement	Base Case	56.7	65.6	74.6
	Post-combustion CO_2 capture plant	112.5	129.4	146.3
€/t CO_2 avoided	Power emissions excluded	103.4	118.1	132.7
	Power emissions from grid included	94.0	107.4	120.7

These results show that the model is sensitive to changes in all operating costs, within the range shown.

(vi) Plant Life

The operating life of the plant has been extended to 40 years to examine the impact on the model results. Table 5-16 shows the results.

Table 5-16: Plant Life Sensitivity

		Base Case	40 yrs
€/t cement	Base Case	65.6	63.4
	Post-combustion CO_2 capture plant	129.4	124.7
€/t CO_2 avoided	Power emissions excluded	118.1	113.5
	Power emissions from grid included	107.4	103.2

The sensitivity analysis shows that extending the life of the plant to 40 years only reduces the cost of CO_2 reduction by around 4%. The CO_2 captured is not discounted in the model and so the increased cost of cement production is spread over an additional 15 years of CO_2 capture.

(vii) Discount Rate

The impact of a ± 2% variation around the base case discount rate of 10% applied in the model has been considered. Table 5-17 shows the results.

Table 5-17: Discount Rate Sensitivities

		8%	Base Case	12%
€/t cement	Base Case	61.1	65.6	70.4
	Post-combustion CO_2 capture plant	119.8	129.4	139.5
€/t CO_2 avoided	Power emissions excluded	108.7	118.1	128.0
	Power emissions from grid included	98.9	107.4	116.4

The results show that the cost of CO_2 avoidance is sensitive to the discount rate used. As future cement production becomes more heavily discounted, the cost per tonne produced in order to recoup the capital expenditure increases.

(viii) Emissions Factor

Where the CO_2 emissions from external power have been included an emissions factor of 0.52 kg/kWh has been used. The impact of using a high emissions factor of 0.745 kg/kWh such as that from a new large coal-fired power plant (IEA GHG, 2004) and a low emissions factor of 0.14 kg/kWh to represent electricity from a low CO_2 source such as a coal-fired plant with 85% capture have been considered. Table 5-18 shows the results.

Table 5-18: Emissions Factor Sensitivities

		Low	Base Case	High
Emissions Factor (kg/kWh)		0.14	0.52	0.745
€/t CO_2 avoided	Power emissions excluded	118.1	118.1	118.1
	Power emissions from grid included	115.0	107.4	103.3

The results show that the cost of CO_2 avoidance is sensitive to the emissions factor when power emissions from the grid are included. It can be seen that as the emissions factor increases from low to high the cost of avoidance decreases.

(ix) Low Sulphur Raw Meal and External Steam Supply Scenario

This study was based on a cement plant located in NE Scotland. Raw meal from Northern UK contains relatively higher levels of sulphides compared to the South of the country so it was necessary to assume that the raw meal used in this study contained high levels of sulphide (0.33% (w/w)). This assumption necessitates the use of FGD to meet the SO_x levels required for efficient absorption of CO_2 with MEA. Modern cement plants which do not have these sulphide minerals in their raw materials can achieve SO_2 emissions below 10 mg/m^3 (EC, 2001). The sensitivity of costs to a low sulphur raw meal and an external source of steam were assessed. The rationale of assuming an external source of steam was that a coal-fired CHP plant would require FGD (even with low-sulphur coal) so in order to gain a substantial cost benefit an external source of steam must be assumed. The scenario was based on the following assumptions:

- The SO_2 emissions from the cement plant were below 10 mg/m^3 (daily average basis and standard conditions of 0C, 101.3 kPa, 10% O2 and dry gas).
- Over-the-fence steam was available at the required temperature and pressure for the process.
- The steam was waste heat and was therefore carbon neutral.
- The cost of the steam was €10/t.
- No power is generated on site.
- CO_2 emissions are 15% of base case emissions.
- Coal and pet-coke consumption is the same as the base case.
- Due to the lack of FGD, CHP plant and reduced volume of CO_2 the electrical demand is 50% of the post-combustion option considered in this Report.
- Water consumption is same as base case.
- Cooling water consumption is 60% of the post-combustion option considered in the Report.
- Installed costs are 60% of the post-combustion option considered in the Report.

- The steam requirement is 1.1 Mt/y.
- Labour costs are 60% of the post-combustion option considered in the Report.
- Administration, rates and insurance are 60% of the post-combustion option considered in the Report.

The cash flow spreadsheet for the scenario are included in Appendix F. Table 5-19 shows the results of the analysis.

Table 5-19: Sensitivities of Costs to a Low Sulphur Raw Meal and External Steam Supply Scenario

	Base Case (No CO_2 Capture)	Post-Combustion Base Case	Low Sulphur Raw Meal and External Steam Supply Scenario
€/t cement	65.6	129.4	97.3
€/t CO_2 avoided (power emissions excluded)	n/a	118.1	51.2
€/t CO_2 avoided (power emissions from grid included)	n/a	107.4	55.2

The results show that the cost of CO_2 avoidance can be significantly reduced if the capital costs for FGD and CHP are avoided and steam is purchased from an external source at the assumed cost of €10/t. This indicates that the co-location of a new-build cement plant with an external source of steam should be investigated if the costs associated with CO_2 capture at the cement plant are to be minimised.

(x) Alternative Solvent for CO_2 Absorption Scenario

This study was based on using MEA for CO_2 absorption. The sensitivity of costs to using an alternative solvent such as MHI's KS-1 solvent which has a steam consumption of about 2.7 MJ/kgCO_2 compared to about 3.7 MJ/kgCO_2 were assessed. The scenario was based on the following assumptions:

- The CO_2 capture rate of the MEA and KS-1 systems are identical.
- The solvent replacement costs for MEA and KS-1 are identical.
- Due to the reduction in size of the CHP plant the power produced is 75% of the post-combustion option considered in the Report.
- 75% less coal is used in the CHP plant compared to the post-combustion option considered in the Report.

- Installed costs are 90% of the post-combustion option considered in the Report.

The cash flow spreadsheet for the scenario are included in Appendix F. Table 5-20 shows the results of the analysis.

Table 5-20: Sensitivities of Costs to an Alternative Solvent for CO2 Absorption Scenario

	Base Case (No CO_2 Capture)	Post-Combustion Capture (with MEA)	Post-Combustion Capture (with KS-1)
€/t cement	65.6	129.4	121.6
€/t CO_2 avoided (power emissions excluded)	n/a	118.1	105.0
€/t CO_2 avoided (power emissions from grid included)	n/a	107.4	95.9

The results show that the cost of CO_2 avoidance can be significantly reduced if alternative solvents with a lower requirement for steam are used. This is due to the capital cost savings resulting from the smaller CHP plant required and the knock-on effects caused by the reduction in flue gases from the CHP plant that require treatment. Improvements in CO_2 absorption solvents will have a strong influence on reducing the costs of CO_2 capture using post-combustion at cement plants.

5.5 Economic Analysis of Oxy-Combustion CO_2 Capture

5.5.1 Capital Costs

The capital cost allocation of the oxy-combustion cement plant described in section 4.5.5 is set out as in Table 5-21. The estimated accuracy of the figures quoted is ± 25%. These figures were derived from budgetary prices provided by suppliers, costs scaled from published information and prices obtained from Mott MacDonald's database of equipment prices.

As stated in section 4.5.2(ii) the effect of reducing the nitrogen content in the precalciner may raise the emissivity of the gas and improve the radiative heat transfer within the precalciner. Hence, for the same fuel firing rate, an oxy-combustion precalciner of smaller dimensions may be able to achieve similar calcination conditions as its air-firing counterpart. The reduction in cost associated with a smaller precalciner has not been considered within this study as a detailed design of the precalciner has not been undertaken. The cost saving will be somewhat minor compared to the overall oxy-combustion cement plant and considering the target budget capital cost estimate of ±25%.

Table 5-21: Capital Cost Allocation for Oxy-Combustion Cement Plant with CO_2 Capture

Cement Plant Process Section	**CAPEX budget €M**	**Total €M**
Raw material preparation		
Limestone crusher	2.50	
Shale crusher	1.25	
Stacker-reclaimer system	2.08	
Raw milling and homogenisation		
Raw mill drive	16.5	
Raw mill fan	0.5	
Raw mill ancillaries	Included above	
Raw mill bag filter	2.0	
Bag filter fan	0.5	
Homogenisation chamber silo	5.28	
Kiln preheater		
Preheater (kiln) feed bin	Included below	
Four stage preheater	2.0	
Preheater (kiln) fan	0.5	
Precalciner preheater		
Preheater (calciner) feed bin	Included below	
Five stage preheater	3.0	
Preheater (calciner) fan	0.7	
Precalciner		
Precalciner	0.5	
Rotary Kiln		
Kiln	12.0	
Clinker Cooler		
Grate cooler	12.0	
Cooler bag filter	2.0	
Cooler bag filter fan	0.5	
Clinker crusher	Included below	
Clinker conveyor	1.5	
Clinker storage silo	9.0	
Export clinker hopper	0.3	
Coal preparation		
Coal milling system	5.0	
Petroleum coke preparation		
Petroleum coke milling system	5.0	
Cement milling		
Mill 1	10.0	
Mill 2	10.0	
Cement packing and loading		
Compartmental silo 1	5.0	
Compartmental silo 2	5.0	
Bag packing plant	3.0	
ORC heat recovery unit		
Electrostatic precipitator	1.5	
Heat recovery heat exchanger	2.4	

Cement Plant Process Section	**CAPEX budget €M**	**Total €M**
ORC heat recovery unit	3.6	
Exhaust gas fan	0.8	
Total mechanical and electrical cost of cement plant		**125.9**
Air Separation Unit (ASU)		
Main air compressors	8.3	
Air purification system	Included above	
Main heat exchanger	Included above	
ASU compander	Included above	
ASU column system	Included above	
Pumps	Included above	
Backup storage vessel	Included above	
Gas-to-gas heat exchanger for oxygen/product	0.7	
Recycle		
Recycle stream fan	0.4	
Gas mixer	0.1	
Gas-to-gas heat exchanger for ‘tertiary air’/recycle	1.0	
Electrostatic precipitator		
Electrostatic precipitator	1.5	
Electrostatic precipitator fan	0.3	
CO_2 compression and purification	7.6	
Venturi scruber		
Indirect contact cooler		
Direct contact seawater cooler		
Compressors		
Heat exchangers		
Flue gas expander		
Total mechanical and electrical cost of additional oxy-combustion plant		**19.9**
Total mechanical and electrical cost of oxy-combustion cement plant		**145.8**

The following must be added to the total mechanical and electrical cost:

- construction costs including mechanical erection, instrument and electrical installation, civil works, buildings and site preparation
- design and engineering costs
- other costs including temporary facilities, training, commissioning, start-up costs and spare parts
- EPC services including contractors home services and construction supervision and freight;
- contingency
- fees
- owners costs.

The breakdown of these costs for the oxy-combustion cement plant are shown in Table 5-22.

Table 5-22: Capital Costs for Oxy-Combustion Cement Plant with CO_2 Capture

Description	Cement Plant €M	Oxy-Combustion Plant €M	Total €M
Mechanical and Electrical Equipment Cost	126	20	146
Design and Engineering	44	7	51
Construction	50	9	59
Other Costs	9	6	15
EPC Services	18	1	19
Installed Costs	**247**	**43**	**290**
Contingency	12	4	17
Fees	5	1	6
Owners Costs	12	2	15
Total Investment Cost	**276**	**51**	**327**

5.5.2 Operating Costs

The variable cost element in the total operating cost is derived from the consumables associated with the plant. Table 5-23 lists the consumables used based on 330 working days per year.

Table 5-23: Variable Operating Costs for Oxy-Combustion Cement Plant with CO_2 Capture

Description	Unit Cost (€/unit)	Consumption	€M/year
Limestone	3/t	1,256,739 t/y	3.77
Shale	1.5/t	286,601 t/y	0.43
Sand	50/t	7,542 t/y	0.38
Iron Oxide	50/t	7,542 t/y	0.38
Gypsum	10/t	40,000 t/y	0.40
Coal	65/t	0.0721 Mt/y	4.68
Petroleum Coke	80/t	0.0271 Mt/y	2.17
Miscellaneous Materials			0.72
Power	0.05/kWh	174,562 MWh/y	8.73
Seawater	0.02/t	11.3 Mt/y	0.25
Cooling water	0.02/t	1.7 Mt/y	0.04
Total variable operating cost			**21.9**

The fixed cost contribution to the operating costs is shown in Table 5-24

Table 5-24: Fixed Operating Costs for Oxy-Combustion Cement Plant with CO_2 Capture

Description	**Cement Plant €M**	**Oxy-Combustion Plant €M**	**Total €M**
Maintenance	9.9	0.9	10.7
Operating labour	3.2	0.8	4.0
Supervision	0.6	0.2	0.8
Administration and general overheads	1.2	0.3	1.4
Local rates	2.5	0.4	2.9
Insurance	2.5	0.4	2.9
Total fixed operating costs	**19.8**	**3.0**	**22.8**

5.5.3 Cash Flow Calculations

The assumptions set out in section 5.2 have been applied to the IEA Greenhouse Gas R&D Programme model which has been adapted to reflect the operations of a cement manufacturing plant. The model projects the cash flows arising from the construction and operation of a plant for a 25 year period and calculates the net increase in costs as a result of employing CO_2 capture technology. Revenues are not considered as these are assumed to be identical for all potential plants. Similarly, any implications for a cost of carbon are omitted as these are not relevant to the study. The cash flow spreadsheet for the base case and oxy-combustion capture are included in Appendix F.

The model assumes that electricity generated on site at the cement plant can be wholly used to offset the consumption of the plant and, where applicable, any excess can be sold to the grid.

A discount rate of 10% is used to calculate the levelised cost of cement manufacture, which represents the cost per tonne of cement produced in current monetary value (at 2008). The results of the model for the base case and for an oxy-combustion plant are shown in Table 5-25.

Table 5-25: Cost of Cement Manufacture

	€/t
No CO_2 capture	65.6
With capture (Oxy-Combustion)	81.6

Although the capital cost of oxy-combustion plant is considerable, it is much lower than that for post-combustion capture, and so the cost of cement production is affected to a lesser degree.

5.5.4 Calculations of CO_2 Cost

The increase in cost per tonne of cement produced has been used to derive a cost per tonne of CO_2 emissions avoided when compared with the base case. Emissions from the base case plant, with no CO_2 capture, were assumed to be 728,422 t CO_2 /yr, or 0.728 tonnes CO_2 per tonne of cement produced. For the oxy-combustion plant, two scenarios have again been considered:

- a conservative scenario, in which only the CO_2 captured from cement manufacture at the site is considered when calculating the cost of reduction;
- an holistic scenario, including both the CO_2 captured from cement manufacture and, as the oxy-combustion cement plant imports more power from the grid than the base case, the CO_2 emissions associated with the additional power imported.

For the second scenario, a CO_2 emissions level of 0.52 kg/kWh for external power has been assumed. Appendix G includes a note on the derivation of this emission factor.

The results for the model are shown in Table 5-26.

Table 5-26: Cost of CO_2 Avoidance

	€ / t CO_2
Power emissions excluded	35.8
Power emissions from grid included	40.2

The cost per tonne of CO_2 emissions avoided is significant but much lower than that for post-combustion. However, in this case, including the impact on CO_2 resulting from the increased power consumption increases the cost per tonne captured. Oxy-combustion uses considerable quantities of imported power which it is important to take into account when assessing the overall impact on CO_2 emissions.

The cost per tonne of CO_2 captured is calculated by dividing the incremental cost per tonne increase in producing cement in an oxy-combustion cement plant by the quantity of CO_2 captured per tonne of cement. For this system the cost per tonne of CO_2 captured is 34.3 €/tCO_2.

5.5.5 Sensitivities

In order to test the robustness of the model results, a number of sensitivity cases for key inputs have been modelled. The key inputs are capital costs, operating costs, plant life, fuel costs and discount rate, and the outcomes are discussed below.

(i) Capital Cost

The impact of a ±25% variation in the capital cost of the whole plant has been considered. The results are shown in Table 5-27. It should be noted that some of the operating costs (e.g. local rates, insurance and maintenance) are calculated as percentages of the installed plant cost. For the purposes of this sensitivity analysis these costs were assumed to remain constant (i.e. the same as for the base case).

Table 5-27: Capital Cost Sensitivities

		-25%	Base Case	+25%
€/t cement	Base Case	58.2	65.6	73.1
	Oxy-combustion CO_2 capture plant	72.4	81.6	90.8
€/t CO_2 avoided	Power emissions excluded	31.8	35.8	39.8
	Power emissions from grid included	35.7	40.2	44.7

Although there is an impact on the model results from variations in the capital cost, the impact is less than that for post-combustion as would be expected given the lower cost of the capture plant.

(ii) Operating Costs

The impact of a ±25% variation in the operating costs of all plant have been considered. The results are shown in Table 5-28. It should be noted that for the purposes of the sensitivity analysis the cost of fuel and power are not included in the operating costs. In the analysis they are assumed to remain at the base case values.

Table 5-28: Operating Cost Sensitivities

		-25%	Base Case	+25%
€/t cement	Base Case	59.4	65.6	71.9
	Oxy-combustion CO_2 capture plant	74.3	81.6	88.9
€/t CO_2 avoided	Power emissions excluded	33.5	35.8	38.0
	Power emissions from grid included	37.7	40.2	42.7

Given that the increase in operating costs when adding oxy-combustion CO_2 capture is less than that for post-combustion, the impact of variations of the operating costs on the model results will also be less. The results are therefore not greatly sensitive to variations in the operating cost assumptions, within the range shown.

(iii) Fuel Costs

Fuel costs are difficult to forecast with any degree of accuracy over the longer-term and can fluctuate significantly. The base case costs of coal and petroleum coke were €65/t (2.51 €/GJ (LHV basis)) and €80/t (2.34 €/GJ (LHV basis)) respectively. The impact of a ± 50% variation in the cost of coal and pet coke has been considered. The results are shown in Table 5-29.

Table 5-29: Fuel Costs Sensitivity

		-50%	Base Case	+50%
€/t cement	Base Case	62.3	65.6	69.0
	Oxy-combustion CO_2 capture plant	78.2	81.6	85.0
€/t CO_2 avoided	Power emissions excluded	35.7	35.8	35.9
	Power emissions from grid included	40.1	40.2	40.3

The table shows that, although the cost of cement is affected, variation in the fuel price has little effect on the cost of CO_2 avoidance through oxy-combustion.

(iv) Power Costs

The impact of a ± 25% variation in the cost of power (purchase and sale) has been considered. The results are shown in Table 5-30.

Table 5-30: Power Costs Sensitivity

		-25%	Base Case	+25%
€/t cement	Base Case	64.6	65.6	66.7
	Oxy-combustion CO_2 capture plant	79.4	81.6	83.8
€/t CO_2 avoided	Power emissions excluded	33.2	35.8	38.4
	Power emissions from grid included	37.2	40.2	43.1

The table shows that the results are not greatly affected by variation in power costs. However, as oxy-combustion capture results in the net import of electricity the cost per tonne of cement or capture increases as the price of power increases due to increases in operating costs.

(v) Compound Operating Costs

The impact of a ±25% variation in all the operating costs (including fuel and power costs) has also been considered. The results are shown in Table 5-31.

Table 5-31: Compound Operating Cost Sensitivities

		-25%	Base Case	+25%
€/t cement	Base Case	56.7	65.6	74.6
	Oxy-combustion CO_2 capture plant	70.4	81.6	92.8
€/t CO_2 avoided	Power emissions excluded	30.9	35.8	40.7
	Power emissions from grid included	34.6	40.2	45.7

The results show that the model is sensitive to variations in the operating cost assumptions, within the range shown.

(vi) Plant Life

The operating life of the plant has been extended to 40 years to examine the impact on the model results. These are shown in Table 5-32.

Table 5-32: Plant Life Sensitivity

		Base Case	40 yrs
€/t cement	Base Case	65.6	63.4
	Oxy-combustion CO_2 capture plant	81.6	78.9
€/t CO_2 avoided	Power emissions excluded	35.8	34.6
	Power emissions from grid included	40.2	38.9

The sensitivity analysis shows that extending the life of the plant reduces the cost of CO_2 reduction as the additional capital cost is spread over the additional 15 years of CO_2 captured.

(vii) Discount Rate

The impact of a ±2% variation around the base case discount rate of 10% applied in the model has been considered. The results are shown in Table 5-33.

Table 5-33: Discount Rate Sensitivities

		8%	Base Case	12%
€/t cement	Base Case	61.1	65.6	70.4
	Oxy-combustion CO_2 capture plant	76.0	81.6	87.5
€/t CO_2 avoided	Power emissions excluded	33.4	35.8	38.4
	Power emissions from grid included	37.5	40.2	43.1

The results show that the cost of CO_2 capture is sensitive to the discount rate used. As future cement production becomes more heavily discounted, the cost per tonne produced in order to recoup the capital expenditure increases. As the capital cost for the capture plant is lower than that for post-combustion, the model results for oxy-combustion are less sensitive to variation in the discount rate applied.

(viii) Emissions Factor

Where the CO_2 emissions from external power have been included an emissions factor of 0.52 kg/kWh has been used. The impact of using a high emissions factor of 0.745 kg/kWh such as that from a new large coal-fired power plant (IEA GHG, 2004) and a low emissions factor of 0.14 kg/kWh to represent electricity from a low CO_2 source such as a coal-fired plant with 85% capture have been considered. Table 5-34 shows the results.

Table 5-34: Emissions Factor Sensitivities

		Low	Base Case	High
Emissions Factor (kg/kWh)		0.14	0.52	0.745
€/t cement	Base Case	65.6	65.6	65.6
	Oxy-combustion CO_2 capture plant	81.6	81.6	81.6
€/t CO_2 avoided	Power emissions excluded	35.8	35.8	35.8
	Power emissions from grid included	36.9	40.2	42.4

The results show that the cost of CO_2 capture is sensitive to the emissions factor when power emissions from the grid are included. It can be seen that as the emissions factor increases from low to high the cost of capture decreases.

(ix) Co-location of Oxy-Combustion Cement Plant with Oxy-Combustion Power Plant Scenario

The scenario of co-locating the oxy-combustion cement plant with an oxy-combustion power plant was examined. The rationale behind examining this sensitivity was the potential cost savings available if the oxygen production and CO_2 compression facilities are consolidated. The scenario was based on the following assumptions:

- Oxygen was available over-the-fence from the oxy-combustion power plant.
- The cost of oxygen was €30/t. This is considered typical for industry.
- The CO_2 produced would be exported for compression and purification at the oxy-combustion power plant.
- The cost of exporting the CO_2 for purification and compression was €10/t.
- The power consumed is 55% of the oxy-combustion option considered in this Report.
- The water and cooling water demand is the same for a co-located plant and the oxy-combustion option considered in this Report.
- Installed costs are 90% of the oxy-combustion option considered in this Report.

The cash flow spreadsheets for the scenario are included in Appendix F. Table 5-19 shows the results of the analysis.

Table 5-35: Sensitivities of Costs to Co-location with an Oxy-Combustion Power Plant

	Base Case **(No CO_2 Capture)**	**Oxy-Combustion Base Case**	**Co-location with an Oxy-Combustion Power Plant**
€/t cement	65.6	81.6	80.8
€/t CO_2 avoided (power emissions excluded)	n/a	35.8	33.9
€/t CO_2 avoided (power emissions from grid included)	n/a	40.2	34.4

Based on the assumed values for oxygen and carbon dioxide there is a decrease in the cost of CO_2 avoidance for a cement plant co-located with an oxy-combustion plant. This scenario is heavily dependant on the values for oxygen and carbon dioxide. The option will become cheaper if lower prices for oxygen and carbon dioxide can be negotiated.

5.6 Sensitivities of Costs to an Asian Developing Country Scenario

The sensitivities of costs to an Asian developing country scenario were assessed. The scenario was based on the following assumptions:

- The capacity of the cement plant is 3 Mt/y of cement. This is typical for modern Asian cement plants.
- The equipment costs (both cement plant and CO_2 capture plant) were assumed to be 60% of the European prices.
- A cost-scale exponent of 0.6 was assumed.
- The labour costs were assumed to be 50% of the European prices.
- The administration, rates and insurance were assumed to be 50% of the European prices.
- All costs for fuel and raw materials were assumed to be the same as for the European scenario.
- The performance of the cement plant and CO_2 capture plant was assumed to be the same as for the European scenario. It is recognised that this is a conservative assumption as there may be some improvements at a larger scale plant e.g. a three times larger CHP would be more efficient.

The cash flow spreadsheets for the base case, post-combustion capture and oxy-combustion capture for the Asian developing country scenario are included in Appendix F. Table 5-36 shows the results of the analysis.

Table 5-36: Sensitivities of Costs to an Asian Developing Country Scenario

	Base Case			**Asian Scenario**		
	No CO_2 Capture	**Post-Combustion**	**Oxy-Combustion**	**No CO_2 Capture**	**Post-Combustion**	**Oxy-Combustion**
€/t cement	65.6	129.4	81.6	37.3	72.2	46.4
€/t CO_2 captured (power emissions excluded)	n/a	118.1	35.8	n/a	64.6	20.4
€/t CO_2 captured (power emissions from grid included)	n/a	107.4	40.2	n/a	58.8	22.9

The cost of cement production is significantly lower in the Asian scenario. This is principally due to the economy of scale of operating a larger plant, the lower equipment costs and the cheaper labour. If carbon capture is applied then the cost of production increases significantly (more for post-combustion than oxy-combustion). However, it should be noted that the cost of cement production is only comparable to the European base case when post-combustion capture is applied.

As in the European base case the cost per tonne of CO_2 captured is significant but much lower for oxy-combustion than for post-combustion. However, when compared to the European base case the costs for capture are significantly lower. This indicates that due to the benefits of economies of scale, lower equipment costs and cheaper labour it is more cost effective to undertake carbon capture at a cement plant in Asia than at a European plant.

6 Retro-fitting Issues

6.1 Introduction

The suitability of each CO_2 capture technology for retrofit to existing modern cement plants has also been assessed. This has taken into account estimates of the additional land area required and the need to make modifications to the existing cement plant equipment.

6.2 Post-Combustion Capture

As post-combustion capture of CO_2 from the exhaust gases of a cement kiln by monoethanolamine (MEA) absorption is an "end-of-pipe" technology then it could be retrofitted to any existing cement kiln. The economic feasibility of that would be dependent on the particular kiln and the costs of overcoming the key constraints on the application of the technology, which are:

- The MEA solvent is degraded by NO_2, SO_2 and O_2 in the exhaust gases.
- There are major space requirements for the absorption and subsequent solvent stripping/regeneration columns.
- There are major steam requirements for absorbent stripping and regeneration
- There are major power requirements for compressing the CO_2 gas.

Whether post-combustion chemical absorption could be economically retrofitted to an existing cement kiln would therefore be determined by a technical audit of the existing cement kiln. The availability of space for the absorption and stripping/regeneration columns would be quickly assessed. Also the availability of sufficient steam or the feasibility of providing the necessary additional steam – with sufficient land area and fuel supply infrastructure for on-site steam-raising or cogeneration.

