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TECHNO-ECONOMIC ANALYSIS OF COAL FIRED OXY-COMBUSTION POWER PLANTS WITH CO2 CAPTURE

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SOMMARIO ESTESO

Le crescenti preoccupazioni sul riscaldamento globale e sulle questioni ambientali spingono a sviluppare tecnologie con basse emissioni di inquinanti e gas serra. Tuttavia l'intera economia è fortemente dipendente dai prezzi dell'energia e l'adozione diffusa di soluzioni caratterezzate da basse emissioni specifiche per unità di potenza prodotta porterebbe ad un incemento dei costi, che potrebbe a sua volta causare un danno rilevante al sistema economico e alla sua competitività.
Per questo motivo, si stanno facendo grossi sforzi per trovare soluzioni economicamente valide e ad impatto ridotto, soprattutto per la produzione stazionaria di potenza che sostenga il carico di base. A tale scopo, un approccio promettente è la cattura dell'anidride carbonica da impianti di produzione di energia elettrica alimentati da combustibile contenente carbonio. In particolare, si possono identificare tre sottoinsiemi: cattura precombustione, postcombustione e ossicombustione. La corrente di CO_2 sequestrata viene poi trasportata e iniettata in un sito di stoccaggio adeguatamente selezionato.
In questa tesi è stata posta l'attenzione sulla cattura tramite ossicombustione. Più specificamente, questo studio si propone di valutare la fattibilità tecnica ed economica di impianti per la produzione di energia elettrica basati sull'ossicombustione del carbone e dotati di sistemi per la cattura dell'anidride carbonica.
L'ossicombustione rappresenta un'opzione interessante, derivata dagli impianti convenzionali a carbone con combustione in aria, caratterizzata da un discreto profilo ambientale e che potrebbe permettere la cattura dell'anidride carbonica con penalità accettabili in termini di rendimento e di costo.
La combustione in ossigeno richiede combustibile, una corrente ricca di ossigeno e un flusso di gas ricircolati. Quest'ultimo è necessario per avere temperature accettabili in camera di combustione. La posizione, il numero e la tipologia (wet-hot, wet-cold o dry) del ricircolo possono variare.
La portata ricca di ossigeno (purezza del 95% vol. o superiore) viene prodotta in un'unità di separazione dell'aria, spesso indicata dall'acronimo ASU. Attualmente, la tecnologia commerciale impiegata a tele scopo è la separazione criogenica.
Due configurazioni (a polverino di carbone o a letto fuido) possono essere adottate per il generatore di vapore. Gli impianti a polverino sono alimentati da combustibile opportunamente asciugato e ridotto in polvere, mentre i generatori di vapore a letto fluido possono essere alimentati direttamente con combustibile ancora umido in pezzi più grandi.
I gas prodotti dalla combustione e non ricircolati sono raffreddati e trattati in modo da rimuovere alcune impurità, che potrebbero danneggiare l'ambiente, il

sito di stoccaggio o l'impianto stesso. Dopo l'ulteriore raffreddamento dei gas, il vapore condensa e l'acqua prodotta può essere eliminata. I gas combusti sono poi ulteriormente trattati per rimuovere le tracce residue di acqua e i non condensabili (questi ultimi sono separati per accrescere la purezza della corrente ricca di CO_2 e per ridurre i consumi dovuti alla compressione) e in seguito compressi. Infine, la corrente ormai diventata liquida può essere trasportata fino ad un sito adatto allo stoccaggio e iniettato.
La piccola corrente di incondensabili è invece liberata in atmosfera. Questa include una porzione di CO_2 piuttosto piccola. Infatti gli impianti ad ossicombustione sono caratterizzati da un elevato tasso di cattura della CO_2.
Riguardo alla configurazione generale d'impianto, gli impianti di produzione elettrica ad ossicombustione sono abbastanza simili agli impianti a carbone convenzionali basati su ciclo a vapore. Oltre alla necessità di una corrente di ricircolo, la principale differenza è la presenza di due sistemi aggiuntivi: l'unità di separazione dell'aria e l'unità di compressione e purificazione, che causano rilevanti penalizzazioni per quanto riguardo il rendimento e il costo.

In questa tesi sono stati simulati e confrontati alcuni impianti di grandi dimensioni. I casi analizzati si distinguevano per presenza di sistemi di abbattimento degli ossidi di zolfo, combustibile (carbone bituminoso o lignite), tipologia del generatore di vapore (a polverino o a letto fluido) e caratteristiche del ricircolo.
In particolare i sei casi realizzati sono:

- caso base (o caso 1): il generatore di vapore è alimentato da polverino di carbone bituminoso a basso contenuto di zolfo. Un ricircolo primario (di gas a basso contenuto di H_2O) viene utilizzato principalmente per asciugare e trasportare il combustibile, mentre un secondo (più caldo e costituito da una corrente contenente vapore) serve a moderare le temperature in camera di combustione. Non è previsto alcun sistema di rimozione degli ossidi di zolfo.
- caso 2: è molto simile al caso base, ma viene impiegato solamente un ricircolo asciutto con il doppio ruolo di mezzo per l'asciugatura e trasporto del carbone e di inerte per limitare le temperature in camera di combustione.
- caso 3: prevede una combustione a letto fluido alimentata da carbone bituminoso, ossigeno e da un ricircolo umido. Non si ricorre a un ricircolo primario, dato che l'asciugatura del carbone non è necessaria. Infatti, il combustibile è solamente macinato in pezzi più grandi rispetto al caso di alimentazione a polverino, prima di essere mandato nella camera di combustione, dove viene inserita una portata di calcare per effettuare una rimozione degli ossidi di zolfo interna alla camera stessa, in cui si forma del gesso, che viene espulso mescolato alle ceneri. La

portata di ricircolo è stabilita per avere un'adeguata concentrazione di ossigeno nella corrente di ossidante.

- caso 4: il generatore di vapore è alimentato da un polverino ricavato da lignite molto umida. Visto che i ricircoli di gas combusti dalle consuete posizioni all'interno dell'impianto non permettono di asciugare sufficientemente il carbone (il contenuto di acqua deve scendere almeno al 15-20%), si ricircolano gas molto caldi (circa 1000°C) prelevati direttamente dalla camera di combustione.
- caso 5: lo stesso combustibile ad alto contenuto di umidità è alimentato ad un generatore di vapore a letto fluido. Lo schema d'impianto è analogo a quello del caso 3.
- caso 6: la configurazione d'impianto è identica a quella del caso base, ma viene effettuata una rimozione di SO_2 dalla corrente di ricircolo primario per ridurre la concentrazione di SO_x nel generatore di vapore e, di conseguenza, nelle correnti in ogni punto dell'impianto.

Il codice GS (Gas-Steam), sviluppato al Politecnico di Milano è stato utilizzato per ottenere una stima dei bilanci di massa e di energia all'interno di tutto l'impianto, con l'eccezione dell'unità dell'unità di compressione e purificazione della CO_2, le cui prestazioni sono state valutate attraverso Aspen Plus® Engineering Suit.
Il codice GS rappresenta i componenti del sistema come moduli che rappresentino il loro reale funzionamento. La struttura completa è vista come interconnesioni dei diversi moduli.
Sono stati impostati molti parametri per replicare le effettive condizioni operative (frazione di gas ricircolati, temperature, concentrazioni, ecc.) degli impianti ad ossicombustione con cattura dell'anidride carbonica così come vengono attualmente concepiti, includendo, tra le altre cose, anche infiltrazioni d'aria e perdite di carico.
La purezza della corrente ricca di CO_2 è stata fissata al 96,5%, con l'eccezione del caso 4, per cui è stata imposta una concentrazione complessiva di incondensabili (ossigeno, azoto e argon) del 4%, data l'elevata concentrazione di SO_2, difficilmente separabile dalla CO_2, nella corrente entrante nell'unità di compressione e purificazione della CO_2. Nel caso 4, la purezza della CO_2 è quindi del 93,0%.
La grande quantità di informazioni ottenuta mediante le simulazioni permette accurate analisi e verfiche dei parametri nei vari punti dell'impianto (specialmente per uqanto riguarda temperatura, pressione e composizione).
In aggiunta, l'output delle simulazioni fornisce dati sul consumo di combustibile, sui parametri operativi e sulle dimensioni dei componenti, che permettono di stimare i costi collegati alla produzione di potenza elettrica con cattura dell'anidride carbonica da impianti di questo tipo.

Si sono quindi impiegati i dati ottenuti per realizzare un'analisi economica per valutare il costo dell'elettricità e il costo della CO_2 evitata per tutti i vari casi analizzati in un contesto generale, rappresentativo delle condizioni tipiche del'Unione Europea e degli USA, e in uno specifico scenario australiano, caratterizzato da alcuni aspetti peculiari, tra cui spiccano il minore costo del combustibile, i maggiori costi d'investimento e l'assenza di sistemi di controllo degli ossidi di azoto e di zolfo nell'impianto di riferimento, visto che sono in vigore normative molto meno restrittive a riguardo.
Il costo dell'energia elettrica può essere visto come somma di tre componenti, ovvero costi d'invesimento, costi del combustibile e costi di O&M (operation and maintenance). Una volta trovati costi d'investimento affidabili per impianti di grossa taglia, questi possono essere aggiustati per prendere in considerazione che si riferiscono a diverse taglie, ricorrendo a opportuni fattori di scala, e/o a diversi anni (nel tempo i costi e i cambi monetari cambiano significativamente ed è necessario utilizzare opportuni indici per confrontare valori da analisi svolte in periodi diversi). E' così possibile stimare i costi totali d'investimento per gli impianti simulati, includendo i costi aggiuntivi di set-up.
Le spese per il combustibile possono essere invece facilmente calcolate, dato che i consumi di carbone sono resi disponibili dall'output del codice GS.
I costi di O&M sono la somma di due componenti: quella fissa, stimata come spesa annuale pari a una quota del costo d'investimento, e quella variabile che è invece dipendente dalla quantità di energia prodotta e viene solitamente espressa in $/MWh.
Durante la preparazione di questo elaborato, è stato preparato un programma di Matlab per stimare anche le prestazioni e il costo di impianti di minori dimensioni e/o dotati di un sistema di abbattimento dell' SO_2, dove, per prima cosa, vengono presi in considerazione gli effetti dell'inserimento di un FGD nell'impianto di grande taglia simulato, sotto l'ipotesi di minima variazione del bilancio di massa e della prevedibilità degli effetti sul bilancio energetico. Infatti, si considera una penalità aggiuntiva in termini di rendimento proporzionale alla portata catturata di SO_2.
I costi di investimento aggiuntivi sono invece calcolati tramite una procedura complessa, mentre quelli di O&M sono proporzionali alla quantità di SO_2 rimossa.
In seguito, è possibile valutare l'effetto dell'effetto di scala, anche se sono stati simulati solo impianti di dimensioni medio-grandi, visto che gli impianti di piccola taglia basati su cicli a vapore avanzati non possono essere economicamente interessanti. A conferma di ciò, non esistono impianti SC con combustione di carbone in aria di piccole dimensioni. Viene assunto che, visto l'intervallo di potenza considerato, il rendimento sia pressoché costante. Per questo motivo, anche i costi legati al combustibile e agli O&M variabili non subiscono grandi variazioni.

D'altra parte, effetti di scala significativi incidono sui costi d'investimento, quindi la relativa quota del costo dell'elettricità è maggiore per impianti più piccoli. Per questo, si ricorre a fattori di scala utilizzati come esponenti del rapporto tra dimensioni caratteristiche.

L'intervallo di rendimento degli impianti considerati (riferito al potere calorifico inferiore) è 30,8%-37,4%. Il valore più alto è ottenuto dall'impianto a letto fluido alimentato con carbone bituminoso, seguito da vicino dal caso base e dall'impianto dotato di FGD. Il caso con ricircolo completamente asciutto e quelli alimentati a lignite umida sono caratterizzate rendimenti inferiori.
I risultati mostrano che i bilanci di massa e di energia per i tre impianti simulati che impiegano polverino di carbone bituminoso come combustibile sono piuttosto simili tra loro, con un vantaggio di rendimento per il caso base.
I costi specifici totali d'investimento (costi di set-up inclusi) nello scenario generale variano dai circa 2600 ai circa 3050 USD/kW$_e$, ovvero il 50-85% in più rispetto a una centrale USC convenzionale a carbone dotata di sistemi di rimozione di SO_x e NO_x, mentre nel contesto australiano l'aumento sarebbe superiore al 90% rispetto ad un analogo impianto di riferimento senza abbattimento di SO_x e NO_x, non richiesto dalla normative australiana.
Nello scenario generico, il costo dell'elettricità è compreso tra i 67,9 e i 71,9 USD/MWh, mentre per l'impianto di riferimento è 49,5 USD/MWh. Da questo punto di vista, le configurazioni con le migliori prestazioni sono associate ai casi 1 e 4, seguiti dagli impianti con generatore di vapore a letto fluido.
Simili considerazioni valgono per il contesto australiano, in cui il costo dell'elettricità è, a seconda dei casi, compreso tra i 91 AUD/MWh e i 102 AUD/MWh, valori simili a quelli del caso generale, tenendo conto del cambio. Aumenta però la componente legata al costo d'investimento e diminuisce quella collegata ai costi variabili di O&M e di combustibile.
Nello scenario generale, il costo atteso della CO_2 evitata si aggira sui 26-31 USD/ton a seconda dei casi, escludendo i costi per il trasporto e lo stoccaggio della CO_2, essendo il valore più basso ottenuto dalla configurazione di base. Anche nel contesto australiano, il caso base offre il migliore risultato economico in termini di CO_2 evitata (62 AUD/ton), ma vi è maggiore differenza tra i vari casi, con i valori massimi associati ai casi in cui viene alimentata lignite (76-77 AUD/ton).
In entrambi i casi, il caso base ha una performance economica migliore del caso analogo con ricircolo completamente asciutto, dato che sia la componente del costo dell'elettricità legata ai costi d'investimento che quella associata al rendimento sono superiori nel caso di ricircolo asciutto. Quest'ultima soluzione è perciò meno interessante, sebbene la differenza non sia notevole.
Inoltre, gli impianti con combustione a letto fluido sono leggeremente meno conveniente del migliore caso a polverino con analoga alimentazione, anche se la differenza è piccola.

Si osserva invece come l'inserimento di un FGD che tratti la piccola corrente del ricircolo primario causi un significativo deterioramento della performance economica, dato che sia i costi d'investimento che i consumi di combustibile aumentano.
E' dunque interessante valutare come le soluzioni a letto fluido possano diventare più vantaggiose di quelle a polverino di carbone, qualora fosse necessario ricorrere a sistemi di rimozione degli ossidi di zolfo, in misura dipendente dai requisiti di efficienza di rimozione.

Riassumendo, questa valutazione preliminare di impianti di ossicombustione alimentati a carbone indica che questa tecnologia potrebbe rappresentare un'opzione valida per la cattura dell'anidride carbonica da impianti di produzione di potenza elettrica. L'impatto ambientale appare abbastanza limitato. In particolare, l'effetto degli inquinanti presenti nei gas liberati in atmosfera è piccolo. Vanno comunque valutate le conseguenze dell'iniezione della CO_2 (insieme a varie impurità) sul bacino di stoccaggio.
A livello economico, si presenta come una soluzione piuttosto competitiva per la produzione di energia elettrica con basse emissioni di gas serra.
Ulteriori informazioni saranno fornite dai test in corso e in programma.
Importanti avanzamenti potrebbero venire dalla ricerca, specialmente da quella nel campo delle unità di separazione dell'aria mediante membrana (OTM) e in quello della chemical looping combustion, che potrebbero rappresentare importanti passi avanti.

ABSTRACT

This study evaluates the technical and economic feasibility of oxyfuel coal-fired power generation with CO_2 capture.
Firstly, the mass and energy balance of six large scale oxyfuel coal-fired plants have been estimated through the GS code, developed by Dipartmento di Energia at Politecnico di Milano, and Aspen Plus® Engineering Suite. The analysed cases differ in SO_x control systems, fuel feed (bituminous coal and lignite), boiler configuration (PC and CFB) and recycle features.
The expected range for LHV efficiency is 30.8-37.4% and the best result is associated with the CFB combustion plant fed by bituminous coal.
Some of the obtained values have been used for the economic analysis, which aims at calculating the cost of electricity and the cost of avoided CO_2 for all the considered cases for a general scenario and for a more specific Australian setting, following the procedure of the CO2CRC adapted to consider this technology. The costs have been estimated even for plants with lower net output and/or equipped with different SO_2 control systems, using approximate correlations.
The cost of electricity in the general scenario is 67.9-71.9 USD/MWh and the cost per tonne of CO_2 avoided is 26-31 USD/ton. These values exclude cost of CO_2 transport and storage. The configuration with the most favourable economic performance results to be the one with PC boiler, wet secondary recycle and black coal feed.
Similar conclusions can be drawn for the Australian setting. Indeed, similar values are attained for the cost of electricity, although the cost of CO_2 avoided is higher, because the COE for the reference plant is lower.
In summary, these preliminary assessment of oxycoal combustion plants indicate that this technology could be a viable approach for carbon capture from power stations, with an acceptable environmental and economic profile.

KEYWORDS: oxyfuel, oxycoal, power, Australia, CCS, coal-fired.

SOMMARIO

Questo studio valuta la fattibilità tecnico-economica di impianti di generazione di potenza elettrica con cattura della CO_2 basati sull'ossicombustione di carbone.
Si sono innanzitutto stimati i bilanci di massa ed energetici di sei differenti impianti di grande taglia tramite il codice GS e Aspen Plus®. I casi analizzati si distinguono tra loro per sistema di rimozione degli ossidi di zolfo, combustibile alimentato (carbone bituminoso o lignite), configurazione del generatore di vapore (a polverino o a letto fluido) e caratteristiche del ricircolo.
Il rendimento (rispetto al PCI) è compreso tra il 30,8% e il 37,4%, essendo il valore massimo ottenuto nel caso di combustione a letto fluido alimentata da carbone bituminoso.
Alcuni dei valori ottenuti dalle simulazioni sono stati utilizzati per l'analisi economica, che mira ad ottenere stime del costo dell'elettricità e del costo della CO_2 evitata per tutti i casi considerati sia nello scenario generale (rappresentativo di UE e USA), che in un contesto specifico australiano, seguendo la procedura del CO2CRC. I costi sono stati stimati anche per impianti più piccoli e/o dotati di desolforatore, tramite correlazioni approssimate.
Il costo dell'elettricità nello scenario generale è compreso tra i 67,9 e i 71,9 USD/MWh e il costo della CO_2 evitata tra i 26 e i 31 USD/ton, escludendo i costi per il trasporto e lo stoccaggio della CO_2. La configurazione caratterizzata dalla migliore performance economica è quella con generatore di vapore alimentato a polverino di carbone bituminoso, con ricircolo secondario umido.
Si possono trarre simili conclusioni per il contesto australiano, dove il costo del'elettricità è simile. Il costo della CO_2 evitata è invece superiore: il costo dell'impianto di riferimento è inferiore.
Riassumendo, questa valutazione preliminare indica che gli impianti a ossicombustione di carbone potrebbero rappresentare un'opzione interessante per la cattura della CO_2, avendo un profilo ambientale ed economico piuttosto favorevole.

PAROLE CHIAVE: ossicombustione, carbone, CO_2, cattura, energia elettrica, Australia.

1 INTRODUCTION

In this opening chapter, a basic description of global warming and its causes will be provided and an overview on the solutions that are being developed to counter it will be reported. In particular, options for carbon capture and storage will be examined.

1.1 GLOBAL WARMING

In the last years scientists have identified a pattern in the trend of the average temperature of the Earth's near-surface air and ocens. This has been increasing since the mid-20th century and is expected to follow the tendency in the next years. This phenomenon is commonly known as global warming.
At present, the majority of the scientific community believes that it is due to human activity. In particular, the main cause would be the increased emission of greenhouse gases (mainly carbon dioxide and methane), that is a result of anthropogenic processes. At the moment, the total input of these gases in the atmosphere is not balanced by an equal reabsorption by means of natural processes such as the chlorophyll photosynthesis.

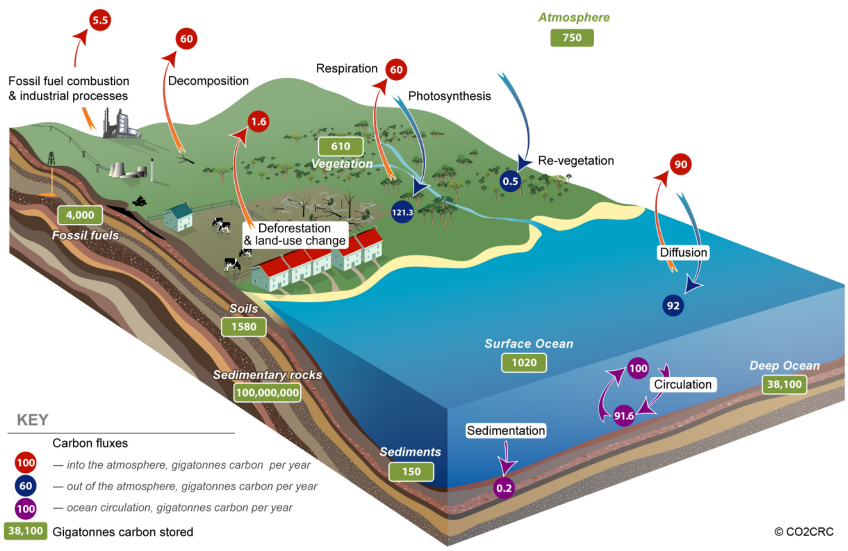

Figure 1. Schematic representation of the carbon cycle, courtesy of the CO2CRC.

The consequence is an increase in the concentration of these gases in the atmosphere, who are responsible for the retention of the heat that reaches the Earth and would be otherwise redirected toward the space.

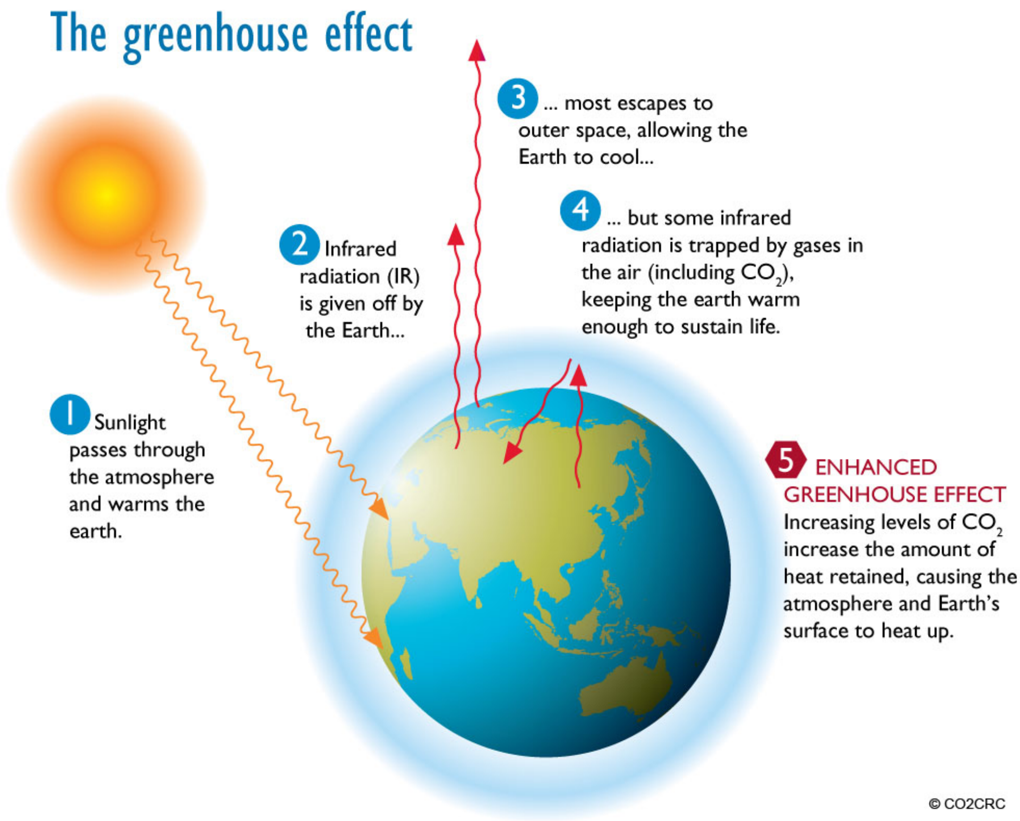

Figure 2. Schematic representation of the greenhouse effect, courtesy of CO2CRC.

The tendency of the temperature is also influenced by some positive feedbacks. the most important is the rise of the amount of water vapor (another significant greenhouse gas) connected to the warming.

The main studies about the global warming have been performed by the Intergovernmental Panel on Climate Change (IPCC), which publishes reports on topics relevant to the implementation of the UN Framework Convention on Climate Change (UNFCCC), basing its assessments on peer reviewed and published scientific literature. Significantly, its conclusions have been endorsed by some of the main scientific societies and academies of science, including all of the national academies of science of the major industrialized countries.

Moreover, IPCC and former vice-President of the USA Al Gore have been jointly awarded the 2007 Nobel Peace Prize.

1.2 ENERGY MARKETS

Two tendencies influence the energy consumption of an economic market. The first one is the positive correlation between the GDP and the energy required by the economic system. Nevertheless this effect is attenuated in the developed countries by their inclination toward energy efficiency, which is fostered by the technologic knowledge and by the necessity to cut expenses, because the cost of labour (and often taxes) make the economic apparatus less competitive.
It is expected that world energy consumption will increase in the next years. In particular, it will probably be stable in the developed countries, however the share required by the developing countries (especially China and India) will likely increase significantly.
Moreover, in the developed countries the historical bent is toward a shift to energy sources and technologies that make possible a reduction in emissions of pollutants and carbon dioxide, such as the renewable energy sources. On the contrary, in the developing countries the need of cheap energy to boost the economic growth is so strong at the moment that cannot be balanced by the concerns about health, local environment and climate change.
So the coal will remain an important item in the medium-term energy statistics, and this will involve an important challange to be faced.
In fact, nowadays global warming is one of the most important issue in the agenda of the major governments and the general attention toward this question is modifying common behaviour of normal people and strategies of the companies.
International treaties have been adopted (example are the Kyoto Protocol and the so-called 20-20-20 European plan) and others have been proposed to reduce, stabilize or control the greenhouse gas emissions and single States are implementing and studying policies that have that task.
Power generation is one of the sectors that will change more because of the new policies aiming at a reduction of greenhouse gas.
Coal is the most used fossil fuel for power generation and it is also the one emitting more specific CO_2 per energy unit produced. It is rensponsible of about 75% of the total CO_2 emissions. To reduce carbon emissions from power sector, a few strategies can be pursued (e. g., plants efficiency, switch to low or zero emissions sources), but these cannot guarantee a considerable decrease with accettable economic penalties.

Coal will likely continue playing a leading role in this sector, because it is cheap, abundant, quite well-distributed and it relies on a mature technology that is improving the pollutant emissions control systems.

1.3 RESPONSES TO GLOBAL WARMING AND CARBON CAPTURE AND STORAGE

The responses to global warming can be divided into three main categories: mitigation of the causes and effects of global warming, adaptation to the changing global environment, and geoengineering to reverse global warming.
Mitigation consists in the reduction of the emissions from human activities such as energy supply, transportation, industry, and agriculture.
Some of the measures proposed for adaption are: water conservation, water rationing, adaptive agricultural practices including diversification, construction of flood defenses, changes to medical care, and interventions to protect threatened species.
Geoengineering is the deliberate modification of Earth's natural environment on a large scale to suit human needs.
Carbon capture and storage (CCS) is a way to mitigate climate change by capturing carbon dioxide (CO_2) produced by the combustion of fossil fuels from large point sources such as power plants and subsequently storing it away safely instead of releasing it into the atmosphere. CCS is considered by the majority of the experts a viable method to control the CO_2 emission with an acceptable impact on the economic system.

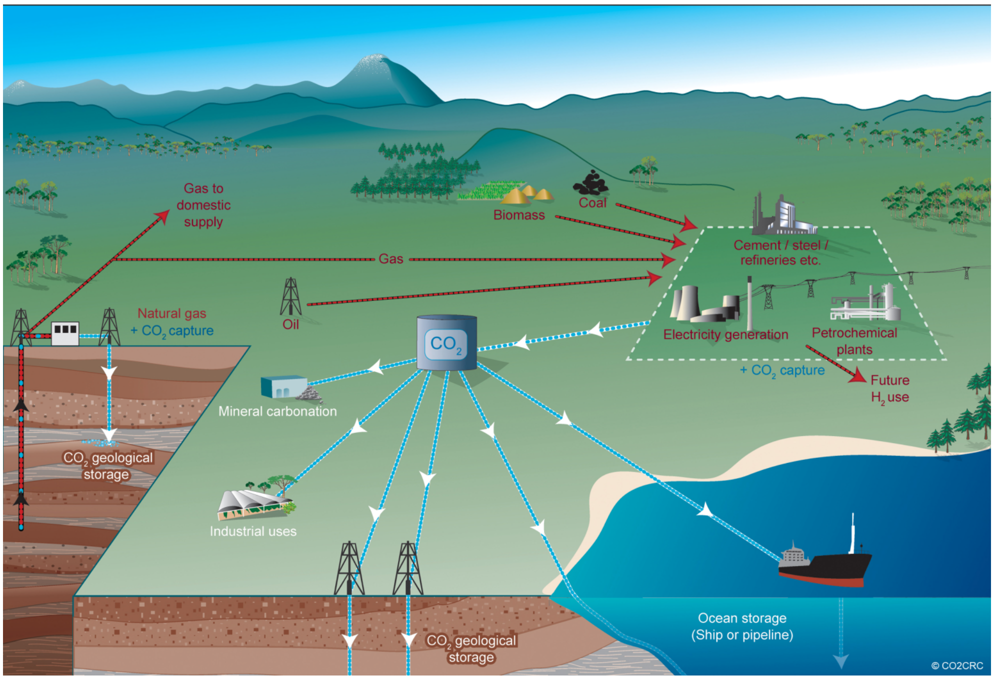

Figure 3. Schematic of the CCS process, courtesy of the CO2CRC.

The application of CCS technologies could reduce CO_2 emissions by approximately 80-90% in comparison to a plant without CCS.
The IPCC estimates that the economic potential of CCS could be between 10% and 55% of the total carbon mitigation effort until year 2100 (Section 8.3.3 of IPCC report [1]).
Carbon capture can be obtained in three different ways: post-combustion, pre-combustion, and oxyfuel combustion. A basic representation of these three processes is shown in the following figure.

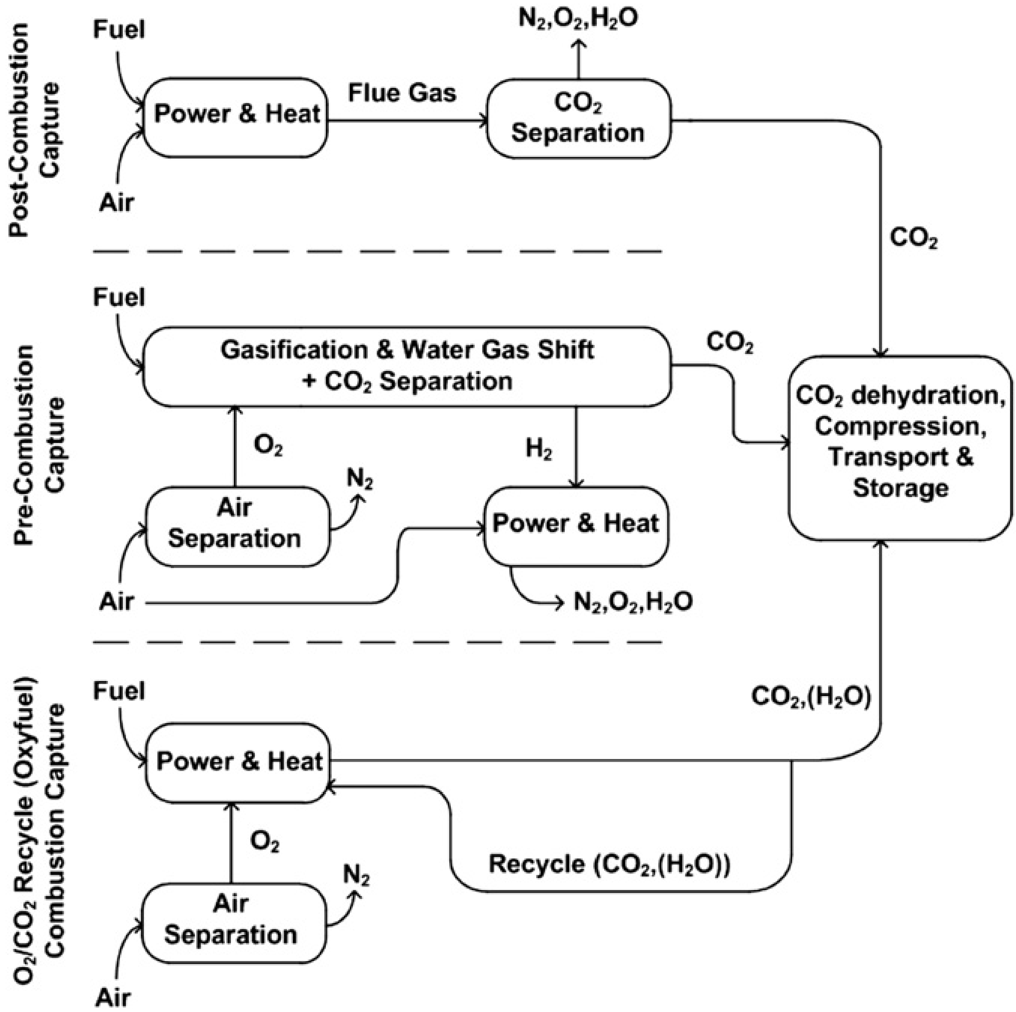

Figure 4. Possible plant configurations for the three main categories of carbon capture technologies. Adopted from [2].

In post-combustion capture, the CO_2 is removed after combustion of the fossil fuel. Several methods can be used to capture the dilute CO_2 from flue gases. The most common one is absorbing the CO_2 through a solvent (usually amine solvents). The CO_2 is therefore released after a change in temperature or pressure. There is a certain experience with some solvents, but no large scale plant is operating. Unfortunately, at the moment, the operating expenses related to the absorber and degraded solvents are still high.

Other post-combustion possibilities, currently being researched, include cryogenically solidifying the CO_2 from the flue gases, or removing CO_2 with an adsorbent solid, or by passing CO_2 through a membrane.

It is possible (and quite easy) to apply this approach to existing plants without significant modifications to the original plant (retrofit).
In pre-combustion capture, the fossil fuel (natural gas or gasified coal) is partially oxidized, for instance in a gasifier. The resulting syngas (CO and H_2) is shifted into CO_2 and more H_2. The resulting CO_2 can be easily separated and stored, whereas the H_2-rich stream can be used to produce electricity (producing mainly vapour as by-product) or treated to supply H_2 as energy vector. It is a industrially proven technology in oil refineries and allows high CO_2 capture efficiencies (>90%).
On the other side, this technology requires high investment costs, is associated with high NO_x emissions, has important problems with flexibility and uses H_2 to feed a gas turbine, requiring significant efforts for its design.
In oxyfuel combustion, the oxidant is oxygen instead of air. The flue gas resulting from the combustion is mainly composed by carbon dioxide and vapour. A relatively pure stream of carbon dioxide is thus attainable by condensing the vapour and separating the water.
An extensive description of oxyfuel plants with capture will be provided later in this work.

1.4 COAL AND RESEARCH ON CCS IN AUSTRALIA

Coal plays a pivotal role in the Australian economy. With regard to the domestic market, more than 80% of the Australian electric production is from coal-fired plants and the burning of coal produces more than 40% of Australia's greenhouse gas emissions.
Coal is also Australia’s largest commodity export. In particular, in the very last years more than 250 millions of tonnes have been exported yearly, for a total value of about 60 billions of Australian dollars (about 45 billions of USD) in 2008-2009. This impressive amount becomes even more weighty, if we consider that the estimated present Australian population is only 22.5 millions.
Australia is currently the world's largest coal exporter. Coal is mainly exported toward Asia (about 89%) and particularly toward Japan (about 40% of the total coal export).
In 2006, Australia had around 72 billion tonnes of identified in situ black coal resources, enough to last about 180 years at current rates of production. Of these, 39.6 billion tonnes were classified as economically recoverable, with over 95% of these resources in New South Wales and Queensland. In the same year, Australia's economically recoverable brown coal (brown coal is the ISO name for low-rank coal) resources were reported to be 37.3 billion tonnes, all of which is in Victoria and with over 90% in the Latrobe Valley.

Since coal is so important for Australian economy, the Australian Government and agencies and companies are engaged in supporting and coordinating the efforts to develop reliable and cheap technologies to reduce the CO_2 emissions from coal-fired plants.
The main example of this committment is probably the Cooperative Research Centre for Greenhouse Gas Technologies (CO2CRC), where the most of this work has been carried out. It represents one of the world's leading collaborative research organisations focused on carbon dioxide capture and geological sequestration.
CO2CRC is an unincorporated joint venture comprising participants from Australian and global industry, universities and other research bodies from Australia and New Zealand, and Australian Commonwealth, State and international government agencies. Its resources come from the Federal Government Cooperative Research Centres Program, other Federal and State Government programs, CO2CRC participants, and wider industry [3].

Figure 5. Current CO2CRC participants.

Not only it examines and develops technological opportunities, but also performs economic and evaluations and backs the uptake of the CCS solutions.
In Australia, many interesting CCS projects are ongoing or being planned. For example, the CO2CRC is engaged in the CO2CRC Otway project, Australia's first demonstration of geosequestration and the world's largest operational monitoring geosequestration research project. With respect to oxyfuel plants with CO_2 capture, the largest plant in the world, the Callide A plant, is currently being tested in Australia. This project made use of the advice from the Cooperative Research Centre for Coal in Sustainable Development (CCSD), that completed its term in 2008.
Other important projects on various areas exploring different solutions for capture and storage are in progress or under study in Australia.

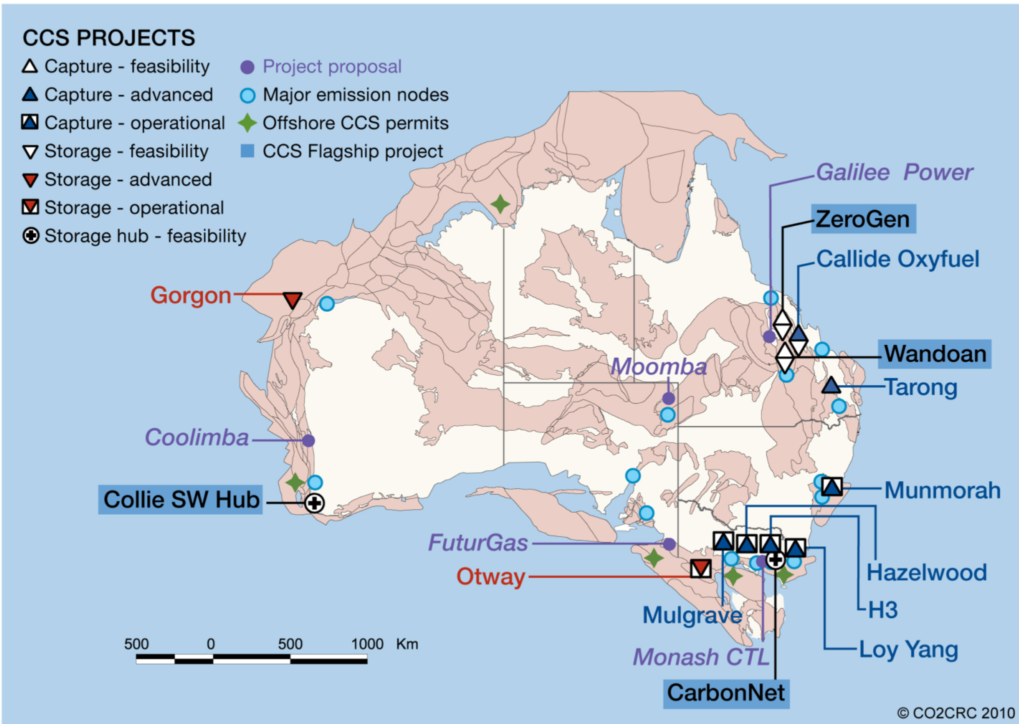

Figure 6. Map of the operational and planned CCS projects in Australia, courtesy of the CO2CRC.

Interestingly, the 2nd International Oxyfuel Combustion Conference will be held in Queensland, Australia from the 12th to 16th of September 2011.

1.5 AIMS OF THE THESIS

This work aims at performing a basic analysis of oxyfuel technology applied to power plants and to build a model to estimate the economic performance of this kind of plant. Simulation tools will be used to analyse mass and energy flows in a few significant cases. Then an economic procedure will be performed and those data obtained by the simulations will be used to estimate the cost of electricity and the cost of CO_2 avoided. Finally, the results will be discussed.

2 OVERVIEW OF OXYFUEL TECHNOLOGY

This chapter aims at providing a general overview on the generic features of oxyfuel plants with carbon capture. Specifically, the major components and systems will be described, evidencing the hurdles that this emerging technology is facing and some issues that are under investigation (such as recycle rates, boiler configuration, pollution control).

2.1 BASIC DESCRIPTION

In the figure below, a schematic of the mass and energy flows in an oxyfuel power plant is provided.

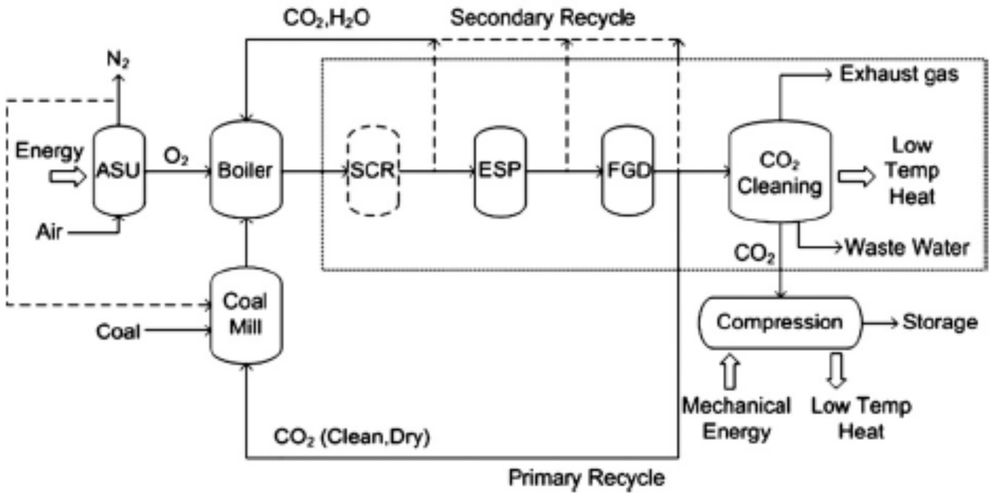

Figure 7. Possible schematic of an oxyfuel power plant. Adopted from [2].

As shown in the figure, the combustion process in an oxyfuel plant requires an oxygen-rich stream, recycle gas and fuel.
The oxygen-rich flow (usually 95% vol. or higher) is obtained by means of the separation of the air in an air separation unit (ASU). In this process, it is produced a nitrogen-rich stream that has no utility in the power plant (except for the particular case of pressurized advanced ASUs, where it is expanded in a gas turbine [4]).
Both the fuel properties and the limitations of steam and metal temperatures of

the boiler require a moderation of the temperatures during combustion and in the downstream heat transfer sections. Moreover, it is useful to achieve a boiler heat transfer profile similar to that of air-fired mode in the case of retrofit. Therefore flue gas recirculation is required. The interdependencies among fuel properties, the amount and temperature of the recycled flue gas, and the resulting oxygen concentration in the combustion atmosphere are being investigated. A brief description of the possible options is provided later in this chapter.
The flue gases exiting the boiler are then treated to remove fly ash and, if it is the case, sulphur oxides, nitrogen oxides, mercury and other pollutants. The amount of impurities in the gas and their effects on the operations are object of various studies, as well as the possible treatments for their removal. The main results are reported later in this work.
The gas that is not recirculated is cooled and the condensed water can be easily segregated. The resulting stream is a mainly composed by CO_2 and can be therefore treated, compressed, transported in pipes and then injected in a fit storage site.
Implementation of oxy-fuel technology requires significant modifications in the plant configuration and is associated with a lower overall net efficiency (8 to 12 percentage point), that corresponds to a 21 to 36% increase in fuel consumption [5].

In summary, carbon capture by means of oxyfuel combustion is a promising technology, that has the potential to become sufficiently mature for large scale plants, if the tests in the pilot plants will give satisfactory results. It should be noticed that most of the technologies and equipment in oxyfuel power plants have already been developed for conventional power plants and chemical plants. They usually need just minor adjustments for the new operation mode and are therefore considered to be quite reliable.
The CO_2 capture process in an oxycombustion plant is fairly easy, because the flue gas is mainly composed of CO_2 and water. The latter one can be readily segregated. If the non-condensables are removed, CO_2 purity can approach 100%. Some carbon dioxide is removed and released with the non-condensables, but the capture efficiency is still high (>90%).
A great advantage of oxyfuel combustion is the good enviromental profile of the combustion process. In fact, the features of the combustion allow a reduction in the production of nitrogen oxides and make the removal of mercury easier, in comparison with the air-fired case.
Moreover, if the co-storage of NO_x, CO, unburned hydrocarbons and SO_x turned to be feasible, the economic implications would be really sizeable and the design would be much more simple.

At present, only several small and laboratory scale oxyfuel plants are active. The most important ones are:

- the Schwarze Pumpe plant, a Vattenfall project [6, 7], that commenced operations in August 2008 and aims to investigate the whole chain of an oxyfuel combustion power plant in a pilot scale (30 MW_{th}). Two recycle streams are used and the performances of air pollution control devices are being tested, as well as different recycle configurations and fuel type feed (black coal, brown coal, biomass). A schematic of the plant design is presented below.

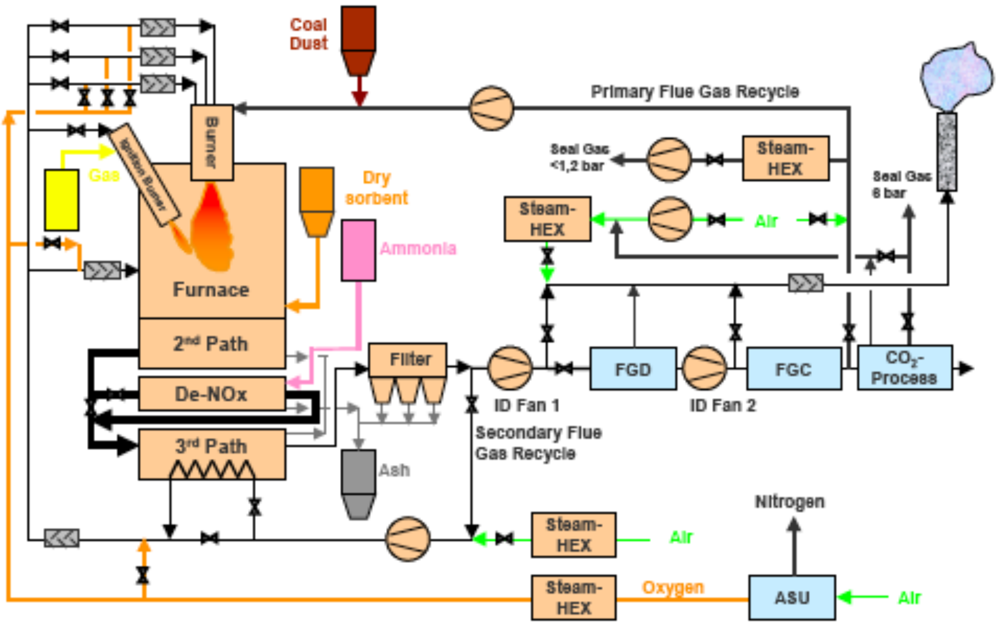

Figure 8. Schematic of the Schwarze Pumpe Plant. Adopted form [7].