Under the case where a 45 MWe CHP plant is used to raise the necessary steam, there will be electrical energy available for export so sufficient grid connection capacity to accommodate the maximum export load would therefore be required.

The potentially greater challenge would be to assess the equipment that would need to be installed to reduce the NO_2, SO_2 and O_2 in the exhaust gases prior to MEA absorption of CO_2. Cement plants with no oxidisable sulphur compounds in their raw materials have negligible SO_2 emissions. The kilns on such cement plants would therefore be ready candidates for retrofitting of this CO_2 capture technology. Where cement plants do use raw materials containing oxidisable sulphur compounds then those that had already been provided with SO_2 scrubbing equipment would also be ready candidates although it is likely that the FGD systems would require some upgrading to achieve the low SO_x levels required for economic CO_2 capture with MEA. Alternatively SO_2 scrubbing equipment could be installed as part of the equipment required for post-combustion CO_2 capture by MEA absorption.

Installation of SCR technology to reduce the NO_2 content of the exhaust gases prior to MEA absorption of CO_2 would need to be undertaken. However, there are currently only two such installations on cement kilns around the world and there is still some uncertainty whether this would reduce the NO_2 content of the exhaust gases sufficiently. This is the only equipment that would directly affect the cement process when it is installed. SCR must be added between the preheater and the raw mill, with additional ammonia sprays necessary in the preheater. This will modify the process and may require additional fuel input to maintain the temperature for the raw mill but it can be integrated.

All current cement kilns have significant oxygen, O_2, content in their exhaust gases. MEA is degraded by oxygen. However, the use of additives can inhibit the oxidation of MEA and so this does not represent a significant obstacle.

The flue gases from a cement plant are generally higher in CO_2 concentration than those from coal or gas fired power plants. This makes the flue gas from a cement plant more amenable to capture of CO_2 than power plants due to the efficiency of absorbing CO_2 from a concentrated gas stream. There may be further economic benefits from reducing air in-leaks in the cement plant to ensure an even more concentrated stream of CO_2 in the flue gas. Hence, retrofitting post-combustion capture of CO_2 by MEA absorption may require major refurbishment and provision of additional equipment to minimise air in-leaks along the kiln exhaust gas flow path. However, given that the capital cost of the absorber tower is a relatively small fraction of the overall cost of capture, air in-leakage is not considered to be a 'showstopper'.

6.3 Oxy-Combustion Capture

Operating the precalciner in an oxy-combustion mode results in fundamental changes to the reaction conditions. Hence, retrofitting oxy-combustion in the precalciner only, and capture of the CO_2 in the exhaust gases from the precalciner only, could only be readily undertaken in certain types of cement plants, dependent on their process and design.

An assumption of this study has been that retrofits for CO_2capture would only be applicable to modern, existing, cement plants. Capture in older, smaller plants was not considered. Modern, existing, cement plants will most likely be equipped with a precalciner. However, in order for the retrofit to be relatively straightforward the precalciner would need to be operated as a separate-line calciner (SLC). It would then be possible to retrofit oxy-combustion to a precalciner cement kiln with a precalciner combustion chamber installed parallel with the exhaust gas riser from the rotary kiln, provided that the kiln was also provided with twin (or multiple) preheater towers. Mott MacDonald notes that operation with a separate-line calciner (SLC) is not currently the preferred technology for cement production.

In assessing the suitability of an existing cement plant for retrofit of this CO_2 capture technology the criteria would be as follows:

- The kiln is provided with a separate line (or separate combustion chamber) precalciner.
- The kiln is provided with twin (or multiple) preheater towers.
- Those twin (or multiple) preheater towers to have separate, dedicated, induced draft fans.
- The solid fuel grinding system to be swept by inert preheater exhaust gases from one (the intended rotary kiln) preheater tower.
- Sufficient space is available alongside the preheater tower, (on the intended precalciner side) for installation of the air-separation unit, CO_2 purification and compression equipment. Mott

MacDonald estimates that for a 1 Mt/y cement plant the area required would be approximately 0.5 ha.

- The electrical energy supply capacity is sufficient for the maximum load to be increased by approximately 15 MW to accommodate the additional power requirements.

Once these criteria have been assessed, it would then be necessary to undertake considerable process modelling and design effort to ensure the technology could be successfully introduced.

It is important to note that one of the key outcomes from this study is the significantly lower cost of employing oxy-combustion technology over post-combustion technology. The criteria developed above could be used to determine the most suitable plant for retrofit of this technology. However, given the cost differential between oxy-combustion and post-combustion capture it may be more economic to undertake a major rebuild of the kiln (i.e. from single to twin preheater towers) and undertake the oxy-combustion retrofit rather than implementing the post-combustion option. The costs of undertaking the retrofit would need to be assessed on a case by case basis.

7 Capture-Ready Cement Plants

7.1 Introduction

The G8 communiqué (Gleaneagles Plan of Action, 2005) raises the question of a definition for 'capture-ready', for which a consensus definition is not currently provided by the literature. A recent study (IEA GHG, 2007a) commissioned by the IEA GHG provides a comprehensive review of the literature, noting that the definition of capture ready can be either primarily technical or economic, and is likely to be ultimately determined by regulators. It is recommended that readers refer to the IEA GHG study for an overview of the relevant literature.

The IEA GHG definition for capture-ready plant outlines the essential considerations that would apply globally to thermal generation plant, irrespective to the specific regulatory and economic context as follows:

"A CO_2 capture ready plant is a plant which can include CO_2 capture when the necessary regulatory or economic drivers are in place. The aim of building plants that are capture ready is to reduce the risk of stranded assets or 'carbon lock-in'.

Developers of capture ready plants should take responsibility for ensuring that all known factors in their control that would prevent installation and operation of CO_2 capture have been identified and eliminated.

This might include:

- *A study of options for CO_2 capture retrofit and potential pre-investments*
- *Inclusion of sufficient space and access for the additional facilities that would be required*
- *Identification of reasonable route(s) to storage of CO_2*

Competent authorities involved in permitting power plants should be provided with sufficient information to be able to judge whether the developer has met these criteria."

The same considerations as developed above for power plant apply equally well to cement plant. Mott MacDonald has reviewed ways of making cement plants 'capture-ready' and the findings are presented in the sections below.

7.2 Post-Combustion Capture

This study has used chemical absorption with monoethanolamine (MEA) for the post-combustion capture of CO_2 from a cement plant. Essentially this is an "end-of-pipe" technology and the requirements for modification of the cement manufacturing process are minimal. As such the technology could be applied to any cement plant. The key constraints on the application of the technology are:

- The MEA solvent is degraded by NO_2, SO_2 and O_2 in the exhaust gases.
- There are major space requirements for the absorption and subsequent solvent stripping/regeneration columns.

- There are major steam requirements for absorbent stripping and regeneration.
- There are major power requirements for compressing the CO_2 gas.

A "capture-ready" cement plant would therefore need to make provision for overcoming these constraints. Overcoming the space constraint would simply require that sufficient space was reserved for the absorption and stripping/regeneration columns as well as additional space for the CO_2 compression and purification and local CO_2 storage and/or pipeline. The power requirement could be met if sufficient electrical energy supply from the grid was available or by a captive power plant. However, a steam supply is also needed so consideration must be given to how this will be provided. In this study it has been assumed that a captive CHP plant would provide the steam and electrical power. In this case the incremental equipment for generation of the additional steam and power would be added at the time the CO_2 capture was implemented and sufficient space for this equipment would need to be allowed for. Location of the new cement plant could also aid with the plant being capture ready. Being close to a power station willing to sell both heat and power or to an industrial provider of steam would be particularly advantageous. Being close to links for CO_2 storage would also help.

The potentially greater challenge is reducing the levels of NO_2 and SO_2 in the exhaust gases from a "capture-ready" cement kiln. Achieving low SO_2 levels in the exhaust gases prior to solvent absorption of CO_2 would be the most readily achieved. Provided that the "capture-ready" cement plant utilised raw materials with no oxidisable sulphur content then there would be negligible SO_2 content in the exhaust gases. If the cement plant did use raw materials containing oxidisable sulphur compounds then installation of wet or dry scrubbing equipment would be sufficient to reduce the SO_2 to levels acceptable for CO_2 absorption by MEA.

Sufficient reduction of the content of NO_2 in the exhaust gases to less than 41 mg/Nm^3 (20 ppmv at 6% O_2) prior to MEA absorption is more problematic. The SCR installation at the Solnhofen cement plant in Germany has demonstrated the ability to reduce NO_x emissions to less than 200 mg/Nm^3 in short-term trials. Based on NO_2 comprising less than 10% of these total NO_x emissions and the reported low absorption rates of NO_x within CO_2 absorption systems (Knudsen *et al.*, 2006 and Iijima *et al.*, 2007) then installation of SCR equipment on the "capture-ready" cement kiln exhaust ahead of the MEA solvent absorption column should be sufficient to reduce the NO_2 content of the exhaust gases sufficiently.

The oxygen content of the exhaust gas from a cement plant would typically be 7~10%. MEA is degraded by oxygen so it is beneficial to have a low oxygen content in the exhaust gases prior to MEA absorption of CO_2. However, the use of additives can inhibit the oxidation of MEA and so this does not represent a significant obstacle.

Although it must be recognised that the flue gases from a cement plant are generally higher in CO_2 concentration than those from coal or gas fired power plants the reduction of air in-leakage in the cement process would be beneficial to the economic capture of CO_2 as it is easier to absorb CO_2 from a concentrated stream. Current cement manufacturing technology utilises the exhaust gases from the cement kiln for drying of raw materials in the raw grinding mill. There are significant air in-leaks in the cement kiln exhaust gas flow path from the kiln inlet, through the preheater, raw mill, gas conditioning and dust collection systems. An ideal "capture-ready" cement plant would deploy additional technologies to minimise these air in-leaks. This would boost the thermal efficiency and productivity of the cement kiln and would therefore be in the interests of the industry. Air in-leaks are minimised in solid fuel grinding systems so it should therefore be feasible to reduce air in-leaks in raw grinding systems, given sufficient incentive to do so. Prevention of the dilution of the exhaust gases by in-leaking air in order that they be most suitable for post combustion capture of CO_2 by MEA absorption would be such an incentive.

7.3 Oxy-Combustion Capture

This study has examined oxy-combustion in the precalciner only and capture of the CO_2 in the exhaust gases from the precalciner only. This requires that the exhaust gases from the precalciner are vented separately from the exhaust gases from the rotary kiln. In order to maintain the thermal efficiency of the cement kiln process the exhaust gases from the precalciner and the rotary kiln are drawn through separate, dedicated preheater towers.

Such separate line calciner (SLC) cement kilns are in the product range of the major cement kiln equipment supplier, FL Smidth, and were regularly provided in the late 1980s and 1990s. Some kilns provided by equipment supplier KHD also operate in this separate line mode. This kiln configuration was perceived to be advantageous as the precalciner was aspirated with pure tertiary air and therefore combustion took place in an atmosphere of 23% (w/w) oxygen.

In an in line calciner (ILC) cement kiln the exhaust gas from the rotary section of the cement kiln is vented through the precalciner. The oxygen content of this vitiated exhaust from the rotary kiln is raised by separately ducting pure tertiary air to the precalciner. Despite this introduction of tertiary air, the combustion efficiency in these in line calciners is compromised by the lower overall oxygen content of the gases sweeping the calciner.

In recent years SLC cement kilns have fallen from favour for a number of reasons:

- Completely separating the kiln exhaust gases from the precalciner precludes using staged combustion in the precalciner as a means to mitigate thermal NOx emissions generated in the rotary section of the kiln.

- The problem of combustion in a depleted oxygen atmosphere in an ILC kiln can be overcome by locating the precalciner combustion chamber parallel with the exhaust gas riser from the rotary kiln. Pure tertiary air is delivered to the precalciner combustion chamber and ducting takes the exhaust gases from the precalciner to the rotary kiln exhaust gas riser. This technology is known as a “downdraft” or Reinforced Suspension Preheater (RSP) type precalciner. In these configurations the precalciner and rotary kiln share the same preheater tower or multiple preheater towers.

- In the event of a power failure with an SLC kiln the burden of material in the precalciner preheater tower falls into the base of precalciner rather than into the rotary kiln.

Installation of a cement plant that is designed to be capture-ready for oxy-combustion in the precalciner only would require twin preheater towers with separate induced draft fans dedicated to each tower. Ideally the SLC configuration would be installed. However, switching from a parallel precalciner combustion chamber to fully separate venting would only require minor preheater ductwork modifications. Inert preheater exhaust gas for solid fuel drying would be drawn from the actual, or intended on switching, rotary kiln preheater tower.

Other ductwork modifications would be required on switching from conventional to precalciner only oxy-combustion operating mode. The tertiary air duct would be redirected from the precalciner to the raw mill via a gas-to-gas heat exchanger to exchange with the recycled precalciner exhaust gas. The precalciner tower exhaust gas duct would be split after the induced draft fan with one duct recirculating exhaust to the precalciner and the other duct leading to the CO_2 purification and compression plant.

A “product” layout of the capture ready cement plant, with the equipment laid out in a straight line in the sequence of operations to manufacture cement, would facilitate this switching between conventional and CO_2 capture operating mode. Tertiary air would be ducted along the length of the rotary kiln from the cooler to the preheater in both modes of operation. Extension of the tertiary air duct from the precalciner to the raw mill behind the preheater tower would not be a major task. Similarly, splitting the exhaust duct from the precalciner tower after the induced draft fan with one duct recirculating to the precalciner would not be overly onerous.

Sufficient space would need to be provided for the air separation unit and CO_2 purification and compression plant and would ideally be installed adjacent to the preheater tower and kiln on the side closest to the precalciner. This study has estimated that the space required for a 1 Mt/y cement plant would be approximately 0.5 ha.

In conventional mode a 1 Mt per year cement plant operating for 330 days per year would require around 16 MW of installed electrical supply capacity. This study has estimated that an oxy-combustion cement plant would require approximately 31 MW of installed electrical supply capacity. A capture–ready oxy-combustion plant would need to make allowance for this additional power requirement by ensuring that a suitable grid connection was available.

8 Information Gaps and R&D Needs

8.1 Introduction

In this section important information gaps are identified and suggestions for research, development and demonstration are made.

For both technologies, there is important potential for realising economies of scale through co-location of cement plant with power plants, or other large emitters, also implementing CCS. This would be particularly true if the capture concept (e.g. oxy- or post-combustion) employed by the co-located plants was the same – allowing shared use of ASUs, MEA units etc. Economies of scale in CO_2 transport and storage are also likely. It is recommended that further quantitative study of the benefits available through co-location be carried out.

8.2 Post-Combustion Capture

With respect to the further development of post-combustion CO_2 capture at cement plants the following suggestions for research, development and demonstration are made:

- The opportunities for combustion of waste MEA in the kiln. This must include research into the influence of trace metals in the waste MEA on the cement product quality.
- Specific investigation into high concentration CO_2 flue gas capture techniques, combined with continuing attention to application of the latest post-combustion capture technologies being developed for power generation applications.
- Specific investigation into the influence of NO_x levels on the performance of MEA.
- Further work and full scale tests must be performed to assess the viability of high dust SCR.
- FGD, SNCR and SCR efficiencies – to meet the stringent levels required to avoid excessive amine waste.
- Further study of hybrid combinations of FGD equipment to meet required SO_x removal efficiency at minimum cost – including combinations of wet limestone, seawater scrubbing and absorbent injection within the cement process.
- Further work to reduce air in-leaks into process equipment resulting in a higher CO_2 concentration in the flue gas stream.
- Alternative CO_2 capture techniques that use less steam, therefore smaller heat requirement and require a smaller CHP.
- Alternative CO_2 capture techniques that require less gas clean up will significantly reduce the capital and operating costs of the plant.
- A pilot scale post-combustion CO_2 capture plant must be built and operated with the cement process to demonstrate that the post-combustion process can be implemented successfully.

- Further study into the opportunities and potential economic savings for co-location of cement plants with power plants particularly where post-combustion capture is to be applied at the power plant.

8.3 Oxy-Combustion Capture

With respect to the further development of oxy-combustion CO_2 capture at cement plants the following suggestions for research, development and demonstration are made:

- The optimal ratio of O_2 to CO_2 within an oxy-combustion cement process should be investigated.
- The heat transfer characteristics of the atmosphere within an oxy-combustion cement process must be understood.
- The kiln wall refractory lining behaviour at different oxygen levels must be determined.
- Demonstrate that clinker produced in a CO_2 environment has similar properties to that produced by conventional means.
- Process chemistry (thermodynamics and equilibrium) for calcination in a CO_2 environment require further investigation.
- Further study of the operation of gas-to-gas heat exchangers at the high temperatures (~800 C) and high dust load required to maximise the heat integration within the oxy-combustion process.
- A pilot scale oxy-combustion cement process must be built and operated to gain insights (e.g. operation of suspension preheaters with O_2/CO_2 atmosphere, determination of optimal precalciner temperature etc.) and demonstrate that the oxy-combustion process can be implemented successfully.
- The feasibility and economics of undertaking oxy-combustion in both the precalciner and the kiln to give the possibility of capturing greater than 90% of the CO_2.
- Further work to reduce air in-leaks into process equipment to improve the feasibility and economics of undertaking oxy-combustion in both the precalciner and the kiln.

It should be noted that during the course of this study Mott MacDonald became aware that cement equipment manufacturers are already undertaking some research and development into the oxy-combustion process. It is expected that the number of research papers relating to this topic will increase in the near future.

9 Comparison between Post-Combustion and Oxy-Combustion for CO_2 Capture at Cement Plants

9.1 Introduction

This section summarises and compares the post-combustion and oxy-combustion solutions for CO_2 capture at cement plants that have been considered in this study.

9.2 Summary

The options for applying post-combustion and oxy-combustion carbon capture at cement plants are summarised in Table 9-1.

Table 9-1: Summary of Oxy-combustion and Post-combustion Cement Plants with CO_2 Capture

	Parameter	Unit	Base Case	Post-Combustion	Oxy-Combustion
Inputs	Coal	t/y	63,317	291,633	72,061
	Petroleum Coke	t/y	32,876	32,876	27,091
Outputs	Clinker produced	t/y	910,000	910,000	910,000
	Cement produced	t/y	1,000,000	1,000,000	1,000,000
	CO_2 captured	t/y	0	1,067,734	465,014
	CO_2 emitted on site	t/y	728,422	188,424	282,853
	Net Power Demand	MWh/y	80,809	-22,734	174,562
Performance	CO_2 associated with net power imports	t/y	42,021	-11,822	90,772
	Net quantity of CO_2 avoided (excluding net power imports)	t/y	N/A	539,998	445,569
	Net quantity of CO_2 avoided (including net power imports)	t/y	N/A	593,841	396,818
	CO_2 avoided, net of power imports	%	-	77	52
Costs	Total Investment Cost	€M	263	558	327
	Net Variable Operating Costs	€M/y	17	31	22
	Fixed Operating Costs	€M/y	19	35	23
	Cost per tonne of CO_2 emissions avoided [i]	€/t	N/A	107.4	40.2
	Costs per tonne of cement product	€/t	65.6	129.4 ($\Delta = +63.8$)	81.6 ($\Delta = +16.0$)
	Cost per tonne of CO_2 captured	€/t	N/A	59.6	34.3

[i] Cost per tonne of CO_2 emissions avoided is calculated by first determining the difference between the CO_2 emissions associated with the design under consideration and the base case. The additional costs incurred (compared to the base case) to achieve this reduction are then evaluated and the result expressed as €/tCO_2.

9.3 Comparison with Other Studies

As the feasibility of capturing CO_2 at cement plants has not been widely investigated or reported in the literature it is difficult to compare the costs determined from this study. The key figures from the Hegerland *et al.* (2006) evaluation of applying post-combustion CO_2 capture as a retrofit at a 1.4 Mt/y cement plant in Norway are reported in Table 9-2. The reported accuracy of the figures is ±35%.

Table 9-2: Conceptual Costs for Retrofitting Post-Combustion CO_2 Capture Reported by Hegerland *et al.* (2006)

Parameter	Norwegian Kroner	Euro
Total equipment cost	255 MNOK	32 €M
Total investment cost	877 MNOK	111 €M
Total variable operating costs	212 MNOK/y	27 €M/y
Fixed operating costs	40 MNOK/y	5 €M/y
Total cost per capture	360 NOK/tonne of CO_2	46 €/tonne of CO_2

On first inspection the figures reported in Table 9-2 appear lower than those determined in this study. However, it should be noted that the option investigated by Hegerland *et al.* (2006) was significantly different to that chosen for this study making a like for like comparison difficult. The most important differences were:

- SNCR was used for NO_x reduction
- seawater scrubbing was used for SO_2 reduction
- natural gas was used to provide the energy for the amine plant
- the plant considered in Norway was larger and therefore offered more economies of scale.

These differences together with the fact that the flue gas outlet from the cement process contained much lower levels of SO_2 (200 mg/Nm^3 of SO_2) account for the differences between the estimated costs.

Mahasenan *et al.* (2005), based on a survey of literature and the typical CO_2 content of the flue gas from cement plants, estimated the cost of capturing CO_2 from the stack of a cement kiln using an amine-based process at about \$50/tonne of CO_2 (approx. €36/tonne of CO_2) plus another \$9/tonne of CO_2 (approx. €6/tonne of CO_2) to compress the CO_2 to pipeline specifications (not fully described). The authors of the study admit that the estimates are approximate and that additional research is needed to better estimate the magnitude and variability of these costs. Based on the findings from this Report and Hegerland *et al.* (2006) it appears that this figure underestimates the costs of providing post-combustion capture at a cement plant.

Mott MacDonald are not aware of any published reports indicating the costs of applying oxy-combustion at a cement plant. However, Zeman (2008) provided the following costs from his investigations into an oxygen-fired kiln:

- The cost of the oxygen plant is estimated at \$3/tCO_2 (approx. €2/tCO_2) captured.
- The cost of the CO_2 compression equipment is estimated at \$7-8/tCO_2 (approx. €5.1/tCO_2) captured.
- The additional electricity costs associated with the production of oxygen and compression of CO_2 are estimated as \$6.85/tCO_2 captured (€4.69/tCO_2 for pulverised coal fuel) and \$5.51/tCO_2 captured (€3.83/tCO_2 for natural gas fuel).

- The minimum cost of capture is estimated at between \$15 and \$18 per tonne of CO_2 captured (€11-13/tCO_2).

Once again, on first inspection the figures reported by Zeman (2008) appear significantly lower than those determined in this study. However, it should be noted that the option investigated by Zeman (2008) considers oxy-combustion being applied to the whole cement plant and that these figures are considered as 'minimum'. Mott MacDonald considers the estimated cost to be conservative.

It should be noted that the quantity of oxygen needed per tonne of CO_2 captured in the case considered in the Report is expected to be less than in the case where oxy-combustion is applied to the whole plant. This is because the majority of the extra CO_2 captured in a total capture case would be fuel-based, which requires oxygen, rather than calcination-based, which does not.

9.4 Comparison between Cement Industry and Power Industry

Previous studies (IEA GHG, 2004) have calculated the cost of avoidance of CO_2 emissions at a USCPF post-combustion power plant using coal at 29.5 \$/tCO_2 (25 €/tCO_2 at exchange rate of the day). This is significantly lower than the estimated cost reported here. Mott MacDonald considers that the main reasons for this difference to be:

- The fact that there is no steam supply available for amine stripping at a cement plant. The costs of providing a CHP plant capable of meeting the steam requirements are significant.

- The high sulphur content in the raw meal considered as part of this study increased the cost of FGD.

- The economies of scale offered by undertaking carbon capture at a large power plant. The plant considered in the IEA GHG study was designed to capture approximately 5.4 Mt/y. The post-combustion cement plant considered in this Report captures approximately 1.1 Mt/y.

- There have been significant rises in plant equipment costs, particularly in the power sector, since the IEA GHG study was carried out in 2004. These cost rises have been primarily driven both by materials costs and high demand for new-build plants. Such high equipment costs both increase the current base plant cost and the penalty associated with implementation of CO_2 capture relative to the original study results.

- There have been significant increases in fuel prices since the IEA GHG study was carried out in 2004. IEA GHG (2004) was based on a coal price of \$1.5/GJ (LHV basis) where as this study used €2.51/GJ (LHV basis) which is approximately \$3.5/GJ (LHV basis).

- Given that application of post-combustion CO_2 capture to cement plant is a relatively new field of study, without existing demonstrations to provide technology learning, the design proposed by this study is conservative relative to power sector examples. Further study would enable refinements to be made to the design approach which would be expected to reduce the cost of CO_2 capture.

Davison (2008) provided updated costs of post-combustion CO_2 capture in power plants using Fluor's Econamine FG+ process which was used as the basis of the IEA GHG study published in 2004 (IEA GHG,2004). Inflating capital costs and capital-related operating costs (insurance, local taxes and maintenance) from early 2004 by 25% in Euro terms and using a coal cost of 65 €/t resulted in a cost of CO_2 emissions avoidance of 39 €/tCO_2 and electricity costs of 59.9 €/MWh without capture and 84.1 €/MWh with capture. This indicates that: (i) the cost of CO_2 emissions avoidance by post-combustion capture is more expensive at a cement plant than at a power plant; and (ii) oxy-combustion capture at a cement plant is more cost effective than post-combustion capture at a power plant.

It should also be recognised that co-location of a post-combustion cement plant with a post-combustion power plant would offer significant cost-saving opportunities as equipment (e.g. FGD, CO_2 absorbers, CO_2 compressors) could be shared. Mott MacDonald suggests that under this scenario the cost of CO_2 avoidance for the cement plant would reduce significantly from the values presented in Table 9-1. The limit of the reduction would be the cost of abatement at the power plant.

Previous studies (IEA GHG, 2005) have calculated the cost of avoidance of CO_2 emission at an oxy-combustion power plant using pulverised coal at 37 \$/tCO_2 (31 €/tCO_2 at exchange rate of the day). This is slightly lower than the estimated cost reported here. In addition to the equipment cost rises noted above, Mott MacDonald considers the main reason for this difference to be economies of scale. The target net power of the coal-fired power plant considered in the IEA GHG study was 500 MWe. The CO_2 captured at a power plant of this scale is approximately 3.3 Mt/y. The oxy-combustion cement plant considered in this Report captures only 0.47 Mt/y. Mott MacDonald would expect the economies of scale offered by the power plant application to reduce the levelised cost of capture.

10 Conclusions

The conclusions from this study are outlined below.

10.1 Post-Combustion CO_2 Capture

10.1.1 Technical Issues

- Economic CO_2 capture using amine solvents requires flue gas impurities to be removed. Post-combustion capture at cement plants will be more technically and commercially favourable when applied at cement plants with low SO_2 and low NO_2 concentrations in the flue gas as this will reduce the costs associated with desulphurisation and deNOx.

- This study used wet scrubbing FGD and SCR to achieve the required levels of SO_2 and NO_2 for economic capture of CO_2 using amines. These technologies are considered conservative approaches and other less capital intensive technologies of reducing levels of SO_2 and NO_2 (e.g. addition of solid sorbents or use of SNCR) may meet the required levels.

- The additional steam requirements for post-combustion CO_2 capture using amine solvents are significant and result in additional CO_2 emissions which require capture themselves. This indicates that post-combustion capture will be most efficient and cost effective if the cement plant is co-located near a pre-existing readily available steam supply e.g. a power station.

10.1.2 Performance

- A post-combustion capture cement plant is capable of capturing the majority of CO_2 emitted from a cement plant including those emissions associated with the auxiliary power generation. This can occur with minimal impact on the conventional cement production process.

- The footprint of a post-combustion capture cement plant producing 1 Mt/y of cement is estimated to be in the range 3.6-4.3 ha.

10.1.3 Environmental

- A post-combustion capture cement plant will achieve emission standards well below current cement plants. This is an added advantage of CO_2 capture.

10.1.4 Economics

- For the post-combustion capture cement plant considered in this study the cost per tonne of CO_2 emissions avoided is €107.4/t.

- For the post-combustion capture cement plant considered in this study the cost delta associated with the capture plant compared to the cost of the plant without capture is €63.7 per tonne of cement product.

10.1.5 Retrofitting

- It is technically feasible to retrofit post-combustion capture using amine absorption to any existing cement plant.
- The key constraints influencing the economic feasibility of applying post-combustion capture as a retrofit solution are:
    - The concentration of NO_2 and SO_x in the flue gases.
    - The space requirements for the capture and compression equipment.
    - The availability of sufficient steam for the amine absorption and regeneration system.
- The cost effectiveness of measures to reduce air in-leaks within the cement process would also need to be evaluated before a post-combustion capture solution was retrofitted at a cement plant.

10.1.6 Capture Ready

- A cement plant that was designed to be capture ready for post-combustion capture using amine would need to make provision for the following:
    - Leaving sufficient space for the flue gas treatment, CO_2 absorption, CO_2 purification and CO_2 compression equipment.
    - Fuel supply capacity, grid supply capacity and land for a power plant capable of generating the additional steam and electricity required by the capture equipment, or else potential for over-the-fence steam purchase from neighbouring cogeneration facilities.
    - Including SO_x and NO_2 abatement technologies within the plant or providing the space and capability to introduce these later.
    - Including technologies to reduce air in-leaks within the cement process.
- It is clear that it would be beneficial if the cement plant was co-located near an existing readily available steam supply (e.g. a power plant) that could be used when the post-combustion capture plant was installed.

10.2 Oxy-Combustion CO_2 Capture

10.2.1 Technical issues

- A number of major technical issues associated with the oxy-combustion process require further research and characterisation if the technology is to be developed further. These include:
    - The influence of the higher flame temperatures possible with combustion in an O_2/CO_2 atmosphere on the design and operation of the precalciner.

  - The influence of the O_2/CO_2 atmosphere on the heat transfer characteristics within the combustion chamber and the effect on the design and operation of the precalciner.
  - The influence of the O_2/CO_2 atmosphere on the feed lifting between cyclone stages in the suspension preheater.
  - Optimising the balance between achieving high temperatures for the cement production process (which are easier to achieve with oxy-combustion) and having to replace the kiln wall lining.
  - Characterising the thermodynamics, kinetics and equilibrium chemistry for an oxy-combustion process.
  - Confirming that the product obtained from the oxy-combustion process meets the required quality parameters.

- A pilot scale oxy-combustion cement process must be built to demonstrate that the oxy-combustion process can be implemented successfully.

- At the time of writing this report Mott MacDonald does not consider any of the technical issues associated with the development of oxy-combustion to be “show-stoppers”. Oxy-combustion CO_2 capture at cement plants is considered feasible although until it is demonstrated at pilot scale it will remain a higher technical risk solution than post-combustion CO_2 capture.

10.2.2 Performance

- Initially, the fact that the majority of CO_2 in a modern cement plant originates within the precalciner offered the possibility of capturing the majority of the CO_2 by focussing on a single process unit. Further research on this possibility and the preparation of a heat and mass balance has indicated that altering the process to focus on the precalciner reduces the quantity of CO_2 produced in the precalciner.

- This study has determined that a CO_2 capture cement plant with oxy-combustion in the precalciner only is capable of capturing 47% of the CO_2 emitted from the plant (excluding those emissions associated with the additional power requirements). This can occur with minimal impact on the kiln operation.

- The footprint of a CO_2 capture cement plant using oxy-combustion capture in the precalciner only and producing 1 Mt/y of cement is estimated to be similar to a conventional cement plant i.e. approximately 15.5 ha.

10.2.3 Environmental

- An oxy-combustion capture cement plant will achieve emission standards well below current cement plants. This is an added advantage of CO_2 capture.

10.2.4 Economics

- For the oxy-combustion capture cement plant considered in this study the cost per tonne of CO_2 emissions avoided is €42.4/t.

- For the oxy-combustion capture cement plant considered in this study the cost delta associated with the capture plant compared to the cost of the plant without capture is €16.8 per tonne of cement product.

10.2.5 Retrofitting

- Key constraints influencing the economic feasibility of applying oxy-combustion capture as a retrofit solution are:
    - The process and design of the existing kiln.
    - The space requirements for the ASU, CO_2 purification and compression equipment.

- The electrical energy supply capacity to be sufficient for the maximum load to be increased by approximately 15 MW to accommodate the additional power requirements (for a 1 Mt/y cement plant).

- The most feasible scenario for retro-fitting oxy-combustion in the precalciner only would be if the existing precalciner operated as a separate line calciner (SLC). SLC operation is not currently the preferred technology within the cement industry.

- However, given the significant cost differential between employing oxy-combustion and post-combustion capture it may be more economic to undertake a major rebuild of the kiln (i.e. from single to twin preheater towers) and undertake the oxy-combustion retrofit rather than implementing the post-combustion option. The costs of undertaking the retrofit would need to be assessed on a case by case basis.

10.2.6 Capture Ready

- Installation of a cement plant that is designed to be capture-ready for oxy-combustion in the precalciner only would require twin preheater towers with separate induced draft fans dedicated to each tower. Ideally the SLC configuration would be installed. However, switching from a parallel precalciner combustion chamber to fully separate venting would only require minor preheater ductwork modifications.

- Sufficient space would need to be provided for the air separation unit and CO_2 purification and compression plant and would ideally be installed adjacent to the preheater tower and kiln on the side closest to the precalciner. This study has estimated that the space required for a 1 Mt/y cement plant would be approximately 0.5 ha.

- A capture–ready oxy-combustion plant would need to make allowance for the additional power requirement by ensuring that a suitable grid connection was available. This study has estimated that a 1 Mt/y cement plant employing CO_2 capture in the precalciner would require an additional installed electrical supply capacity of approximately 15 MW.

10.3 Overall

- The most significant air in-leaks at a cement plant occur in the raw mill, preheater and kiln. Technological developments to reduce these air in-leaks should be encouraged as they will improve the efficiency and costs of carbon capture using either oxy-combustion or post-combustion methods.

- Auxiliary power for post-combustion capture should be met by combustion of a low sulphur fuel as this will reduce the desulphurisation costs associated with the flue gas from the energy plant.

- Combustion of biomass to generate the auxiliary power for post-combustion will reduce the carbon footprint of the plant.

- Until more research and development has been undertaken on the oxy-combustion solution for cement production, post-combustion capture will remain the lowest technical risk capture solution.

- For the case considered in this study, oxy-combustion capture at cement plants offers the lowest cost solution for capturing CO_2 at new-build cement plants. However, there are a number of fundamental technical issues which will need to be further investigated for this option to be successfully deployed.

- This study has focussed on oxy-combustion capture within the precalciner only. The option for the whole plant to operate in an oxy-combustion mode should also be pursued particularly if higher capture rates are required.

- The option of co-locating a post-combustion capture (or oxy-combustion capture) cement plant with a post-combustion capture (or oxy-combustion capture) power station offers significant cost-saving opportunities.

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Appendix A Coal Specification

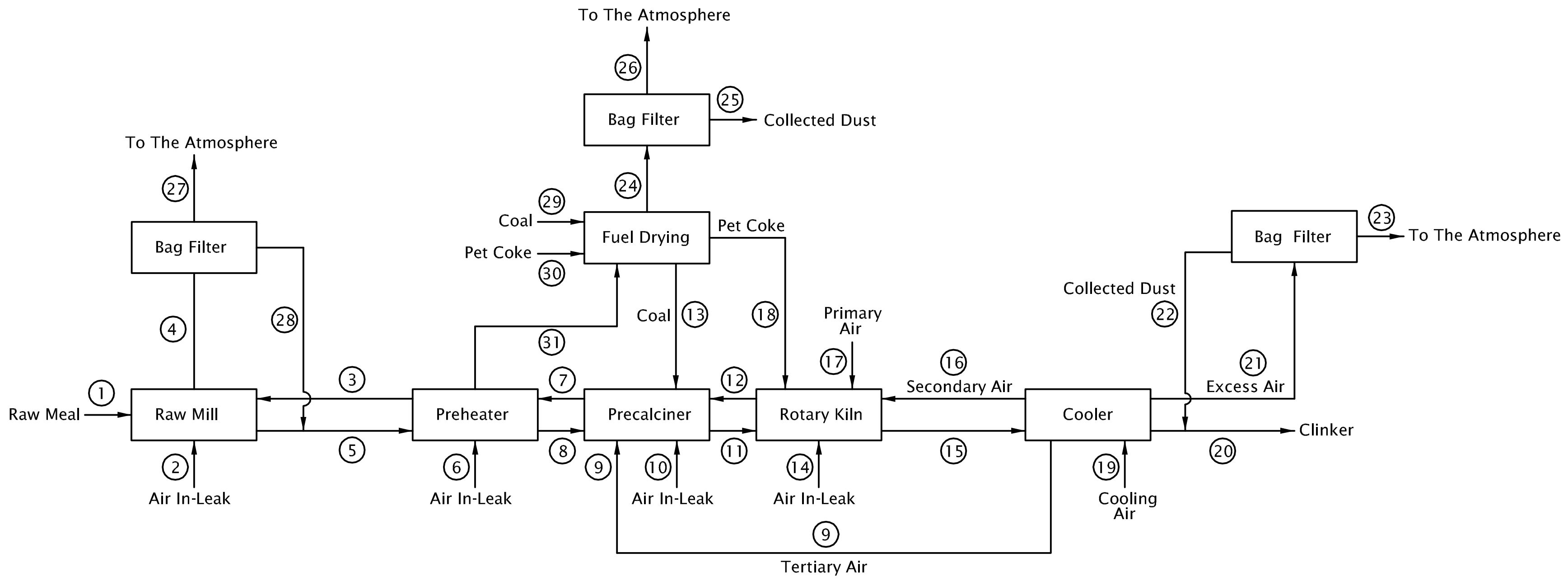
Proximate analysis:	**weight %**
Coal (dry, ash-free)	78.3
Ash	12.2
Moisture	9.5
Ultimate analysis:	**weight %**
Carbon	82.5
Hydrogen	5.6
Oxygen	9.0
Nitrogen	1.8
Sulphur	1.1
Chlorine	0.03
Ash analysis:	**weight %**
SiO_2	50.0
Al_2O_3	30.0
TiO_2	2.0
Fe_2O_3	9.7
CaO	3.9
MgO	0.4
Na_2O	0.1
K_2O	0.1
P_2O_5	1.7
SO_3	1.7
Gross CV (HHV)	27.06 MJ/kg
Net CV (LHV)	25.87 MJ/kg
Hardgrove Index	45
Ash fusion point (reducing atmosphere)	1350°C

The coal specification is based on an open-cut coal from Eastern Australia.

Appendix B Petroleum Coke Specification

Proximate analysis:	**weight %**
Petroleum coke (dry, ash-free)	97.6
ash	0.4
moisture	2.0
Ultimate analysis:	**weight %**
Carbon	87.6
Hydrogen	3.7
Oxygen	1.7
Nitrogen	1.5
Sulphur	5.5
Net CV (LHV)	34.16 MJ/kg

Appendix C PFD and Stream Compositions for Base Case Option



Notes

Key to symbols

Rev	Date	Drawn	Description	Ch'k'd	App'd
P1	13.02.08	AK	Preliminary Issue		

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Client

International Energy Association
Greenhouse Gas R&D Programme
(IEA GHG) CO2 Capture in the Cement Industry

Title

Base Case Cement Plant

Designed			Eng.Chk.	
Drawn	AK		Coordination	
Dwg.Chk.			Approved	

Scale: NTS

Project: 234996

CAD file: 234996-BC-FS01

Status: Pre

Drawing No: 234996/BC/FS01

Rev: P1

CO2 capture in the cement industry

Post-combustion

Refer to drawing no. 234996/PC/FS02

Spreadsheet designed by Simon Turner 13/06/2007
Spreadsheet reviewed by Pete Lilley 17/08/2007
Spreadsheet checked by Duncan Barker 21/11/2007

Mass flowrates

STREAM NUMBER		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31
STREAM NAME		**Raw meal**	**Air Inleak (raw mill)**	**Preheater Exit Gas to Raw Mill**	**Raw Mill Exit Gas**	**Raw Meal Exit**	**Air Inleak (Preheater)**	**Precalciner Exit Gas**	**Preheated Raw Meal**	**Tertiary Air**	**Air Inleaks to precalciner**	**Precalcined raw meal**	**Rotary Kiln Exit Gas**	**Dry Coal**	**Air Inleaks to Kiln**	**Clinker**	**Secondary Air**	**Primary Air**	**Dry Pet Coke**	**Cooling Air**	**Cooled Clinker**	**Excess Air**	**Collected Dust**	**Air**	**Exit Gas from Fuel Drying**	**Collected Dust post FD**	**Bag Filter Exit Gas post FD**	**Bag Filter Exit Gas post raw mill**	**Collected Dust post raw mill**	**Wet Coal**	**Wet Pet Coke**	**Preheater Exit Gas to Fuel Drying**
COMPONENT		**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**
Carbon dioxide	kg/s		0.01	22.99	22.99		0.00	25.54		0.01	0.00		4.35		0.00		0.01	0.00		0.03		0.02		0.02	2.55		2.55	22.99				2.55
Hydrogen	kg/s													0.10					0.04											0.10	0.04	0.20
Nitrogen	kg/s		11.99	27.94	39.93		3.12	27.93		13.70	3.61		10.61	0.03	1.20		8.28	1.13	0.02	50.79		28.80		28.80	3.10		3.10	39.93		0.03	0.02	3.10
Oxygen	kg/s		3.69	1.77	5.46		0.96	1.10		4.21	1.11		0.31	0.16	0.37		2.55	0.35	0.02	15.61		8.85		8.85	0.20		0.20	5.46		0.16	0.02	0.20
Argon	kg/s		0.20	0.47	0.68		0.05	0.47		0.23	0.06		0.18		0.02		0.14	0.02		0.86		0.49		0.49	0.05		0.05	0.68				0.05
Carbon	kg/s													1.43					0.98											1.43	0.98	
Nitrogen dioxide	kg/s			0.14	0.14			0.16					0.05												0.02		0.02	0.14				0.02
Sulphur dioxide	kg/s			0.17	0.06	0.11		0.00	0.11			0.15	0.00			0.27					0.27				0.02		0.02	0.06				0.02
Raw meal	kg/s	48.06				50.38			60.61			35.60																				
Clinker	kg/s															29.98					28.48											
Sulphur	kg/s	0.16				0.16			0.06			0.06		0.02		0.06			0.06		0.06									0.02	0.06	
Ammonia	kg/s																															
Limestone	kg/s																															
Gypsum	kg/s																															
MEA	kg/s																															
Ash	kg/s											0.27		0.27		0.28			0.01		0.28									0.27	0.01	
Dust	kg/s			3.28	0.96	0.96		13.88	0.96	0.40		1.37	4.79	0.01		1.98	0.24				2.83	0.85	0.85	0.00	0.01	0.01	0.00	0.00	0.96		0.01	0.36
Total Dry Flow	kg/s	48.22	15.88	56.77	70.22	51.62	4.13	69.08	61.75	18.56	4.79	37.46	20.30	2.02	1.59	32.57	11.22	1.50	1.13	67.29	31.92	39.02	0.85	38.17	5.96	0.01	5.94	69.25	0.96	2.01	1.13	6.50
Water	kg/s	4.19	0.07	1.78	5.52	0.52	0.02	1.43		0.08	0.02		0.44	0.02	0.01		0.05	0.01	0.01	0.31		0.17		0.17	0.40		0.40	5.52		0.21	0.02	0.20
Phase	-	Solid	Gas	Gas	Gas	Solid	Gas	Gas	Solid	Gas	Gas	Solid	Gas	Solid	Gas	Solid	Gas	Gas	Solid	Gas	Solid	Gas	Solid	Gas	Gas	Solid	Gas	Gas	Solid	Solid	Solid	Gas
Temperature	°C	9	9	330	110	100	9	900	780	908	9	870	1025	100	9	1350	1025	9	100	9	59	279	279	279	130	130	130	110	110	9	9	330
Pressure	bar(a)	-	1.013	<1.013	<1.013	-	1.013	<1.013	-	1.013	1.013	-	<1.013	-	1.013	-	1.013	1.013	-	1.013	-	1.013	-	1.013	<1.013	-	<1.013	<1.013	-	-	-	<1.013
Total Wet flow	kg/s	52.41	15.96	58.54	75.74	52.14	4.15	70.51	61.75	18.64	4.81	37.46	20.73	2.04	1.60	32.57	11.27	1.50	1.14	67.60	31.92	39.19	0.85	38.34	6.36	0.01	6.34	74.77	0.96	2.22	1.15	6.70
Total Stream Enthaply	MJ/hr	-2327796	-4644	-1059804	-1079820	-2242116	-1332	-1253196	-2455236	57456	-1260	-1448100	-276588	-8028	-576	-1166616	32616	-576	-2052	-19944	-1280160	-5616	-12384	22752	-105588	-18	-105588	-1038384	-41436	-16668	-1116	-
Total Stream Enthaply	MJ/s	-646.61	-1.29	-294.39	-299.95	-622.81	-0.37	-348.11	-682.01	15.96	-0.35	-402.25	-76.83	-2.23	-0.16	-324.06	9.06	-0.16	-0.57	-5.54	-355.6	-1.56	-3.44	6.32	-29.33	-0.005	-29.33	-288.44	-11.51	-4.63	-0.31	-

Note

For molar flows and heat capacities the clinker is assumed to be 100% CaO
For molar flows and heat capacities the raw meal is assumed to be 100% limestone
For molar flows and heat capacities the Ash is assumed to be 100% SiO_2
For molar flows and heat capacities the Coal is assumed to be carbon
For molar flows and heat capacities the Pet Coke is assumed to be carbon
For molar flows and heat capacities the Dust in streams 3, 4, 5, 8 and 28 are assumed to be 100% $CaCO_2$
For molar flows and heat capacities the Dust in streams 7, 9, 11, 12, 15, 16, 20, 21 and 22 are assumed to be 100% CaO
For molar flows and heat capacities the Dust in streams 13, 24, 25 and 28 are assumed to be 100% Coal
The enthalpy for all liquid and gaseous components was calculated in CHEMCAD using SRK equation of state method.
The enthalpy for all solid components was calculated in CHEMCAD using latent heat method.
Solid components were: coal, pet coke, calcium carbonate, calcium oxide, ash

CO2 capture in the cement industry

Post-combustion

Refer to drawing no. 234996/PC/FS02

Spreadsheet designed by Simon Turner 13/06/2007
Spreadsheet reviewed by Pete Lilley 17/08/2007
Spreadsheet checked by Duncan Barker 21/11/2007

Mass compositions

STREAM NUMBER		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31
STREAM NAME		**Raw meal**	**Air Inleak (raw mill)**	**Preheater Exit Gas to Raw Mill**	**Raw Mill Exit Gas**	**Raw Meal Exit**	**Air Inleak (Preheater)**	**Precalciner Exit Gas**	**Preheated Raw Meal**	**Tertiary Air**	**Air Inleaks to precalciner**	**Precalcined raw meal**	**Rotary Kiln Exit Gas**	**Dry Coal**	**Air Inleaks to Kiln**	**Clinker**	**Secondary Air**	**Primary Air**	**Dry Pet Coke**	**Cooling Air**	**Cooled Clinker**	**Excess Air**	**Collected Dust**	**Air**	**Exit Gas from Fuel Drying**	**Collected Dust post FD**	**Bag Filter Exit Gas post FD**	**Bag Filter Exit Gas post raw mill**	**Collected Dust post raw mill**	**Wet Coal**	**Wet Pet Coke**	**Preheater Exit Gas to Fuel Drying**
COMPONENT		**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**	**% (w/w)**
Carbon dioxide	% (w/w) dry		0.05	40.49	32.75		0.05	36.97		0.04	0.05		21.45		0.05		0.04	0.05		0.05		0.04		0.05	42.88		42.98	33.20				39.26
Hydrogen	% (w/w) dry													4.82					3.68											4.85	3.67	3.04
Nitrogen	% (w/w) dry		75.47	49.22	56.86		75.47	40.43		73.83	75.47		52.28	1.55	75.47		73.83	75.47	1.49	75.47		73.83		75.47	52.11		52.24	57.65		1.56	1.49	47.72
Oxygen	% (w/w) dry		23.20	3.12	7.77		23.20	1.60		22.70	23.20		1.53	7.75	23.20		22.70	23.20	1.69	23.20		22.70		23.20	3.30		3.31	7.88		7.79	1.69	3.02
Argon	% (w/w) dry		1.28	0.83	0.96		1.28	0.69		1.25	1.28		0.89		1.28		1.25	1.28		1.28		1.25		1.28	0.88		0.89	0.98				0.81
Carbon	% (w/w) dry													71.07					87.16											71.38	86.86	
Nitrogen dioxide	% (w/w) dry			0.25	0.20			0.23					0.27												0.26		0.26	0.20				0.24
Sulphur dioxide	% (w/w) dry			0.30	0.08	0.22		0.00	0.18			0.40	0.00			0.83					0.85				0.32		0.32	0.08				0.29
Raw meal	% (w/w) dry	99.67				97.61			98.15			95.05																				
Clinker	% (w/w) dry															92.05					89.23											
Sulphur	% (w/w) dry	0.33				0.31			0.10			0.17		0.95		0.20			5.47		0.20									0.95	5.32	
Ammonia	% (w/w) dry																															
Limestone	% (w/w) dry																															
Gypsum	% (w/w) dry																															
MEA	% (w/w) dry																															
Ash	% (w/w) dry											0.72		13.42		0.85			0.50		0.87									13.48	0.50	
Dust	% (w/w) dry			5.79	1.37	1.87		20.09	1.56	2.18		3.65	23.59	#REF!		6.07	2.18				8.86	2.18	100.00	0.00	0.24	100.00	0.00	0.00	100.00		0.48	5.61
Total	% (w/w) dry	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	#REF!	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
Water	% (w/w) wet	8.00	0.46	3.04	7.29	1.01	0.46	2.03		0.45	0.46		2.10	1.00	0.46		0.45	0.41	0.97	0.46		0.44		0.45	6.29		6.30	7.38		9.50	1.94	2.95

CO2 capture in the cement industry

Post-combustion

Refer to drawing no. 234996/PC/FS02

Spreadsheet designed by Simon Turner 13/06/2007
Spreadsheet reviewed by Pete Lilley 17/08/2007
Spreadsheet checked by Duncan Barker 21/11/2007

Molar flowrates (dry)

STREAM NUMBER		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31
STREAM NAME		**Raw meal**	**Air Inleak (raw mill)**	**Preheater Exit Gas to Raw Mill**	**Raw Mill Exit Gas**	**Raw Meal Exit**	**Air Inleak (Preheater)**	**Precalciner Exit Gas**	**Preheated Raw Meal**	**Tertiary Air**	**Air Inleaks to precalciner**	**Precalcined raw meal**	**Rotary Kiln Exit Gas**	**Dry Coal**	**Air Inleaks to Kiln**	**Clinker**	**Secondary Air**	**Primary Air**	**Dry Pet Coke**	**Cooling Air**	**Cooled Clinker**	**Excess Air**	**Collected Dust**	**Air**	**Exit Gas from Fuel Drying**	**Collected Dust post FD**	**Bag Filter Exit Gas post FD**	**Bag Filter Exit Gas post raw mill**	**Collected Dust post raw mill**	**Wet Coal**	**Wet Pet Coke**	**Preheater Exit Gas to Fuel Drying**
COMPONENT	**MW (g/mol)**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**
Carbon dioxide	44.01		0.17	522.30	522.47		0.04	580.29		0.19	0.05		98.91		0.02		0.11	0.02		0.70		0.40		0.40	58.03		58.03	522.47				58.03
Hydrogen (H_2)	2.02													48.30					20.58											48.30	20.58	97.94
Nitrogen (N_2)	28.01		427.97	997.32	1425.29		111.27	996.86		489.07	128.98		378.80	1.12	42.80		295.68	40.32	0.60	1813.01		1028.25		1028.25	110.81		110.81	1425.29		1.12	0.60	110.81
Oxygen (O_2)	32.00		115.17	55.33	170.50		29.94	34.51		131.62	34.71		9.72	4.89	11.52		79.57	10.85	0.60	487.91		276.72		276.72	6.15		6.15	170.50		4.89	0.60	6.15
Argon	39.95		5.09	11.86	16.95		1.32	11.86		5.82	1.53		4.51		0.51		3.52	0.48		21.56		12.23		12.23	1.32		1.32	16.95				1.32
Carbon	12.01													119.44					81.77											119.44	81.77	
Nitrogen dioxide	46.01			3.06	3.06			3.41					1.17												0.34		0.34	3.06				0.34
Sulphur dioxide	64.06			2.68	0.91	1.77		0.00	1.77			2.37	0.00			4.24					4.24				0.30		0.30	0.91				0.30
Raw meal	100.90	476.28				499.33			600.72			352.85																				
Clinker	56.08															534.54					507.81											
Sulphur	32.06	4.96				4.96			1.99			1.99		0.60		1.99			1.92		1.99									0.60	1.88	
Ammonia	17.03																															
Limestone	100.90																															
Gypsum	172.17																															
MEA	61.08																															
Ash	88.64											3.06		3.06		3.12			0.06		3.12									3.06	0.06	
Dust	100.90			32.56	9.56	9.55		137.56	9.55	4.01		13.56	47.45	#REF!		19.60	2.42				28.01	8.42	8.42	0.01	0.14	0.14	0.00	0.01	9.55	0.09	0.05	3.62
Total dry molar flow	mol/s	481.24	548.39	1625.12	2148.74	515.61	142.58	1764.49	614.02	630.71	165.27	373.81	540.55	#REF!	54.84	563.48	381.31	51.67	105.53	2323.18	545.17	1326.02	8.42	1317.60	177.09	0.14	176.95	2139.19	9.55	177.49	105.54	278.51
Water	18.015	232.73	4.05	98.64	306.33	29.09	1.05	79.46		4.63	1.21		24.21	1.13	0.40		2.79	0.34	0.61	17.08		9.66		9.66	22.18		22.18	306.33		11.71	1.24	10.96
Total wet molar flow	mol/s	713.98	552.44	1723.76	2455.07	544.70	143.64	1843.94	614.02	635.33	166.49	373.81	564.76	#REF!	55.24	563.48	384.09	52.01	106.15	2340.26	545.17	1335.68	8.42	1327.27	199.27	0.14	199.13	2445.52	9.55	189.20	106.78	289.47
Molecular weight	g/mol	73.40	28.89	33.96	30.85	95.73	28.89	38.24	100.57	29.34	28.89	100.20	36.71	#REF!	28.89	57.79	29.34	28.89	10.72	28.89	58.54	29.34	100.90	28.89	31.90	100.90	31.85	30.58	100.90	11.74	10.80	23.15