- the Alliance (Ohio) plant, a Babcock & Wilcox project [8], which is aiming at verifying the process parameters for full-scale design. The basic design of this plant is reported in the figure below.

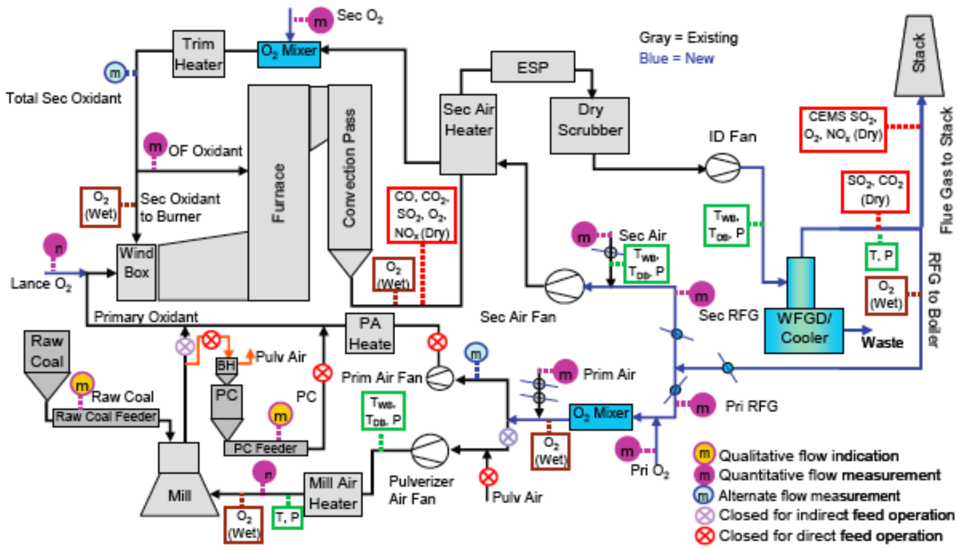

Figure 9. Schematic of the Alliance power plant [8].

Satisfying results regarding combustion, pollutants, air-leakage, switch between air- and oxy-fired mode are being published by groups working on these and other operating plants, proving that oxyfuel technology is feasible and encouraging larger scale testing sites.
Other plants are starting the operations soon. In particular, the following projects are note-worthy:

- the Callide A plant, a CS Energy project, that will be the largest operating oxyfuel plant (30 MW_e). One of the goals is demonstrating the near zero emission feasibility.
- the Ciuden project, a Spanish project, that will study a small scale oxyfuel CFB boiler.
- The PowerGen clean-coal power program and carbon dioxide storage network in Illinois, that will cost more than one billion USD (one billion USD has been awarded by the US energy department in mid 2010) to retrofit the conventional coal-fired Ameren's Meredosia Unit 4. The plant is expected to become operative within 2015 and is expected to supply 275 MW_e.

2.2 *POSSIBLE CONFIGURATIONS OF THE PLANT*

There are lots of uncertainties about the design and performance of oxy-fuel plants. Several possible plant configurations have been proposed. We can identify several main systems, as shown in the figure below: the air separation unit (ASU), the boiler, the steam turbine cycle, the flue gas (FG) cleaning and recycle and the CO_2 processing unit (CPU).

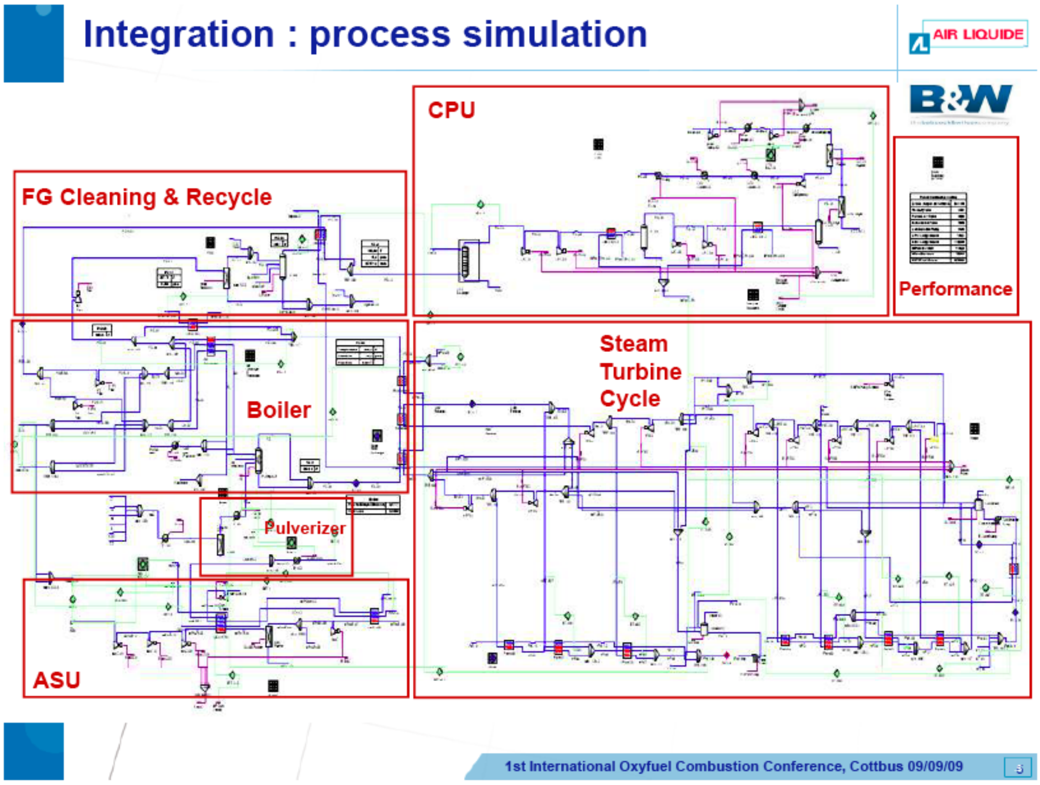

Figure 10. Schematic of an oxy-fuel power plant.

The ASU produces the required oxygen-rich stream and can provide low temperature heat to the preheaters of the steam cycle. Several technologies are being investigated for the ASU. The most mature one is the cryogenic compression, while promising solutions could emerge in the future, such as the oxygen transport membranes (OTMs, also called ion transport membranes) or the advanced ASU configuration [4]. They would hopefully guarantee better performances and lower costs. Many studies are under way in this field, since the ASU is important in building both capital and operational expenses and some researchers believe that it is possible to achieve sizeable savings.

Two main options exist for the boiler, which are the pulverized coal (PC) and the circulating fluidized bed (CFB) configurations, with different implications (especially on the coal drying process), performances and emissions. Obviously the boiler tranfers the large majority of the heat used in the steam cycle section for steam production.
The steam turbine cycle section is very similar to the one in the air-fired case about which a substantial literature exists.
Most of the uncertainties about the design of the plant concern the flue gas clening and recycle section, mainly linked to issues about polluting emissions, coal drying process, heat recovery and corrosion. The problem of the pollutants is particularly important to choose configuration of the plant. Some researchers argue for efficacy, safety and political acceptability of co-storage of NO_x and SO_x with the CO_2-rich gas flow [1]. Further studies are needed to confirm the feasibility of this solution, that would result in a simplification of the FG cleaning and recycle section and in consistent savings.
If the co-storage will turn to be impossible and the SO_x and/or NO_x emissions without control are exceeding the limits, two main possibilities are being considered: the abatment of the polluting emissions in the FG cleaning and recycle section by means of an FGD unit and a SCR unit or the sour compression, a promising procedure being developed by Air Products [9], that would be performed in the CO_2 processing unit. The former solution is considered as reliable because of the unusual composition of the flue gas (although some doubts are risen about it [2]), whereas the latter is still under evaluation.
In the CO_2 processing unit (CPU), water, non-condensables and other compounds with boiling point higher than CO_2 and a certain amount of CO_2 itself (depending on technology) are removed and the CO_2-rich stream is compressed [10]. Three different streams are obtained: waste water (that is usually quite acidic and must be treated before disposal), vented gas (CO_2, Ar, N_2, O_2, CO, others) and the CO_2-rich stream to be stored. In this section some low-temperature heat is also obtained from the cooling of the stream and from the intercooled compressors.
The separation of non-condesables can be obtained by means of simple flash units or using a distillation column. A conceptual schematic of a conventional CPU is shown in the figure below.

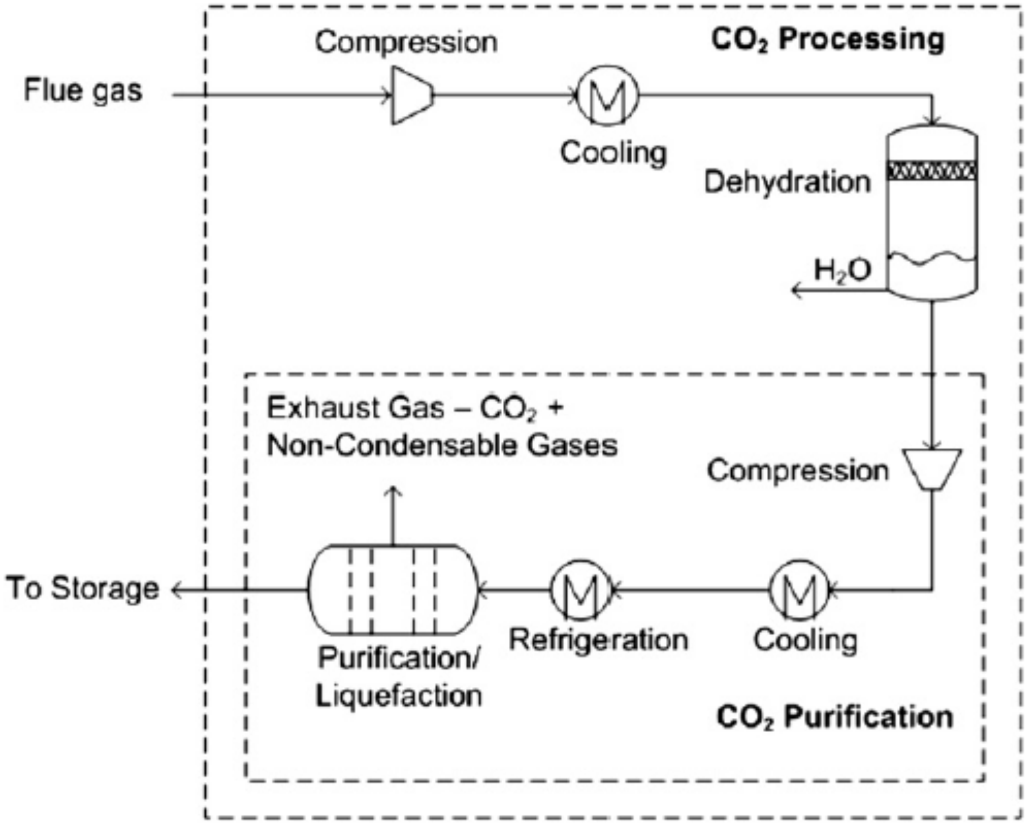

Figure 11. Possible configuration for the CPU [2]. Adapted from [10].

It is noteworthy that Air Products Inc. has proposed a process, the so-called sour compression, that removes both non-condensable impurities as well as Hg, essentially all SO_x and about 90% of the NO_x through the production of waste water containing sulfuric and nitric acid. Technical feasibility tests and economic evaluation of this solution are ongoing.

Heat exchangers are used throughout the plant to recover heat to be used mainly in the coal drying process and in the steam turbine cycle, leading to an increase in the efficiency. The heat integration level is chosen on the basis of an economic optimization. Indeed, if the recovery rate is increased, the fuel consumption (and therefore the operational expenses) decrease, but a larger investment is needed.

In the following sections, more specific descriptions are provided for the main components/systems and for the issues concerning recirculation and CO_2 purity.

2.3 ASU

The current best available commercial technology for air separation is the cryogenic distillation unit. Because of mechanical limitations, the larger designed units can produce 7000 t/d of oxygen [11] (although this limit could be raised to 10000 t/d in the future [12]), while the largest working units can yield 4500 t/d. These production rates are not sufficient to supply the amount of oxygen required by a hypothetical large size oxyfuel plant (with a net output in the range of 650-750 MW_e), so 2 or maybe even 3 parallel units would be necessary, at least. Indeed, such a plant would require about 15000 ton/day of oxygen.
In the 1990s, cycles for the production of low purity oxygen at 95% vol. were extensively developed for 2 applications: gasification and oxygen enrichment of blast furnace vent streams.
The cycles developed in the 1990s were not adapted for oxy-combustion, mainly because they were optimized to produce relatively high pressure oxygen (in the 5-80 bar range) and sometimes to co-produce a nitrogen-rich stream.
In the last years, ASU development plans to develop an Air Separation Unit optimized for oxycombustion have been started.

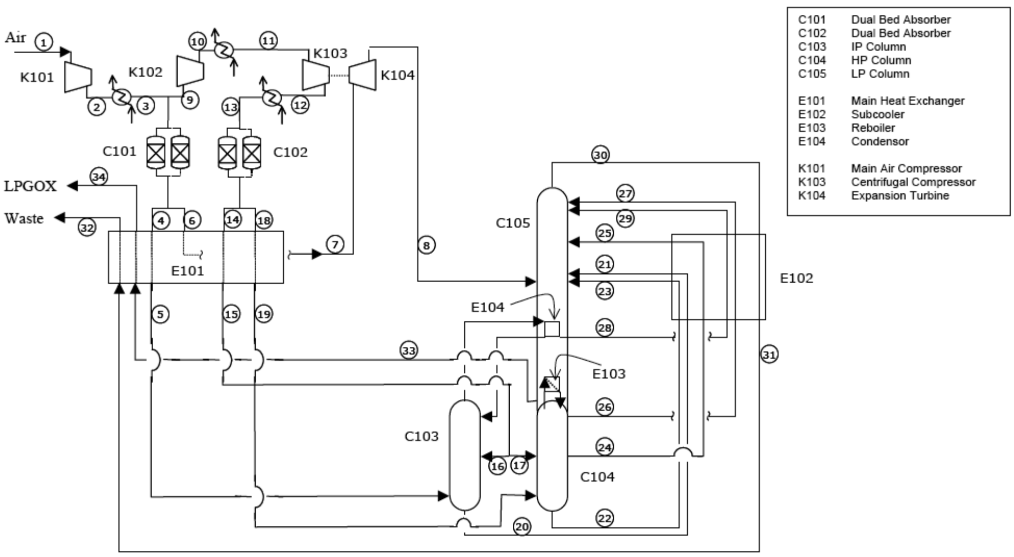

Figure 12. Schematic of a cryogenic air separation unit, adopted from [11].

A schematic of a 95% vol., 1.6 bar O_2 plant for use in an oxyfuel coal fired boiler is reported in the figure above [11, 12]. An air tream is compressed to 3.5 bar and part of it is further compressed to 5.3 bar and used to evaporate the liquid O_2 product stream. The heat recovered from the adiabatic compressors is used for condensate and boiler feed water preheating. The intermediate pressure column (C103) operates at 3.2 bar and produces nitrogen which boils low purity O_2, whereas the HP one (C104) operates at 5 bar pressure boiling higher purity O_2 at the base of the LP column. Part of the condensed N_2 from the condenser/reboiler E103 provides the top reflux stream to the LP column, while impure reflux N_2 from the condenser E104 provides an intermediate reflux stream. Finally, the LP column (C105) has 4 feed points, two reflux points and two reboilers.

The cost of produced O_2 is up to 70% power and 30% capital costs. In recent years capital and operating costs have been reduced through improvements in process and equipment design. Thus far, ASU designs requiring as low as 160 kWh/t have been developed [4]. However, as a general rule, the consumption of the ASU is chosen on the basis of a trade off between capital and operating expenses. The reported capital cost for an economically optimized ASU is about 4000 $/kW of ASU consumption.

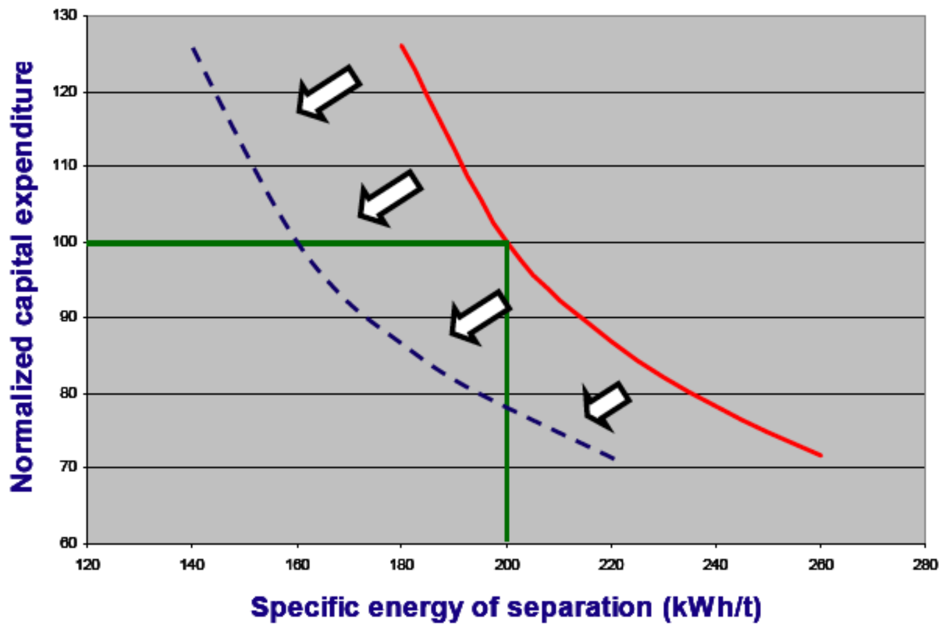

Figure 13. Tradeoff between capital and operating expenses in ASU design [4].

This consumption level is associated with an efficiency penalty of 6-9% [1, 13]. Possible improvements can be obtained by means of an ASU integration with the power plant (by using the low temperature heat from intercooled ASUs for

boiler water feed preheating) or by other improved cryogenic ASU technologies [4, 12].
In some cases, a reduction of about 10% in power consumption of the ASU could be achieved, according to studies performed by Air Liquide, as shown in the figure above, tranferring heat from the intercooled compressor(s) to the steam cycle. Anyway gains are very dependent on the conditions of the specific plant, such as ambient conditions, efficiency of the steam cycle, cooling system (dry versus wet), coal type (water and sulfur content) etc..

An alternative very promising development could the oxygen (or ion) membrane technology (OTM, or ITM). It uses mixed metal oxide ceramic selective membranes operating at >700°C integrated with a high temperature and pressure air source (usually a gas turbine). In the figure below, an oxyfuel plant with integrated OTM ASU is represented.

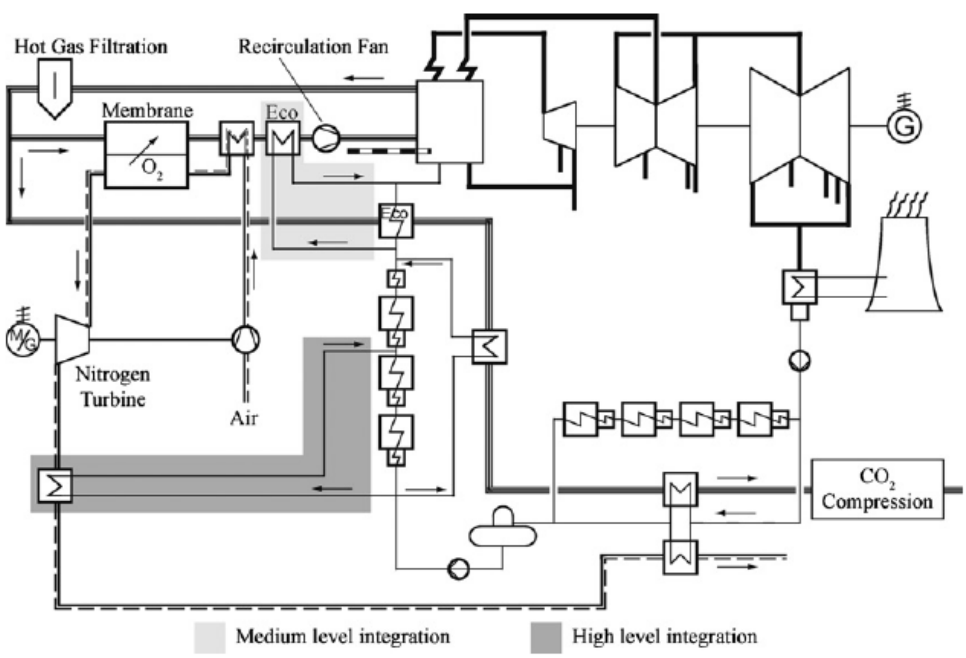

Figure 14. Schematic of an oxyfuel plant equipped with an integrated OTM ASU, adopted from [14].

Important features of OTM systems are:
- Feed and product must be free from particulates or solid products which would block the membrane surfaces.
- The membrane oxide material must be thermally stable and have stable crystal structures at operating temperature and over the entire oxygen

partial pressure operating range.
Suitable materials typically have a perovskite or a brownmillerite structure. The oxygen separation process can be divided into these main steps: dissociation of the O_2 molecule into two O_2^- ions at the feed side, bulk diffusion of the ions through the membrane and recombination of the ions at the permeate side. The ions transport mechanism is based on the presence of oxygen vacancies in the membrane sieve.
This technology could represent a breakthrogh in the development of oxyfuel plants, if it turned to be technically and economically efficient. To date, no membranes satisfied the requirements, but researchers envisage possible interesting advances.
Preliminary economic evaluations indicate that the capital cost associated with OTM-based solutions is lower than in a plant with cryogenic ASU.

It is useful to recall that the ASU is fundamental for the plant operation, so availability (that is estimated to be higher than 98% [11]), start-up speed and the performance at reduced load are important factors that should be taken into account.
In particular, long start-up, shut-down and ramping procedures are associated with the cyogenic ASUs, whereas the new-built coal fired power stations would probably require a ramping capability of 5% flow/minute at all operating levels. If long periods of unavailability are not acceptable different solutions can be considered and adequate high reliability O_2 production with constant purity are to be installed. The most simple solution is the installation of more parallel units: the availability improves, but the capital costs increases. It is also possible to design the plant in order to allow it to work in air-fired mode without capture, although this would increase the cost of the plant and could not be economically convenient and is not considered as a promising possibility [12]. Otherwise an external backup source of oxygen connected by pipes or a storage of liquid oxygen (with an instant demand vaporizer system) able to provide the quantity of oxygen that cannot be provided by the ASU (a few tens of tonnes) can be installed, but this solution would imply more costs and safety risks, as most of the materials become inflamable in oxygen-ernriched mixtures in the usual atmospheric conditions.
The most likely solution is the adoption of a few of the listed methods. The problem is less critical for OTM ASUs, which are expected to ramp more quickly (about 2%/min).
For economic reasons it should be estimated whether the massive amount of almost pure nitrogen at low temperature could be used in the plant or externally.