CO2 capture in the cement industry

Post-combustion

Refer to drawing no. 234996/PC/FS02

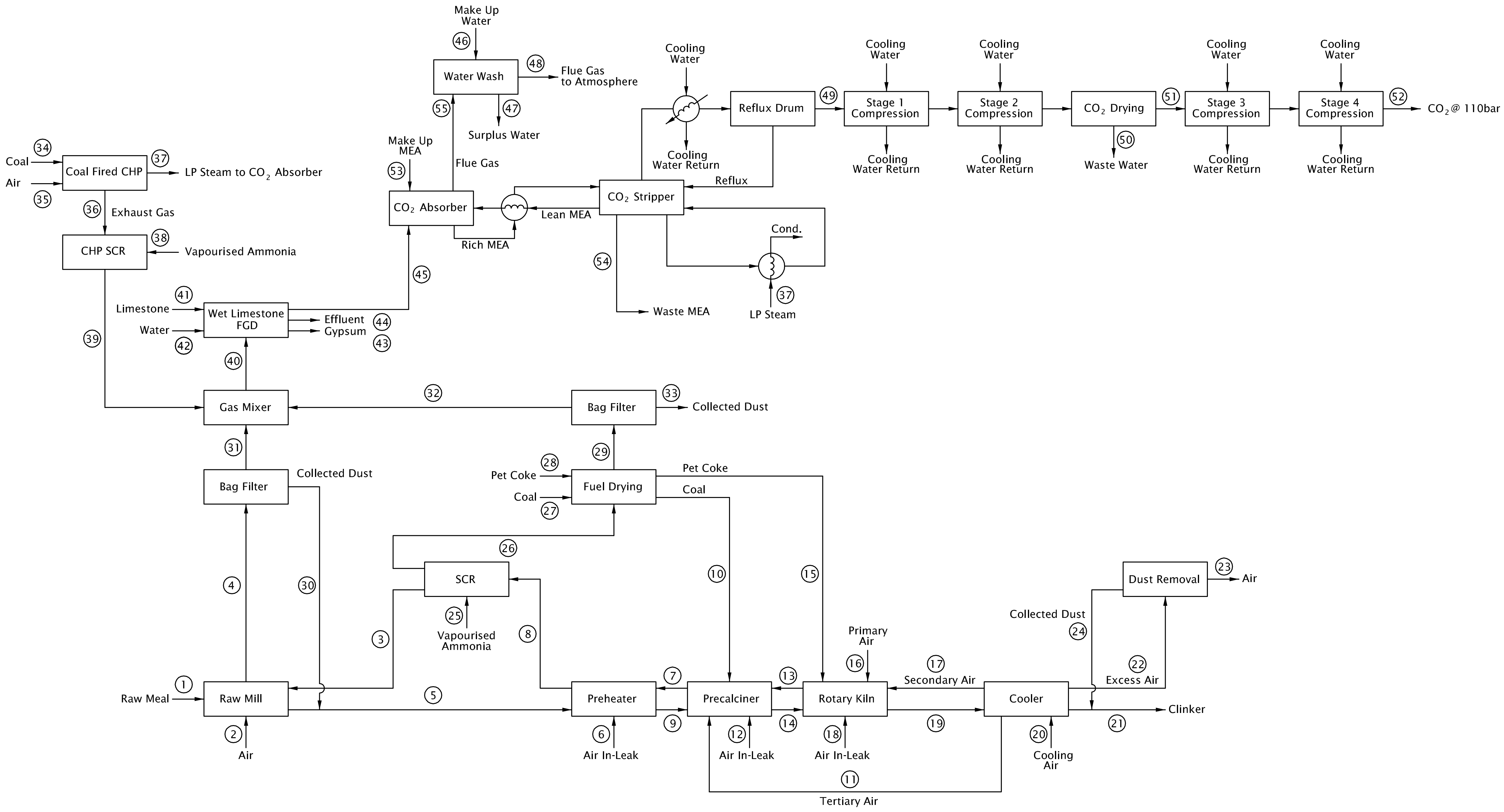
Spreadsheet designed by	Simon Turner	13/06/2007
Spreadsheet reviewed by	Pete Lilley	17/08/2007
Spreadsheet checked by	Duncan Barker	21/11/2007

Molar Compositions (dry)

STREAM NUMBER		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
STREAM NAME		Raw meal	Air Inleak (raw mill)	Preheater Exit Gas to Raw Mill	Raw Mill Exit Gas	Raw Meal Exit	Air Inleak (Preheater)	Precalciner Exit Gas	Preheated Raw Meal	Tertiary Air	Air Inleaks to precalciner	Precalcined raw meal	Rotary Kiln Exit Gas	Dry Coal	Air Inleaks to Kiln	Clinker
COMPONENT		mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)
Carbon dioxide	mol% (dry)		0.03	32.14	24.32		0.03	32.89		0.03	0.03		18.30		0.03	
Hydrogen	mol% (dry)													#REF!		
Nitrogen	mol% (dry)		78.04	61.37	66.33		78.04	56.50		77.54	78.04		70.08	#REF!	78.04	
Oxygen	mol% (dry)		21.00	3.40	7.93		21.00	1.96		20.87	21.00		1.80	#REF!	21.00	
Argon	mol% (dry)		0.93	0.73	0.79		0.93	0.67		0.92	0.93		0.83		0.93	
Carbon	mol% (dry)													#REF!		
Nitrogen dioxide	mol% (dry)			0.19	0.14			0.19					0.22			
Sulphur dioxide	mol% (dry)			0.16	0.04	0.34		0.00	0.29			0.63	0.00			0.75
Raw meal	mol% (dry)	98.97				96.84			97.83			94.39				
Clinker	mol% (dry)															94.86
Sulphur	mol% (dry)	1.03				0.96			0.32			0.53		#REF!		0.35
Ammonia	mol% (dry)															
Limestone	mol% (dry)															
Gypsum	mol% (dry)															
MEA	mol% (dry)															
Ash	mol% (dry)											0.82		#REF!		0.55
Dust	mol% (dry)			2.00	0.44	1.85		7.80	1.55	0.64		3.63	8.78	#REF!		3.48
Total	mol% (dry)	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	#REF!	100.00	100.00

STREAM NUMBER		16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31
STREAM NAME		Secondary Air	Primary Air	Dry Pet Coke	Cooling Air	Cooled Clinker	Excess Air	Collected Dust	Air	Exit Gas from Fuel Drying	Collected Dust post FD	Bag Filter Exit Gas post FD	Bag Filter Exit Gas post raw mill	Collected Dust post raw mill	Wet Coal	Wet Pet Coke	Preheater Exit Gas to Fuel Drying
COMPONENT		mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)
Carbon dioxide	mol% (dry)	0.03	0.03		0.03		0.03		0.03	32.77		32.80	24.42				20.84
Hydrogen	mol% (dry)			19.50											27.21	19.50	35.17
Nitrogen	mol% (dry)	77.54	78.04	0.57	78.04		77.54		78.04	62.57		62.62	66.63		0.63	0.57	39.79
Oxygen	mol% (dry)	20.87	21.00	0.56	21.00		20.87		21.00	3.47		3.47	7.97		2.76	0.56	2.21
Argon	mol% (dry)	0.92	0.93		0.93		0.92		0.93	0.74		0.74	0.79				0.47
Carbon	mol% (dry)			77.49											67.30	77.48	
Nitrogen dioxide	mol% (dry)									0.19		0.19	0.14				0.12
Sulphur dioxide	mol% (dry)					0.78				0.17		0.17	0.04				0.11
Raw meal	mol% (dry)																
Clinker	mol% (dry)					93.15											
Sulphur	mol% (dry)			1.82		0.36									0.34	1.78	
Ammonia	mol% (dry)																
Limestone	mol% (dry)																
Gypsum	mol% (dry)																
MEA	mol% (dry)																
Ash	mol% (dry)			0.06		0.57									1.72	0.06	
Dust	mol% (dry)	0.64				5.14	0.64	100.00	0.00	0.08	100.00	0.00	0.00	100.00	0.05	0.05	1.30
Total	mol% (dry)	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00

Appendix D PFD and Stream Compositions for Post-Combustion Option



Notes

Key to symbols

Rev	Date	Drawn	Description	Ch'k'd	App'd
P1	11.01.08	AK	Preliminary Issue		

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Client

International Energy Association
Greenhouse Gas R&D Programme
(IEA GHG) CO2 Capture in the Cement Industry

Title

Post - Combustion CO_2 Capture Cement Plant

Designed			Eng.Chk.	
Drawn	AK		Coordination	
Dwg.Chk.			Approved	

Scale
NTS

Project 234996
CAD file 234996-PC-FS02

Status
Pre

Drawing No
234996/PC/FS02

Rev
P1

CO2 capture in the cement industry

Post-combustion

Refer to drawing no. 234996/PC/FS02

Spreadsheet designed by	Simon Turner	13/06/2007
Spreadsheet reviewed by	Pete Lilley	17/08/2007
Spreadsheet checked by	Duncan Barker	21/11/2007

Mass flowrates

STREAM NUMBER		**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**	**27**
STREAM NAME		**Raw meal**	**Air Inleak (raw mill)**	**SCR Exit Gas**	**Raw Mill Exit Gas**	**Raw Meal Exit**	**Air Inleak (Preheater)**	**Precalciner Exit Gas**	**Preheater Exit Gas**	**Preheated Raw Meal**	**Dry Coal**	**Tertiary Air**	**Air Inleaks to precalciner**	**Rotary Kiln Exit Gas**	**Precalcined raw meal**	**Dry Pet Coke**	**Primary Air**	**Secondary Air**	**Air Inleaks to Kiln**	**Clinker**	**Cooling Air**	**Cooled Clinker**	**Excess Air**	**Air**	**Collected Dust**	**Vapourised Ammonia**	**SCR Exit Gas**	**Wet Coal**
COMPONENT		**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**
Carbon dioxide	kg/s		0.01	22.36	22.37		0.00	25.54	25.54			0.01	0.00	4.35			0.00	0.01	0.00		0.03		0.02	0.02		0.00	2.55	
Hydrogen	kg/s										0.10					0.04												0.10
Nitrogen	kg/s		11.99	28.24	40.23		3.12	27.93	31.04		0.03	13.70	3.61	10.61		0.02	1.13	8.28	1.20		50.79		28.81	28.81		1.10	3.23	0.03
Oxygen	kg/s		3.69	2.02	5.70		0.96	1.10	1.97		0.16	4.21	1.11	0.31		0.02	0.35	2.55	0.37		15.61		8.86	8.86		0.34	0.23	0.16
Argon	kg/s		0.20	0.48	0.68		0.05	0.47	0.53			0.23	0.06	0.18			0.02	0.14	0.02		0.86		0.49	0.49		0.02	0.05	
Carbon	kg/s										1.43					0.98												1.43
Nitrogen dioxide	kg/s			0.00	0.00			0.16	0.16					0.05													0.00	
Sulphur dioxide	kg/s			0.17	0.06	0.11		0.00	0.19	0.11				0.00	0.15					0.27		0.27					0.02	
Raw meal	kg/s	48.06				50.39				60.62					35.61													
Clinker	kg/s																			29.98		28.48						
Sulphur	kg/s	0.16				0.16				0.06	0.02				0.06	0.06				0.06		0.06						0.02
Ammonia	kg/s			0.00																						0.08	0.00	
Limestone	kg/s																											
Gypsum	kg/s																											
MEA	kg/s																											
Ash	kg/s										0.27				0.27	0.01				0.28		0.28						0.27
Dust	kg/s			3.29	0.96	0.96		13.88	3.65	0.96		0.40		4.79	1.37			0.24		1.98		2.83	0.85	0.00	0.85			
Total Dry Flow	kg/s	48.22	15.88	56.55	70.00	51.62	4.13	69.08	63.07	61.76	2.01	18.56	4.79	20.30	37.46	1.13	1.50	11.22	1.59	32.57	67.30	31.92	39.02	38.17	0.85	1.54	6.08	2.01
Water	kg/s	4.19	0.07	1.84	5.58	0.52	0.02	1.43	1.97		0.02	0.08	0.02	0.44		0.01	0.01	0.05	0.01		0.31		0.17	0.17		0.01	0.21	0.21
Phase	-	Solid	Gas	Gas	Gas	Solid	Gas	Gas	Gas	Solid	Solid	Gas	Gas	Gas	Solid	Solid	Gas	Gas	Gas	Solid	Gas	Solid	Gas	Gas	Solid	Liquid	Gas	Solid
Temperature	°C	9	9	310	110	100	9	900	330	780	100	908	9	1025	870	100	9	1025	9	1350	9	59	279	279	279	9	310	9
Pressure	bar(a)	-	1.013	<1.013	<1.013	-	1.013	<1.013	<1.013	-	-	1.013	1.013	<1.013	-	-	1.013	1.013	1.013	-	1.013	-	1.013	1.013	-	10	<1.013	-
Total Wet flow	kg/s	52.42	15.96	58.39	75.59	52.15	4.15	70.51	65.05	61.76	2.03	18.64	4.81	20.73	37.46	1.14	1.50	11.27	1.60	32.57	67.61	31.92	39.20	38.35	0.85	1.55	6.29	2.22
Total Stream Enthaply	MJ/hr	-2332137.6	-4644	-940716	-1061748	-2245968	-1224	-1253952	-1060236	-2458440	-8028	57456	-1260	-276588	-1450548	-2052	-576	-13608	-576	-1169388	-19944	-1283256	-7416	22464	-28368	-1548	-91728	-16668
Total Stream Enthaply	MJ/s	-647.816	-1.29	-261.31	-294.93	-623.88	-0.34	-348.32	-294.51	-682.9	-2.23	15.96	-0.35	-76.83	-402.93	-0.57	-0.16	-3.78	-0.16	-324.83	-5.54	-356.46	-2.06	6.24	-7.88	-0.43	-25.48	-4.63

Mass flowrates

STREAM NUMBER		**28**	**29**	**30**	**31**	**32**	**33**	**34**	**35**	**36**	**37**	**38**	**39**	**40**	**41**	**42**	**43**	**44**	**45**	**46**	**47**	**48**	**49**	**50**	**51**	**52**	**53**	**54**
STREAM NAME		**Wet Pet Coke**	**Exit Gas from Fuel Drying**	**Collected Dust post raw mill**	**Bag Filter Exit Gas post raw mill**	**Bag Filter Exit Gas post FD**	**Collected Dust post FD**	**Coal for CHP**	**Air**	**CHP Exit Gas**	**LP Steam**	**Vapourised Ammonia**	**CHP SCR Exit Gas**	**Gas Mixer Exit Gas**	**Limestone**	**Water**	**Gypsum from FGD**	**Effluent from FGD**	**Exit Gas from FGD**	**Make up water**	**Surplus Water**	**Flue Gas to Atmosphere**	**CO2 to drying**	**Waste water from drying**	**Dry CO2**	**Compressed CO2**	**Make Up MEA**	**Waste MEA**
COMPONENT		**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**	**kg/s**
Carbon dioxide	kg/s		2.55		22.37	2.55			0.03	18.99		0.00	18.99	43.91					44.06			6.61	37.45		37.45	37.45		0.00
Hydrogen	kg/s	0.04						0.35																				
Nitrogen	kg/s	0.02	3.23		40.23	3.23		0.11	53.13	53.13		2.61	56.00	99.45					99.45			99.45						
Oxygen	kg/s	0.02	0.23		5.70	0.23		0.56	16.33	0.00		0.80	0.80	6.74					6.68			6.68						
Argon	kg/s		0.05		0.68	0.05			0.90	0.90		0.04	0.95	1.68					1.68			1.68						
Carbon	kg/s	0.98						5.17																				
Nitrogen dioxide	kg/s		0.00		0.00	0.00				0.37			0.00	0.00					0.00			0.00						
Sulphur dioxide	kg/s		0.02		0.06	0.02				0.14			0.14	0.21					0.00			0.00						
Raw meal	kg/s																											
Clinker	kg/s																											
Sulphur	kg/s	0.06						0.07																				
Ammonia	kg/s		0.00			0.00				0.00		0.18		0.00					0.00			0.00						
Limestone	kg/s														0.45			0.11										
Gypsum	kg/s																0.57											
MEA	kg/s																										0.08	0.08
Ash	kg/s	0.01						0.98		0.98																		
Dust	kg/s	0.01	0.01	0.96	0.00	0.00	0.01							0.00					0.00			0.00						
Total Dry Flow	kg/s	1.13	6.10	0.96	69.04	6.08	0.01	7.25	70.39	74.50	0.00	3.64	76.88	152.00	0.45	0.00	0.57	0.11	151.88	0.00	0.00	114.43	37.45	0.00	37.45	37.45	0.08	0.08
Water	kg/s	0.02	0.41		5.58	0.41		0.76	0.32	4.22	56.17	0.02	4.53	10.52		4.04		3.92	10.52	108.81	107.23	11.58	0.53	0.53	0.00	0.00	0.25	0.25
Phase	-	Solid	Gas	Solid	Gas	Gas	Solid	Solid	Gas	Gas	Gas	Liquid	Gas	Gas	Solid	Liquid	Solid	Liquid	Gas	Liquid	Liquid	Gas	Gas	Liquid	Gas	Gas	Liquid	Liquid
Temperature	°C	9	130	110	110	130	130	100	9	400	140	9	310	140	9	9	59	50	50	9	27	27	27	27	27	43	9	45
Pressure	bar(a)	-	<1.013	-	<1.013	<1.013	-	-	1.013	1.013	3.5	10	1.013	1.013	-	-	-	-	1.013	1.38	2.76	1.013	1.62	1.013	12	110	-	-
Total Wet flow	kg/s	1.15	6.51	0.96	74.62	6.50	0.01	8.01	70.71	78.72	56.17	3.66	81.40	162.52	0.45	4.04	0.57	4.03	162.40	108.81	107.23	126.00	37.97	0.53	37.45	37.45	0.34	0.34
Total Stream Enthaply	MJ/hr	-1116	-106092	-41436	-1020312	-106056	-18	-62640	-20628	-725220	-2777688	-3132	-748728	-1920024	-19548	-233028	-21960	-228492	-1979136	-6532452	-6409044	-877716	-1277280	-30852	-1250928	-1274688	-1260	-1224
Total Stream Enthaply	MJ/s	-0.31	-29.47	-11.51	-283.42	-29.46	-0.005	-17.4	-5.73	-201.45	-771.58	-0.87	-207.98	-533.34	-5.43	-64.73	-6.1	-63.47	-549.76	-1814.57	-1780.29	-243.81	-354.8	-8.57	-347.48	-354.08	-0.35	-0.34

Note

For molar flows and heat capacities the clinker is assumed to be 100% CaO
For molar flows and heat capacities the raw meal is assumed to be 100% limestone
For molar flows and heat capacities the Ash is assumed to be 100% SiO_2
For molar flows and heat capacities the Coal is assumed to be carbon
For molar flows and heat capacities the Pet Coke is assumed to be carbon
For molar flows and heat capacities the Dust in streams 3, 4, 5, 8, 9 and 30 are assumed to be 100% $CaCO_2$
For molar flows and heat capacities the Dust in streams 7, 11, 13, 14, 17, 19, 21, 22 and 24 are assumed to be 100% CaO
For molar flows and heat capacities the Dust in streams 28, 29 and 33 are assumed to be 100% Coal
The enthalpy for all liquid and gaseous components was calculated in CHEMCAD using SRK equation of state method.
The enthalpy for all solid components was calculated in CHEMCAD using latent heat method.
Solid components were: coal, pet coke, calcium carbonate, calcium oxide, ash, gypsum

CO2 capture in the cement industry

Post-combustion

Refer to drawing no. 234996/PC/FS02

Spreadsheet designed by Simon Turner 13/06/2007
Spreadsheet reviewed by Pete Lilley 17/08/2007
Spreadsheet checked by Duncan Barker 21/11/2007

Mass compositions

STREAM NUMBER		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27
STREAM NAME		Raw meal	Air Inleak (raw mill)	SCR Exit Gas	Raw Mill Exit Gas	Raw Meal Exit	Air Inleak (Preheater)	Precalciner Exit Gas	Preheater Exit Gas	Preheated Raw Meal	Dry Coal	Tertiary Air	Air Inleaks to precalciner	Rotary Kiln Exit Gas	Precalcined raw meal	Dry Pet Coke	Primary Air	Secondary Air	Air Inleaks to Kiln	Clinker	Cooling Air	Cooled Clinker	Excess Air	Air	Collected Dust	Vapourised Ammonia	SCR Exit Gas	Wet Coal
COMPONENT		% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)
Carbon dioxide	% (w/w) dry		0.05	39.54	31.95		0.05	36.97	40.49			0.04	0.05	21.45			0.05	0.04	0.05		0.05		0.04	0.05		0.04	41.98	
Hydrogen	% (w/w) dry										4.85					3.68												4.85
Nitrogen	% (w/w) dry		75.47	49.94	57.47		75.47	40.42	49.22		1.56	73.83	75.47	52.28		1.49	75.47	73.83	75.47		75.47		73.83	75.47		71.68	53.02	1.56
Oxygen	% (w/w) dry		23.20	3.57	8.15		23.20	1.60	3.12		7.79	22.70	23.20	1.53		1.69	23.20	22.70	23.20		23.20		22.70	23.20		22.04	3.79	7.79
Argon	% (w/w) dry		1.28	0.84	0.97		1.28	0.69	0.83			1.25	1.28	0.89			1.28	1.25	1.28		1.28		1.25	1.28		1.22	0.90	
Carbon	% (w/w) dry										71.38					87.16												71.38
Nitrogen dioxide	% (w/w) dry			0.00	0.00			0.23	0.25					0.27													0.00	
Sulphur dioxide	% (w/w) dry			0.30	0.08	0.21		0.00	0.30	0.18				0.00	0.40					0.82		0.84					0.31	
Raw meal	% (w/w) dry	99.67				97.61				98.16					95.06													
Clinker	% (w/w) dry																			92.06		89.24						
Sulphur	% (w/w) dry	0.33				0.31				0.10	0.95				0.17	5.47				0.20		0.20						0.95
Ammonia	% (w/w) dry			0.00																						5.02	0.00	
Limestone	% (w/w) dry																											
Gypsum	% (w/w) dry																											
MEA	% (w/w) dry																											
Ash	% (w/w) dry										13.48				0.72	0.50				0.85		0.87						13.48
Dust	% (w/w) dry			5.81	1.38	1.87		20.09	5.79	1.56		2.18		23.59	3.65			2.18		6.07		8.86	2.18	0.00	100.00			
Total	% (w/w) dry	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00
Water	% (w/w) wet	8.00	0.46	3.15	7.39	1.01	0.46	2.03	3.04		1.00	0.45	0.46	2.10		0.97	0.41	0.45	0.46		0.46		0.44	0.45		0.43	3.34	9.50

Mass compositions

STREAM NUMBER		28	29	30	31	32	33	34	35	36	37	38	39	40	41	42	43	44	45	46	47	48	49	50	51	52	53	54
STREAM NAME		Wet Pet Coke	Exit Gas from Fuel Drying	Collected Dust post raw mill	Bag Filter Exit Gas post raw mill	Bag Filter Exit Gas post FD	Collected Dust post FD	Coal for CHP	Air	CHP Exit Gas	LP Steam	Vapourised Ammonia	CHP SCR Exit Gas	Gas Mixer Exit Gas	Limestone	Water	Gypsum from FGD	Effluent from FGD	Exit Gas from FGD	Make up water	Surplus Water	Flue Gas to Atmosphere	CO2 to drying	Waste water from drying	Dry CO2	Compressed CO2	Make Up MEA	Waste MEA
COMPONENT		% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)	% (w/w)
Carbon dioxide	% (w/w) dry		41.88		32.40	41.98			0.05	25.49		0.04	24.70	28.89					29.01			5.78	100.00		100.00	100.00		
Hydrogen	% (w/w) dry	3.67						4.85																				
Nitrogen	% (w/w) dry	1.49	52.90		58.27	53.02		1.56	75.47	71.31		71.68	72.85	65.43					65.48			86.91						
Oxygen	% (w/w) dry	1.69	3.78		8.26	3.79		7.79	23.20	0.00		22.04	1.04	4.43					4.40			5.84						
Argon	% (w/w) dry		0.89		0.99	0.90			1.28	1.21		1.22	1.23	1.11					1.11			1.47						
Carbon	% (w/w) dry	86.86						71.38																				
Nitrogen dioxide	% (w/w) dry		0.00		0.00	0.00				0.50			0.00	0.00					0.00			0.00						
Sulphur dioxide	% (w/w) dry		0.31		0.08	0.31				0.18			0.18	0.14					0.00			0.00						
Raw meal	% (w/w) dry																											
Clinker	% (w/w) dry																											
Sulphur	% (w/w) dry	5.32						0.95																				
Ammonia	% (w/w) dry		0.00			0.00				0.00		5.02		0.00					0.00			0.00						
Limestone	% (w/w) dry														100.00			100.00										
Gypsum	% (w/w) dry																100.00											
MEA	% (w/w) dry																										100.00	100.00
Ash	% (w/w) dry	0.50						13.48		1.31																		
Dust	% (w/w) dry	0.48	0.23	100.00	0.00	0.00	100.00							0.00					0.00			0.00						
Total	% (w/w) dry	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	0.00	100.00	100.00	100.00	100.00	0.00	100.00	100.00	100.00	0.00	0.00	100.00	100.00	0.00	100.00	100.00	100.00	100.00
Water	% (w/w) wet	1.94	6.33		7.48	6.35		9.50	0.46	5.36	100.00	0.43	5.56	6.47		100.00		97.22	6.48	100.00	100.00	9.19	1.39	100.00	0.00	0.00	75.00	75.00

CO2 capture in the cement industry

Post-combustion

Refer to drawing no. 234996/PC/FS02

Spreadsheet designed by Simon Turner 13/06/2007
Spreadsheet reviewed by Pete Lilley 17/08/2007
Spreadsheet checked by Duncan Barker 21/11/2007

Molar flowrates (dry)

STREAM NUMBER		**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**	**27**
STREAM NAME		**Raw meal**	**Air Inleak (raw mill)**	**SCR Exit Gas**	**Raw Mill Exit Gas**	**Raw Meal Exit**	**Air Inleak (Preheater)**	**Precalciner Exit Gas**	**Preheater Exit Gas**	**Preheated Raw Meal**	**Dry Coal**	**Tertiary Air**	**Air Inleaks to precalciner**	**Rotary Kiln Exit Gas**	**Precalcined raw meal**	**Dry Pet Coke**	**Primary Air**	**Secondary Air**	**Air Inleaks to Kiln**	**Clinker**	**Cooling Air**	**Cooled Clinker**	**Excess Air**	**Air**	**Collected Dust**	**Vapourised Ammonia**	**SCR Exit Gas**	**Wet Coal**
COMPONENT	**MW (g/mol)**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**
Carbon dioxide	44.01		0.17	508.06	508.22		0.04	580.29	580.34			0.19	0.05	98.91			0.02	0.11	0.02		0.70		0.40	0.40		0.02	58.04	
Hydrogen (H_2)	2.02										48.30					20.58												48.30
Nitrogen (N_2)	28.01		427.97	1008.07	1436.03		111.27	996.86	1108.14		1.12	489.07	128.98	378.80		0.60	40.32	295.68	42.80		1813.22		1028.46	1028.46		39.40	115.15	1.12
Oxygen (O_2)	32.00		115.17	63.10	178.27		29.94	34.51	61.48		4.89	131.62	34.71	9.72		0.60	10.85	79.57	11.52		487.96		276.77	276.77		10.60	7.21	4.89
Argon	39.95		5.09	11.95	17.04		1.32	11.86	13.18			5.82	1.53	4.51			0.48	3.52	0.51		21.57		12.23	12.23		0.47	1.36	
Carbon	12.01										119.44					81.77												119.44
Nitrogen dioxide	46.01			0.00	0.00			3.41	3.41					1.17													0.00	
Sulphur dioxide	64.06			2.61	0.89	1.72		0.00	2.98	1.72				0.00	2.32					4.19		4.19					0.30	
Raw meal	100.90	476.36				499.41				600.79					352.91													
Clinker	56.08																			534.66		507.92						
Sulphur	32.06	4.96				4.96				1.99	0.60				1.99	1.92				1.99		1.99						0.60
Ammonia	17.03			0.00																						4.54	0.00	
Limestone	100.90																											
Gypsum	172.17																											
MEA	61.08																											
Ash	88.64										3.06				3.06	0.06				3.12		3.12						3.06
Dust	100.90			32.56	9.56	9.55		137.57	36.18	9.55		4.01		47.45	13.56			2.42		19.60		28.02	8.43	0.01	8.42			
Total dry molar flow	mol/s	481.33	548.39	1626.34	2150.01	515.64	142.58	1764.50	1805.69	614.05	177.40	630.71	165.27	540.55	373.83	105.53	51.67	381.31	54.84	563.55	2323.45	545.24	1326.29	1317.87	8.42	55.03	182.06	177.40
Water	18.015	232.77	4.05	102.24	309.97	29.10	1.05	79.46	109.61		1.13	4.63	1.21	24.21		0.61	0.34	2.79	0.40		17.08		9.67	9.67		0.37	11.68	11.71
Total wet molar flow	mol/s	714.10	552.44	1728.58	2459.98	544.74	143.64	1843.95	1915.29	614.05	178.53	635.33	166.49	564.76	373.83	106.15	52.01	384.09	55.24	563.55	2340.53	545.24	1335.95	1327.54	8.42	55.40	193.74	189.12
Molecular weight	g/mol	73.40	28.89	33.78	30.73	95.73	28.89	38.24	33.96	100.57	11.37	29.34	28.89	36.71	100.21	10.72	28.89	29.34	28.89	57.79	28.89	58.54	29.34	28.89	100.90	27.91	32.49	11.74

Molar flowrates (dry)

STREAM NUMBER		**28**	**29**	**30**	**31**	**32**	**33**	**34**	**35**	**36**	**37**	**38**	**39**	**40**	**41**	**42**	**43**	**44**	**45**	**46**	**47**	**48**	**49**	**50**	**51**	**52**	**53**	**54**
STREAM NAME		**Wet Pet Coke**	**Exit Gas from Fuel Drying**	**Collected Dust post raw mill**	**Bag Filter Exit Gas post raw mill**	**Bag Filter Exit Gas post FD**	**Collected Dust post FD**	**Coal for CHP**	**Air**	**CHP Exit Gas**	**LP Steam**	**Vapourised Ammonia**	**CHP SCR Exit Gas**	**Gas Mixer Exit Gas**	**Limestone**	**Water**	**Gypsum from FGD**	**Effluent from FGD**	**Exit Gas from FGD**	**Make up water**	**Surplus Water**	**Flue Gas to Atmosphere**	**CO2 to drying**	**Waste water from drying**	**Dry CO2**	**Compressed CO2**	**Make Up MEA**	**Waste MEA**
COMPONENT	**MW (g/mol)**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**	**mol/s**
Carbon dioxide	0.00		58.04		508.22	58.04			0.74	431.44		0.04	431.48	997.74					1001.07			150.16	850.91		850.91	850.91		
Hydrogen (H_2)	0.00	20.58						174.17																				
Nitrogen (N_2)	0.00	0.60	115.15		1436.03	115.15		4.03	1896.45	1896.45		93.23	1999.08	3550.26					3550.26			3550.26						
Oxygen (O_2)	78.09	0.60	7.21		178.27	7.21		17.63	510.36	0.00		25.09	25.09	210.57					208.90			208.90						
Argon	0.03		1.36		17.04	1.36			22.55	22.55		1.11	23.66	42.07					42.07			42.07						
Carbon	0.93	81.77						430.70																				
Nitrogen dioxide	0.86		0.00		0.00	0.00				8.06			0.00	0.00					0.00			0.00						
Sulphur dioxide	0.00		0.30		0.89	0.30				2.15			2.15	3.34					0.00			0.00						
Raw meal	1.03																											
Clinker	0.91																											
Sulphur	0.00	1.88						2.15																				
Ammonia	0.52		0.00			0.00				0.00		10.74		0.00					0.00			0.00						
Limestone	1.03														4.45			1.11										
Gypsum	0.00																3.34											
MEA	4.19																										1.38	1.38
Ash	0.00	0.06						11.02		11.02																		
Dust	1.03	0.05	0.14	9.55	0.01	0.00	0.14							0.01					0.01			0.01						
Total dry molar flow	mol/s	105.54	182.20	9.55	2140.46	182.06	0.14	639.71	2430.10	2371.67	0.00	130.21	2481.46	4803.98	4.45	0.00	3.34	1.11	4802.31	0.00	0.00	3951.40	850.91	0.00	850.91	850.91	1.38	1.38
Water	100.003	1.24	22.89		309.97	22.89		42.23	17.86	234.26	3118.12	0.88	251.26	584.12		224.19		217.52	584.12	6039.84	5952.22	642.53	29.21	29.21	0.00	0.00	14.02	14.02
Total wet molar flow	mol/s	106.78	205.09	9.55	2450.43	204.95	0.14	681.94	2447.96	2605.93	3118.12	131.09	2732.71	5388.09	4.45	224.19	3.34	218.63	5386.43	6039.84	5952.22	4593.93	880.12	29.21	850.91	850.91	15.40	15.40
Molecular weight	g/mol	10.80	31.75	100.90	30.45	31.70	100.90	11.74	28.89	30.21	18.02	27.91	29.79	30.16	100.90	18.02	172.17	18.44	30.15	18.02	18.02	27.43	43.15	18.02	44.01	44.01	21.87	21.87

CO2 capture in the cement industry

Post-combustion

Refer to drawing no. 234996/PC/FS02

Spreadsheet designed by Simon Turner 13/06/2007
Spreadsheet reviewed by Pete Lilley 17/08/2007
Spreadsheet checked by Duncan Barker 21/11/2007

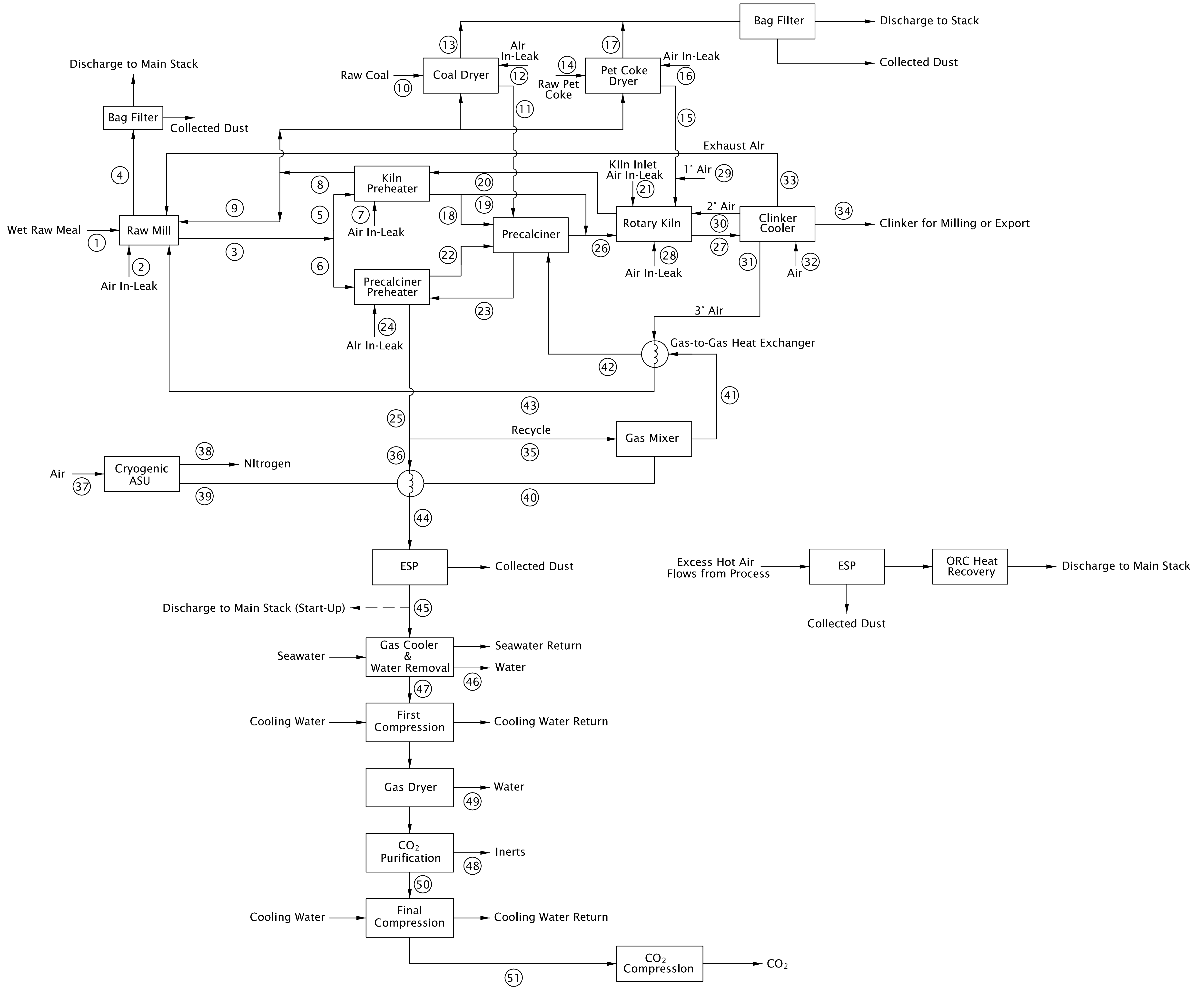
Molar Compositions (dry)

STREAM NUMBER		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27
STREAM NAME		Raw meal	Air Inleak (raw mill)	SCR Exit Gas	Raw Mill Exit Gas	Raw Meal Exit	Air Inleak (Preheater)	Precalciner Exit Gas	Preheater Exit Gas	Preheated Raw Meal	Dry Coal	Tertiary Air	Air Inleaks to precalciner	Rotary Kiln Exit Gas	Precalcined raw meal	Dry Pet Coke	Primary Air	Secondary Air	Air Inleaks to Kiln	Clinker	Cooling Air	Cooled Clinker	Excess Air	Air	Collected Dust	Vapourised Ammonia	SCR Exit Gas	Wet Coal
COMPONENT		mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)
Carbon dioxide	mol% (dry)		0.03	31.24	23.64		0.03	32.89	32.14			0.03	0.03	18.30			0.03	0.03	0.03		0.03		0.03	0.03		0.03	31.88	
Hydrogen	mol% (dry)										27.23					19.50												27.23
Nitrogen	mol% (dry)		78.04	61.98	66.79		78.04	56.50	61.37		0.63	77.54	78.04	70.08		0.57	78.04	77.54	78.04		78.04		77.54	78.04		71.60	63.25	0.63
Oxygen	mol% (dry)		21.00	3.88	8.29		21.00	1.96	3.40		2.76	20.87	21.00	1.80		0.56	21.00	20.87	21.00		21.00		20.87	21.00		19.27	3.96	2.76
Argon	mol% (dry)		0.93	0.73	0.79		0.93	0.67	0.73			0.92	0.93	0.83			0.93	0.92	0.93		0.93		0.92	0.93		0.85	0.75	
Carbon	mol% (dry)										67.33					77.49												67.33
Nitrogen dioxide	mol% (dry)			0.00	0.00			0.19	0.19					0.22													0.00	
Sulphur dioxide	mol% (dry)			0.16	0.04	0.33		0.00	0.16	0.28				0.00	0.62					0.74		0.77					0.16	
Raw meal	mol% (dry)	98.97				96.85				97.84					94.40													
Clinker	mol% (dry)																			94.87		93.16						
Sulphur	mol% (dry)	1.03				0.96				0.32	0.34				0.53	1.82				0.35		0.36						0.34
Ammonia	mol% (dry)			0.00																						8.25	0.00	
Limestone	mol% (dry)																											
Gypsum	mol% (dry)																											
MEA	mol% (dry)																											
Ash	mol% (dry)										1.72				0.82	0.06				0.55		0.57						1.72
Dust	mol% (dry)			2.00	0.44	1.85		7.80	2.00	1.56		0.64		8.78	3.63			0.64		3.48		5.14	0.64	0.00	100.00			
Total	mol% (dry)	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00

Molar Compositions (dry)

STREAM NUMBER		28	29	30	31	32	33	34	35	36	37	38	39	40	41	42	43	44	45	46	47	48	49	50	51	52	53	54
STREAM NAME		Wet Pet Coke	Exit Gas from Fuel Drying	Collected Dust post raw mill	Bag Filter Exit Gas post raw mill	Bag Filter Exit Gas post FD	Collected Dust post FD	Coal for CHP	Air	CHP Exit Gas	LP Steam	Vapourised Ammonia	CHP SCR Exit Gas	Gas Mixer Exit Gas	Limestone	Water	Gypsum from FGD	Effluent from FGD	Exit Gas from FGD	Make up water	Surplus Water	Flue Gas to Atmosphere	CO2 to drying	Waste water from drying	Dry CO2	Compressed CO2	Make Up MEA	Waste MEA
COMPONENT		mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)	mol% (dry)
Carbon dioxide	mol% (dry)		31.85		23.74	31.88			0.03	18.19		0.03	17.39	20.77					20.85			3.80	100.00		100.00	100.00		
Hydrogen	mol% (dry)	19.50						27.23																				
Nitrogen	mol% (dry)	0.57	63.20		67.09	63.25		0.63	78.04	79.96		71.60	80.56	73.90					73.93			89.85						
Oxygen	mol% (dry)	0.56	3.96		8.33	3.96		2.76	21.00	0.00		19.27	1.01	4.38					4.35			5.29						
Argon	mol% (dry)		0.75		0.80	0.75			0.93	0.95		0.85	0.95	0.88					0.88			1.06						
Carbon	mol% (dry)	77.48						67.33																				
Nitrogen dioxide	mol% (dry)		0.00		0.00	0.00				0.34			0.00	0.00					0.00			0.00						
Sulphur dioxide	mol% (dry)		0.16		0.04	0.16				0.09			0.09	0.07					0.00			0.00						
Raw meal	mol% (dry)																											
Clinker	mol% (dry)																											
Sulphur	mol% (dry)	1.78						0.34																				
Ammonia	mol% (dry)		0.00			0.00				0.00		8.25		0.00					0.00			0.00						
Limestone	mol% (dry)														100.00			100.00										
Gypsum	mol% (dry)																100.00											
MEA	mol% (dry)																										100.00	100.00
Ash	mol% (dry)	0.06						1.72		0.46																		
Dust	mol% (dry)	0.05	0.08	100.00	0.00	0.00	100.00							0.00					0.00			0.00						
Total	mol% (dry)	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	0.00	100.00	100.00	100.00	100.00	0.00	100.00	100.00	100.00	0.00	0.00	100.00	100.00	0.00	100.00	100.00	100.00	100.00

Appendix E PFD and Stream Compositions for Oxy-Combustion Option



Notes

ESP - Electrostatic Precipitator
ASU - Air Separation Unit
ORC - Organic Rankine Cycle

Key to symbols

Rev	Date	Drawn	Description	Ch'k'd	App'd
P1	10.01.08	AK	Preliminary Issue		

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Client

International Energy Association
Greenhouse Gas R&D Programme
(IEA GHG) CO2 Capture in the Cement Industry

Title

Oxy - Combustion CO_2 Capture Cement Plant
(1Mt/yr)

Designed			Eng.Chk.	
Drawn	AK		Coordination	
Dwg.Chk.			Approved	

Scale NTS
Project 234996
CAD file
Status Pre
Drawing No 234996/OC/FS03
Rev P1

CO_2 capture in the cement industry

Oxy-combustion option

Refer to drawing no. 234996/OC/FS03

Spreadsheet designed by	Duncan Barker 13/06/2007
Spreadsheet reviewed by	Peter Lilley 12/07/2007
Spreadsheet checked by	Simon Turner 13/12/2007

Mass flowrates

STREAM ID		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
STREAM NAME		Wet raw meal	Raw mill air in-leaks	Dried raw meal	Exit gases from raw mill	Dry raw meal to kiln preheater	Dry raw meal to precalciner preheater	Kiln preheater air in-leaks	Kiln preheater exit gases	Kiln preheater exit gases to raw mill	Raw coal	Dried coal	Coal dryer air in-leaks	Exit gas from coal dryer	Raw pet coke	Dried pet coke	Pet coke dryer air in-leaks	Exit gas from pet coke dryer	Kiln preheater to precalciner	Kiln preheater to kiln	Kiln exit gas to kiln preheater	Kiln inlet air in-leaks	Precalciner preheater to precalciner	Precalciner to precalciner preheater	Preclaciner preheater air in-leaks	Precalciner preheater exit gas	Feed to kiln
COMPONENT																											
Ash	kg/s										0.28	0.28			0.00	0.00											
Carbon	kg/s										1.66	1.66			0.82	0.82											
Hydrogen	kg/s										0.11	0.11			0.03	0.03											
Nitrogen	kg/s		12.04		75.77			1.13	13.03	11.02	0.18	0.18	0.24	1.67	0.02	0.02	0.24	0.81			9.49	2.41		2.14	2.00	4.14	
Oxygen	kg/s		3.70		21.09			0.35	1.41	1.19	0.04	0.04	0.07	0.23	0.01	0.01	0.07	0.14			0.32	0.74		0.80	0.62	1.42	
Sulphur	kg/s	0.16									0.02	0.02			0.05	0.05											
Carbon dioxide	kg/s		0.01		7.21			0.00	8.48	7.17			0.00	0.93			0.00	0.37			5.46	0.00		34.84	0.00	34.84	
Argon	kg/s		0.20		1.29			0.02	0.22	0.19			0.00	0.03			0.00	0.01			0.16	0.04		0.46	0.03	0.49	
Nitrogen dioxide	kg/s				0.03				0.03	0.03				0.00				0.00			0.03			0.16		0.16	
Sulphur dioxide	kg/s				0.06				0.20	0.17				0.02				0.01			0.11			0.12		0.12	
Raw meal	kg/s	48.67		51.03		18.33	32.59												25.55	6.39			58.84				51.81
Clinker	kg/s																										
Dust	kg/s								0.93	0.78	0.00	0.10		0.00	0.00	0.04		0.00			17.55			27.90		1.65	
Total dry flow	kg/s	48.67	15.96	51.03	105.45	18.33	32.59	1.49	24.29	20.55	2.29	2.38	0.32	2.89	0.93	0.97	0.32	1.35	25.55	6.39	33.12	3.19	58.84	66.42	2.66	42.82	51.81
Water	kg/s	4.23	0.00	0.52	4.18	0.19	0.33	0.00	0.52	0.44	0.24	0.02	0.00	0.27	0.02	0.01	0.00	0.03	0.00	0.00	0.33	0.00	0.00	2.07	0.00	2.07	0.00
Phase	-	Solid	Gas	Solid	Gas	Solid	Solid	Gas	Gas	Gas	Solid	Solid	Gas	Gas	Solid	Solid	Gas	Gas	Solid	Solid	Gas	Gas	Solid	Gas	Gas	Gas	Solid
Temperature	°C	9	9	100	110	100	100	9	330	330	9	100	9	70	9	100	9	70	870	870	1100	9	870	900	9	330	900
Pressure	bar(a)	-	1.013	-	<1.013	-	-	1.013	<1.013	<1.013	-	-	1.013	-	-	-	1.013	<1.013	-	-	-	1.013	-	<1.013	1.013	<1.013	-
Total wet flow	kg/s	52.90	15.96	51.54	109.62	18.52	32.92	1.49	24.81	20.99	2.53	2.41	0.32	3.16	0.95	0.98	0.32	1.38	25.55	6.39	33.45	3.19	58.84	68.49	2.66	44.89	51.81
Total Stream Enthaply	MJ/hr	-2360088	-1260	-2240280	-400392	-800676	-1423188	-86.4	-313776	-265032	-18504	-8640	-18	-42408	-972	-1692	-18	-13320	-996264	-249156	-775116	-187.2	-2294352	-2132280	-154.8	-1244736	-2001384
Total Stream Enthaply	MJ/s	-655.58	-0.35	-622.3	-111.22	-222.41	-395.33	-0.024	-87.16	-73.62	-5.14	-2.4	-0.005	-11.78	-0.27	-0.47	-0.005	-3.7	-276.74	-69.21	-215.31	-0.052	-637.32	-592.3	-0.043	-345.76	-555.94

Mass flowrates

STREAM ID		27	28	29	30	31	32	33	34	35	36	37	38	39	40	41	42	43	44	45	46	47	48	49	50	51
STREAM NAME		Kiln exit	Kiln exit air in-leaks	Primary air	Secondary air	Tertiary air	Inlet air to air cooler	Exhaust air from clinker cooler	Cooled clinker	Recycled gas	Purge stream	Air to ASU	Waste gas from ASU	Oxygen from ASU	Heated oxygen to gas mixer	Recycled gas with oxygen	Heated recycled gas with oxygen	Cooled tertiary air to raw mill	Cooled purge stream	Cooled and dedusted purge stream	Water from cooler	Cooled CO_2 rich waste stream	Inerts from gas clean up	Water from CO_2 purification	CO_2 rich stream for compression	Compressed CO_2 rich stream
COMPONENT																										
Ash	kg/s																									
Carbon	kg/s																									
Hydrogen	kg/s																									
Nitrogen	kg/s		1.20	1.01	7.27	20.21	59.98	32.50		2.04	2.11	18.30	18.20	0.10	0.10	2.14	2.14	20.21	2.11	2.11		2.11	1.68		0.43	0.43
Oxygen	kg/s		0.37	0.31	2.24	6.21	18.44	9.99		0.70	0.72	5.63	0.11	5.51	5.51	6.21	6.21	6.21	0.72	0.72	0.00	0.72	0.64	0.00	0.09	0.09
Sulphur	kg/s																									
Carbon dioxide	kg/s		0.00	0.00	0.00	0.01	0.04	0.02		17.13	17.71	0.01	0.01	0.00	0.00	17.13	17.13	0.01	17.71	17.71	0.00	17.71	1.40	0.00	16.31	16.31
Argon	kg/s		0.02	0.02	0.12	0.34	1.02	0.55		0.24	0.25	0.31	0.09	0.22	0.22	0.46	0.46	0.34	0.25	0.25		0.25	0.21		0.04	0.04
Nitrogen dioxide	kg/s									0.08	0.08					0.08	0.08		0.08	0.08	0.00	0.08	0.06	0.00	0.02	0.02
Sulphur dioxide	kg/s									0.06	0.06					0.06	0.06		0.06	0.06	0.05	0.01	0.00	0.01	0.00	0.00
Raw meal	kg/s																									
Clinker	kg/s	33.51							31.92																	
Dust	kg/s				1.60	1.60				0.81	0.84					0.81	0.81	1.60	0.84	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total dry flow	kg/s	33.51	1.60	1.34	11.23	28.37	79.47	43.06	31.92	21.05	21.77	24.25	18.42	5.83	5.83	26.89	26.89	28.37	21.77	20.93	0.05	20.88	3.98	0.01	16.89	16.89
Water	kg/s	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.02	1.05	0.11	0.11	0.00	0.00	1.02	1.02	0.00	1.05	1.05	0.79	0.26	0.00	0.26	0.00	0.00
Phase	-	Solid	Gas	Gas	Gas	Gas	Gas	Gas	Solid	Gas	Gas	Gas	Gas	Gas	Gas	Gas	Gas	Gas	Gas	Gas	Liquid	Gas	Gas	Liquid	Gas	Liquid
Temperature	°C	1350	9	9	1125	795	9	270	59	330	330	9	16	16	310	325	775	404	248	248	35	13	7	20	7	43
Pressure	bar(a)	-	1.013	1.013	<1.013	<1.013	1.013	<1.013	-	<1.013	<1.013	1.013	1.2	1.6	<1.013	<1.013	<1.013	<1.013	<1.013	<1.013	1.013	1.013	1.013	1.013	9.33	110
Total wet flow	kg/s	33.51	1.60	1.34	11.23	28.37	79.47	43.06	31.92	22.07	22.82	24.36	18.53	5.83	5.83	27.90	27.90	28.37	22.82	21.99	0.84	21.14	3.98	0.27	16.89	16.89
Total Stream Enthaply	MJ/hr	-1214748	-93.6	-79.2	-17424	18180	-5940	38052	-1298052	-612072	-632664	-7092	-6264	-180	5616	-606564	-553860	-25920	-639936	-604080	-45864	-584568	-45144	-15048	-526500	-533484
Total Stream Enthaply	MJ/s	-337.43	-0.026	-0.022	-4.84	5.05	-1.65	10.57	-360.57	-170.02	-175.74	-1.97	-1.74	-0.05	1.56	-168.49	-153.85	-7.2	-177.76	-167.8	-12.74	-162.38	-12.54	-4.18	-146.25	-148.19

Note

Moisture content within air inleaks has been ignored
For molar flows and heat capacities the clinker is assumed to be 100% CaO
For molar flows and heat capacities the raw meal is assumed to be 100% limestone
For molar flows and heat capacities the Ash is assumed to be 100% SiO_2
For molar flows and heat capacities the Coal is assumed to be carbon
For molar flows and heat capacities the Pet Coke is assumed to be carbon
For molar flows and heat capacities the Dust in streams 8, 9, 25, 35, 36, 41, 42 and 44 are assumed to be 100% $CaCO_2$
For molar flows and heat capacities the Dust in streams 20, 23, 30, 31 and 43 are assumed to be 100% CaO
For molar flows and heat capacities the Dust in streams 11 and 15 are assumed to be 100% Coal
The enthalpy for all liquid and gaseous components was calculated in CHEMCAD using SRK equation of state method.
The enthalpy for all solid components was calculated in CHEMCAD using latent heat method.
Solid components were: coal, pet coke, calcium carbonate, calcium oxide, ash

CO_2 capture in the cement industry

Oxy-combustion option

Refer to drawing no. 234996/OC/FS03

Spreadsheet designed by	Duncan Barker 13/06/2007
Spreadsheet reviewed by	Peter Lilley 12/07/2007
Spreadsheet checked by	Simon Turner 13/12/2007