2.4 BOILER CONFIGURATION

Two possible configurations are being considered for the boiler:

- pulverised coal (PC) boiler, as shown in the figure below. If a PC configuration is adopted, the stored coal is dried by means of the primary recycle stream and ground in a mill. The small coal particles are therefore injected in the furnace and burned. The basic idea of a firing system using pulverised fuel is to use the whole volume of the furnace for the combustion of solid fuels. The bottom ash is removed at the furnace bottom.

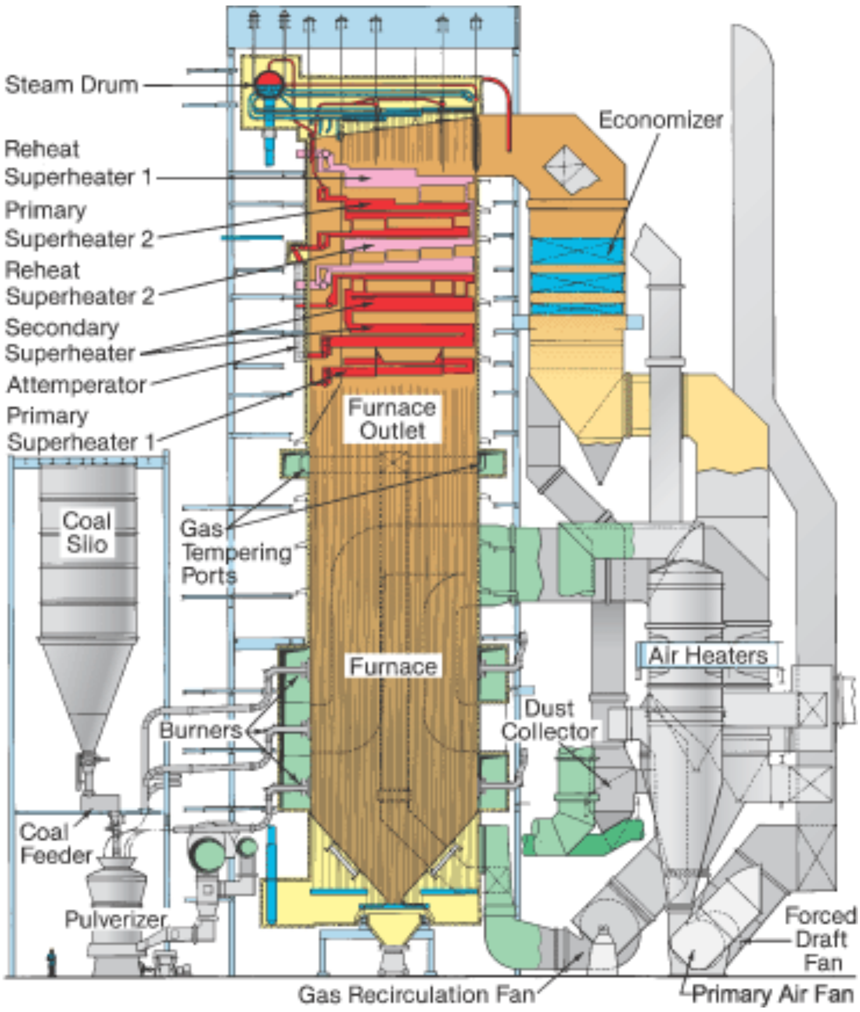

Figure 15. Schematic of a PC boiler and its auxiliaries (Babcock&Wilcox).

- circulating fluidized bed (CFB), that is schematically represented in the figure below. The solid fuel is injected in larger pieces (<25 mm) and is suspended on upward-blowing jets of air during the combustion process. A turbolent mixing of gas and solids, that helps chemical reactions and heat transfer, is obtained. This process warrants uniform furnace temperature and heat flux profile [15]. CFB plants can be fed with a wider range of fuel than PC plants. Virtually no drying process is needed, because of the long solid residence time. With respect to the PC plant, CFB plants show higher forced fans consumption, but lower induced fan consumption and no need (or reduced load) of FGD. In fact, in-furnace SO_2 removal can be performed by introducing limestone and extracting gypsum together with the bottom ash. Eventually slightly higher net efficiency is obtained for CFB boiler plants (about 0.3% gain [16]), which also allow good emission control for NO_x, CO and HC [17].

Despite the higher efficiency, CFB plants emit slightly higher specific CO_2 emissions, because of the CO_2 stream generated from $CaCO_3$ for sulphur absorption decomposition.

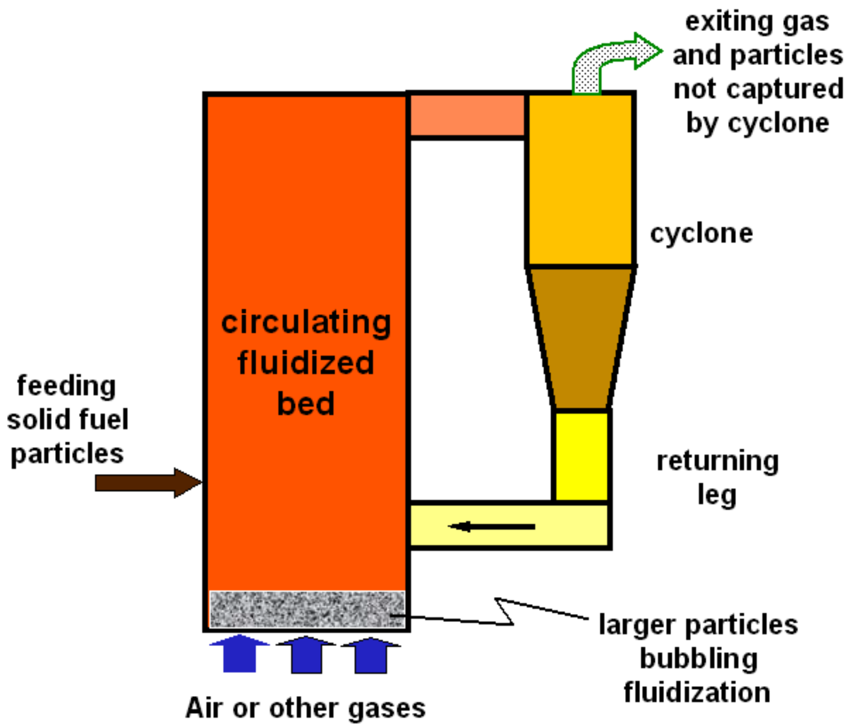

Figure 16. Schematic representation of a CFB boiler.

A more detailed discussion about SO_x and NO_x production in PC and CFB boilers is reported later in this chapter.
Economically, the CFB configuration seems to be superior, as it allows savings in both capital expenses (especially for the boiler, that is smaller) and, possibly, in fuel expenses. For example, a 63% reduction in furnace volume, a 36% lowering in furnace wall area, a 25% in hot loop heat exchange area and up to 8% of boiler island cost reduction have been estimated for 60% O_2 concentration with respect to PC [16].
The decrease in variable fuel costs is instead due to the possibility of using cheaper fuels and to the slightly better expected efficiency associated with CFB plants.
Other advantages of a CFB boiler are reduced risk of corrosion, fuel flexibility (enabling even co-combustion, for example with biofuels) and decrease in space requirements (due to smaller boiler and secondary flue gas cleaning systems) [16, 18], which can result in further economic gains.
On the other side, an economic disadvantage is that the economical value of the solids derived from CFB is lower than in the PC case: the ash and gypsum streams are not collected separately.
Nevertheless the CFB technology is much less mature than the PC one and only one CFB supecritical unit is active today (Łagisza, 460 MW_e [19]), that is giving good results in terms of availability and efficiency (45%) [15, 16]. According to the suppliers and developers no important hurdles should be able to stop a further scale up to 800 MW_e [16, 17].

2.5 EFFECTS OF IMPURITIES

Power plants must respect the requirements imposed by the competent agencies. Nowadays the main species generally controlled in US and European power plants by means of air pollution control devices are: fly ash, NO_x, SO_x.
No regulation exists about mercury emissions at the moment, but this aspect is worth to be pondered, because it rightly attracts attention from the public and has significant environmental relevance.
In Australia, no limits regarding NO_x and SO_x are set by the regulator and no abatement device is adopted in large power plants and the rules are not expected to change in the short-mid term.
In an oxyfuel plant, flue gas cleaning is no longer only used for air emission control, that actually represents a minor issue. Indeed, for a given impurity, the interactions within various downstream processes such as CO_2 compression, transport and storage must be carefully considered for the design of the flue gas cleaning processes [20]. On top of that, it is important to meet the operating requirements for the correct functioning of the components of the plant, with

particular attention to the boiler and some pipes.
The purity requirements in the plant and in the downstream sections and the real costs of the control systems have not been exactly identified yet. This aspect will play a key role in determining whether oxyfuel plants are an economically competitive option for CO_2 capture. In particular, the more the impurities demonstrate to compromize the functioning of the plant and of the downstream activities the more pollutions control systems are needed, adding additional costs and energy penalties [21]. Alternatively, the downstream systems must be designed to be able to handle high concentrations of impurities [22]. For example, if corrosive compounds are present, the pipelines must be made of more expensive materials or substituted more frequently.
A summary of the impacts of the different compounds on the plant and its emission is reported in the following table.

Table 1. Expected impacts of impurities and proposed control strategies, adopted from [20].

Component	Major impacts			Cleaning options
	On boiler systems	On CO_2 capture (including downstream processes)	On air emissions (after the flue gas cleaning and downstream processes)	
H_2O	Fuel handling, acid dew-pint (corrosion)	Corrosion, erosion (water droplet)	No	Condensation, dehydration
O_2	No significant change is expected compared to air-firing	Corrosion, non-condensable gas, oxidising hydrocarbon in EOR applications, change in redox conditions leading to dissolution/precipitation of minerals in storage sites	No	Physical separation, catalytic oxidisation, avoid air in-leakage
Ar	No	(Non-condensable gas) increased transport and storage volume and 2-phase behaviour.	No	Physical separation in ASU or in CO2 compression train, avoid air in-leakage
N_2	NO_x formation	(Non-condensable gas) increased transport and storage volume and the risk of 2-phase behaviour during transport/storage	No	Physical separation, avoid air in-leakage
SO_2	Corrosion, convert to SO_3	Corrosion, changing redox or pH conditions, precipitation of sulphur compounds in storage sites	Little	Wet/dry scrubbing, co-capture with CO_2
SO_3	Corrosion, fouling concerns	Corrosion	Little	Wet scrubbing
NO	Convert to NO_2	(Non-condensable gas) somewhat on gas purification	Reduced in mass, but maybe with higher concentration	Combustion control, oxidisation then wet scrubbing. Treatment in venting stream. SCR as final option.
NO_2	Corrosion	Corrosion	No or very little	Wet scrubbing
CO	Efficiency, maybe corrosion	(Non-condensable gas) somewhat on gas purification	No or very little	Physical separation
Cl	Low temperature corrosion	Corrosion	No	Wet scrubbing
F	Low temperature corrosion	Corrosion	No	Wet scrubbing
PM (ash)	Erosion, fouling	Erosion, fouling	No	ESP, fabric filters, wet scrubbing

The concentrations of flue gas components will not simply be defined by fuel properties and combustion. In fact, the concentrations of impurities may be significantly affected by flue gas recirculation and involved cleaning processes. It is also important to determine the changes required to adapt conventional pollution control systems or the downstream components to the new situation and the impacts on their costs.

The performance of the ESP in an oxyfuel plant is quite different compared to the one in an air-fired plant, as a potential change in the size distribution of the fly ash from oxyfuel combustion has been observed. Moreover the flue gas composition could influence the ion producion rate within the ESP, leading to a

variation in collection efficiency. It is likely that the ESP has to be positioned before recycling, to limit their accumulation in the furnace and erosion.
SO_x production, expressed in kg/kWh, in a PC oxyfired plant are related to the nature of the fuel and slightly increase in comparison with the air-fired case as a result of the decrease in the efficiency and the concentration of the SO_x in the flue gas is higher, because the nitrogen ballast is absent. In addition, an higher conversion rate of SO_2 to other sulphur species has been observed [23]. In particular, the table below indicates that higher oxygen concentrations result in a higher conversion of SO_2 to SO_3, causing an increase in the acid dew point, and higher levels of sulphation reactions in the ash.

Table 2. Comparison of SO_x emissions in air- and oxy-fired plants, adopted from [23].

Measured sulphur gas products.

	Coal A		Coal B		Coal C	
	Air	Oxy-fuel	Air	Oxy-fuel	Air	Oxy-fuel
SO_2 ppm, dry	178	552	387	1296	508	1612
SO_2 mg/MJ	150	112	319	219	416	291
SO_3 ppm, dry	2	7	9	18	3	11
SO_3 conversion	1.1%	1.3%	2.3%	1.4%	0.6%	0.7%
Acid Dew Point, °C[a]	115.8	137.0	133.4	147.5	120.2	141.7

None of the main configurations proposed for oxyfuel plants seems to bring about excessive SO_x emissions in the vented gas, because, as CO_2 and SO_2 have similar characteristics, it is hard to segregate SO_2 together with the other compounds. So, if necessary, other approaches are necessary.
Economically, the most attractive solution is the co-storage of CO_2 and SO_2. However, as stated previously, it is likely that a limit wil be necessary, since high SO_x concentration can cause problems during the transportation and storage processes. On top of it, uncertainties exist about the regulation that will be adopted.
If SO_x control is needed, two main ways are viable: flue gas desulphurization and sour compression.
The sour compression is a process being developed by Air Products to incorporate SO_x removal (and NO_x removal) into the compression stage of CO_2 purification [4], taking advantage of the higher partial pressures of the pollutants.
Tests are ongoing to evaluate the performance of a conventional FGD in the oxyfuel case. The first published results are revealing no noticeable change [8].
Moreover, the impact of desulfurization on economics is being investigated.

Recent studies state it is comparable to the cost of a conventional FGD for the same output [24], with a decrease in capital expenses and an increase in operating costs. However, a slight increase in oxygen and water consumption is expected.
If a high-sulphur coal is used, an FGD before the recycle could be necessary to limit the corrosion problems in the recycle ducts [25]. According to EPRI [26], if the SO_2 concentration in the flue gas is higher than 3000 ppmv, the recirculated flue gas (RFG) should first pass through an FGD unit which follows the particulate removal device, such as an ESP or a baghouse. Similar concentrations are obtained with coals with a sulphur content higher than 2%. Moreover, it is worth saying that direct SO_x removal with limestone addition or other calcium derivatives has interestingly shown a greater desulphurisation capacity in oxy-fuel.
The table below describe all the possible interventions to abate SO_x that can be adopted in an oxyfuel plant.

Table 3. Strategies to mitigate and control SO_x emissions, adopted from [23].

Mitigation and Control Strategies for SO_x.

Operation/ remediation	Location on Fig. 1	Notes
Use of low sulphur coal	Coal Supply	- Affects market value of low S coals. - Limited by low S coal availability. - Blending options dependant on process.
Use of high calcium coals	Coal Supply	- sulphur retention is coal/ash dependant
Wall soot blowing	Furnace, convection pass	Frequency may change with oxy-fuel conditions as ash deposition rate undefined
Limestone injection	Furnace	- Injection port placement essential. - Increases solid loading on ESP - Availability and price of calcium feed material
Sulphur scrubber	Gas recycle stream	Reduces overall SO_x level in furnace
Condenser + Cooler	Primary Feed Gas	Necessary if primary feed gas drops below acid dew point
Sulphur scrubber	Prior to compression	Avoids corrosion in compression circuit, may avoid higher fabrication costs
SO_x removal in compression	CO_2 compression circuit	Material selection needs to include acid resistance

In comparison with air-fired plants, the combustion conditions in a PC oxyfired plant allow a significant reduction (about 60-70%) in NO_x production [27, 28], that can be attributed to the absence of thermal NO_x. However, the concentration of NO_x in the flue gas of an oxy-coal combustion plant can be up to about two times to that of an equivalent air-coal combustion plant, because the flow of the flue gas is lower.
The increased concentration of SO_x in the flue gas could reduce the performance of the SCR, if needed. The sour gas compression abates NO_x as well as SO_x and represents a promising opportunity to reduce the risk of corrosion in the downstream section.
CFB combustion allows definetely lower SO_x presence in the flue gas, if in-furnace capture through limestone injection is performed. On top of it, even primary thermal NO_x emissions are probably lower than in the PC case, because of the relatively low and uniform temperature in the boiler [17].
Importantly, primary NO_x emissions can be controlled through good mixing and effective staging of the oxidant feed.

With regard to mercury emissions, tests showed an increase in the oxidized Hg/elemental Hg ratio during oxycombustion with coal. Oxidized Hg is more efficiently captured in the baghouse and FGD unit. This represents a significant environmental advantage and, possibly, a design simplification in the long-term (and, consequently, even an economic saving), since mercury emissions are likely to be regulated in future.

With respect to transportation the greatest concern involves the water content in the CO_2 stream. In fact, water and CO_2 can interact and cause so called sweet corrosion or form hydrates. Concurrent presence of water and SO_2 (or H_2S) in the CO_2 stream increase the risk of sulfuric acid corrosion. For these reasons, the water concentration in the CO_2-rich stream must be very low. The flue gas is dehydrated to a dew point 5°C below the temperature required for transport conditions to avoid corrosion problems. Others report no risk of corrosion at a dewpoint of less than 60°C [21].
Condensation alone would not probably be sufficient. Consequently a drying process is necessary, for example absorption in a recyclable dehydrant (triethylene glycol) in combination with the last compression step.

2.6 RECIRCULATION

Tipically, between 60 and 80% of the flue gas is recycled to the PC boiler [2, 29]. A primary and a secondary recycle can be distinguished. The possible positions of the recycles is represented in the figure below.

The primary recycle is used to dry and transport the fuel and is analogous to the primary air in air-fired plants. It is estimated that it should be about 20% of the total amount of combustion gases (recycle+O_2 flows) [1]. The secondary flow is instead equivalent to secondary, tertiary and overfire (if needed) air and is necessary to control the combustion temperature and preserve the integrity of the boiler.
To preserve an ESP and a flue gas fan, the recycled gas temperature should be between 200 and 350 °C.
A general consensus exists about the position of the primary recycle stream. It should be cooled, scrubbed, dried, reheated to about 250 °C (or more) and then injected into the mill [11].
The reheating is necessary, because the recycled stream must be able to dry the coal, while the scrubbing avoid corrosion problems in the mill, due to condensation of sulfuric acid.

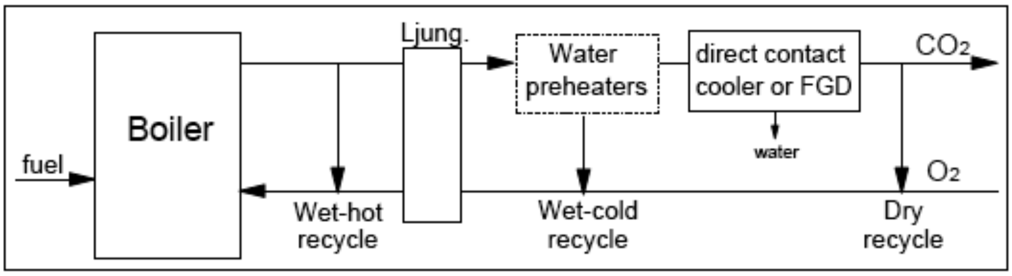

Figure 17. Simplified representation of the possible recycle positions in oxy-fuel power plants. Adopted from [30].

A great number of configurations (with different recycle rates and different positions where the gas are recycled) have been proposed. They result in different consumption of the recirculation fan, size of the heat exchangers and of the air pollution control systems and corrosion effects on ducts and components. Three possible approaches are possible:

- wet-hot recycle: part of the flue gas exiting the economiser is recirculated before gas-gas heater exchanger. This configurarion assures the maximum efficiency, but an ESP working in inappropriate conditions (about 300 °C) is needed, to protect the recirculation fan.
- Wet-cold recycle: a share of flue gas is recirculated after gas-gas exchanger, before water condensation begins. The efficiency is slightly lower and a larger preheater is larger than in the wet-hot case, but particulates removal system and recirculation fan work in more appropriate conditions.
- Dry recycle: the secondary recycle stream is recirculated after cooling and water condensation. Because of the increased amount of heat

transferred at lower temperature, an efficiency decay is inevitable, but the recycle fan consumption is reduced as the recirculated gas are cold. If a FGD is installed, the lower SO_2 concentration in flue gas involve a reduced corrosion in furnace and in low temperature heat exchangers. FGD itself (if needed) and low temperature heat exchangers must be larger, with negative consequences on the economic performance.

Moreover, the actual configuration of the recycles influence temperature, water content and concentration of pollutants (e.g, NO_x, SO_x, particulates) in all the points of the plant.

2.6.1 LIGNITE CASE

Conventional pulverised coal preparation/combustion arrangement for oxyfuel plants fed by brown coal is similar to that found in current pulverised coal plants burning high-moisture (for example about 60% by weight) brown coals (principally in Germany and Australia).

In oxyfuel plants fed by low-rank fuel, even a conventional all-dry recycle stream would not be able to provide enough enthalpy to sufficiently reduce the moisture level.

Indeed, for high-moisture subbituminous coal or lignite, which can produce highly reactive dust when dried, it is preferable to withdraw furnace gas from the boiler for use as drying agent [11]. For this arrangement, gas pressure in the mill is reduced by an exhauster fan situated between the mill classifier and the burner.

Indeed, if the fuel is brown coal, coal drying is usually obtained through recycle of hot gas (around 1000°C) directly from the furnace that contacts raw coal as it falls down a drying shaft into a beater mill acting also as circulating fan. Gas temperature falls rapidly in the shaft, and drying is completed in the mill to give an exit temperature of about 120ºC or higher and a coal moisture content of approximately 15% by weight. Indeed, the coal moisture level must be reduced to 10-20% to allow correct combustion in a PC boiler.

The maximum exit temperature of gas exiting the mill depends upon the drying agent: it is 200°C if recycled gas is used, whereas it is 100°C if an equivalent drying process is performed by means of preheated air [31].

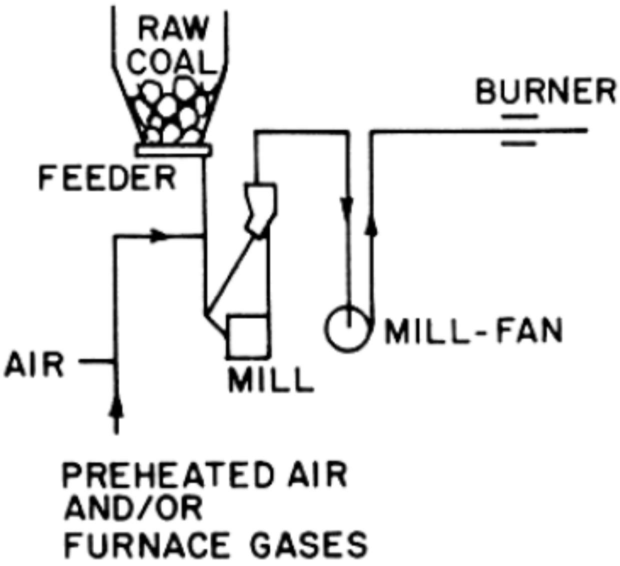

Figure 18. Schematic of conventional drying process for high-moisture fuel, adopted from [31].

A cyclone may be inserted in the pulverised coal/flue gas riser to inject a fuel-rich mixture at the main burner level, promoting stable ignition. The moisture-laden remainder flue gases are injected at a higher level, after combustion is established. The oxygen for combustion, apart from the small amount in the recycle flow, is all supplied by the secondary oxidant stream, which must be mixed with the primary coal/flue gas stream before combustion.

Two recycle streams would probably be used: a "very hot" one used to dry the coal and a hot one (that could even be in excess sometimes) to control the combustion temperature.

Coal drying by means of furnace gas recirculation improves the grindability of the coal and may assist combustion properties, although it does not increase the plant efficiency.

Other innovative methods are being investigated for the drying systems of air-fired plants with low-rank fuel feed and could be suitable for oxyfuel combustion plants too. They are supposed to improve the global efficiency and reduce pollutant and greenhouse gas emissions with subsequent economic benefits that would exceed the additional costs of the drying system.

In particular, other drying schemes of particular note have been proposed [31, 32]: the WTA system (a German acronym for "fluidized-bed dryer with integrated waste heat recovery"), the MTE (Mechanical Thermal Expression) system (Germany/Australia), the Steam Drying (SD) and the Hydrothermal Dewatering (HTD) and others.

2.7 CO_2 PURITY

The exact requirements for the quality of the CO_2 stream for different storage scenarios are not fully clarified yet. A number of suggestions for purity requirements can be found in the literature. Limits for individual species are set for the following reasons: health, safety and enviromental aspects in case of leakage, economic considerations, minimum miscibility pressure, risk of corrosion, materials integrity, legal aspects, storage integrity, operational issues, etc.. Some of the quality specifications expected in different studies are reported in the table below.

Table 4. Suggested CO_2 quality specifications from different sources. Adopted from [2].

Parameter	Modest quality, aquifer storage	High quality, on-shore storage	U.S. Specifications	Saline formation
	Anheden et al. [76]		Lee and Miller [77]	Fout [78]
Pressure	110 bar	110 bar	–	150 bar
Temperature	50 °C	50 °C	<50 °C	–
CO_2	>96 vol%	>96 vol%	>95%	Not limited[b]
H_2O	<500 ppm	<50 ppm	<480 ppmv	150 ppmv
N_2, Ar	<4 vol%[a]	–	<4%	Not limited[b]
O_2	<4 vol%[a]	<100 ppm	<10 ppm	<100 ppmv
SO_2	<200 mg/Nm³	<50 mg/Nm³	–	<3 vol%
H_2S	–	–	<10–200 ppm	<1.3 vol%
NO_x	–	–	–	Uncertain
NH_3	–	–	–	Not limited
CO	–	–	–	Not limited
CH_4	–	–	–	<0.8 vol%
HC's	–	–	<5%	<5 vol%
H_2	–	–	–	Uncertain
Glycol	–	–	<0.04 ppmv	–

[a] Sum of N_2, O_2, and Ar should be <4 vol%.
[b] No limit but the impacts on compression power and equipment cost need to be considered.

However, precise requirements will be probably determined for each case of capture and storage [2].
With respect to economic considerations, the preferable option is to co-store as many impurities as possible. This is not true for the non-condensables, for which an optimum concentration exists, since their costorage causes an increase in energy and reservoir size for the storage as well as in capital and operating expenses for the transport, while their separation involves capital and energy penalties. This is the reason why non-condensables are removed in the CPU. Usually the set cumulative concentration of oxygen, nitrogen and argon is lower than 4%.
The purity of the O_2-rich stream provided by the ASU should be identified with a trade off between the costs of transport and liquefaction and the cost of air separation. If no air ingress is considered, an oxygen purity of 97.5% is associated with the lowest overall power consumption [2], as shown in the figure below. Production of O_2 purities higher than 97.5% otherwise requires the more difficult O_2/Ar separation with up to double the number of separation stages in a distillation column [12] and are therefore avoided.

Some studies that take into account air leakages report optimum purities of about 95% [2]. At an oxygen purity of 95% (that can be obtained by cryogenic separation), the main impurities are argon (3-4% mol.) and nitrogen (1-2% mol) [25].

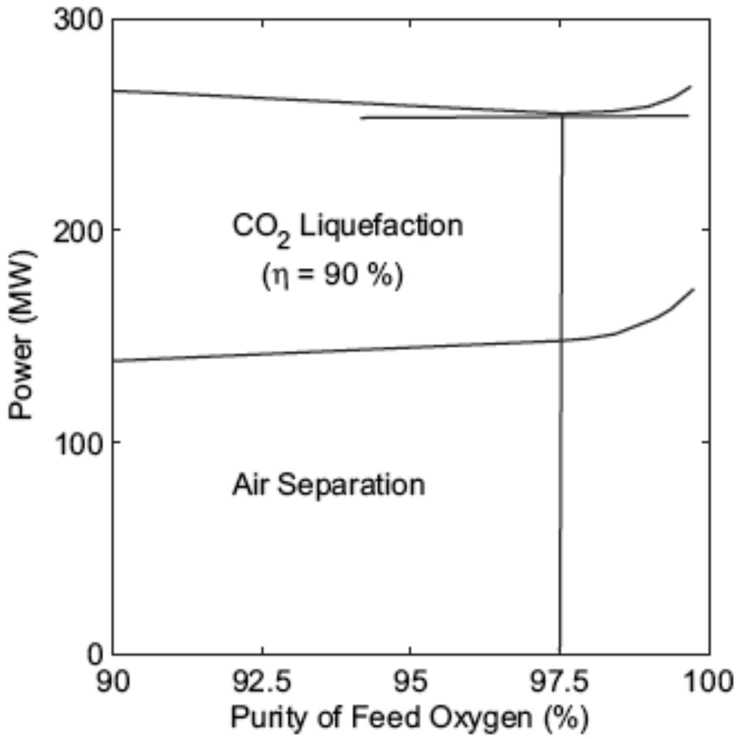

Figure 19. Relationship between the purity of oxygen and the consumpion for air separation and CO_2 compression, in absence of air in-leakage. Adopted from [2].

Technically, two other main issues deserve close examination.
Firstly, the risk of corrosion and of structural changes should be limited.
Secondly, the technological moderate removal abilities and the possibility of CO_2 leakage should be taken into account.
SO_2 is the compound that receives the largest amount of interest with respect to the effect of contaminants on the structure of storage formations [2] and the possibility of its co-storage together with CO_2 is being studied. This is a key issue for design and economics of oxyfuel plants. A concentration limit in the CO_2-rich stream has not been defined yet.
Another aspect regarding the concentrated CO_2 stream is the legislative classification. This will depend on the content of contaminants such as H_2S, sulphur oxides, NO_x, hydrocarbons, etc.. If it was classified as an hazardous

waste, additional difficulties (location of specifically suitable storage sites and/or need of further purification and efficiency penalties) would be faced.

2.8 POSSIBLE FUTURE DEVELOPMENTS

Possible future developments could decrease significantly the cost of capture. In particular, one of the most interesting possibilities is the adoption of OTM technology for the ASU [12], that could be a breakthrogh in the economics of oxyfuel plants with capture [33]. Each company engaged in the development of OTM systems has a unique design approach [12]. The claimed reductions in energy and/or costs are really sizeable, especially for highly integrated plants, but still need to be confirmed in normal working conditions. Non integrated OTM ASUs are expected to consume less than 150 kWh/tO_2, while, for a highly integrated one, a recent study predicts a consumption of 26 kWh/tO_2, that would represent a 85% decrease from the present cryogenic ASU consumption and could lead to a net efficiency higher than 40% [14]. Most of the studies expect a consumption of less than 100 kWh/tO_2, which is well below values for the most efficient cryogenic air separation.
Some of the proposed solutions include combustion of a small natural gas mass flow to increase the temperature of the gas entering gas, to reach the working conditions for the OTM ASU. That quantity represents another source of CO_2, that has to be taken into account and, if convenient, processed to limit CO_2 emissions. Otherwise the capture efficiency would decrease.
In the short term, another possible advancement will likely come from the development of new more efficient cryogenic ASUs.

Another possible future development could be the adoption of chemical looping, that is schematically represented in the figure below. The main concept of chemical looping combustion is to split combustion of a hydrocarbon or carbonaceous fuel into separate oxidation and reduction reactions by introducing an oxygen carrier to circulate between two reactors, where the temperature is in the 800-1200°C range.

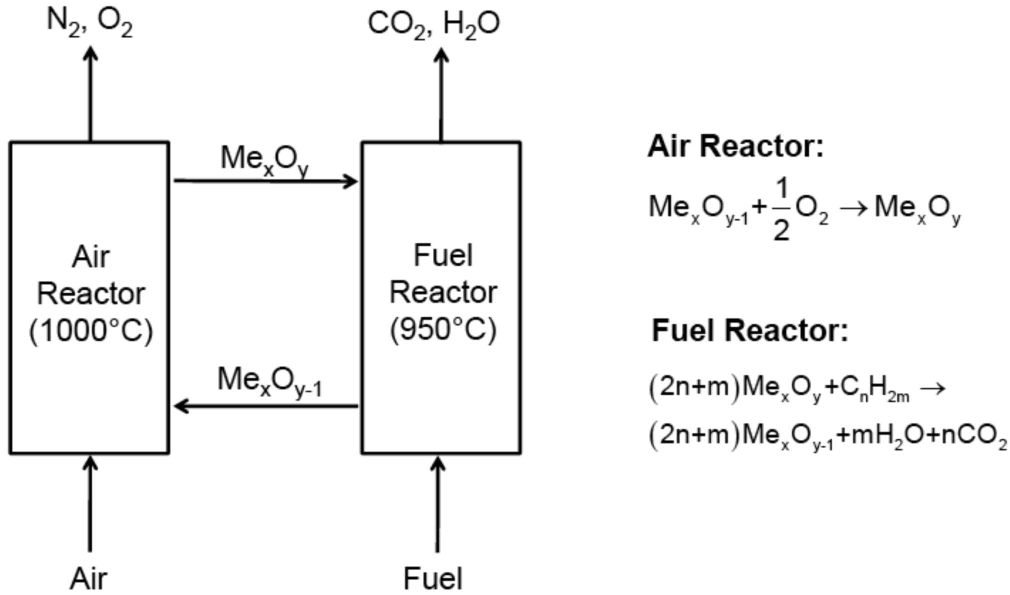

Figure 20. Schematic of the chemical looping process for coal, adopted from [34].

A critical issue is the long-term mechanical and chemical stability of the particles that have to undergo repeated cycles of oxidation and reduction, to minimize the make-up requirement. Other problems are the incomplete coal conversion (a carbon stripper is in fact needed [34]) and the mechanical wear in some components.
If the ongoing early-stage tests are successful, chemical looping combustion could represent an interesting option for future oxyfuel plants. It would guarantee a very low efficiency penalty for CO_2 capture, low NO_x specific emission and the elimination of the air separation unit.
In economic terms, this technology would probably result in an increase in a lower cost of electricity, but it still needs to be proven in larger test facilities.

Costs for first-of-a-kind commercial demonstration plants in general are usually higher than those of subsequent plants, for example because of design conservatism and high design. CO_2 capture technologies are in different stages of development, which has implications for their relative costs.
Excluding the market-related effects, the costs of power generation with CCS are eventually expected to decrease as a result of ‘learning-by-doing’ [35], as shown in the figure below, and of development of improved capture processes, although costs may reach a peak for the first commercial demonstration plants.

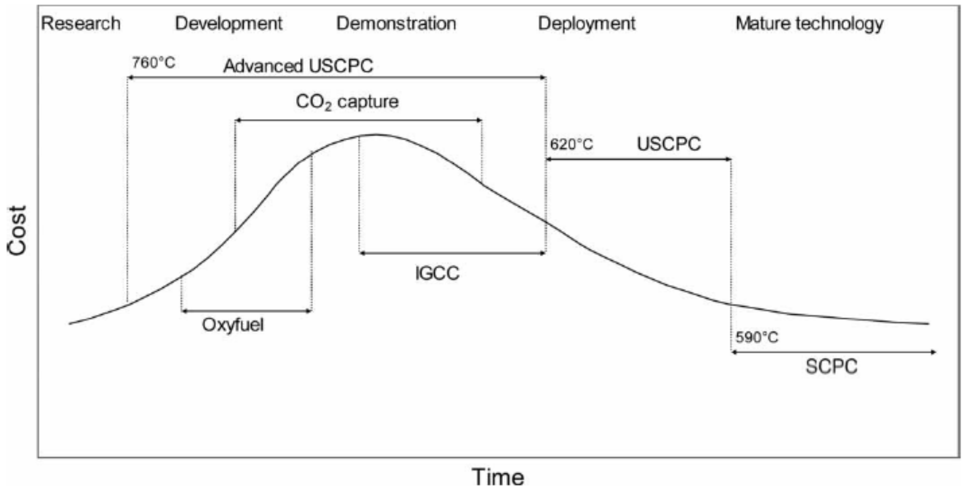

Figure 21. Trend of the costs of different technologies at different development stages (adopted from [36]).

An option to improve the capture efficiency is the capture by means of chemical absorption (or other technologies) performed on the gas flow that would be otherwise vented. These processes would allow the recovery of CO_2 from the stream at a cost that could be economically interesting because the concentration of CO_2 in that stream is quite high. In this way, the capture rate would approach 100%.

A separate evaluation of possible downstream systems could be performed to determine the cost of avoided CO_2, indicating the feasibility of this solution.

This additional system could even be associated with a lower cost for capture than the upstream oxyfuel plant. In this case, it would partly decrease the cost per tonne of CO_2 avoided for the integrated system.

If this approach will turn to be valid, the non-condensables removal efficiency has to be identified by a minimization of the cost of CO_2 avoidance for the combined plant, obviously taking into account the composition limits for the transport and storage operations. For this reason, the optimum composition of the gas available for storage could change.

3 METHODOLOGY, MODELS AND SIMULATION TOOLS

In this chapter, the approach, the assumptions and the tools used to perform the techno-economic analysis are briefly described.

3.1 CONCEPTUAL APPROACH

In this work, a simulation of the mass and energy flows has been performed using GS code for a certain number of interesting cases, for which it can give satisfactory results with acceptable efforts and without changing the main features of the plant.
A simulation for an oxyfuel base case with CO_2 capture has been provided by the Dipartmento di Energia at Politecnico di Milano. It represents a plant with cryogenic ASU, secondary wet recycle and direct drying system through primary dry recycle. It considers heat exchangers (including the ones used for the coal drying process) to take into account the heat recovery, even from the intercooled compression in the ASU and from the CPU. The primary recycle stream flows through an FGD, to avoid excessive corrosion in the boiler, but also in the CO_2-rich stream to be compressed.
The net power output of that simulated plants is about 700 MW_e. This size results from imposing the discharge velocity (255 m/s) at the exit of the last turbine stage and the height of the blades of that section. Thus it represents an estimate of the maximum possible size of the single-unit plant for the specific configuration.
The base case has been modified to exclude the FGD, since there is no realistic risk of corrosion in the boiler. Moreover, though it is not very important for oxyfuel plants, it is worthy of note that Australian power plants are not equipped with SO_x and NO_x control devices, nor is a change in policy foreseen about this issue. This represents a relevant difference compared to the European regulation, that is supposed to force the operators to adopt the best available technology (BAT). Since the separation of SO_2 in the CPU is not practicable, it is not likely that SO_2 emissions would be excessive, so this aspect could seem irrelevant, but it is a fact that there is no familiarity with FGDs in Australia.
Adoption and use of SO_x (and NO_x) control systems in oxyfuel plants should not be taken for granted anyway. As discussed in the previous chapter, NO_x emissions are expected to be lower than in air-fired mode and SO_x emissions are not much higher. Moreover, co-storage [1] or "sour-compression" [4] seem to be viable opportunities, that would reduce or cancel the need for those other

pollution control devices, although their technical and economical performances are still under investigation. On top of it, it is still questionable if co-storage is politically acceptable [2].
Some authors reckon that SCR and FGD are needed in any case to meet the requirements at air-firing conditions during start-up and shut-down [2, 11]. This idea is not convincing as the investment would be so sizeable that start-ups and shut-downs using more expensive but less polluting fuels (such as natural gas), if technically feasible, would probably be more more convenient and able to meet the regulatory requirements.
These reasons incited changes in the GS simulated plant (in particular the switching off of the modules representing the pollution control devices and their losses) to convert it to meet less strict requirements on SO_2. An adaptation to the more pressing demands has been implemented in a Matlab programme using approximate correlations, that will affect only the energy and the economic analyses. This procedure is based on the assumptions that the flow rates are only slighly modified by the changes and that the added capital and O&M expenses and the resulting efficiency penalty can be easily calculated to a first approximation, if the SO_x (and, in case, NO_x) mass flows to be abated are known.

It must be noticed that the power output of each simulated plant depends on the particular plant configuration. In fact, the imposed condition are the last stage discharge velocity (set equal to 255 m/s) and the size of the blades of the last stage of the turbine, not the power output. This choice is due to the structure of the GS code and has the advantage to make possible to estimate the maximum possible net output of a single unit.
The GS base case has been therefore modified during the thesis writing to consider different configurations.
GS code enables to calculate the mass and energy flows according to the model and to check whether the thermodynamic conditions in the different points of the plant are realistic and acceptable. These values can be used to calculate the fuel consumption and the size of the components or systems, that can be employed in the estimate of the expenses.
The list of significant values is completed by using Aspen Plus® Engineering Suite, to estimate mass and energy flows in the compression and purification unit. After the thermodynamic analysis is over, an economic assessment of the case studies can be performed.
An important step is the evaluation of the list of items (components or systems) that can be taken into account to calculate the capital expenses. Basing on previous works available in literature [11, 28, 37] and on the availability of data, the list for this work is:

- coal handling and feed

- feed water and miscellaneous
- boiler
- sorbent handling and feed
- FGD
- dust removal
- steam turbine
- ash and spent sorbent handling
- accessory electric plant
- instrumentation and control
- ASU
- CO_2 purification and compression
- civil works and piping (BOP)

At this point, it is possible to calculate the different costs for the maximum-size plants and to estimate the COE and the cost of avoided CO_2 for each case, following the procedure that is described in detail later in this chapter, for both the general and the Australian case.
It is important to provide the economic results for different plant sizes and to take into account different strategies for SO_x control, so a MATLAB programme, which uses the obtained database and several correlations, has been built to supply at least an approximate estimate for downscaled and/or FGD-equipped plants.
Several results have been collected and commented. Comparisons among the different cases or against literature have been made and a sensitivity analysis has been performed.
Finally, some conclusions about the potential of oxyfuel combustion to become a commercial viable way to reduce the CO_2 emissions from coal-fired power plants are presented.

3.2 GENERAL ASSUMPTIONS AND REFERENCE POWER PLANTS

There is no general agreement about the reference conventional plant without CO_2 capture that is compared to the plant with CO_2 capture that we are analysing. In particular, a debate exists on the parameter that should be in common with the CCS plant (net output, gross output or thermal input), if any (for example the possibility of a comparison between the largest designable single-unit plants exists). It is also possible to compare the performance of a plant with CO_2 capture against the average performance of new-build coal-fired plants (or even plants fed by fossil fuels in general).
Moreover, the issue is made more complex by the presence of different regulations in different locations. For example, the reference plant for the

Australian setting would not equipped with NO_x and SO_x control, whereas a refenrence plant in Europe would be equipped with FGD and SCR.
This is why two different reference plants have been taken into account, either with and without SO_x and NO_x pollution control devices.
The approach must be different than the one used for post-combustion plants, where the introduction of the CCS system does not modify heavily the upstream power system. On the contrary, the components of an oxyfuel plant with CO_2 capture are different than in a conventional air fired plant. For this reason, the comparison must be made between the two complete plants and cannot be made on an incremental basis.
An analysis of conventional plants is beyond the goals of this work and only an approximate economic analysis is performed. Data found in literature for advanced large scale coal-fired plants will be used as reference for the technical and economic performance, although mass and energy flows have been obtained. In particular, data from large plants both with and without pollution control devices have been collected and eventually adjusted to take into account the different year (using the CEPCI index for capital costs) and/or currency and the changes in the price of coal. Specifically, only the cost of electricity and the specific CO_2 emissions are eventually needed to evaluate the cost of avoided CO_2.
These two values are different in the general and in the Australian case, because in the latter no pollution control is taken into account, resulting in decreased COE and slightly inferior specific CO_2 emissions (at least theoretically, with regard to the issue of the reference plant).
The specific CO_2 emissions can be calculated once a realistic energy efficiency is set and a specific coal is chosen (LHV and C % wt. are at least needed). The efficiency is taken equal to the one warranted by a new conventional advanced USC plant, whereas the South African coal used in the simulations of oxy-fired plants has been considered (64.4 %wt., LHV=24.62 MJ/kg).
All the parameters used in the economic analysis (such as coal price and discount rate) to calculate the COE are set in accordance with the ones in the oxyfuel case.

3.3 SIMULATED CASES

In this work, the GS code has been used to build a database that is therefore used for the economic analysis. Different configurations has been simulated, as shown in the figure below, depending on the possible combination of the following choices:

- secondary recycle position: a wet recycle and a all-dry recycle have been considered.

- boiler configuration: both CFB and PC boiler have been evaluated.
- type of fuel: a South-African black coal and an Australian brown coal have been used.
- presence of an FGD: the base case has been run both with and without SO_x control.

Simulated cases

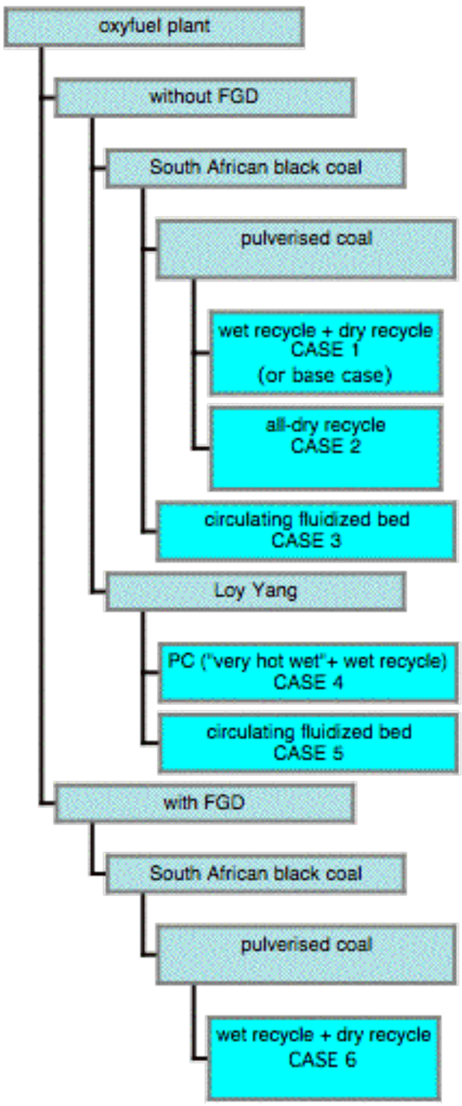

Figure 22. List of simulated cases.

For every configuration represented in the figure above, a GS simulation has been launched. It should be noticed that the plant size has not been changed in the GS simulations and that only one configuration is run with SO_x control. For each configuration, the following data are collected for the simulated plant:

- oxygen mass flow
- mass flow and composition of the stream treated in the CPU
- fuel consumption
- ash input
- SO_x amount in the flue gas flow
- Net power output
- Gross power output

3.4 GS CODE

In this work, mass and energy balances and overall performances are calculated by using GS (gas-steam cycles) computer code, developed at Dipartimento di Energia of Politecnico di Milano.
It was originally designed to analyse gas-steam cycles for power production, but is presently used for a wide range of complex systems such as combined cycles, integrated gasification combined cycles, waste-to-energy plants, fossil fuel-fired plants with CO_2 capture, hybrid cycle with fuel cells and nuclear plants.
An accurate guide to the code is the GS user manual [38]. Here, a brief description will be provided, in order to explain the procedure used to obtain the mass and energy balances.
GS represents the components of the system as modules that can be set to depict their real performance. The system is constructed as an interconnection of different modules.
The modules considered for this work are:

1) compressor
2) combustor
3) heat exchanger
4) mixer
5) splitter
6) oxygen separation plant
7) shaft
8) steam cycle

Firstly, the programme reads the input file and checks the consistency of the data lines. Then a first iteration is run using the inputs data provided by the user. Subsequently, the input data for the iterations are calculated according to the assigned order. Once the system to be calculated is defined and the coherence of the component characteristics and their interconnections are verified, the code

sequentially calculates mass, energy and atomic species balances of all the plant components until convergence is reached. For each iteration mass and energy mass balances are worked out. The iterative process continues until a stable convergence is reached. At that time, the programme will calculate the components excluded in the previous iterations. Finally, the output file with the global results is printed.
A resuming diagram representing the GS code procedure is reported in the figure below.