Mass compositions

STREAM NUMBER		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25
STREAM NAME		**Wet raw meal**	**Raw mill air in-leaks**	**Dried raw meal**	**Exit gases from raw mill**	**Dry raw meal to kiln preheater**	**Dry raw meal to precalciner preheater**	**Kiln preheater air in-leaks**	**Kiln preheater exit gases**	**Kiln preheater exit gases to raw mill**	**Raw coal**	**Dried coal**	**Coal dryer air in-leaks**	**Exit gas from coal dryer**	**Raw pet coke**	**Dried pet coke**	**Pet coke dryer air in-leaks**	**Exit gas from pet coke dryer**	**Kiln preheater to precalciner**	**Kiln preheater to kiln**	**Kiln exit gas to kiln preheater**	**Kiln inlet air in-leaks**	**Precalciner preheater to precalciner**	**Precalciner to precalciner preheater**	**Preclaciner preheater air in-leaks**	**Precalciner preheater exit gas**
COMPONENT																										
Ash	% (w/w) dry	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	12.20	11.72	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon	% (w/w) dry	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	72.44	69.56	0.00	0.00	87.60	83.99	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen	% (w/w) dry	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	4.92	4.72	0.00	0.00	3.70	3.55	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen	% (w/w) dry	0.00	75.47	0.00	71.86	0.00	0.00	75.47	53.64	53.64	7.90	7.59	75.47	57.95	1.70	1.63	75.47	60.44	0.00	0.00	28.66	75.47	0.00	3.22	75.47	9.67
Oxygen	% (w/w) dry	0.00	23.20	0.00	20.00	0.00	0.00	23.20	5.79	5.79	1.58	1.52	23.20	7.92	1.50	1.44	23.20	10.09	0.00	0.00	0.96	23.20	0.00	1.21	23.20	3.32
Sulphur	% (w/w) dry	0.33	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.97	0.93	0.00	0.00	5.50	5.27	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon dioxide	% (w/w) dry	0.00	0.05	0.00	6.84	0.00	0.00	0.05	34.90	34.90	0.00	0.00	0.05	32.28	0.00	0.00	0.05	27.70	0.00	0.00	16.48	0.05	0.00	52.45	0.05	81.36
Argon	% (w/w) dry	0.00	1.28	0.00	1.22	0.00	0.00	1.28	0.91	0.91	0.00	0.00	1.28	0.98	0.00	0.00	1.28	1.03	0.00	0.00	0.49	1.28	0.00	0.69	1.28	1.15
Nitrogen dioxide	% (w/w) dry	0.00	0.00	0.00	0.03	0.00	0.00	0.00	0.13	0.13	0.00	0.00	0.00	0.12	0.00	0.00	0.00	0.10	0.00	0.00	0.09	0.00	0.00	0.24	0.00	0.37
Sulphur dioxide	% (w/w) dry	0.00	0.00	0.00	0.05	0.00	0.00	0.00	0.82	0.82	0.00	0.00	0.00	0.76	0.00	0.00	0.00	0.65	0.00	0.00	0.32	0.00	0.00	0.18	0.00	0.29
Raw meal	% (w/w) dry	100.00	0.00	100.00	0.00	100.00	100.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	100.00	100.00	0.00	0.00	100.00	0.00	0.00	0.00
Clinker	% (w/w) dry	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dust	% (w/w) dry	0.00	0.00	0.00	0.00	0.00	0.00	0.00	3.81	3.81	0.00	4.28	0.00	0.00	0.00	4.19	0.00	0.00	0.00	0.00	53.00	0.00	0.00	42.00	0.00	3.84
Total	% (w/w) dry	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100
Water	% (w/w) wet	8.00	0.00	1.00	3.81	1.00	1.00	0.00	2.09	2.09	9.50	1.00	0.00	8.60	2.00	1.00	0.00	2.34	0.00	0.00	1.00	0.00	0.00	3.02	0.00	4.61

Mass compositions

STREAM NUMBER		26	27	28	29	30	31	32	33	34	35	36	37	38	39	40	41	42	43	44	45	46	47	48	49	50	51
STREAM NAME		**Feed to kiln**	**Kiln exit**	**Kiln exit air in-leaks**	**Primary air**	**Secondary air**	**Tertiary air**	**Inlet air to air cooler**	**Exhaust air from clinker cooler**	**Cooled clinker**	**Recycled gas**	**Purge stream**	**Air to ASU**	**Waste gas from ASU**	**Oxygen from ASU**	**Heated oxygen to gas mixer**	**Recycled gas with oxygen**	**Heated recycled gas with oxygen**	**Cooled tertiary air to raw mill**	**Cooled purge stream**	**Cooled and dedusted purge stream**	**Water from cooler**	**Cooled CO2 rich waste stream**	**Inerts from gas clean up**	**Water from CO2 purification**	**CO2 rich stream for compression**	**Compressed CO2 rich stream**
COMPONENT																											
Ash	% (w/w) dry	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon	% (w/w) dry	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen	% (w/w) dry	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen	% (w/w) dry	0.00	0.00	75.47	75.47	64.75	71.23	75.47	75.47	0.00	9.67	9.67	75.47	98.82	1.74	1.74	7.95	7.95	71.23	9.67	10.06	0.00	10.08	42.09	0.00	2.54	2.54
Oxygen	% (w/w) dry	0.00	0.00	23.20	23.20	19.90	21.90	23.20	23.20	0.00	3.32	3.32	23.20	0.61	94.53	94.53	23.10	23.10	21.90	3.32	3.45	0.01	3.46	15.95	0.02	0.51	0.51
Sulphur	% (w/w) dry	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon dioxide	% (w/w) dry	0.00	0.00	0.05	0.05	0.04	0.05	0.05	0.05	0.00	81.36	81.36	0.05	0.07	0.00	0.00	63.71	63.71	0.05	81.36	84.61	1.63	84.82	35.19	4.05	96.58	96.58
Argon	% (w/w) dry	0.00	0.00	1.28	1.28	1.10	1.21	1.28	1.28	0.00	1.15	1.15	1.28	0.51	3.73	3.73	1.71	1.71	1.21	1.15	1.20	0.00	1.20	5.18	0.00	0.27	0.27
Nitrogen dioxide	% (w/w) dry	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.37	0.37	0.00	0.00	0.00	0.00	0.29	0.29	0.00	0.37	0.38	1.77	0.38	1.59	2.91	0.10	0.10
Sulphur dioxide	% (w/w) dry	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.29	0.29	0.00	0.00	0.00	0.00	0.22	0.22	0.00	0.29	0.30	96.59	0.05	0.00	93.02	0.00	0.00
Raw meal	% (w/w) dry	100.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Clinker	% (w/w) dry	0.00	100.00	0.00	0.00	0.00	0.00	0.00	0.00	100.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dust	% (w/w) dry	0.00	0.00	0.00	0.00	14.21	5.62	0.00	0.00	0.00	3.84	3.84	0.00	0.00	0.00	0.00	3.01	3.01	5.62	3.84	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	% (w/w) dry	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100	100
Water	% (w/w) wet	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	4.61	4.61	0.46	0.60	0.00	0.00	3.65	3.65	0.00	4.61	4.79	93.60	1.24	0.00	96.00	0.00	0.00

Note

Moisture content within air inleaks has been ignored
For molar flows and heat capacities the clinker is assumed to be 100% CaO
For molar flows and heat capacities the raw meal is assumed to be 100% limestone
For molar flows and heat capacities the Ash is assumed to be 100% SiO_2
For molar flows and heat capacities the Coal is assumed to be carbon
For molar flows and heat capacities the Pet Coke is assumed to be carbon
For molar flows and heat capacities the Dust in streams 8, 9, 25, 35, 36, 41, 42 and 44 are assumed to be 100% $CaCO_2$
For molar flows and heat capacities the Dust in streams 20, 23, 30, 31 and 43 are assumed to be 100% CaO
For molar flows and heat capacities the Dust in streams 11 and 15 are assumed to be 100% Coal
The enthalpy for all liquid and gaseous components was calculated in CHEMCAD using SRK equation of state method.
The enthalpy for all solid components was calculated in CHEMCAD using latent heat method.
Solid components were: coal, pet coke, calcium carbonate, calcium oxide, ash

CO_2 capture in the cement industry

Oxy-combustion option

Refer to drawing no. 234996/OC/FS03

Spreadsheet designed by	Duncan Barker 13/06/2007
Spreadsheet reviewed by	Peter Lilley 12/07/2007
Spreadsheet checked by	Simon Turner 13/12/2007

Molar flowrates (dry)

STREAM NUMBER		1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
STREAM NAME		**Wet raw meal**	**Raw mill air in-leaks**	**Dried raw meal**	**Exit gases from raw mill**	**Dry raw meal to kiln preheater**	**Dry raw meal to precalciner preheater**	**Kiln preheater air in-leaks**	**Kiln preheater exit gases**	**Kiln preheater exit gases to raw mill**	**Raw coal**	**Dried coal**	**Coal dryer air in-leaks**	**Exit gas from coal dryer**	**Raw pet coke**	**Dried pet coke**	**Pet coke dryer air in-leaks**	**Exit gas from pet coke dryer**	**Kiln preheater to precalciner**	**Kiln preheater to kiln**	**Kiln exit gas to kiln preheater**	**Kiln inlet air in-leaks**	**Precalciner preheater to precalciner**	**Precalciner to precalciner preheater**	**Preclaciner preheater air in-leaks**	**Precalciner preheater exit gas**	**Feed to kiln**
COMPONENT	MW (g/mol)	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s
Ash	88.64	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	3.15	3.15	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon	12.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	137.95	137.95	0.00	0.00	67.92	67.92	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen (H_2)	2.016	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	55.78	55.78	0.00	0.00	17.09	17.09	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen (N_2)	28.02	0.00	429.82	0.00	2704.13	0.00	0.00	40.23	464.95	393.35	6.45	6.45	8.60	59.74	0.56	0.56	8.60	29.05	0.00	0.00	338.76	85.96	0.00	76.30	71.52	147.82	0.00
Oxygen (O_2)	32.00	0.00	115.70	0.00	659.16	0.00	0.00	10.83	43.92	37.16	1.13	1.13	2.31	7.15	0.44	0.44	2.31	4.25	0.00	0.00	9.96	23.14	0.00	25.14	19.25	44.39	0.00
Sulphur	32.06	5.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.69	0.69	0.00	0.00	1.60	1.60	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon dioxide	44.01	0.00	0.18	0.00	163.90	0.00	0.00	0.02	192.59	162.93	0.00	0.00	0.00	21.19	0.00	0.00	0.00	8.48	0.00	0.00	124.00	0.04	0.00	791.62	0.03	791.65	0.00
Argon	39.94	0.00	5.11	0.00	32.18	0.00	0.00	0.48	5.53	4.68	0.00	0.00	0.10	0.71	0.00	0.00	0.10	0.35	0.00	0.00	4.03	1.02	0.00	11.53	0.85	12.38	0.00
Nitrogen dioxide	46.01	0.00	0.00	0.00	0.57	0.00	0.00	0.00	0.68	0.57	0.00	0.00	0.00	0.07	0.00	0.00	0.00	0.03	0.00	0.00	0.68	0.00	0.00	3.44	0.00	3.44	0.00
Sulphur dioxide	64.06	0.00	0.00	0.00	0.89	0.00	0.00	0.00	3.11	2.63	0.00	0.00	0.00	0.34	0.00	0.00	0.00	0.14	0.00	0.00	1.67	0.00	0.00	1.92	0.00	1.92	0.00
Raw meal	100.09	486.29	0.00	509.82	0.00	183.15	325.60	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	255.32	63.83	0.00	0.00	587.90	0.00	0.00	0.00	517.66
Clinker	56.08	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dust	100.09	0.00	0.00	0.00	0.00	0.00	0.00	0.00	9.25	7.83	0.00	1.02	0.00	0.00	0.00	0.41	0.00	0.00	0.00	0.00	175.38	0.00	0.00	278.74	0.00	16.44	0.00
Total dry molar flow		491.30	550.82	509.82	3560.84	183.15	325.60	51.56	720.04	609.15	205.15	206.17	11.02	89.20	87.61	88.01	11.02	42.29	255.32	63.83	654.48	110.16	587.90	1188.68	91.66	1018.05	517.66
Water	18.016	234.92	0.00	28.61	231.98	10.28	18.27	0.00	28.77	24.34	13.33	1.34	0.00	15.09	1.05	0.54	0.00	1.79	0.00	0.00	18.50	0.00	0.00	114.91	0.00	114.91	0.00
Total wet molar flow		726.22	550.82	538.43	3792.82	193.43	343.87	51.56	748.81	633.49	218.48	207.51	11.02	104.29	88.66	88.56	11.02	44.08	255.32	63.83	672.97	110.16	587.90	1303.59	91.66	1132.96	517.66
Molecular weight	g/mol	72.85	28.97	95.73	28.90	95.73	95.73	28.97	33.13	33.13	11.57	11.59	28.97	30.31	10.72	11.08	28.97	31.29	100.09	100.09	49.71	28.97	100.09	52.54	28.97	39.63	100.09

Molar flowrates (dry)

STREAM NUMBER		27	28	29	30	31	32	33	34	35	36	37	38	39	40	41	42	43	44	45	46	47	48	49	50	51
STREAM NAME		**Kiln exit**	**Kiln exit air in-leaks**	**Primary air**	**Secondary air**	**Tertiary air**	**Inlet air to air cooler**	**Exhaust air from clinker cooler**	**Cooled clinker**	**Recycled gas**	**Purge stream**	**Air to ASU**	**Waste gas from ASU**	**Oxygen from ASU**	**Heated oxygen to gas mixer**	**Recycled gas with oxygen**	**Heated recycled gas with oxygen**	**Cooled tertiary air to raw mill**	**Cooled purge stream**	**Cooled and dedusted purge stream**	**Water from cooler**	**Cooled CO2 rich waste stream**	**Inerts from gas clean up**	**Water from CO2 purification**	**CO2 rich stream for compression**	**Compressed CO2 rich stream**
COMPONENT	MW (g/mol)	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s	mol/s
Ash	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon	12.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen (H_2)	0.03	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen (N_2)	28.02	0.00	42.98	36.21	259.56	721.18	2140.52	1159.77	0.00	72.67	75.15	653.17	649.55	3.63	3.63	76.30	76.30	721.18	75.15	75.15	0.00	75.15	59.83	0.00	15.32	15.32
Oxygen (O_2)	32.00	0.00	11.57	9.75	69.87	194.12	576.17	312.18	0.00	21.82	22.57	175.82	3.52	172.30	172.30	194.12	194.12	194.12	22.57	22.57	0.00	22.57	19.86	0.00	2.71	2.71
Sulphur	32.06	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon dioxide	44.01	0.00	0.02	0.02	0.11	0.30	0.90	0.49	0.00	389.19	402.46	0.28	0.28	0.00	0.00	389.19	389.19	0.30	402.46	402.46	0.02	402.44	31.85	0.01	370.58	370.58
Argon	39.94	0.00	0.51	0.43	3.09	8.58	25.47	13.80	0.00	6.08	6.29	7.77	2.33	5.44	5.44	11.53	11.53	8.58	6.29	6.29	0.00	6.29	5.17	0.00	1.13	1.13
Nitrogen dioxide	46.01	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.69	1.75	0.00	0.00	0.00	0.00	1.69	1.69	0.00	1.75	1.75	0.02	1.73	1.37	0.01	0.35	0.35
Sulphur dioxide	64.06	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.94	0.97	0.00	0.00	0.00	0.00	0.94	0.94	0.00	0.97	0.97	0.81	0.16	0.00	0.16	0.00	0.00
Raw meal	100.09	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Clinker	56.08	597.58	0.00	0.00	0.00	0.00	0.00	0.00	569.12	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dust	100.09	0.00	0.00	0.00	15.94	15.94	0.00	0.00	0.00	8.08	8.36	0.00	0.00	0.00	0.00	8.08	8.08	15.94	8.36	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total dry molar flow		597.58	55.08	46.41	348.57	940.14	2743.06	1486.24	569.12	500.49	517.55	837.04	655.67	181.37	181.37	681.86	681.86	940.14	517.55	509.19	0.86	508.34	118.07	0.18	390.09	390.09
Water	18.016	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	56.49	58.42	6.15	6.15	0.00	0.00	56.49	56.49	0.00	58.42	58.42	43.81	14.60	0.00	14.60	0.00	0.00
Total wet molar flow		597.58	55.08	46.41	348.57	940.14	2743.06	1486.24	569.12	556.98	575.97	843.19	661.82	181.37	181.37	738.35	738.35	940.14	575.97	567.61	44.67	522.94	118.07	14.78	390.09	390.09
Molecular weight	g/mol	56.08	28.97	28.97	32.22	30.18	28.97	28.97	56.08	39.63	39.63	28.89	28.00	32.16	32.16	37.79	37.79	30.18	39.63	38.74	18.88	40.43	33.73	18.54	43.29	43.29

Note

Moisture content within air inleaks has been ignored
For molar flows and heat capacities the clinker is assumed to be 100% CaO
For molar flows and heat capacities the raw meal is assumed to be 100% limestone
For molar flows and heat capacities the Ash is assumed to be 100% SiO_2
For molar flows and heat capacities the Coal is assumed to be carbon
For molar flows and heat capacities the Pet Coke is assumed to be carbon
For molar flows and heat capacities the Dust in streams 8, 9, 25, 35, 36, 41, 42 and 44 are assumed to be 100% $CaCO_2$
For molar flows and heat capacities the Dust in streams 20, 23, 30, 31 and 43 are assumed to be 100% CaO
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The enthalpy for all liquid and gaseous components was calculated in CHEMCAD using SRK equation of state method.
The enthalpy for all solid components was calculated in CHEMCAD using latent heat method.
Solid components were: coal, pet coke, calcium carbonate, calcium oxide, ash

CO_2 capture in the cement industry

Oxy-combustion option

Refer to drawing no. 234996/OC/FS03

Spreadsheet designed by	Duncan Barker 13/06/2007
Spreadsheet reviewed by	Peter Lilley 12/07/2007
Spreadsheet checked by	Simon Turner 13/12/2007

Molar compositions (dry)

STREAM NUMBER		1	2	3	4	5	6	7	8	9	10	11	12	13
STREAM NAME		**Wet raw meal**	**Raw mill air in-leaks**	**Dried raw meal**	**Exit gases from raw mill**	**Dry raw meal to kiln preheater**	**Dry raw meal to precalciner preheater**	**Kiln preheater air in-leaks**	**Kiln preheater exit gases**	**Kiln preheater exit gases to raw mill**	**Raw coal**	**Dried coal**	**Coal dryer air in-leaks**	**Exit gas from coal dryer**
COMPONENT														
Ash	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.53	1.53	0.00	0.00
Carbon	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	67.24	66.91	0.00	0.00
Hydrogen	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	27.19	27.06	0.00	0.00
Nitrogen	mol% (dry)	0.00	78.03	0.00	75.94	0.00	0.00	78.03	64.57	64.57	3.14	3.13	78.03	66.97
Oxygen	mol% (dry)	0.00	21.00	0.00	18.51	0.00	0.00	21.00	6.10	6.10	0.55	0.55	21.00	8.01
Sulphur	mol% (dry)	1.02	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.34	0.33	0.00	0.00
Carbon dioxide	mol% (dry)	0.00	0.03	0.00	4.60	0.00	0.00	0.03	26.75	26.75	0.00	0.00	0.03	23.75
Argon	mol% (dry)	0.00	0.93	0.00	0.90	0.00	0.00	0.93	0.77	0.77	0.00	0.00	0.93	0.80
Nitrogen dioxide	mol% (dry)	0.00	0.00	0.00	0.02	0.00	0.00	0.00	0.09	0.09	0.00	0.00	0.00	0.08
Sulphur dioxide	mol% (dry)	0.00	0.00	0.00	0.03	0.00	0.00	0.00	0.43	0.43	0.00	0.00	0.00	0.38
Raw meal	mol% (dry)	98.98	0.00	100.00	0.00	100.00	100.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Clinker	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dust	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	1.28	1.28	0.00	0.49	0.00	0.00
Total	mol% (dry)	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00

STREAM NUMBER		14	15	16	17	18	19	20	21	22	23	24	25
STREAM NAME		**Raw pet coke**	**Dried pet coke**	**Pet coke dryer air in-leaks**	**Exit gas from pet coke dryer**	**Kiln preheater to precalciner**	**Kiln preheater to kiln**	**Kiln exit gas to kiln preheater**	**Kiln inlet air in-leaks**	**Precalciner preheater to precalciner**	**Precalciner to precalciner preheater**	**Preclaciner preheater air in-leaks**	**Precalciner preheater exit gas**
COMPONENT													
Ash	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon	mol% (dry)	77.53	77.17	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen	mol% (dry)	19.51	19.42	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen	mol% (dry)	0.64	0.64	78.03	68.70	0.00	0.00	51.76	78.03	0.00	6.42	78.03	14.52
Oxygen	mol% (dry)	0.50	0.50	21.00	10.04	0.00	0.00	1.52	21.00	0.00	2.11	21.00	4.36
Sulphur	mol% (dry)	1.82	1.81	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon dioxide	mol% (dry)	0.00	0.00	0.03	20.05	0.00	0.00	18.95	0.03	0.00	66.60	0.03	77.76
Argon	mol% (dry)	0.00	0.00	0.93	0.82	0.00	0.00	0.62	0.93	0.00	0.97	0.93	1.22
Nitrogen dioxide	mol% (dry)	0.00	0.00	0.00	0.07	0.00	0.00	0.10	0.00	0.00	0.29	0.00	0.34
Sulphur dioxide	mol% (dry)	0.00	0.00	0.00	0.32	0.00	0.00	0.25	0.00	0.00	0.16	0.00	0.19
Raw meal	mol% (dry)	0.00	0.00	0.00	0.00	100.00	100.00	0.00	0.00	100.00	0.00	0.00	0.00
Clinker	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dust	mol% (dry)	0.00	0.46	0.00	0.00	0.00	0.00	26.80	0.00	0.00	23.45	0.00	1.62
Total	mol% (dry)	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00

Molar compositions (dry)

STREAM NUMBER		26	27	28	29	30	31	32	33	34	35	36	37	38
STREAM NAME		**Feed to kiln**	**Kiln exit**	**Kiln exit air in-leaks**	**Primary air**	**Secondary air**	**Tertiary air**	**Inlet air to air cooler**	**Exhaust air from clinker cooler**	**Cooled clinker**	**Recycled gas**	**Purge stream**	**Air to ASU**	**Waste gas from ASU**
COMPONENT														
Ash	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen	mol% (dry)	0.00	0.00	78.03	78.03	74.46	76.71	78.03	78.03	0.00	14.52	14.52	78.03	99.07
Oxygen	mol% (dry)	0.00	0.00	21.00	21.00	20.04	20.65	21.00	21.00	0.00	4.36	4.36	21.00	0.54
Sulphur	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon dioxide	mol% (dry)	0.00	0.00	0.03	0.03	0.03	0.03	0.03	0.03	0.00	77.76	77.76	0.03	0.04
Argon	mol% (dry)	0.00	0.00	0.93	0.93	0.89	0.91	0.93	0.93	0.00	1.22	1.22	0.93	0.36
Nitrogen dioxide	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.34	0.34	0.00	0.00
Sulphur dioxide	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.19	0.19	0.00	0.00
Raw meal	mol% (dry)	100.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Clinker	mol% (dry)	0.00	100.00	0.00	0.00	0.00	0.00	0.00	0.00	100.00	0.00	0.00	0.00	0.00
Dust	mol% (dry)	0.00	0.00	0.00	0.00	4.57	1.70	0.00	0.00	0.00	1.62	1.62	0.00	0.00
Total	mol% (dry)	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00

STREAM NUMBER		39	40	41	42	43	44	45	46	47	48	49	50	51
STREAM NAME		**Oxygen from ASU**	**Heated oxygen to gas mixer**	**Recycled gas with oxygen**	**Heated recycled gas with oxygen**	**Cooled tertiary air to raw mill**	**Cooled purge stream**	**Cooled and dedusted purge stream**	**Water from cooler**	**Cooled CO2 rich waste stream**	**Inerts from gas clean up**	**Water from CO2 purification**	**CO2 rich stream for compression**	**Compressed CO2 rich stream**
COMPONENT														
Ash	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Hydrogen	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Nitrogen	mol% (dry)	2.00	2.00	11.19	11.19	76.71	14.52	14.76	0.00	14.78	50.67	0.00	3.93	3.93
Oxygen	mol% (dry)	95.00	95.00	28.47	28.47	20.65	4.36	4.43	0.02	4.44	16.82	0.04	0.69	0.69
Sulphur	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Carbon dioxide	mol% (dry)	0.00	0.00	57.08	57.08	0.03	77.76	79.04	2.34	79.17	26.97	5.73	95.00	95.00
Argon	mol% (dry)	3.00	3.00	1.69	1.69	0.91	1.22	1.24	0.00	1.24	4.38	0.00	0.29	0.29
Nitrogen dioxide	mol% (dry)	0.00	0.00	0.25	0.25	0.00	0.34	0.34	2.43	0.34	1.16	3.93	0.09	0.09
Sulphur dioxide	mol% (dry)	0.00	0.00	0.14	0.14	0.00	0.19	0.19	95.21	0.03	0.00	90.30	0.00	0.00
Raw meal	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Clinker	mol% (dry)	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Dust	mol% (dry)	0.00	0.00	1.19	1.19	1.70	1.62	0.00	0.00	0.00	0.00	0.00	0.00	0.00
Total	mol% (dry)	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00	100.00

Note

Moisture content within air inleaks has been ignored
For molar flows and heat capacities the clinker is assumed to be 100% CaO
For molar flows and heat capacities the raw meal is assumed to be 100% limestone
For molar flows and heat capacities the Ash is assumed to be 100% SiO_2
For molar flows and heat capacities the Coal is assumed to be carbon
For molar flows and heat capacities the Pet Coke is assumed to be carbon
For molar flows and heat capacities the Dust in streams 8, 9, 25, 35, 36, 41, 42 and 44 are assumed to be 100% $CaCO_2$
For molar flows and heat capacities the Dust in streams 20, 23, 30, 31 and 43 are assumed to be 100% CaO
For molar flows and heat capacities the Dust in streams 11 and 15 are assumed to be 100% Coal
The enthalpy for all liquid and gaseous components was calculated in CHEMCAD using SRK equation of state method.
The enthalpy for all solid components was calculated in CHEMCAD using latent heat method.
Solid components were: coal, pet coke, calcium carbonate, calcium oxide, ash

Appendix F Economic Analysis Spreadsheets

Version 1
Date 11 January 2008

IEA GREENHOUSE GAS R&D PROGRAMME

Cement Manufacturing Plant - No CO_2 Capture

Production		
Cement	1.0	Mt/y
Power generation	0.0	MWh/y
CO_2 emissions	728,422	t/y
Consumption @ 90% load factor		
Coal	63,317	t/y
Pet Coke	32,876	t/y
Power	80,809	MWh/y
Water	0.24	Mt/y
Cooling water	0	Mt/y
CO_2 emissions	0.728	t/t
CO_2 emissions		
CO_2 emissions - power generation	0.520	kg/kWh
CO_2 emissions including power	0.770	t/t