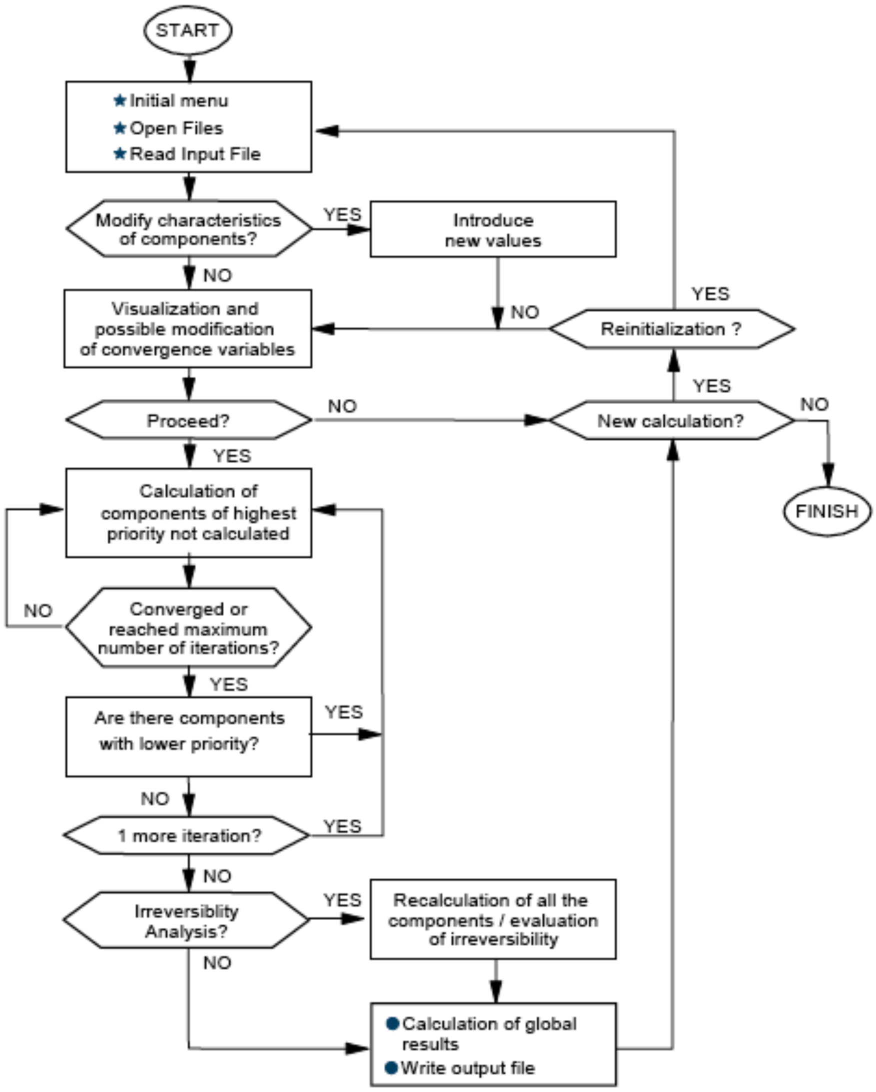

Figure 23. GS code programme procedure [38].

In addition to the flexibility related to its modular structure, the distinctive strength of the code lies in its capability to predict reasonably well the performances of key plant components (such as the steam turbine and the air separation unit) according to built-in models and correlations, especially for turbomachines.
The properties data file used in the performed simulations includes the following 20 chemical components: Ar, CH_4, CO, COS, CO_2, C_2H_2, C_2H_4, C_2H_6, C_3H_8, H_2, H_2O, H_2S, NH_3, N_2, O_2, SO_2, S, H_2O(L), C(S), Ash(S).
Coal and heavy fuels have a very complex composition, that is often given as ultimate analysis composition. An equivalent fuel mixture with the same atomic composition and the same LHV and HHV of the given fuel is used, allowing handling every kind of fuel.

3.4.1 ASSUMPTIONS FOR GS SIMULATIONS

A list of the general assumptions accepted for the GS simulation is reported:

- ISO ambient conditions are assumed. Pressure, temperature, humidity and air composition are reported in the following table:

Table 5. Assumptions for the air, adopted from [30].

Pressure, bar	1.01325
Temperature, °C	15
Relative humidity, %	60
Dry composition	% vol.
N_2	78.09
O_2	20.95
Ar	0.93
CO_2	0.03

- Pure water/steam is the only compound treated as real fluid. For its evaluation S.I. tables are used [39]. All the mixtures are handled as ideal fluids whose thermodynamic properties are calculated by means of NASA polynomials [40].
- Gas composition at reactors outlet is determined by assuming chemical equilibrium, calculated with the model originally developed by Reynolds [41], implemented in the code.
- Correlations used to predict the performances of some components are based on data coming from current commercial machinery. They are handled even for unconventional machinery. This is a strong assumption that should be verified case by case.

- CO_2 compression and liquefaction consists in a simple intercooled compression up to 110 bars and a final cooling to 20°C. Subsequently this stream is pumped to the final pressure of 150 bars.
- Convergence is considered as reached when both the following conditions are verified: mass flow rate, temperature and pressure differences of each stream between two successive iterations are less than 0.01% and mass and energy balance of each component is respected with a maximum error of 0.005% [30].
- Most of the assumptions used in the simulations performed in this work are borrowed by previous simulations run by members of the Politecnico di Milano. In particular, the assumptions used by Dr. Matteo Romano in his PhD thesis for steam cycle plants have been very precious. They are reported in the following table:

Table 6. Assumptions for PC and CFB steam cycle plants, adopted from [30].

	PC	CFB
Pressure drops		
Boiler and windbox, kPa	3	16
Convective pass, kPa	2	2
Air preheater (hot and cold sides), kPa	1	1
FGD, kPa	2	-
Fabric filter, kPa	1.5	1.5
Gas to gas heat exchanger (hot and cold side), kPa	1	-
Combustion and heat exchangers		
Losses for unburned C, % of input LHV		1
Thermal losses, % of heat transferred	0.7	0.7
Ca/S in boiler	-	2
Sulphur captured in boiler, %	-	90
Oxygen in flue gas, %vol.	3.5	3.5
Gas temperature at economizer outlet, °C	350	350
Gas temperature at air preheater outlet, °C	130	130

Steam cycle		
SH/RH live steam pressure, bar	247/53.75	
SH/RH live steam temperature, °C	600/610	
Condensing pressure, bar	0.042	
Steam side pressure loss in ECO+EVA+SH, %	20	
Steam side pressure loss in RH, %	7	
Number of preheaters	7	
Steam side regenerative bleedings pressure loss, %	2	
Preheaters steam-feedwater pinch point ΔT, °C	3	
BFW temperature, °C	310	
Turbine rotation speed, RPM	3000	
Number of IP/LP flows	2/4	
Turbine mechanical efficiency, %	99.5	
Electric generator efficiency, %	99.0	
Turbopump		
Turbine inlet/outlet pressure, bar	6.21/0.05	
Turbine adiabatic efficiency, %	80	
Pump adiabatic efficiency, %	85	
Mechanical efficiency, %	98.8	
Auxiliaries		
Fans efficiency, %	80	80
Pulverizers and coal handling, kJe/kgcoal	50	30
Precipitators and ash handling, kJe/kgash	200	100[a]
Limestone handling, kJe/kg$CaCO_3$	0[b]	90
Power for heat rejection, MJe/MJth	0.01	0.01
FGD auxiliaries, MJe/kgSO_2	5.34	-

[a] including spent sorbent
[b] included in FGD auxiliaries

3.5 ASPEN PLUS

The commercial tool Aspen Plus® Engineering Suite, release 2004.1 has been used to simulate the performance of the compression and purification unit, since the GS code cannot reproduce the real gas effect and some other properties (such as gas solubility).
The simulated process is the updated version of the one reported in the figure below, which performs the separarion in two low temperature flashes. The main upgrade is the addition of three inter-heated turbines for the non-condensables at the outlet of the second flash to produce electric power and to increase the refrigerating potential.

The purity of the CO_2-rich stream and the outlet pressure have been set to 150 bar and 96.5% vol., respectively, when it was possible.
Indeed, in the lignite-fed PC plant the SO_2 concentration is quite relevant and the separation of a large quantity of SO_2 would lead to a significant release of CO_2 and, consequently, to a lower capture efficiency. This happens because SO_2 and CO_2 have similar characteristics, making the parting in a flash difficult.
In this case, the goal of the purification process was the achievement of a cumulative concentration of argon, oxygen and nitrogen lower than 4% [2].
The quantity of SO_2 in the gas stream could be too high for co-storage, although the limit depends on the specific storage site and has not been fully clarified yet. Therefore, if SO_2 turned to be excessive, since the flash process is not effective against SO_2, alternative technologies would be considered to abate SO_2.
Specifically, flue gas desulfurization and sour compression are the two main opportunities, although the latter is still being researched.
A study [21] proposes to use seawater as removal medium for desulfurization in the CPU. Sulphur dioxide forms, together with seawater, sulphites and after aeration sulphites are converted into sulphates. Sulphates are considered harmless for the environment; in order to bring pH near to neutrality more seawater is added, the liquid stream rich in sulphates can be discharged in the ocean. The proposed process would be performed at 20 atm in order to have fairly small plant dimensions.

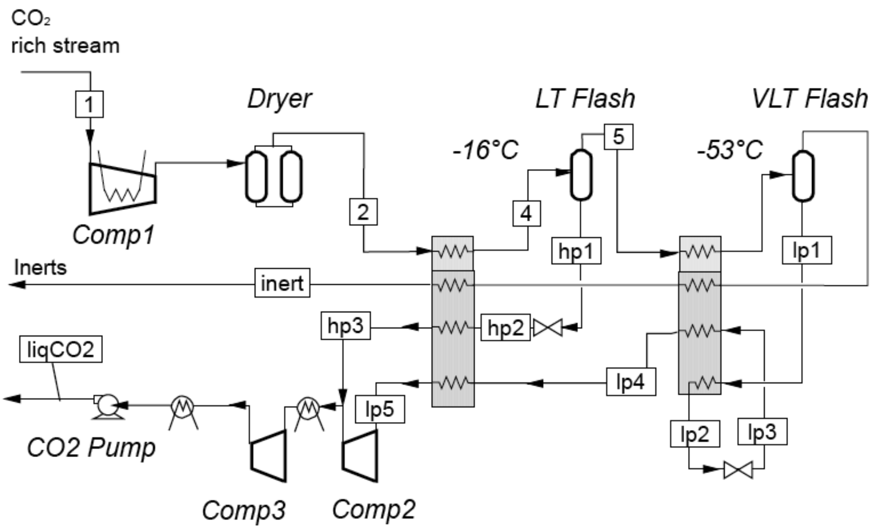

Figure 24. Simplified scheme of the compression and purification process, adopted from [37].

The simulation of the compression and purification process require the thermodynamic conditions and the composition of the inlet gas flow to be performed.
From this simulation, several data can be obtained:

- composition of the CO_2-rich stream
- energy penalty
- mass flow of vented CO_2 (and therefore percentage of captured CO_2)

With reference to this last aspect, it should be mentioned that the percentage of captured CO_2 could possibly be notably increased, if CO_2 absorbtion was performed on the mixture of non-condensables and CO_2, that would otherwise be vented.

3.6 PROCEDURE FOR THE ECONOMIC ANALYSIS

After the simulation of the process, the estimated mass and energy flows are available for the large simulated plants. It is then possible to perform an economic analysis to assess the expected cost of electricity for each configuration.
To calculate the total cost of CCS, we must add up capital, operating and abandonment costs. After having estimated these, we phase them over time and calculate the present value. By dividing it by the present value of the annual CO_2 avoided, we obtain the cost of a tonne of CO_2 avoided, where the CO_2 avoided is the difference between the amount of CO_2 emitted without CCS and the amount of CO_2 emitted with CCS.
To calculate the capital costs, the procedure used is the one developed by the CO_2CORC research group, as explained in their Methodology Report [42].
Generally speaking, we can see the total capital costs as the sum of the following categories:

- costs of equipment and material
- international freight (if applicable)
- local freight
- construction and installation costs
- engineering and project management costs
- owners' costs

The Equipment and Materials Costs **(a)** are Free-On-Board (or FOB) cost of importing equipment and materials from overseas or the cost of purchasing the equipment. To calculate the capital expenses, a set of specific capital expenses (with the associated reference size) expressed in a fit currency (in this work 2009 USD or 2009 AUD) and scaling factors is necessary.
To estimate the capital cost of each item, several data are needed:

- reference specific costs (connected to a certain size and to a certain year). They can be found in literature or in commercial data released by the supplier. The cost must be specific in reference to the measure that is most associable to the effects of the scale economies (for example the oxygen mass flow for the ASU and the thermal power for the heat excangers). It must be noticed that analogous data must be compared to choose a cost as reference value, if no exact values are not known (for example no data are provided by the supplier). So it is important to understand if the analysed values are FOB costs, installed costs or total costs (inclusive of start-up costs). Obviously, they are expressed in a certain currency and refer to a specific year and size of the component. So values from different sources cannot be immediately compared unless they are properly adjusted. That can be done, at least in an approximate way, if the following data are available;
- scaling factors (SF), that allow us to scale the specific capital cost to describe the required component size . They are used to take into account the scale economies. In fact, they are used in formulas of the kind: $C = C_0 * (\frac{S}{S_0})^{SF}$. A value for each item is chosen after analysing the different values available in literature. The scaling factors are usually in the 0.6-0.9 range for components or systems in both PC and CFB power plants [17]. The progress ratios (PR) have the same aim, but they are used in a different way.
- Chemical Engineering Plant Cost Index (CEPCI), that are used to account for different equipment pricing years. In particular the ratio between the CEPCI in the reference period and the CEPCI in the period we want to refer the final result to (usually the most recent available). Introduced in 1963, the index has not been changed in its structure, but many underlying details have been updated. It is a weighted sum of several sub-indexes and each of these is a weighted sum of components. The sub-indexes are: equipment (that is made up by heat exchangers and tanks; process machinery; pipes, valves and fittings; process instruments; pumps and compressors; electrical equipment; structural supports and miscellaneous), buildings, engineering & supervision, construction and labour. The use of the index is relevant only if the costs are expressed in US dollars. Obviously the result is not exact, but it provides a better estimate. The recourse to this or other analogous indexes is quite common to several studies on plants with carbon capture in general [36] and specifically on oxyfuel plants [28].

Table 7. Average values of the CEPCI index in the last few years.

CEPCI 1999	390.6
CEPCI 2000	394.1
CEPCI 2001	394.3
CEPCI 2002	395.6
CEPCI 2003	402
CEPCI 2004	444.2
CEPCI 2005	468.2
CEPCI 2006	499.6
CEPCI 2007	525
CEPCI 2008	575.4
CEPCI 2009	521.9

- the currency exchange rate in the reference period and in the most recent period taken into account.

Costs expressed in three different currencies are considered: US dollars (USD), Australian dollars (AUD) and Euros (EUR). The currency exchanges rates, as shown in the figure below, and even the purchasing power of the currencies itself changed drastically in the last few years.

Figure 25 . Historical graph of the EUR/USD exchange rate (in the last ten years).

The main currency for this work is the 2009 US dollar. The US dollar is more suitable for our purposes, because it allows us to use the CEPCI index, that allows us to translate old costs into present costs.
For example, if an available value is expressed in 2004 Euros, the following steps are needed:

- the cost is converted into 2004 USD, using the 2004 currency exchange rate.

- the cost is converted into present (or recent) USD, using the CEPCI index.
- the cost is converted into present (or recent) AUD and EUR, using the present (or recent) currency exchange rate.

Once all the available values are expressed in the same currency (for example 2009 USD), they can be compared. The best available reference data or average values calculated taking into account different reliable sources will be used. After that, the present results in the other currencies are estimated by means of the currency exchange rate of the most period being considered (and/or using a translation factor if it is necessary). Moreover, if the values from different sources are dispersed, it can be taken into consideration for a sensitivity analysis.

When specific data are not available, the International Freight Cost **(b)**, if needed, can be estimed as 10% of the FIB cost.

The Local Freight Cost **(c)** depends on the project, because it is related to the loading port (for imported items) or to the point of purchase (for items purchased locally). It can be estimated as the 5% of the FIB cost, underlining that a range of possible values is quite wide.

The costs including freight **(d)**, or CIF, are the sum of the FOB cost plus the cost of international freight (if appropriate) plus the cost of local freight. Inother words, it is the sum of **(a) + (b) + (c).**

These values are identified, after reviewing different available economic studies on oxyfuel plants [1, 17, 28, 37, 43].

Table 8. Resuming table of the specific capital expenses used in this work for the components, with their associated base size, maximum size and scaling factor.

Reference specific CAPEXes (cost and related size)	Scale parameter	Specific CAPEX		Base size	Unit	Max size	Scaling factor
Coal Handling and Feed	Coal flow rate	12.7	USD/ (kg/day)	3000	tpd	5000	0.76
Sorbent Handling and Feed	Captured Sulphur	106.7	USD/ (kg/day)	100	tpd	200	0.76
Feed Water and Misc	Power input	41.3	USD/kWth	1200	MW_{th}	-	0.7
Boiler	Power input	165.3	USD/kWth	1200	MW_{th}	-	0.66
FGD	Captured Sulphur	600.0	USD/ (kg/day)	100	tpd	200	0.76
Dust Removal	Power input	34.7	USD/kWth	1200	MW_{th}	1500	0.76
Steam Turbine	Gross power output	226.7	USD/kW_e	500	MW_e	-	0.88
Ash and spent Sorbent Handling	Ash input	97.3	USD/ (kg/day)	300	tpd	650	0.76
Accessory Electric Plant	Gross Electric Power	76.0	USD/kW_e	600	MW_e	-	0.7
Instrumentation and Control	Gross Electric Power	40.0	USD/kW_e	600	MW_e	-	0.7
ASU	Pure O_2 production	24.8	USD/ (kg/day)	5000	tpd	9000	0.82
CPU	CPU electric consumption	1494.6	USD/kW_e	13	MW_e	-	0.76
Civil works and piping (BOP)	Gross Electric Power	240.0	USD/kW_e	500	MW_e	-	0.76

Construction and Installation Costs **(e)** are estimated for the single components using different methods.

The Engineering, Procurement and Construction costs **(f)**, often called EPC or base plant costs or direct costs, are the sum of the cost including freight and the construction and installation costs. So they are the sum of **(d)** and **(e).**

Engineering and project management costs **(g)** are the costs of designing and overseeing the construction of the project. They are generally estimated as the 15% of the base plant cost **(f)**.

The owners'costs **(i)** are the cost of obtaining approvals (including enviromental approvals and land purchase) and of negotiations and legal processes. They are

estimated as the 7% of the EPC costs **(i)**. This is the assumption reccomended by the IEA.
The total EPCO costs **(j)** are the sum of EPC costs plus owners' costs. In other words, they are the sum of **(h)** + **(i)**.
Contingency **(k)** accounts the costs of miscellaneous items not included in **(j)**. It is set to 10% of the total EPCO costs **(j)**.
The total project capital cost **(l)** is the sum of the total EPCO costs and contingency, or, that is the same, the sum of **(j)** and **(k)**.

Table 9. Resume of the rule of thumb for the estimation of the setup costs.

		Capital cost elements	Nominal value
Equipment costs	A	Process Equipment Cost (PEC)	Sum of all process equipment
	B	General facilities	10-20% PEC
		Total Equipment Cost (TEC)	A+B
Set up costs	C	Instrumentation	15% TEC
	D	Piping	20% TEC
	E	Electrical	7 % TEC
	F	Total Installed Cost	A + B+ C+ D +E
	G	Start-up costs	8% TIC
	H	Engineering	5% TIC
	I	Owners costs	7% (F + G + H)
	J	Engineering, procurement, construction and owner's cost (EPCO)	F + G + H + I
	K	Project Contingency	10% EPCO
		TOTAL CAPITAL COST (CAPEX)	J + K

For each system (such as the boiler), the total capital cost can be calculated in this way:

$$(C_{\max\_power})_i = \frac{(C_{S_0})_i}{(S_0)_i} * (S_0)_i * (\frac{(S_{\max\_power})_i}{(S_0)_i})^{SF_i} \tag{3.1}$$

where:
$\frac{(C_{ref\_power})_i}{(S_0)_i}$ is the specific capital cost for the item i. It can be found in literature and is associated with the size $(S_0)_i$;

SF_i is the scaling factor for a certain system. It depends on the value $(S_0)_i$ chosen to represent the size;
$(S_{max\_power})_i$ is the size of the system given by the simulation.
The trend is not continous, if the number of necessary parallel units changes. In fact, a different procedure could be needed for the following systems: fuel handling, sorbent handling, ash removal and handling, FGD and, importantly, ASU. These are the ones which could be unable to statisfy the requirements, if a single train is adopted. In that case, two or more parallel trains are therefore needed. Two trains could be not sufficient for the ASU, because each of the two would have to supply 7000-8000 tonnes of O_2 per day, that is more than the production of the largest available ASU, but could be provided by newly designed ASUs, if the expectations of the suppliers are confirmed.
If the parallel trains are of the same size, the following formula is used:

$$(C_{max\_power})_i = n^f * \frac{(C_{S_0})_i}{(S_0)_i} * (S_0)_i * (\frac{(\frac{S_{max\_power}}{n})_i}{(S_0)_i})^{SF_i} \quad (3.2)$$

where n is the minimum number of identical components to satisfy the requirements and f is a scaling factor, set to 0.9, that takes into account the positive effect of multiple installation of identical components. Indeed, two units usually share auxiliary equipment, installation labour and engineering costs, particularly fo uncommon components and cost less than two of them, being developed and installed independently.
This is the approach used in this work to estimate capital expenses, because it is simple and realistic. However, theoretically, this is not the best approach, as it does not make maximum use of the scale economies.
In this sense, the best solution would be the adoption of the minimum number of full-scale systems and only one component to fill the gap. For these components, the formula can be:

$$(C_{max\_power})_i = m * \frac{(C_{ref\_power})_i}{(S_0)_i} * (S_0)_i * (\frac{(S_{max\_size})_i}{(S_0)_i})^{SF_i} - \frac{(C_{ref\_power})_i}{(S_0)_i} * (S_0)_i *$$

$$*(\frac{(S_{max\_power} - m * S_{max\_size})_i}{(S_0)_i})^{SF_i} \quad (3.3)$$

where m is number of full-scale components needed and S_{max_size} is the size of those full-scale components.

Total capital expenses are the sum of the total project capital costs of the single

items.
The operating costs of a CCS project have fixed and variable components.
The classification and amount of those expenses are often hard to define for power plants. In fact, the task is very complex for a plant whose characteristics are very dependant on many factors (for example fuel feed, location and regulation). On top of it, the design of an oxyfuel plant is not totally defined yet, because some technical aspects still remain unclear.
Generally speaking, fixed operating costs include direct labour, administration and maintenance expenses.
On the other side, variable operating costs are composed of fuel (the most prevalent item), waste disposal (fly and bottom ash), water, labour, maintenance and miscellaneous. In the case FGD is operating, the list takes into account also FGD make-up water, reagent (lime or limestone) feed, steam and gypsum disposal (this last one can be a credit). If a DeNOx is installed, the associated items have to be considered: reagent (ammonia) feed and catalyst.
In this work, operating costs will frequenly be divided into fuel and non-fuel components, because two different fuel types are regarded.
For generic cases and as defaults in economic models, the CO2CRC uses the rules of thumb to estimate the operating costs.
For generic cases, as a rule of thumb the real costs of abandonment are estimated as 25% of the real EPC in the CO2CRC Methodolgy Report.
Costs of abandonment are ignored in this study, mainly because their infuence on the COE can be neglected.
In this analysis, even the effect of taxation has been disregarded as well: that is beyond the goal of this study.
The procedure to calculate the cost of electricity is then performed following the discounted cash flow method proposed by the CO2CRC research group [42], adapted to this process. It should be taken into account that the effect of taxation is neglected in this thesis, since it provides estimates for real costs before tax.
The calculations have been performed on Excel worksheets realised for this application, adopting the assumptions reported below.