Capital Cost		**Million €**	
Installed costs		235.2	
Average contingencies	5.0%	11.8	
Owners costs	7.0%	16.5	
Total capital cost		263.5	
Decommissioning cost		0	
Stocks			
Fuel		15	days
Limestone		7	days
Other raw materials		15	days
Working Capital			
Fuel		0.3	
Limestone		0.1	
Other raw materials		0.1	
Total working capital		0.4	

Operating Costs at load factor	90%
	Million €/year
Fuel	6.7
Power	4.0
Limestone	3.7
Other raw materials	1.6
Water	0.02
Cooling water	0.0
Consumables	0.7
Total Variable Costs	16.8
Fixed costs	
Maintenance	9.4
Labour	3.8
Administration, rates & insurance	5.9
Total fixed	19.1
Total Operating Costs	35.9

Economic parameters		
Discount rate	10%	%
Base year	2008	%
Coal price	65.00	€/t
Pet coke price	80.00	€/t
Power Purchase Price	50.00	€/MWh
Power Sale Price	50.00	€/MWh
Water price	0.10	€/t
Cooling water price	0.02	€/t

Results summary		
Cement production cost	65.650	€/t

CASH FLOW ANALYSIS

Million $	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**	**2018**	**2019**	**2020**	**2021**
Year	**000**	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**
Load Factor				60%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%
Equivalent yearly hours				5256	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884
Expediture Factor	0%	20%	45%	35%										
Cement Production	0.0	0.0	0.0	0.7	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Revenues														
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	4.5	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7
Power	0.0	0.0	0.0	2.7	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0
Limestone	0.0	0.0	0.0	2.5	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7
Other raw materials	0.0	0.0	0.0	1.0	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6
Water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Cooling water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Consumables	0.0	0.0	0.0	0.5	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Maintenance	0.0	0.0	0.0	6.3	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4
Labour	0.0	0.0	0.0	2.6	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8
Administration, rates & insurance	0.0	0.0	0.0	3.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9
Fixed Capital Expenditures	0.0	52.7	118.6	92.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	0.0	52.7	118.6	116.6	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9
Total Cash Flow (cumulated)	0.0	52.7	171.3	287.9	323.8	359.7	395.7	431.6	467.5	503.4	539.4	575.3	611.2	647.2

Million $	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**	**2032**	**2033**	**2034**	**2035**	**2036**
Year	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	
Equivalent yearly hours	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	
Expediture Factor															
Cement Production	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	0.0
Revenues															
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs															
Fuel	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	0.0
Power	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	0.0
Limestone	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	0.0
Other raw materials	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	0.0
Water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Cooling water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Consumables	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.0
Maintenance	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	0.0
Labour	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	0.0
Administration, rates & insurance	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	0.0
Fixed Capital Expenditures	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	0.0
Total Cash Flow (cumulated)	683.1	719.0	755.0	790.9	826.8	862.7	898.7	934.6	970.5	1006.5	1042.4	1078.3	1114.3	1150.2	1150.2

Version 1
Date 11 January 2008

IEA GREENHOUSE GAS R&D PROGRAMME

Cement Manufacturing Plant - Post Combustion CO_2 Capture

Production		
Cement	1.0	Mt
Power generation	356,400	MWh/y
CO_2 emissions	188,424	t/y
Consumption @ 90% load factor		
Coal	291,633	t/y
Pet Coke	32,876	t/y
Power	333,665	MWh/y
Water	0.390	Mt/y
Cooling Water	4.4	Mt/y
CO_2 emissions	0.188	t/t
CO_2 emissions including power	0.177	t/t
Reference plant data		
CO_2 emissions	0.728	t/t
Cement production cost	65.6	€/t
CO_2 emissions including power	0.770	t/t

Capital Cost		**Million €**	
Installed costs		491.6	
Average contingencies	6.5%	31.8	
Owners costs	7.0%	34.4	
Total capital cost		557.8	
Decommissioning cost		0	
Stocks			
Fuel		15	days
Limestone		7	days
Other raw materials		15	days
Working Capital			
Fuel		0.9	
Limestone		0.1	
Other raw materials		0.2	
Total working capital		1.2	

Operating Costs at load factor	90%
	Million €/year
Fuel	21.6
Power	0.0
Limestone	3.8
Other raw materials	6.0
Water	0.04
Cooling water	0.1
Consumables	0.7
Total Variable Costs	32.2
Fixed costs	
Maintenance	16.8
Labour	6.7
Administration, rates & insurance	11.8
Total fixed	35.4
Total Operating Costs	67.6

Economic parameters		
Discount rate	10%	%
Base year	2008	
Coal price	65.00	€/t
Pet coke price	80.00	€/t
Power Purchase Price	50.00	€/MWh
Power Sale Price	50.00	€/MWh
Water price	0.10	€/t
Cooling water price	0.02	€/t
CO_2 emissions for external power	0.52	tCO_2/MWh

Results summary		
Cement production cost	129.4	€/t
Emission avoidance cost Excluding Power	118.1	€/t CO_2
Emission avoidance cost Including Power	107.4	€/t CO_2
CO_2 emissions avoided, including power	77%	

CASH FLOW ANALYSIS
Million $

	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017	2018	2019	2020	2021
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11
Load Factor				60%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%
Equivalent yearly hours				5256	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884
Expediture Factor	0%	20%	45%	35%										
Cement Production	0.0	0.0	0.0	0.7	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Revenues														
Power generation	0.0	0.0	0.0	0.8	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1
Operating Costs														
Fuel	0.0	0.0	0.0	14.4	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6
Power	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Limestone	0.0	0.0	0.0	2.5	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8
Other raw materials	0.0	0.0	0.0	4.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0
Water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Cooling water	0.0	0.0	0.0	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
Consumables	0.0	0.0	0.0	0.5	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Maintenance	0.0	0.0	0.0	11.2	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8
Labour	0.0	0.0	0.0	4.5	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7
Administration, rates & insurance	0.0	0.0	0.0	7.9	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8
Fixed Capital Expenditures	0.0	111.6	251.0	195.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	1.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	0.0	111.6	251.0	240.7	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5
Total Cash Flow (cumulated)	0.0	111.6	362.6	603.3	669.8	736.2	802.7	869.2	935.6	1002.1	1068.6	1135.0	1201.5	1267.9

	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2035	2036
Year	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	
Equivalent yearly hours	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	
Expediture Factor															
Cement Production	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	0.0
Revenues															
Power generation	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	0.0
Operating Costs															
Fuel	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	0.0
Power	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Limestone	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	0.0
Other raw materials	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	0.0
Water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Cooling water	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.0
Consumables	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.0
Maintenance	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	0.0
Labour	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	0.0
Administration, rates & insurance	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	0.0
Fixed Capital Expenditures	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	0.0
Total Cash Flow (cumulated)	1334.4	1400.9	1467.3	1533.8	1600.3	1666.7	1733.2	1799.7	1866.1	1932.6	1999.0	2065.5	2132.0	2198.4	2198.4

IEA GREENHOUSE GAS R&D PROGRAMME

Cement Manufacturing Plant - Oxy Combustion CO_2 Capture

Version 1
Date 11 January 2008

Production		
Cement	1.0	Mt
Power generation	5,548.0	MWh/y
CO_2 emissions	282,853	t/y
Consumption @ 90% load factor		
Coal	72,061	t/y
Pet Coke	27,091	t/y
Power	180,109	MWh/y
Water	11.30	Mt/y
Cooling Water	1.7	Mt/y
CO_2 emissions	0.283	t/t
CO_2 emissions including power	0.374	t/t
Reference plant data		
CO_2 emissions	0.728	t/t
Cement production cost	65.6	€/t
CO_2 emissions including power	0.770	t/t

Capital Cost		**Million €**	
Installed costs		290.0	
Average contingencies	5.8%	16.7	
Owners costs	7.0%	20.3	
Total capital cost		327.0	
Decommissioning cost		0	
Stocks			
Fuel		15	days
Limestone		7	days
Other raw materials		15	days
Working Capital			
Fuel		0.3	
Limestone		0.1	
Other raw materials		0.1	
Total working capital		0.4	

Operating Costs at load factor	90%
	Million €/year
Fuel	6.9
Power	8.7
Limestone	3.8
Other raw materials	1.6
Sea Water	0.3
Cooling water	0.04
Consumables	0.7
Total Variable Costs	21.9
Fixed costs	
Maintenance	10.7
Labour	4.8
Administration, rates & insurance	7.2
Total fixed	22.8
Total Operating Costs	44.7

Economic parameters		
Discount rate	10%	%
Base year	2008	%
Coal price	65.00	€/t
Pet coke price	80.00	€/t
Power Purchase Price	50.00	€/MWh
Power Sale Price	50.00	€/MWh
Water price	0.10	€/t
Cooling water price	0.02	€/t
CO_2 emissions for external power	0.52	tCO_2/MWh

Results summary		
Cement production cost	81.6	€/t
Emission avoidance cost Excluding Power	35.8	€/t CO_2
Emission avoidance cost Including Power	40.2	€/t CO_2
CO_2 emissions avoided, including power	52%	

CASH FLOW ANALYSIS

Million $

	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017	2018	2019	2020	2021	2022
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11	12
Load Factor				60%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%
Equivalent yearly hours				5256	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884
Expediture Factor	0%	20%	45%	35%											
Cement Production	0.0	0.0	0.0	0.7	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Revenues															
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs															
Fuel	0.0	0.0	0.0	4.6	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9
Power	0.0	0.0	0.0	5.8	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7
Limestone	0.0	0.0	0.0	2.5	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8
Other raw materials	0.0	0.0	0.0	1.1	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6
Water	0.0	0.0	0.0	0.2	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3
Cooling water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Consumables	0.0	0.0	0.0	0.5	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Maintenance	0.0	0.0	0.0	7.2	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7
Labour	0.0	0.0	0.0	3.2	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8
Administration, rates & insurance	0.0	0.0	0.0	4.8	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2
Fixed Capital Expenditures	0.0	65.4	147.2	114.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	0.0	65.4	147.2	144.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7
Total Cash Flow (cumulated)	0.0	65.4	212.6	357.2	402.0	446.7	491.4	536.1	580.9	625.6	670.3	715.0	759.8	804.5	849.2

	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2035	2036
Year	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	
Equivalent yearly hours	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	
Expediture Factor														
Cement Production	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	0.0
Revenues														
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	0.0
Power	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	0.0
Limestone	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	0.0
Other raw materials	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	0.0
Water	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.0
Cooling water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Consumables	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.0
Maintenance	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	0.0
Labour	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	0.0
Administration, rates & insurance	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	0.0
Fixed Capital Expenditures	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	0.0
Total Cash Flow (cumulated)	893.9	938.7	983.4	1028.1	1072.9	1117.6	1162.3	1207.0	1251.8	1296.5	1341.2	1385.9	1430.7	1430.7

IEA GREENHOUSE GAS R&D PROGRAMME

Cement Manufacturing Plant - No CO_2 Capture

Version 1
Date 11 January 2008

Production		
Cement	1.0	Mt/y
Power generation	0.0	MWh/y
CO_2 emissions	728,422	t/y
Consumption @ 90% load factor		
Coal	63,317	t/y
Pet Coke	32,876	t/y
Power	80,809	MWh/y
Water	0.24	Mt/y
Cooling water	0	Mt/y
CO_2 emissions	0.728	t/t
CO_2 emissions		
CO_2 emissions - power generation	0.520	kg/kWh
CO_2 emissions including power	0.770	t/t

Capital Cost		**Million €**	
Installed costs		235.2	
Average contingencies	5.0%	11.8	
Owners costs	7.0%	16.5	
Total capital cost		263.5	
Decommissioning cost		0	
Stocks			
Fuel		15	days
Limestone		7	days
Other raw materials		15	days
Working Capital			
Fuel		0.3	
Limestone		0.1	
Other raw materials		0.1	
Total working capital		0.4	

Operating Costs at load factor	90%
	Million €/year
Fuel	6.7
Power	4.0
Limestone	3.7
Other raw materials	1.6
Water	0.0
Cooling water	0.0
Consumables	0.7
Total Variable Costs	16.8
Fixed costs	
Maintenance	9.4
Labour	3.8
Administration, rates & insurance	5.9
Total fixed	19.1
Total Operating Costs	35.9

Economic parameters		
Discount rate	10%	%
Base year	2008	%
Coal price	65.00	€/t
Pet coke price	80.00	€/t
Power Purchase Price	50.00	€/MWh
Power Sale Price	50.00	€/MWh
Water price	0.10	€/t
Cooling water price	0.02	€/t

Results summary		
Cement production cost	63.448	€/t

CASH FLOW ANALYSIS
Million $

	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017	2018	2019	2020	2021	2022
Year	**000**	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**	**12**
Load Factor				60%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%
Equivalent yearly hours				5256	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884
Expediture Factor	0%	20%	45%	35%											
Cement Production	0.0	0.0	0.0	0.7	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Revenues															
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs															
Fuel	0.0	0.0	0.0	4.5	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7
Power	0.0	0.0	0.0	2.7	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0
Limestone	0.0	0.0	0.0	2.5	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7
Other raw materials	0.0	0.0	0.0	1.0	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6
Water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Cooling water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Consumables	0.0	0.0	0.0	0.5	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Maintenance	0.0	0.0	0.0	6.3	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4
Labour	0.0	0.0	0.0	2.6	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8
Administration, rates & insurance	0.0	0.0	0.0	3.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9
Fixed Capital Expenditures	0.0	52.7	118.6	92.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly	0.0	52.7	118.6	116.6	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9
Total Cash Flow (cumulated)	0.0	52.7	171.3	287.9	323.8	359.7	395.7	431.6	467.5	503.4	539.4	575.3	611.2	647.2	683.1

	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2035	2036
Year	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%
Equivalent yearly hours	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884
Expediture Factor														
Cement Production	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Revenues														
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7
Power	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0
Limestone	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7
Other raw materials	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6
Water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Cooling water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Consumables	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Maintenance	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4
Labour	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8
Administration, rates & insurance	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9
Fixed Capital Expenditures	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9
Total Cash Flow (cumulated)	719.0	755.0	790.9	826.8	862.7	898.7	934.6	970.5	1006.5	1042.4	1078.3	1114.3	1150.2	1186.1

	2037	2038	2039	2040	2041	2042	2043	2044	2045	2046	2047	2048	2049	2050
Year	**27**	**28**	**29**	**30**	**31**	**32**	**33**	**34**	**35**	**36**	**37**	**38**	**39**	**40**
Load Factor	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%
Equivalent yearly hours	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884
Expediture Factor														
Cement Production	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Revenues														
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7
Power	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0
Limestone	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7
Other raw materials	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6
Water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Cooling water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Consumables	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Maintenance	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4	9.4
Labour	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8
Administration, rates & insurance	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9	5.9
Fixed Capital Expenditures	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9	35.9
Total Cash Flow (cumulated)	1222.1	1258.0	1293.9	1329.8	1365.8	1401.7	1437.6	1473.6	1509.5	1545.4	1581.4	1617.3	1653.2	1689.1

IEA GREENHOUSE GAS R&D PROGRAMME

Cement Manufacturing Plant - Post Combustion CO$_2$ Capture

Version 1
Date 11 January 2008

Production		
Cement	1.0	Mt
Power generation	356,400	MWh/y
CO₂ emissions	188,424	t/y
Consumption @ 90% load factor		
Coal	291,633	t/y
Pet Coke	32,876	t/y
Power	333,665	MWh/y
Water	0.390	Mt/y
Cooling Water	4.4	Mt/y
CO₂ emissions	0.188	t/t
CO₂ emissions including power	0.177	t/t
Reference plant data		
CO₂ emissions	0.728	t/t
Cement production cost	63.4	€/t
CO₂ emissions including power	0.770	t/t

Capital Cost		**Million €**	
Installed costs		491.6	
Average contingencies	6.5%	31.8	
Owners costs	7.0%	34.4	
Total capital cost		557.8	
Decommissioning cost		0	
Stocks			
Fuel		15	days
Limestone		7	days
Other raw materials		15	days
Working Capital			
Fuel		0.9	
Limestone		0.1	
Other raw materials		0.2	
Total working capital		1.2	

Operating Costs at load factor	90%
	Million €/year
Fuel	21.6
Power	0.0
Limestone	3.8
Other raw materials	6.0
Water	0.0
Cooling water	0.1
Consumables	0.7
Total Variable Costs	32.2
Fixed costs	
Maintenance	16.8
Labour	6.7
Administration, rates & insurance	11.8
Total fixed	35.4
Total Operating Costs	67.6

Economic parameters		
Discount rate	10%	%
Base year	2008	
Coal price	65.00	€/t
Pet coke price	80.00	€/t
Power Purchase Price	50.00	€/MWh
Power Sale Price	50.00	€/MWh
Water price	0.10	€/t
Cooling water price	0.02	€/t
CO₂ emissions for external power	0.52	tCO₂/MWh

Results summary		
Cement production cost	124.7	€/t
Emission avoidance cost Excluding Power	113.5	€/t CO₂
Emission avoidance cost Including Power	103.2	€/t CO₂
CO₂ emissions avoided, including powe	77%	

CASH FLOW ANALYSIS

Million $	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017	2018	2019	2020	2021	2022
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11	12
Load Factor				60%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%
Equivalent yearly hours				5256	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884
Expediture Factor	0%	20%	45%	35%											
Cement Production	0.0	0.0	0.0	0.7	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Revenues															
Power generation	0.0	0.0	0.0	0.8	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1
Operating Costs															
Fuel	0.0	0.0	0.0	14.4	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6
Power	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Limestone	0.0	0.0	0.0	2.5	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8
Other raw materials	0.0	0.0	0.0	4.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0
Water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Cooling water	0.0	0.0	0.0	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
Consumables	0.0	0.0	0.0	0.5	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Maintenance	0.0	0.0	0.0	11.2	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8
Labour	0.0	0.0	0.0	4.5	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7
Administration, rates & insurance	0.0	0.0	0.0	7.9	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8
Fixed Capital Expenditures	0.0	111.6	251.0	195.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	1.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	0.0	111.6	251.0	240.7	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5
Total Cash Flow (cumulated)	0.0	111.6	362.6	603.3	669.8	736.2	802.7	869.2	935.6	1002.1	1068.6	1135.0	1201.5	1267.9	1334.4

Million $	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2035	2036
Year	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%
Equivalent yearly hours	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884
Expediture Factor														
Cement Production	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Revenues														
Power generation	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1
Operating Costs														
Fuel	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6
Power	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Limestone	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8
Other raw materials	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0
Water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Cooling water	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
Consumables	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Maintenance	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8
Labour	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7
Administration, rates & insurance	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8
Fixed Capital Expenditures	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5
Total Cash Flow (cumulated)	1400.9	1467.3	1533.8	1600.3	1666.7	1733.2	1799.7	1866.1	1932.6	1999.0	2065.5	2132.0	2198.4	2264.9

Million $	2037	2038	2039	2040	2041	2042	2043	2044	2045	2046	2047	2048	2049	2050
Year	27	28	29	30	31	32	33	34	35	36	37	38	39	40
Load Factor	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%
Equivalent yearly hours	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884
Expediture Factor														
Cement Production	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Revenues														
Power generation	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1
Operating Costs														
Fuel	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6	21.6
Power	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Limestone	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8
Other raw materials	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0
Water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Cooling water	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
Consumables	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Maintenance	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8	16.8
Labour	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7
Administration, rates & insurance	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8
Fixed Capital Expenditures	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5	66.5
Total Cash Flow (cumulated)	2331.4	2397.8	2464.3	2530.7	2597.2	2663.7	2730.1	2796.6	2863.1	2929.5	2996.0	3062.4	3128.9	3195.4

IEA GREENHOUSE GAS R&D PROGRAMME

Cement Manufacturing Plant - Oxy Combustion CO $_2$ Capture

Version 1
Date 11 January 2008

Production		
Cement	1.0	Mt
Power generation	5548.0	MWh/y
CO₂ emissions	282,853	t/y
Consumption @ 90% load factor		
Coal	72,061	t/y
Pet Coke	27,091	t/y
Power	180,109	MWh/y
Water	11.3	Mt/y
Cooling Water	1.7	Mt/y
CO₂ emissions	0.283	t/t
CO₂ emissions including power	0.374	t/t
Reference plant data		
CO₂ emissions	0.728	t/t
Cement production cost	63.4	€/t
CO₂ emissions including power	0.770	t/t

Capital Cost		**Million €**	
Installed costs		290.0	
Average contingencies	5.8%	16.7	
Owners costs	7.0%	20.3	
Total capital cost		327.0	
Decommissioning cost		0	
Stocks			
Fuel		15	days
Limestone		7	days
Other raw materials		15	days
Working Capital			
Fuel		0.3	
Limestone		0.1	
Other raw materials		0.1	
Total working capital		0.4	

Operating Costs at load factor	90%
	Million €/year
Fuel	6.9
Power	8.7
Limestone	3.8
Other raw materials	1.6
Sea Water	0.3
Cooling water	0.0
Consumables	0.7
Total Variable Costs	21.9
Fixed costs	
Maintenance	10.7
Labour	4.8
Administration, rates & insurance	7.2
Total fixed	22.8
Total Operating Costs	44.7

Economic parameters		
Discount rate	10%	%
Base year	2008	%
Coal price	65.00	€/t
Pet coke price	80.00	€/t
Power Purchase Price	50.00	€/MWh
Power Sale Price	50.00	€/MWh
Water price	0.10	€/t
Cooling water price	0.02	€/t
CO₂ emissions for external power	0.52	tCO₂/MWh

Results summary		
Cement production cost	78.9	€/t
Emission avoidance cost Excluding Power	34.6	€/t CO₂
Emission avoidance cost Including Power	38.9	€/t CO₂
CO₂ emissions avoided, including powe	52%	

CASH FLOW ANALYSIS
Million $

	2008	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**	**2018**	**2019**	**2020**	**2021**
Year	**000**	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**
Load Factor				60%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%
Equivalent yearly hours				5256	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884
Expediture Factor	0%	20%	45%	35%										
Cement Production	0.0	0.0	0.0	0.7	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Revenues														
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	4.6	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9
Power	0.0	0.0	0.0	5.8	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7
Limestone	0.0	0.0	0.0	2.5	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8
Other raw materials	0.0	0.0	0.0	1.1	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6
Sea Water	0.0	0.0	0.0	0.2	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3
Cooling water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Consumables	0.0	0.0	0.0	0.5	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Maintenance	0.0	0.0	0.0	7.2	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7
Labour	0.0	0.0	0.0	3.2	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8
Administration, rates & insurance	0.0	0.0	0.0	4.8	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2
Fixed Capital Expenditures	0.0	65.4	147.2	114.5	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	0.0	65.4	147.2	144.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7
Total Cash Flow (cumulated)	0.0	65.4	212.6	357.2	402.0	446.7	491.4	536.1	580.9	625.6	670.3	715.0	759.8	804.5

	2022	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**	**2032**	**2033**	**2034**	**2035**	**2036**
Year	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%
Equivalent yearly hours	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884
Expediture Factor															
Cement Production	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Revenues															
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs															
Fuel	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9
Power	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7
Limestone	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8
Other raw materials	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6
Sea Water	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3
Cooling water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Consumables	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Maintenance	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7
Labour	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8
Administration, rates & insurance	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2
Fixed Capital Expenditures	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7
Total Cash Flow (cumulated)	849.2	893.9	938.7	983.4	1028.1	1072.9	1117.6	1162.3	1207.0	1251.8	1296.5	1341.2	1385.9	1430.7	1475.4

	2037	**2038**	**2039**	**2040**	**2041**	**2042**	**2043**	**2044**	**2045**	**2046**	**2047**	**2048**	**2049**	**2050**
Year	**27**	**28**	**29**	**30**	**31**	**32**	**33**	**34**	**35**	**36**	**37**	**38**	**39**	**40**
Load Factor	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%
Equivalent yearly hours	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884
Expediture Factor														
Cement Production	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Revenues														
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9
Power	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7	8.7
Limestone	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8
Other raw materials	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6
Sea Water	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3
Cooling water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Consumables	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Maintenance	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7	10.7
Labour	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8
Administration, rates & insurance	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2
Fixed Capital Expenditures	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7	44.7
Total Cash Flow (cumulated)	1520.1	1564.8	1609.6	1654.3	1699.0	1743.7	1788.5	1833.2	1877.9	1922.6	1967.4	2012.1	2056.8	2101.6

IEA GREENHOUSE GAS R&D PROGRAMME

Cement Manufacturing Plant - Post Combustion CO_2 Capture (Low Sulphur Scenario)

Version 1
Date 11 January 2008

Production		
Cement	1.0	Mt
Power generation	0	MWh/y
CO_2 emissions	109,263	t/y
Consumption @ 90% load factor		
Coal	63,317	t/y
Pet Coke	32,876	t/y
Power	166,833	MWh/y
Water	0.240	Mt/y
Cooling Water	2.6	Mt/y
CO_2 emissions	0.109	t/t
CO_2 emissions including power	0.196	t/t
Reference plant data		
CO_2 emissions	0.728	t/t
Cement production cost	65.6	€/t
CO_2 emissions including power	0.770	t/t

Capital Cost		**Million €**	
Installed costs		295.0	
Average contingencies	6.5%	19.1	
Owners costs	7.0%	20.6	
Total capital cost		334.7	
Decommissioning cost		0	
Stocks			
Fuel		15	days
Limestone		7	days
Other raw materials		15	days
Working Capital			
Fuel		0.3	
Limestone		0.1	
Other raw materials		0.2	
Total working capital		0.6	

Operating Costs at load factor	90%
	Million €/year
Fuel	6.7
Power	8.3
Steam	11
Limestone	3.7
Other raw materials	6.0
Water	0.02
Cooling water	0.1
Consumables	0.7
Total Variable Costs	36.7
Fixed costs	
Maintenance	11.8
Labour	4.0
Administration, rates & insurance	7.1
Total fixed	22.9
Total Operating Costs	59.6

Economic parameters		
Discount rate	10%	%
Base year	2008	
Coal price	65.00	€/t
Pet coke price	80.00	€/t
Power Purchase Price	50.00	€/MWh
Power Sale Price	50.00	€/MWh
Water price	0.10	€/t
Cooling water price	0.02	€/t
Steam price	10	€/t
CO_2 emissions for external power	0.52	tCO_2/MWh

Results summary		
Cement production cost	97.3	€/t
Emission avoidance cost Excluding Power	51.2	€/t CO_2
Emission avoidance cost Including Power	55.2	€/t CO_2
CO_2 emissions avoided, including power	75%	