3.7 ECONOMIC ASSUMPTIONS

The results will be expressed in 2009 USD and can be converted in 2009 EUR and 2009 AUD through the respective currency exchange rates: 1 EUR=1.39 USD and 1 AUD=0.75 USD.
Capital expenses have been estimated by means of the procedure described earlier in this chapter, including scale adjustment through scaling factors and addition of set-up costs.
The assumed availability of the oxyfuel plant is assumed equal to 85%, as most

of the analogous examined studies do [1, 13, 28, 37, 43]. This value is also used in the IEA Greenhouse Gas Standard Economic Spreadsheet and assessment criteria and supported by the high values coming from active commercial plants, including CFB [17]. No study expects the load factor to be very different from that value, since oxyfuel plants would be expected to operate at base load.
A higher load factor has been assumed for the reference plants (90%), after reviewing the most recent statistics.
The discount rate has been set to 7% real (discount rate minus inflation rate) per year, that is the value indicated by the CO2CRC methodology report [42]. This is much higher than the return rates on typical government and corporate bonds, to allow for the effects of taxation and the higher return rates necessary to compensate for investment risks. The discount rate has a significant impact on the cost of electricity generation.
According to the most recent OECD data, the long-term interest rate is forecasted to be in the 4-5% range in the Euro area and the US and about 6% in Australia [44].
The standard power plant economic and assessment criteria introduced by IEA schedule a 10% discount rate and this approach is common to many studies [1, 13, 28, 37, 43], that assume a discount rate in the 7-10% range, although some of them perform a sensitivity analysis using 5% [11, 13, 43].
Sometimes, the interest rate is reported. However the approach is equivalent, because for every annual effective interest rate, there is a corresponding annual effective discount rate, given by the following formula:

$$i = \frac{d}{1-d} \tag{3.4}$$

The Discounted Cash Flow (DCF) method is considered the most useful, at least theoretically, because it estimates the effective future cash flows and it allows to calculate the investment return rate associated with the project if the electricity price is known (or estimated). Moreover, it is possible to make different assumptions for different years.
However, many other studies resort to the capital charge rate (CCR) method, that is analogous to the DCF method in several cases. This analysis is not one of those, mainly because the first year load factor is lower than the one assumed for other years.
Nevertheless, good estimates for an equivalent capital charge rate can be reported. It ranges from about 8% to about 14% of the total plant cost, depending on the chosen definition of TPC, that can include (or not) owners costs, start-up costs, contingencies and interests during construction.
Under the commonly chosen definitions, the CCR would be about 12-14%, which is slighly lower than other values adopted in literature, mainly because it does not include the effect of taxation, that is neglected in this analysis and/or

different assumptions about the financing strategy.
Cost estimates in real terms before tax do not require assumptions about inflation, though after tax cost estimates do.
For the general case, the black coal price is estimated at 2 USD/GJ, in line with the values adopted with the recent analysed studies, whereas the brown coal price is set to 1,2 USD/GJ on the basis of reviewed data (in particular [43]) and information from the CO2CRC research group.
Australia is a country with large coal production that is partially shielded from energy markets [45]. For the Australian case, values provided by the CO2CRC research group have been used. In particular, the prices were 1.1 and 0.7 AUD/GJ respectively for black and brown coal.
The coal price is relatively volatile, so it is better to examine the effects of possible price fluctuations by means of a sensitivity analysis, as in most of the analysed studies.
The expected project life is 25 years, that is the value usually assumed by the CO2CRC for advanced plants with capture and coincides with the life time adopted by many authors [11, 13, 37, 46], whereas other studies assume a 20-year life time [28, 43]. Only the ENCAP, a European project, study supposes that the expected life could be longer (40 years in the base case), but it performs a sentisivity analysis for a 25-year life time [47].
Different procedures are applied to evaluate the ratio between total costs and installed costs. They bring to different estimates for the total costs, that are very important in order to define the cost of electricity, causing uncertainty in defining the expected cost of electricity. In this study, the conservative value used by the CO2CRC (33%) [42] is employed. In fact, for example, sizeable costs are likely to be necessary for adapting the designing of components to the new configuration and unforeseen event can easily occur. The analysed studies assign different weight to the sum of the set up costs (sum of start-up, engineering, owners and project contingency costs), ranging from 17% to 33% of the TIC. To avoid confusion, total costs have been used to compare different studies.
A large number of the studies evaluating the economics of CCS uses a capital charge rate procedure to estimate the expected cost of electricity, using CCR in the 10-17% range. This strategy is surely useful for preliminary analyses, but it seems inappropriate at a later stage, because it limits the options (for example it does not allow to consider different assumptions for different years). In this work, a realistic cash flow is simulated, in order to build a procedure to evaluate the real cost of electricity. Inevitably, some of the expected values used will turn to be impraticable, but they can be changed easily, once the procedure is defined.
The significance of non-fuel operating costs is uncertain, as well as the shares of it that can be traced respectively to the fixed and to the variable costs. Rules of thumb are used to estimate non-fuel operating costs.

In this work, for general cases, the fixed operating costs are estimated as 3% of total capital costs, that falls in the middle of the range of normally adopted values (1.0-4.8%) [11, 13, 28, 46]. The variable O&M costs are set to 5 USD/MWh for the PC base case without FGD, taking into account the expenses for the water and for ash removal. They increase to 7 USD/MWh in the CFB case, because an in-furnace capture is performed. Compared to values available in literature, the chosen numbers are in the same magnitude. For specific cases where better data are available, the best data available will be used. However, the operating variable O&M costs are just estimated, because they depend on a number of factors, such as ash composition, location, availability of water and consumables, etc..

In the Australian setting, the variable O&M expenses are historically lower than overseas, so they are set equal to 5 AUD/MWh for the PC base case without FGD and to 7 AUD/MWh for the CFB cases. On the other hand, the capital expenses are higher, because several components are manifactured overseas and transported to Australia. Moreover, the domestic suppliers of other components have limited production and they cannot fully expoit the scale economies. For these reasons, an additional translation factor (other than the currency exchange rate) is needed to take into account this economic penalty. In this work, it is set to 1.3 on the total capital expenses.

Finally, expected variable O&M costs for the reference plant in the general setting are 6 USD/MWh, whereas they are 4 AUD/MWh for the reference plant in the Australian scenario.

3.8 DOWNSCALING AND SO_2 REMOVAL

In this work, different plants have been simulated. Each one of these is characterized by a certain configuration, where the term “configuration” is here used in reference to a certain combination of SO_x control (absent), fuel type, boiler mode, secondary recycle and position, since the size is fixed by the conditions imposed on the last stage of the turbine.

Conceptually, all the collected GS output data are a function of the configuration, as long as we are considering the simulated size plant without pollution control. Eventually, the variable operating and the variable costs of the simulated plant are a function of the configuration.

$$c_{\max\_power,no\_FGD} = f(configuration) \tag{3.5}$$

$$C_{\max\_power,no\_FGD} = f(configuration) \tag{3.6}$$

In this thesis, capital and operating expenses are calculated for the simulated large plants without FGD and for their analogous plants of different size and/or with an installed FGD. Firstly, an analysis is performed following the procedure explained in the next chapter and the components of the COE are calculated. It is therefore necessary to modify the results obtained for the large plant without pollution control to adapt them for different size and/or different SO_x removal strategies.
Regarding the presence of an FGD, its cost can be estimated since we know the composition and characteristics of the flue gas and the limit of the emissions.
Before scaling the values, the introduction of pollution control must be considered in a simplified way. Indeed, the variable (fuel and non-fuel) costs increase and the item “FGD” of the capital (and fixed operating) costs assumes a finite value.

$$c_{\max\_power,w\_FGD} = c_{\max\_power,no\_FGD} + f(SOx - SOx_{allowed}) \tag{3.7}$$

$$C_{\max\_power,w\_FGD} = C_{\max\_power,no\_FGD} + f(SOx - SOx_{allowed}) \tag{3.8}$$

Moreover, the presence of the FGD causes a slight increase in CO_2 emissions, associated with the release of carbon dioxide due to the calcination reaction.

$$CO2_{\max\_power,w\_FGD} = CO2_{\max\_power,no\_FGD} + f(SOx - SOx_{allowed}) \tag{3.9}$$

The O_2 consumption does not change, since air can be used to supply the required oxygen to a reaction tank that is not in contact with the flue gas stream.
These correlations take into account the main expected effects of the insertion of an FGD. This approach is not completely correct, as additional modifications in the mass flows (and therefore even in the energy balance) can be foreseen. However, the changes are probably minimal, even because the share of sulphur in the fed coals is low, so the approximation is likely to be acceptable.
The net output decreases only because of the additional O_2 request and to the electricity required by the FGD.

$$Max\_power_{w\_FGD} = Max\_power_{no\_FGD} - f(SOx - SOx_{allowed}) \tag{3.10}$$

With reference to the size, the specific capical cost of the components for plants of different size can be assessed by means of scaling factors, that express the effect of the scale economy [48]. It should be reminded that the net output of the simulated plant is almost equal (or at least close) to the maximum output of a single unit, so it is only possible to downscale it.
The share of the COE that is associated with the investment cost is calculated in the following way:

$$(COE.CAPEX)_i = (COE.CAPEX_{Max\_power})_i * (\frac{output}{output_{Max\_power}})^{SF_i - 1} \qquad (3.11)$$

Importantly, the presence of more parallel units has to be taken into account, if it is the case.
An analogous method is used to estimate the share linked to the fixed expenses:

$$COE.fOPEX = COE.fOPEX_{Max\_power} * (\frac{\sum (COE.CAPEX)_i}{\sum (COE.CAPEX_{Max\_power})_i}) \qquad (3.12)$$

It should be taken into account that the capital expenses (and therefore the fixed O&M costs) do not always vary continuously: there are discontinuities when the number of required parallel trains changes.
The variable shares of the COE are instead adjusted by means through the ratio $(\frac{Max\_efficiency}{Efficiency})$.

$$COE.vOPEX = COE.vOPEX_{Max\_power} * (\frac{Efficiency_{Max\_power}}{Efficiency}) \qquad (3.13)$$

$$COE.fuel = COE.fuel_{Max\_power} * (\frac{Efficiency_{Max\_power}}{Efficiency}) \qquad (3.14)$$

Obviously the total COE is obtained by the sum of the four calculated components.
Finally, the cost per tonne of avoided CO_2, usually expressed in \$/ton, is calculated through the following formula:

$$Avoided\_Cost = \frac{COE_{with\_removal} - COE_{w/o\_removal}}{Emissions_{w/o\_removal} - Emissions_{with\_removal}} \qquad (3.15)$$

The procedure described in this paragraph is implemented in a Matlab programme, whose description is provided later in this chapter.

3.9 ICCSEM OVERVIEW

ICCSEM is a computational economic model to estimate the total cost of CCS being developed by the CO2CRC research group at the University of New South

Wales, that is built to estimate the economic performances of different plants with carbon capture systems.
It examines both the capture and the storage processes. Specifically, the capture component of the model aims to develop, design and describe key aspects of major CO_2 capture technologies.
A beta version of the model has been issued for an internal review by CO2CRC sponsors, and is in the process of commercialization.
Oxyfuel plants were not included in the model and this thesis is committed to the formulation of a section that takes into account this technology for greenfield power plants.

3.10 PROGRAMME TO ESTIMATE OXYFUEL PLANTS ECONOMICS

The Matlab programme, that has been elaborated to become integrating part of ICCSEM, is intended to give an approximate estimate of the most relevant values for the plant whose characteristics are different from the simulated ones. In particular, it provides an estimate for different sizes (net outputs) and different SO_2 emissions requirements, for each configuration.

Table 10. List of the studied configurations.

without FGD					with FGD
South African black coal			Loy Yang brown coal		SA bl. Coal
PC		CFB	PC	CFB	PC
wet sec. recycle	all-dry recycle				
Case 1	Case 2	Case 3	Case 4	Case 5	Case 6

The user can choose the following parameters:
- Configuration
- need of SO_x pollution control or not
- in case of SO_x control, allowed maximum SO_x concentration in the flue gas (expressed in mg/Nm3 6% O_2 dry) after the . It should be taken into account that this concentration limit cannot be compared to the one set for air-fired plants, because the exiting volumetric flow is lower. A comparison in terms of specific emissions would make more sense.
- required plant net power output.

The database of the values from the simulations include:
- plant net efficiency
- SO_x emissions
- captured CO_2 mass flow (and purity)
- purity of the stored mass flow
- net power output
- gross power
- O_2 consumption
- Composition of the flue gas
- estimated capital costs
- cost of electricity (total and components)
- scaling factors

Moreover, several values available from literature are needed, such as CAPEX, energy consumption and variable expenses related to the FGD.
The capital costs of an FGD plant are considerably influenced by market conditions and other factors, for example, geographical location and the amount of preparatory site work required. In addition, the costs of FGD plant also depend on technical factors such as volume of flue gas to be treated; concentration of SO_2 in the flue gas; desulphurisation efficiency required; quality of the by-products produced; other environmental constraints (for example, permitted waste water discharges); the need or otherwise for flue gas reheat; the degree of reliability and redundancy required; design life [49].

In particular, the gas mass flow entering the FGD in the oxy-fired case is lower than in an air-fired plant. Previous studies consider an investment cost for the FGD reduced to 60% of the corresponding FGD for the reference plant [43], suggested by presonal communications with the suppliers.
A complex procedure developed by the EPA in 2000 to estimate the capital costs for the FGD has been used. The input parameters are the SO_2 and sorbent mass flows, the volumetric flows of the entering and exiting the FGD, the FGD energy consumption and other minor parameters.
Given that the volumetric flows and all the other values are taken into account, it should give accurate results.
However, it is conceived as a procedure to estimate costs for FGDs in air-fired plants. In this case, an oxidation system provides an ambient air stream to the absorber reaction tank.
If the same approach is adopted for oxyfuel plants, it is possible to supply air or oxygen. In the former possibility, the gas stream entering the FGD is partially diluted by the oxidation stream (increasing the load on the non-condensables removal system), whereas, in the latter case, oxygen is produced by the ASU, that supplies a small extra stream for this reason.
Another approach is probably more efficient: the slurry stream is taken from the absorber tower to a different tank, where the calcium sulfite ($CaSO_3 \bullet H_2O$) formed by the SO_2 removal process to be oxidized to calcium sulfate ($CaSO_4 \bullet 2H_2O$). Air can be used for the oxidation process.
Unfortunately, when the procedure for FGD cost estimate is performed using the data from reports, the calculated costs differ significantly from the ones expected in the report itself, although the results were adjusted by means of adequate indexes. In particular, the calculated expenses are less than 70% of the reported values. Partially, this difference could be due to an undervaluation of the costs that has been recently fixed when more data from real plants became available. The ratio between reported and calculated values is about 1.5 for air-fired plants and 1.8 for oxyfuel plants, which used oxygen for oxidation.
On the other side, the FGD market is still experiencing a cost redution referable to the effect of the experience curves. For all these reasons, it is impossible to provide an exact value and it is also difficult to provide an estimate.
It seems appropriate to use that procedure and correct the obtained estimates for capital costs by multiplying them by a coefficient.
After having analysed different sources, the specific installed capital cost for an FGD in the base case has been set to the value calculated by the procedure for the specific plants multiplied by a coefficient, equal to 1.5.
Actually, capital cost of the FGD depends on the SO_2 removal efficiency. For the purpose of this thesis, a >95% removal requirement is set, that is very conservative for low sulphur coal.
In 2007, the average capital cost for installing an FGD was 316 USD/kW [50].

The FGD electric consumption is closely related to the amount of removed SO_2 and accounts for 5.34 MJ/kgSO_2 [30].
According to Srivastava et al. [51], the estimated operating variable costs associated with SO_2 removal in a LSFO FGD (mainly related towaste ponding cost and reagent cost) are 0.0627 $/kg, including a gypsum byproduct credit.
The programme carries out the procedure, that has been described previously in this chapter. Firsly, the programme loads the data relative to the selected configuration. Then, it adjusts CAPEX (and consequently fixed OPEX), variable operating expenses and energy balance taking into account the presence and the load of the FGD.
At this point, the dowscaling is performed. CAPEX and, consequently, fixed O&M are scaled by means of the plant scaling factor. The amount of emitted CO_2 is changed (supposing a limestone feed) if an FGD is required. It changes proportionally to the amount of removed SO_2. The specific emission per unit of produced electricity increases even because of the worsening of the efficiency.
O_2 production in the ASU does not change, beacause it is supposed that the required oxygen is supplied as part of an air stream fed into the reaction tank.

In addition to the increase in capital costs, it is important to take into account that usually smaller plants have lower efficiency than analogous larger plants. On the economic side, this difference causes an increase in specific fuel costs and in the cost of electricity in general.
Unfortunately, no one of the analysed studies examined this issue for super critical and ultra super critical coal-based cycles in general and for oxyfuel plants in particular, although it is a well-known effect, mainly due to the decrease in efficiency in some of the most important components for the energy balance (for example turbine and boiler).
However, a minimum size exists for advanced commercial plants. In fact, they are so complex that the realization of a small plant is not economically attractive. The smallest commercial coal-fired USC plants provide about 400 MW_e net output. In this range, the effect of the size on the performance of the components is not very sizeable and efficiency is not significanly affected [52].
Since the goal of this study is the elaboration of a simplified and flexible programme to estimate the cost of avoided CO_2 without running a simulation for each possible case, the only viable solution is the adoption of an approximation. Therefore we can assume that, in a certain range of electric output, the efficiency is constant even for the proposed oxyfuel plants.
It is reasonable to set the lower value for that range to 350 MW_e, corresponding to a gross output of more than 400 MW_e. It should be noticed that this imprecise simplification has little impact on the variation in the cost of electricity. Indeed, the variable components of it do not increase as much as the ones related to the capital costs.

Total fuel and variable O&M costs are calculated by scaling them linearly. Therefore the variable operating costs are calculated by the simple formula $c=c_0*S/S_{ref}$ for all the components. Then the associated components of the COE do not vary.

4 INFLUENCING PARAMETERS

The final goal of this work is the evaluation of the cost per tonne of CO_2 avoided in a plant equipped with a carbon capture & storage (CCS) system for different plant solutions.
Several design parameters with a large influence on the thermodynamic and economic performance can be identified. Some of them are modificable in simulations and/or considered. They are:

- size
- boiler configuration (PC vs. CFB)
- recycle rate and configuration of the recirculation flows
- adoption of technologies for the control of the pollutants
- fuel feed

Different solutions have been studied varying these parameters duly.
Other important variables (for which further studies are needed) are:

- type and performance of the ASU
- adoption of co-storage of SO_x and NO_x with the CO_2-rich stream
- adoption of sour compression

Unfortunately, relevant uncertainties exist around these issues and their implications on mass and energy flows, so it was impossible to include them in the analysis. On top of it, no sufficient economic data were available.

4.1 SIZE

At the moment, no large power plant is running under oxyfuel conditions. In fact, only several test facilities are operating in a oxy-fuel mode. The largest operating plants are: Callide A (CS Energy, 30 MW_e), Schwarze Pumpe (Vattenfall, 30 MW_{th}), Alliance (Babcock & Wilcox, 30 MW_{th}), Lacq (Total, 30 MW_{th}), Ciuden (Endesa, 30+20 MW_{th}).
The size of these plants is not comparable to the typical ones of the recently-built air-fired groups, that are usually designed to produce more than 800 MW_e (net) to exploit the scale economies. In Australia, smaller plants are operating. However, their net electric output is usually close to 500 MW_e.
The influence of the scale economy is significant for the capital cost of the power plants, affecting deeply the specific capital expenses ($/kW) and therefore the cost for tonne of CO_2 avoided.
For this kind of plants, the influence of the scale is usually weighted by means of an equipment scaling factor with an exponent [48] that is usually in the 0.6-0.9 range [37, 48].

As stated previously, to a first approximation, total variable operating costs are almost proportional to the power output, if the configuration is fixed and we are considering large plants. Obviously, an effect of the scale on the efficiency of the components also exists and should be considered, especially for small plants. In summary, large scale oxyfuel plants would be the ones associated with the best preformances and therefore are more interesting.
In this work, only large plants are simulated. However, the effects of scale have been investigated in an approximate way.

4.2 RECYCLES

The pros and cons of the concepts underlying different recycle configurations have been described previously. In this work, an all-dry recycle case has been evaluated for the PC oxyfuel plant using black coal, in addition to the base case with dry primary recycle and wet secondary recycle. This variant has been studied in order to consider the effect of a change in recycle configuration on the performance of the plant.
Usually a wet-hot secondary recycle slightly improves the efficiency as compared to wet-cold recycle, although particulates removal system and recirculation fan work in less appropriate conditions.
Both wet-cold and wet-hot solutions could be evaluated for the secondary recycle. However, the overall performance would not change significantly, so a single configuration could be evaluated.
A secondary very hot recycle has been used for the PC case using lignite as fuel to allow a proper drying, as described previously, in addition to a hot recycle, which is used to guarantee suitable combustion temperatures.
For CFB cases, a single configuration has been taken into account, which is the commonly accepted one, although uncertainties exist about the recycle rate.

4.3 BOILER CONFIGURATION

Besides the pulverized coal (PC) case, the circulating fluidized bed (CFB) configuration has been evaluated. This latter technology is much more rare in working plants and the upscaling is still ongoing, but it is very promising, because it has several advantages over the PC one, as discussed previously. Economically, the impact should be evident. The main advandages of a CFB boiler are: lower specific investment expenses [16], better plant efficiency and fuel flexibility.
In order to switch from a PC boiler to a CFB boiler configuration, several changes are to be implemented:

- the recycle rate can be decreased: the solids act like a ballast and help the temperature control. It is thus possible to obtain a lower energy consumption by the recirculation fan, a smaller sectional area of the boiler and a reduction in the size of ducts and heat exchangers fo the recycle stream.
- in-furnace sulfur capture by injecting limestone and extracting gypsum can be performed. The consequent advantages are a decreased corrosion and a lower lower load for SO_x control systems (if they are needed).
- more uniform and lower temperature in the furnace. Importantly this, together with the lower CO_2 levels, causes a lower CO concentration in the flue gas, that, in turn, brings about a decrease in the corrosion in the ducts downstream.
- coal pulverization is not necessary.
- the coal feed can probably be wet. In that case the coal drying process would not be needed and primary recycle can be avoided.
- the pressure losses in the boiler increase.

In this work, both the configurations are taken into account and simulated.

4.4 POLLUTION CONTROL

In the PC cases (except for one), no pollution control devices other than ash removal are present. Several reasons back this decision:

- the oxyfuel combustion process naturally reduces the amount of pollutants such as NO_x in the flue gas.
- The opportunity of the co-storage of the pollutants is under investigation and could prove its feasibility in the near future.
- Recent studies identified a promising technology to remove nitrogen and sulphur compounds together with condensed water [9].

In Australia, no limits regarding NO_x and SO_x concentration in the vented gas exist.

A variant of the base case including SO_x removal has been considered. It can be foreseen that the mass balance is not affected heavily by the change. At the same time, a slight energy penalty is unavoidable.

In the CFB cases, in-furnace sulphur capture has been performed, since it is quite cheap and easy. Moreover, it could result in a simplification of the design of the flue gas treatment section, because the risk of corrosion is reduced and the compression phase is less complicated.