CASH FLOW ANALYSIS
Million $

	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2035	2036
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor				60%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	
Equivalent yearly hours				5256	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	
Expediture Factor	0%	20%	45%	35%																									
Cement Production	0.0	0.0	0.0	0.7	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	0.0
Revenues																													
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs																													
Fuel	0.0	0.0	0.0	4.5	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	6.7	0.0
Power	0.0	0.0	0.0	5.6	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	8.3	0.0
Steam	0.0	0.0	0.0	7.3	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	11.0	0.0
Limestone	0.0	0.0	0.0	2.5	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	3.7	0.0
Other raw materials	0.0	0.0	0.0	4.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	0.0
Water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Cooling water	0.0	0.0	0.0	0.0	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.0
Consumables	0.0	0.0	0.0	0.5	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.0
Maintenance	0.0	0.0	0.0	7.9	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	11.8	0.0
Labour	0.0	0.0	0.0	2.7	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	4.0	0.0
Administration, rates & insurance	0.0	0.0	0.0	4.7	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	0.0
Fixed Capital Expenditures	0.0	66.9	150.6	117.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	0.0	66.9	150.6	157.5	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	59.6	0.0
Total Cash Flow (cumulated)	0.0	66.9	217.5	375.0	434.6	494.2	553.8	613.3	672.9	732.5	792.1	851.7	911.3	970.9	1030.5	1090.0	1149.6	1209.2	1268.8	1328.4	1388.0	1447.6	1507.1	1566.7	1626.3	1685.9	1745.5	1805.1	1805.1

Version 1
Date 11 January 2008

IEA GREENHOUSE GAS R&D PROGRAMME

Cement Manufacturing Plant - Oxy Combustion CO_2 Capture (Co-location with Power Plant)

Production		
Cement	1.0	Mt
Power generation	5,548.0	MWh/y
CO_2 emissions	282,853	t/y
Consumption @ 90% load factor		
Coal	72,061	t/y
Pet Coke	27,091	t/y
Power	99,060	MWh/y
Sea Water	0.24	Mt/y
Cooling Water	0.0	Mt/y
CO_2 emissions	0.283	t/t
CO_2 emissions including power	0.331	t/t
Reference plant data		
CO_2 emissions	0.728	t/t
Cement production cost	65.6	€/t
CO_2 emissions including power	0.770	t/t

Capital Cost		**Million €**
Installed costs		261.0
Average contingencies	5.8%	15.0
Owners costs	7.0%	18.3
Total capital cost		294.3
Decommissioning cost		0

Stocks		
Fuel	15	days
Limestone	7	days
Other raw materials	15	days

Working Capital	
Fuel	0.3
Limestone	0.1
Other raw materials	0.1
Total working capital	0.4

Operating Costs at load factor	90%
	Million €/year
Fuel	6.9
Power	4.7
Limestone	3.8
Oxygen	5.0
Carbon dioxide	2.8
Other raw materials	1.6
Sea Water	0.0
Cooling water	0.00
Consumables	0.7
Total Variable Costs	25.4
Fixed costs	
Maintenance	10.4
Labour	4.7
Administration, rates & insurance	7.0
Total fixed	22.1
Total Operating Costs	47.6

Economic parameters		
Discount rate	10%	%
Base year	2008	%
Coal price	65.00	€/t
Pet coke price	80.00	€/t
Power Purchase Price	50.00	€/MWh
Power Sale Price	50.00	€/MWh
Sea Water price	0.02	€/t
Cooling water price	0.02	€/t
CO_2 emissions for external power	0.52	tCO_2/MWh
Oxygen Price	30	€/t
CO_2 Export Price	10	€/t

Results summary		
Cement production cost	80.8	€/t
Emission avoidance cost Excluding Power	33.9	€/t CO_2
Emission avoidance cost Including Power	34.4	€/t CO_2
CO_2 emissions avoided, including power	57%	

CASH FLOW ANALYSIS
Million $

	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017	2018	2019	2020	2021
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11
Load Factor				60%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%
Equivalent yearly hours				5256	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884
Expediture Factor	0%	20%	45%	35%										
Cement Production	0.0	0.0	0.0	0.7	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0
Revenues														
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	4.6	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9
Power	0.0	0.0	0.0	3.1	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7
Limestone	0.0	0.0	0.0	2.5	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8
Oxygen	0.0	0.0	0.0	3.3	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0
Carbon dioxide	0.0	0.0	0.0	1.9	2.8	2.8	2.8	2.8	2.8	2.8	2.8	2.8	2.8	2.8
Other raw materials	0.0	0.0	0.0	1.1	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6
Water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Cooling water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Consumables	0.0	0.0	0.0	0.5	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Maintenance	0.0	0.0	0.0	7.0	10.4	10.4	10.4	10.4	10.4	10.4	10.4	10.4	10.4	10.4
Labour	0.0	0.0	0.0	3.1	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7
Administration, rates & insurance	0.0	0.0	0.0	4.7	7.0	7.0	7.0	7.0	7.0	7.0	7.0	7.0	7.0	7.0
Fixed Capital Expenditures	0.0	58.9	132.4	103.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	0.0	58.9	132.4	135.1	47.6	47.6	47.6	47.6	47.6	47.6	47.6	47.6	47.6	47.6
Total Cash Flow (cumulated)	0.0	58.9	191.3	326.4	374.0	421.6	469.2	516.7	564.3	611.9	659.4	707.0	754.6	802.2

	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2035	2036
Year	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	
Equivalent yearly hours	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	
Expediture Factor															
Cement Production	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	0.0
Revenues															
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs															
Fuel	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	6.9	0.0
Power	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	0.0
Limestone	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	0.0
Oxygen	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	5.0	0.0
Carbon dioxide	2.8	2.8	2.8	2.8	2.8	2.8	2.8	2.8	2.8	2.8	2.8	2.8	2.8	2.8	0.0
Other raw materials	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	1.6	0.0
Water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Cooling water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Consumables	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.0
Maintenance	10.4	10.4	10.4	10.4	10.4	10.4	10.4	10.4	10.4	10.4	10.4	10.4	10.4	10.4	0.0
Labour	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	0.0
Administration, rates & insurance	7.0	7.0	7.0	7.0	7.0	7.0	7.0	7.0	7.0	7.0	7.0	7.0	7.0	7.0	0.0
Fixed Capital Expenditures	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	47.6	47.6	47.6	47.6	47.6	47.6	47.6	47.6	47.6	47.6	47.6	47.6	47.6	47.6	0.0
Total Cash Flow (cumulated)	849.7	897.3	944.9	992.5	1040.0	1087.6	1135.2	1182.7	1230.3	1277.9	1325.5	1373.0	1420.6	1468.2	1468.2

Version 1
Date 11 January 2008

IEA GREENHOUSE GAS R&D PROGRAMME

Cement Manufacturing Plant - Post Combustion CO_2 Capture (KS-1 Scenario)

Production		
Cement	1.0	Mt
Power generation	267,300	MWh/y
CO2 emissions	195,341	t/y
Consumption @ 90% load factor		
Coal	234,554	t/y
Pet Coke	32,786	t/y
Power	250,929	MWh/y
Water	0.390	Mt/y
Cooling Water	4.5	Mt/y
CO2 emissions	0.195	t/t
CO2 emissions including power	0.187	t/t
Reference plant data		
CO2 emissions	0.728	t/t
Cement production cost	65.6	€/t
CO2 emissions including power	0.770	t/t

Capital Cost		**Million €**
Installed costs		442.4
Average contingencies	6.5%	28.6
Owners costs	7.0%	31.0
Total capital cost		502.0
Decommissioning cost		0
Stocks		
Fuel		15 days
Limestone		7 days
Other raw materials		15 days
Working Capital		
Fuel		0.7
Limestone		0.1
Other raw materials		0.2
Total working capital		1.1

Operating Costs at load factor	90%
	Million €/year
Fuel	17.9
Power	0.0
Limestone	3.8
Other raw materials	6.0
Water	0.04
Cooling water	0.1
Consumables	0.7
Total Variable Costs	28.5
Fixed costs	
Maintenance	17.7
Labour	7.1
Administration, rates & insurance	12.5
Total fixed	37.2
Total Operating Costs	65.8

Economic parameters		
Discount rate	10%	%
Base year	2008	
Coal price	65.00	€/t
Pet coke price	80.00	€/t
Power Purchase Price	50.00	€/MWh
Power Sale Price	50.00	€/MWh
Water price	0.10	€/t
Cooling water price	0.02	€/t
CO_2 emissions for external power	0.52	tCO2/MWh

Results summary		
Cement production cost	121.6	€/t
Emission avoidance cost Excluding Power	105.0	€/t CO2
Emission avoidance cost Including Power	95.9	€/t CO2
CO2 emissions avoided, including power	76%	

CASH FLOW ANALYSIS
Million $

	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017	2018	2019	2020	2021	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2035	2036
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor				60%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	
Equivalent yearly hours				5256	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	
Expediture Factor	0%	20%	45%	35%																									
Cement Production	0.0	0.0	0.0	0.7	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	0.0
Revenues																													
Power generation	0.0	0.0	0.0	0.5	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.8	0.0
Operating Costs																													
Fuel	0.0	0.0	0.0	11.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	17.9	0.0
Power	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Limestone	0.0	0.0	0.0	2.5	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	3.8	0.0
Other raw materials	0.0	0.0	0.0	4.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	6.0	0.0
Water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Cooling water	0.0	0.0	0.0	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.0
Consumables	0.0	0.0	0.0	0.5	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.0
Maintenance	0.0	0.0	0.0	11.8	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	0.0
Labour	0.0	0.0	0.0	4.7	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	7.1	0.0
Administration, rates & insurance	0.0	0.0	0.0	8.3	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	0.0
Fixed Capital Expenditures	0.0	100.4	225.9	175.7	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	1.1	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	0.0	100.4	225.9	220.1	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	65.0	0.0
Total Cash Flow (cumulated)	0.0	100.4	326.3	546.4	611.3	676.3	741.2	806.2	871.2	936.1	1001.1	1066.0	1131.0	1195.9	1260.9	1325.9	1390.8	1455.8	1520.7	1585.7	1650.6	1715.6	1780.5	1845.5	1910.5	1975.4	2040.4	2105.3	2105.3

Version 1
Date 11 January 2008

IEA GREENHOUSE GAS R&D PROGRAMME

Cement Manufacturing Plant - No CO_2 Capture (Asian Scenario)

Production		
Cement	3.0	Mt/y
Power generation	0.0	MWh/y
CO_2 emissions	2,185,266	t/y
Consumption @ 90% load factor		
Coal	189,951	t/y
Pet Coke	98,628	t/y
Power	242,427	MWh/y
Water	0.7	Mt/y
Cooling water	0.0	Mt/y
CO_2 emissions	0.728	t/t
CO_2 emissions		
CO_2 emissions - power generation	0.520	kg/kWh
CO_2 emissions including power	0.770	t/t

Capital Cost		**Million €**
Installed costs		272.8
Average contingencies	5.0%	13.7
Owners costs	7.0%	19.1
Total capital cost		305.6
Decommissioning cost		0
Stocks		
Fuel	15	days
Limestone	7	days
Other raw materials	15	days
Working Capital		
Fuel	0.8	
Limestone	0.2	
Other raw materials	0.2	
Total working capital	1.2	

Operating Costs at load factor	90%
	Million €/year
Fuel	20.2
Power	12.1
Limestone	11.2
Other raw materials	4.7
Water	0.07
Cooling water	0.0
Consumables	2.2
Total Variable Costs	50.5
Fixed costs	
Maintenance	12.2
Labour	5.8
Administration, rates & insurance	8.8
Total fixed	26.8
Total Operating Costs	77.3

Economic parameters		
Discount rate	10%	%
Base year	2008	%
Coal price	65.00	€/t
Pet coke price	80.00	€/t
Power Purchase Price	50.00	€/MWh
Power Sale Price	50.00	€/MWh
Water price	0.10	€/t
Cooling water price	0.02	€/t

Results summary		
Cement production cost	37.274	€/t

CASH FLOW ANALYSIS

Million $	**2008**	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**	**2018**	**2019**	**2020**	**2021**
Year	**000**	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**
Load Factor				60%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%
Equivalent yearly hours				5256	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884
Expediture Factor	0%	20%	45%	35%										
Cement Production	0.0	0.0	0.0	2.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0
Revenues														
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	13.5	20.2	20.2	20.2	20.2	20.2	20.2	20.2	20.2	20.2	20.2
Power	0.0	0.0	0.0	8.1	12.1	12.1	12.1	12.1	12.1	12.1	12.1	12.1	12.1	12.1
Limestone	0.0	0.0	0.0	7.5	11.2	11.2	11.2	11.2	11.2	11.2	11.2	11.2	11.2	11.2
Other raw materials	0.0	0.0	0.0	3.1	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7
Water	0.0	0.0	0.0	0.0	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
Cooling water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Consumables	0.0	0.0	0.0	1.4	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2
Maintenance	0.0	0.0	0.0	8.1	12.2	12.2	12.2	12.2	12.2	12.2	12.2	12.2	12.2	12.2
Labour	0.0	0.0	0.0	3.8	5.8	5.8	5.8	5.8	5.8	5.8	5.8	5.8	5.8	5.8
Administration, rates & insurance	0.0	0.0	0.0	5.9	8.8	8.8	8.8	8.8	8.8	8.8	8.8	8.8	8.8	8.8
Fixed Capital Expenditures	0.0	61.1	137.5	107.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	1.2	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	0.0	61.1	137.5	159.7	77.3	77.3	77.3	77.3	77.3	77.3	77.3	77.3	77.3	77.3
Total Cash Flow (cumulated)	0.0	61.1	198.7	358.4	435.7	512.9	590.2	667.5	744.7	822.0	899.3	976.6	1053.8	1131.1

Million $	**2022**	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**	**2032**	**2033**	**2034**	**2035**	**2036**
Year	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	
Equivalent yearly hours	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	
Expediture Factor															
Cement Production	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	0.0
Revenues															
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs															
Fuel	20.2	20.2	20.2	20.2	20.2	20.2	20.2	20.2	20.2	20.2	20.2	20.2	20.2	20.2	0.0
Power	12.1	12.1	12.1	12.1	12.1	12.1	12.1	12.1	12.1	12.1	12.1	12.1	12.1	12.1	0.0
Limestone	11.2	11.2	11.2	11.2	11.2	11.2	11.2	11.2	11.2	11.2	11.2	11.2	11.2	11.2	0.0
Other raw materials	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	4.7	0.0
Water	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.0
Cooling water	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Consumables	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	0.0
Maintenance	12.2	12.2	12.2	12.2	12.2	12.2	12.2	12.2	12.2	12.2	12.2	12.2	12.2	12.2	0.0
Labour	5.8	5.8	5.8	5.8	5.8	5.8	5.8	5.8	5.8	5.8	5.8	5.8	5.8	5.8	0.0
Administration, rates & insurance	8.8	8.8	8.8	8.8	8.8	8.8	8.8	8.8	8.8	8.8	8.8	8.8	8.8	8.8	0.0
Fixed Capital Expenditures	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	77.3	77.3	77.3	77.3	77.3	77.3	77.3	77.3	77.3	77.3	77.3	77.3	77.3	77.3	0.0
Total Cash Flow (cumulated)	1208.4	1285.6	1362.9	1440.2	1517.4	1594.7	1672.0	1749.3	1826.5	1903.8	1981.1	2058.3	2135.6	2212.9	2212.9

Version 1
Date 11 January 2008

IEA GREENHOUSE GAS R&D PROGRAMME

Cement Manufacturing Plant - Post Combustion CO_2 Capture (Asian Scenario)

Production		
Cement	3.0	Mt
Power generation	1,069,200	MWh/y
CO_2 emissions	565,272	t/y
Consumption @ 90% load factor		
Coal	874,899	t/y
Pet Coke	98,628	t/y
Power	1,000,995	MWh/y
Water	1.2	Mt/y
Cooling Water	13.1	Mt/y
CO_2 emissions	0.188	t/t
CO_2 emissions including power	0.177	t/t
Reference plant data		
CO_2 emissions	0.728	t/t
Cement production cost	37.3	€/t
CO_2 emissions including power	0.770	t/t

Capital Cost		**Million €**	
Installed costs		570.2	
Average contingencies	6.5%	36.9	
Owners costs	7.0%	39.9	
Total capital cost		647.0	
Decommissioning cost		0	
Stocks			
Fuel		15	days
Limestone		7	days
Other raw materials		15	days
Working Capital			
Fuel		2.7	
Limestone		0.2	
Other raw materials		0.7	
Total working capital		3.6	

Operating Costs at load factor	90%
	Million €/year
Fuel	64.8
Power	0.0
Limestone	11.3
Other raw materials	18.1
Water	0.12
Cooling water	0.3
Consumables	2.2
Total Variable Costs	96.7
Fixed costs	
Maintenance	22.1
Labour	10.1
Administration, rates & insurance	17.7
Total fixed	49.9
Total Operating Costs	146.7

Economic parameters		
Discount rate	10%	%
Base year	2008	
Coal price	65.00	€/t
Pet coke price	80.00	€/t
Power Purchase Price	50.00	€/MWh
Power Sale Price	50.00	€/MWh
Water price	0.10	€/t
Cooling water price	0.02	€/t
CO_2 emissions for external power	0.52	tCO_2/MWh

Results summary		
Cement production cost	72.2	€/t
Emission avoidance cost Excluding Power	64.6	€/t CO_2
Emission avoidance cost Including Power	58.8	€/t CO_2
CO_2 emissions avoided, including power	77%	

CASH FLOW ANALYSIS

Million $

	2008	**2009**	**2010**	**2011**	**2012**	**2013**	**2014**	**2015**	**2016**	**2017**	**2018**	**2019**	**2020**	**2021**
Year	**000**	**00**	**0**	**1**	**2**	**3**	**4**	**5**	**6**	**7**	**8**	**9**	**10**	**11**
Load Factor				60%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%
Equivalent yearly hours				5256	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884
Expediture Factor	0%	20%	45%	35%										
Cement Production	0.0	0.0	0.0	2.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0
Revenues														
Power generation	0.0	0.0	0.0	2.3	3.4	3.4	3.4	3.4	3.4	3.4	3.4	3.4	3.4	3.4
Operating Costs														
Fuel	0.0	0.0	0.0	43.2	64.8	64.8	64.8	64.8	64.8	64.8	64.8	64.8	64.8	64.8
Power	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Limestone	0.0	0.0	0.0	7.6	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3
Other raw materials	0.0	0.0	0.0	12.1	18.1	18.1	18.1	18.1	18.1	18.1	18.1	18.1	18.1	18.1
Water	0.0	0.0	0.0	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
Cooling water	0.0	0.0	0.0	0.2	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3
Consumables	0.0	0.0	0.0	1.4	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2
Maintenance	0.0	0.0	0.0	14.7	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1
Labour	0.0	0.0	0.0	6.7	10.1	10.1	10.1	10.1	10.1	10.1	10.1	10.1	10.1	10.1
Administration, rates & insurance	0.0	0.0	0.0	11.8	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7
Fixed Capital Expenditures	0.0	129.4	291.1	226.4	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	3.6	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	0.0	129.4	291.1	325.6	143.3	143.3	143.3	143.3	143.3	143.3	143.3	143.3	143.3	143.3
Total Cash Flow (cumulated)	0.0	129.4	420.5	746.1	889.4	1032.7	1176.0	1319.2	1462.5	1605.8	1749.1	1892.3	2035.6	2178.9

	2022	**2023**	**2024**	**2025**	**2026**	**2027**	**2028**	**2029**	**2030**	**2031**	**2032**	**2033**	**2034**	**2035**	**2036**
Year	**12**	**13**	**14**	**15**	**16**	**17**	**18**	**19**	**20**	**21**	**22**	**23**	**24**	**25**	**26**
Load Factor	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	
Equivalent yearly hours	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	
Expediture Factor															
Cement Production	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	0.0
Revenues															
Power generation	3.4	3.4	3.4	3.4	3.4	3.4	3.4	3.4	3.4	3.4	3.4	3.4	3.4	3.4	0.0
Operating Costs															
Fuel	64.8	64.8	64.8	64.8	64.8	64.8	64.8	64.8	64.8	64.8	64.8	64.8	64.8	64.8	0.0
Power	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Limestone	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	0.0
Other raw materials	18.1	18.1	18.1	18.1	18.1	18.1	18.1	18.1	18.1	18.1	18.1	18.1	18.1	18.1	0.0
Water	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.0
Cooling water	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.3	0.0
Consumables	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	0.0
Maintenance	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	22.1	0.0
Labour	10.1	10.1	10.1	10.1	10.1	10.1	10.1	10.1	10.1	10.1	10.1	10.1	10.1	10.1	0.0
Administration, rates & insurance	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	17.7	0.0
Fixed Capital Expenditures	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	143.3	143.3	143.3	143.3	143.3	143.3	143.3	143.3	143.3	143.3	143.3	143.3	143.3	143.3	0.0
Total Cash Flow (cumulated)	2322.2	2465.4	2608.7	2752.0	2895.3	3038.5	3181.8	3325.1	3468.4	3611.6	3754.9	3898.2	4041.5	4184.7	4184.7

IEA GREENHOUSE GAS R&D PROGRAMME

Cement Manufacturing Plant - Oxy Combustion CO_2 Capture (Asian Scenario)

Version 1
Date 11 January 2008

Production		
Cement	3.0	Mt
Power generation	16,644	MWh/y
CO_2 emissions	848,559	t/y
Consumption @ 90% load factor		
Coal	216,183	t/y
Pet Coke	81,273	t/y
Power	540,327	MWh/y
Water	33.9	Mt/y
Cooling Water	5.1	Mt/y
CO_2 emissions	0.283	t/t
CO_2 emissions including power	0.374	t/t
Reference plant data		
CO_2 emissions	0.728	t/t
Cement production cost	37.3	€/t
CO_2 emissions including power	0.770	t/t

Capital Cost		**Million €**	
Installed costs		336.4	
Average contingencies	5.8%	19.4	
Owners costs	7.0%	23.5	
Total capital cost		379.3	
Decommissioning cost		0	
Stocks			
Fuel		15	days
Limestone		7	days
Other raw materials		15	days
Working Capital			
Fuel		0.8	
Limestone		0.2	
Other raw materials		0.2	
Total working capital		1.3	

Operating Costs at load factor	90%
	Million €/year
Fuel	20.6
Power	26.2
Limestone	11.3
Other raw materials	4.8
Sea Water	0.7
Cooling water	0.11
Consumables	2.2
Total Variable Costs	65.8
Fixed costs	
Maintenance	12.5
Labour	7.2
Administration, rates & insurance	10.9
Total fixed	30.5
Total Operating Costs	96.3

Economic parameters		
Discount rate	10%	%
Base year	2008	%
Coal price	65.00	€/t
Pet coke price	80.00	€/t
Power Purchase Price	50.00	€/MWh
Power Sale Price	50.00	€/MWh
Sea Water price	0.02	€/t
Cooling water price	0.02	€/t
CO_2 emissions for external power	0.52	tCO_2/MWh

Results summary		
Cement production cost	46.4	€/t
Emission avoidance cost Excluding Power	20.4	€/t CO_2
Emission avoidance cost Including Power	22.9	€/t CO_2
CO_2 emissions avoided, including power	52%	

CASH FLOW ANALYSIS
Million $

	2008	2009	2010	2011	2012	2013	2014	2015	2016	2017	2018	2019	2020	2021
Year	000	00	0	1	2	3	4	5	6	7	8	9	10	11
Load Factor				60%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%
Equivalent yearly hours				5256	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884
Expediture Factor	0%	20%	45%	35%										
Cement Production	0.0	0.0	0.0	2.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0
Revenues														
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs														
Fuel	0.0	0.0	0.0	13.7	20.6	20.6	20.6	20.6	20.6	20.6	20.6	20.6	20.6	20.6
Power	0.0	0.0	0.0	17.5	26.2	26.2	26.2	26.2	26.2	26.2	26.2	26.2	26.2	26.2
Limestone	0.0	0.0	0.0	7.5	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3
Other raw materials	0.0	0.0	0.0	3.2	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8
Water	0.0	0.0	0.0	0.5	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7
Cooling water	0.0	0.0	0.0	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1
Consumables	0.0	0.0	0.0	1.4	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2
Maintenance	0.0	0.0	0.0	8.3	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5
Labour	0.0	0.0	0.0	4.8	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2
Administration, rates & insurance	0.0	0.0	0.0	7.2	10.9	10.9	10.9	10.9	10.9	10.9	10.9	10.9	10.9	10.9
Fixed Capital Expenditures	0.0	75.9	170.7	132.8	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	1.3	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	0.0	75.9	170.7	198.2	96.3	96.3	96.3	96.3	96.3	96.3	96.3	96.3	96.3	96.3
Total Cash Flow (cumulated)	0.0	75.9	246.5	444.7	541.0	637.3	733.6	829.9	926.2	1022.4	1118.7	1215.0	1311.3	1407.6

	2022	2023	2024	2025	2026	2027	2028	2029	2030	2031	2032	2033	2034	2035	2036
Year	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26
Load Factor	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	90%	
Equivalent yearly hours	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	7884	
Expediture Factor															
Cement Production	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	3.0	0.0
Revenues															
Power generation	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Operating Costs															
Fuel	20.6	20.6	20.6	20.6	20.6	20.6	20.6	20.6	20.6	20.6	20.6	20.6	20.6	20.6	0.0
Power	26.2	26.2	26.2	26.2	26.2	26.2	26.2	26.2	26.2	26.2	26.2	26.2	26.2	26.2	0.0
Limestone	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	11.3	0.0
Other raw materials	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	4.8	0.0
Water	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.7	0.0
Cooling water	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.1	0.0
Consumables	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	2.2	0.0
Maintenance	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	12.5	0.0
Labour	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	7.2	0.0
Administration, rates & insurance	10.9	10.9	10.9	10.9	10.9	10.9	10.9	10.9	10.9	10.9	10.9	10.9	10.9	10.9	0.0
Fixed Capital Expenditures	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Working Capital	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Decommissioning Cost	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
Total Cash Flow (yearly)	96.3	96.3	96.3	96.3	96.3	96.3	96.3	96.3	96.3	96.3	96.3	96.3	96.3	96.3	0.0
Total Cash Flow (cumulated)	1503.9	1600.1	1696.4	1792.7	1889.0	1985.3	2081.6	2177.8	2274.1	2370.4	2466.7	2563.0	2659.3	2755.5	2755.5

Appendix G Note on Emission Factors

Emission factors throughout the assessment have been sourced using published reference sources and are noted within the text. It should be remembered that all emission factors are subject to a degree of uncertainty and therefore the calculations derived should be considered as approximations. Emission factors for energy use, for example, are subject to the exact operating regimes and fuel type, while transport emissions vary between vehicle type, and usage patterns such as the vehicle speed.

In order to determine the emission factor for average grid electrical consumption (and to estimate avoided emissions), emission factors supplied by Defra were used in combination with the energy statistics presented in the Digest of UK Energy Statistics (DUKES) to derive an average grid emission factors. This was done by calculating the sum of the emissions of CO_2 from each type of power generation (coal, natural gas, oil etc.) divided by the total electricity supplied to the grid (net). In this way the individual efficiencies of each type of power generation are also accounted for. The overall emission factor for the grid was calculated as 520 kg/MWh supplied to the grid. The derivation of this emission factor is presented in Table G-1.

Table G-1: Derivation of Grid Emission Factor (2006 Base Data)

Generation Method	Coal	Oil	Gas	Nuclear	Renewable	Other	Total
Fuel Consumed GWh	378805	15075	328515	213661	38896	20865	
Emission Factor (t/GWh)	320	270	190	0	0	320	
kTonnes CO_2 emitted	121,218	4,070	62,418	0	0	6,677	194,383

Total electricity supplied to the grid (2006) = 376793 GWh

Tonnes CO_2/GWh = 194,383,000 tonnes CO_2 / 376793 GWh = 520 kg/MWh