4.5 FUEL

Two different coal types have been evaluated: a South African black coal and an Australian brown coal from the Lao Yang site.
The South African coal has characteristics that are similar to the ones of a fine Australian black coal, especially if we compare two dried samples. The description and the equivalent fuel for the SA coal are reported in the table below.

Table 11. South African coal real composition (%wt.), equivalent composition (%vol.) and Heating Values.

	%vol.
CH_4	2.18
CO	1.97
CO_2	2.40
C_2H_2	0.00
C_2H_4	0.00
C_2H_6	0.72
C_3H_8	6.09
H_2	3.74
H_2S	0.00
NH_3	0.01
N_2	0.93
O_2	0.70
S	0.47
H_2O (L)	9.02
C (S)	68.55
Ash (S)	3.22
	% wt.
C	64.44
H	3.95
N	1.49
O	7.40
S	0.85
Moisture	9.20
Ash	9.20
LHV, MJ/kg	24.62
HHV, MJ/kg	25.71

This coal is dried in the PC cases, so that it can reach a 5% moisture level before being fed to the combustion process, whereas in a CFB boiler the residence time is much longer. That means that there is no need to dry the coal before feeding it into the furnace.
Australia has very large reserves of high-moisture brown coal, so it is interesting to evaluate even a plant fed by lignite. As explained previously, the drying process is quite different in this case, so the GS code has to undergo some modifications to become more representative of the plant equipped with a PC boiler and fed by brown coal.
The characteristics of the brown coal used in this work, that are the ones of a sample from the Lao Yang site, are reported in the table below:

Table 12. Lao Yang coal composition and heating values.

		Concentration, % weight
Lao Yang brown coal	C	25.92
	S	0.11
	N	0.23
	H	1.82
	O	9.23
	Ash	0.68
	H_2O	62.00
	LHV (MJ/kg)	NA
	HHV (MJ/kg)	10.055

4.6 GS CODE MODIFICATIONS

In this section, the modifications that have been carried out to describe new configurations are briefly reported. A descriptive approach is used here to make this chapter understandable for those who are not familiar to the GS code (and in particular to its versions used for oxyfuel plants).
However, it should be taken into account that working with GS code has been one of the most important and challenging parts of this thesis.
As stated previously, a base case simulation was provided and then modified to consider different configurations.

4.6.1 POLIMI CASE TO BASE CASE

To build the base case, the pollution control systems for sulphur oxides have been switched off. In particular the partial SO_2 split and the loss related to the FGD in the primary recycle are excluded.
In the figure below, a schematic of the simulated base case plant is shown.

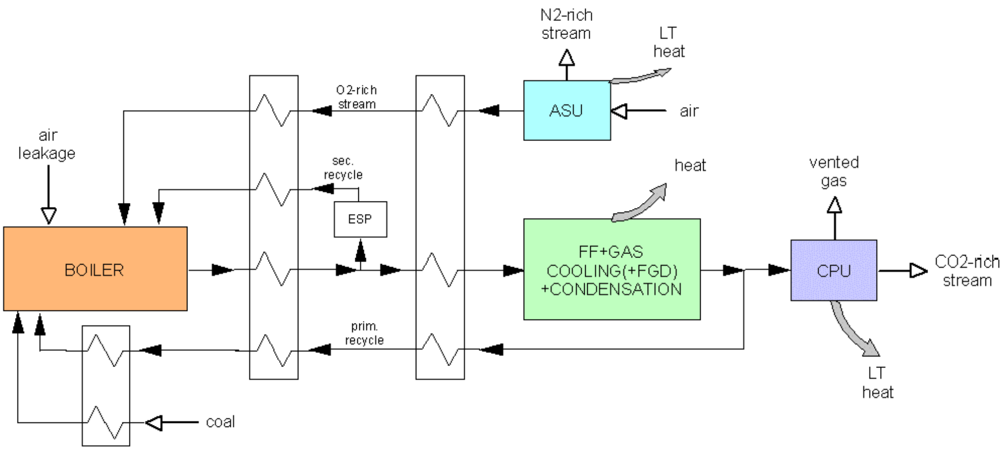

Figure 26. Schematic of the simulated base case oxyfuel plant (gas side).

The primary recycle rate is selected to allow the required drying process of coal (residual moisture level in the dried coal is 5%), whereas the secondary recycle rate is set to have adequate adiabatic flame temperature (2000°C). The pemperature at the boiler outlet is instead 350°C.
The temperatures of the O_2-rich stream and of the primary recycle stream at the boiler furnace inlet are 150°C and 80°C, respectively.
Oxygen production is defined to give a certain O_2 concentration (2.5% mol.) at boiler outlet. ASUs consumption has been set to 200 kWhel/kg of impure oxygen and makes some low temperature heat (40 kJth/kg of impure oxygen).
The performance of the compression and purification unit is not evaluated through GS code, because it is simulated on Aspen Plus, but it is taken into account that the CPU provides some low temperature heat (450 kJth/kg of impure CO_2).

In addition to the heat recovery from the ASUs and the CPU, the gas cooling section provides heat to the steam cycle. More specifically, part of the feed water is preheated in the gas cooler. This solution reduces the load on the middle rigenerative preheaters, so the need of steam extraction from the IP and LP turbines decreasses.
Moreover, two heat exhangers providing heat from the main stream to the ASUs and to the recycle streams were taken into account. However, the high-temperature one has been deactivated, because this configuration is the most efficient and requires lower expenses for heat exchangers, though operating conditions are more critical for some components and the long term reliability of this approach has not been proven yet. Nevertheless, the adoption of wet-cold recycle instead of wet-hot recycle would have minor impacts on the overall performance of the plant and, in case, could be considered through a simplified incremental approach.
Air in-leakages have been considered throughout the plant (especially in the boiler, but also in the FF and in the ESP), as well as pressure losses (bolier, heat exchangers on both gas and steam-water sides, ESP, FF).
Ash removal efficiency has been set to 85% for the ESP treating the secondary recycle stream and to a virtual 100% for the fabric filter treating the main stream.
Temperatures and pressures at several points in the plant (especially in the steam cycle) are imposed to replicate the real plant operating conditions. The steam turbine discharge velocity is set to 255 m/s and therefore, since the end section is also fixed, the volumetric flow exiting the LP turbines is set.
Pulverization and coal handling require 50 kJel/kg of dry coal, whereas ash hadling consumption is 200 kJ/kg. Heat for coal heating (about 90 kJ/kg) and heat from coal milling (about 20 kJ/kg) have been taken into account.

4.6.2 BASE CASE TO ALL-DRY CASE

To evaluate the all-dry recycle case, the split associated with the secondary recycle is removed and the primary split fraction is increased. The recycle rate is chosen chosen to guarantee appropriate combustion conditions.

4.6.3 BASE CASE TO CFB CASE

To describe the CFB case with the GS code, the following changes are introduced to the base case:

- the exchangers representing the coal drying process must be “switched off”.

- The recycle is totally wet, as there is no more need to use dry gas to dry the coal.
- The recycle rate is set to attain an adequate (50%) O_2 concentration in the oxidant stream.

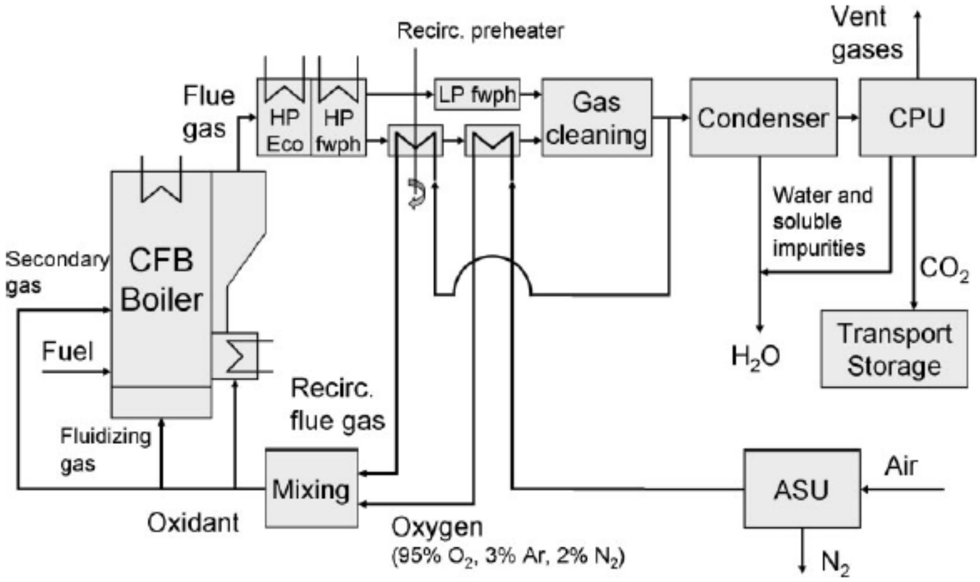

Figure 27. Schematic of an oxyfuel power plant, adopted from [15].

- No heat from coal milling is available and no heat for coal heating is required.
- Pulverization and coal handling consumption is changed. In particular the consumption is reduced to 30 kJ/kg of coal from 30 kJ/kg in PC cases.
- The ash handling consumption is reduced to 100 kJ/kgash. In the PC configurations the demand is double.
- A virtual new component is necessary to account the limestone handling consumption, that is estimated at 100 kJ/kg of $CaSO_4$, because actually gypsum is mixed with ash.
- To take into account in-furnace SO_2 removal, three new components are inserted: a SO_2 splitter, an O_2 splitter and a CO_2 mixer. They represent the inlets and outlets in the gas stream due to the direct sulfatation reaction ($CaCO_3$+SO_2+1/2O_2→$CaSO_4$+CO_2).
- $CaCO_3$ consumption is related to the SO_2 concentration in the gas exiting the boiler. For example, a Ca/S ratio equal to 2 can be imposed.
- Pressure losses are modified. In particular, the pressure losses in the heat exchangers representing the boiler increase on the hot size (0.14 bar).

- Ash treatment consumption should comprise $CaCO_3$ handling (even if its presence is not considered in the simulation). We add a new component to take into account those losses, which are set to 90 kJ/kg$CaCO_3$.

4.6.4 BASE CASE TO LIGNITE-FIRED PC CASE

In this case, the main modifications occurred are:

- the alimentation is changed. So the boiler is fed by a dried lignite.
- To dry the lignite to a moisture level acceptable for the combustion, even an all-dry recycle would not be enough. In fact, it is impossible to use high-moisture fuel. The maximum acceptable value is about 18%. In this simulation, coal is dried to 15%. The traditional drying process is switched off. Only a significant very hot wet recycle (conceptually bled from the boiler at a certain temperature) is taken into account. This is a simple process, explained previously, that is used in the majority of the plants fed by very high-moisture solid coals, although it is not efficient and causes a sizeable efficiency penalty.
- The primary dry recycle does not exist, whereas the secondary wet recycle is used to mitigate the temperature in the boiler.

4.6.5 CFB BASE CASE TO LIGNITE-FIRED CFB CASE

To describe the CFB case using lignite as fuel, the coal composition is modified to consider the values associated with the brown coal, instead of the ones related to the South African black coal.

5 RESULTS

In the following paragraphs, the thermodynamic and economic results for the different cases will be reported and commented and a comparison with the most relevant analogous studies on the subject will be showed.

5.1 MASS BALANCE

The chosen configuration and a the data resulting from simulations are reported and commented. Detailed information is provided especially for the base case plant design.

5.1.1 MASS BALANCE FOR THE BASE CASE

A schematic of the oxyfuel base case is reported, as well as tables reporting the detailed features of the mass flows at several relevant points.

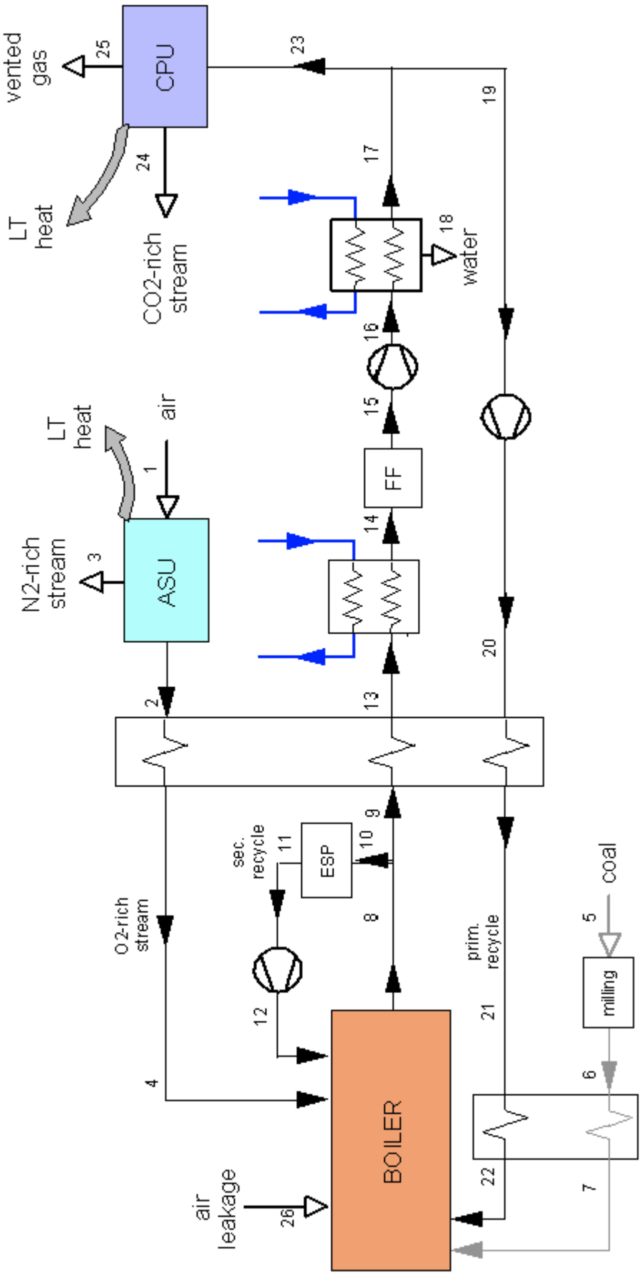

Figure 28. Schematic of the gas side of the oxyfuel base case plant.

Table 13. Main thermodynamic features at several points of the plant (gas side).

N	gas/ fuel/ liq.	G (kg/s)	T (°C)	P (bar)	W (kg/kmol)
1	g	720.17	15.0	1.01	28.851
2	g	174.16	15.0	1.60	32.160
3	g	546.01	15.0	1.60	27.934
4	g	174.16	150.0	1.50	32.160
5	f	82.28	15.0	2.00	16.556
6	f	82.28	48.4	2.00	16.556
7	f	82.28	71.1	2.00	16.556
8	g	841.29	350.0	0.99	37.751
9	g	423.65	350.0	0.99	37.751
10	g	417.65	350.0	0.99	37.751
11	g	417.02	347.0	0.98	37.442
12	g	417.02	356.7	1.04	37.442
13	g	423.65	243.3	0.98	37.751
14	g	423.65	190.0	0.98	37.751
15	g	422.16	188.3	0.96	37.407
16	g	422.16	198.6	1.05	37.407
17	g	389.53	30.0	1.03	41.114
18	w	32.626	30.0	1.03	18.000
19	g	164.00	30.0	1.03	41.114
20	g	164.00	34.2	1.08	41.114
21	g	164.00	222.4	1.07	41.114
22	g	164.00	80.0	1.04	39.321
23	g	225.53	30.0	1.03	41.114
24	l	188.5		150	
25	g	37.0		1.01	31.409
26	g	38.26	15.0	1.01	28.851

As stated previously, the mass flows entering the boiler are: dried fuel, O_2-rich stream, primary and secondary recycle and air leakage (5% of the other mass flows as default). The flue gas exiting the boiler are associated with the significant enthalpy of a large mass flow at 350°C. Part of it is used for combustion temperature mitigation (secondary recycle), after ash removal in an ESP.

The rest of the flue gas is then cooled to 190°C and the obtained heat is used to preheat oxygen (from 15°C to 150°C) and primary recycle stream (from 34°C to 222°C) and also to preheat the water in the steam cycle, reducing the need of steam for rigeneration. Ash is then removed in a fabric filter and the stream is compressed by a fan, before further cooling and water condensation at 30°C. Part of the resulting gas is sent to the CPU and the remaining share is recirculated, compressed by a fan, reheated to 222°C and used to dry and

transport coal. The other share is sent to the CPU, in which it is purified and compressed.
Interestingly, air in-leakages in the ESP and FF and pressure losses throughout the plant have been considered.

Table 14. Composition of the mass flows at several points of the plant (gas side).

% vol.	Ar	CO_2	H_2O	N_2	O_2	SO_2	Ash
1	0.920	0.030	1.034	77.282	20.733	0.000	0.000
2	3.024	0.000	0.000	1.976	95.000	0.000	0.000
3	0.337	0.038	1.321	98.146	0.158	0.000	0.000
4	3.024	0.000	0.000	1.976	95.000	0.000	0.000
5							
6				fuel: see composition above			
7							
8	2.569	67.944	19.601	6.325	2.500	0.336	0.726
9	2.569	67.944	19.601	6.325	2.500	0.336	0.726
10	2.569	67.944	19.601	6.325	2.500	0.336	0.726
11	2.563	67.487	19.482	7.276	2.750	0.334	0.108
12	2.563	67.487	19.482	7.276	2.750	0.334	0.108
13	2.569	67.944	19.601	6.325	2.500	0.336	0.726
14	2.569	67.944	19.601	6.325	2.500	0.336	0.726
15	2.566	67.562	19.504	7.282	2.752	0.334	0.000
16	2.566	67.562	19.504	7.282	2.752	0.334	0.000
17	3.056	80.476	4.118	8.674	3.278	0.398	0.000
18			100.00				
19	3.056	80.476	4.118	8.674	3.278	0.398	0.000
20	3.056	80.476	4.118	8.674	3.278	0.398	0.000
21	3.056	80.476	4.118	8.674	3.278	0.398	0.000
22	2.819	74.230	11.560	8.001	3.024	0.367	0.000
23	3.056	80.476	4.118	8.674	3.278	0.398	0.000
24	0.836	96.460	0.000	1.251	0.944	0.509	0.000
25	11.180	21.982	19.188	35.838	11.821	0.000	0.000
26	0.920	0.030	1.034	77.282	20.733	0.000	0.000

The main compound in the flue gas is carbon dioxide. Its concentration is 80.5% after water condensation and 96.5% at the CPU outlet. The high concentration in the vented gas could make CO_2 post-combustion capture from this small stream interesting, because of the high CO_2 partial pressure.
The most relevant other compounds are water (before condensation), nitrogen, oxygen and argon. The concentration of these last three components depend on O_2 purity and, more importantly, on the significance of air in-leakages.

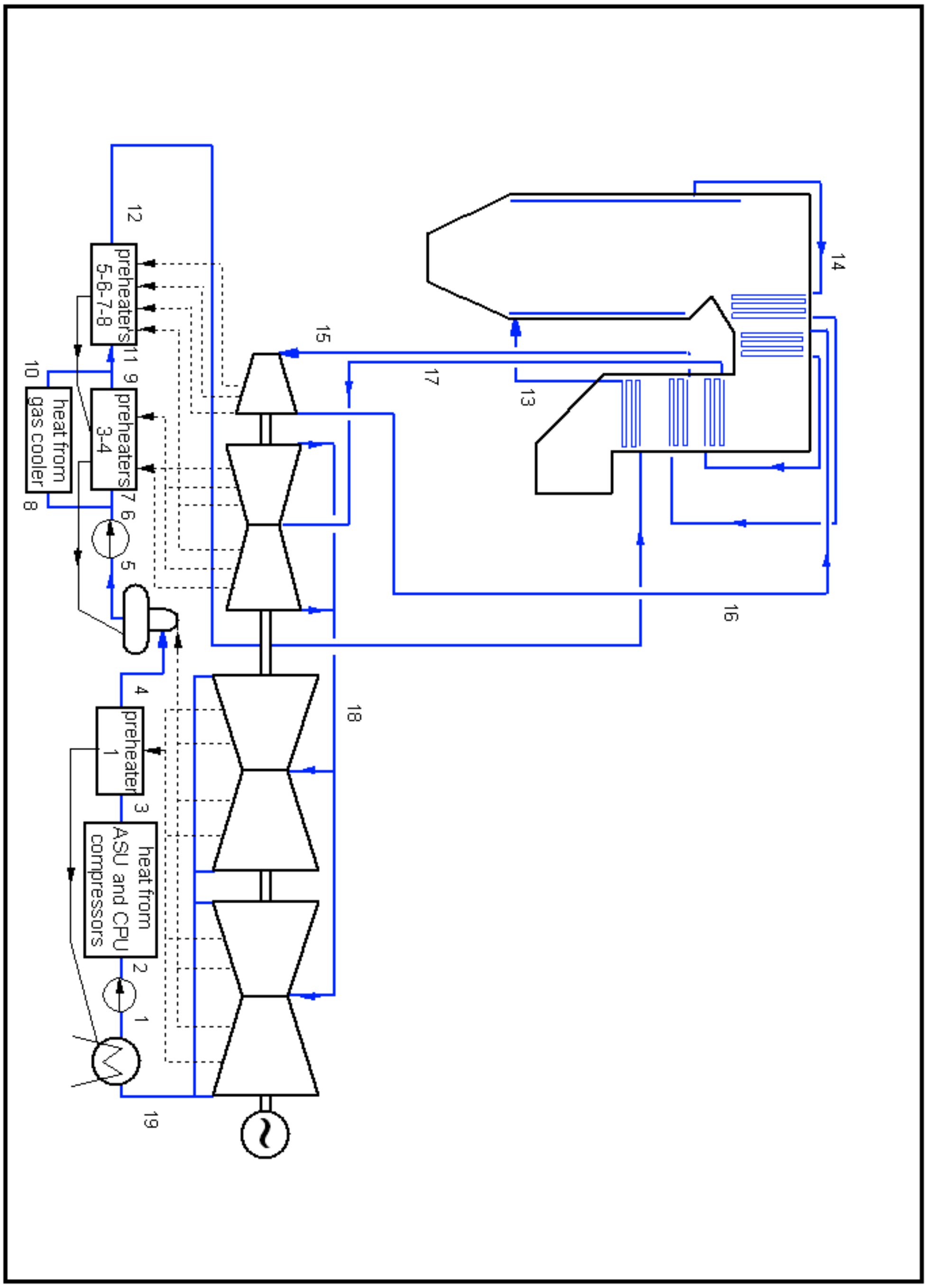

Figure 29. Schematic of the steam-water side of the oxyfuel base case plant.

Table 15. Main thermodynamic features at several points of the plant (steam-water side).

Number	Point	Destination	water/ steam	G (kg/s)	T (°C)	P (bar)
1			w	510.88	29.0	0.04
2			w	510.88	29.1	9.00
3			w	512.72	89.81	7.700
4			w	512.72	120.28	6.700
5			w	736.17	149.53	4.700
6			w	736.17	154.66	347.000
7			w	636.17	154.66	347.000
8			w	100.44	154.66	347.000
	8A	to preheater 4	w	636.17	180.53	346.000
9			w	636.17	211.43	345.000
10			w	100.44	211.0	347.00
11			w	736.17	211.37	347.00
	11A	to preheater 6	w	736.61	244.11	344.000
	11B	to preheater 7	w	736.61	268.07	343.000
	11C	to preheater 8	w	736.61	293.66	342.000
12			w	736.61	314.97	341.000
13			w	736.61	360.0	337.00
14			s	736.61	400.0	317.00
15			s	736.61	598.0	279.00
16			s	611.14	339.1	55.50
17			s	611.14	608.0	50.00
18			s	483.76		62.000
19			s/w (x=0.8923)	434.70	29.0	0.0400
Spillamenti						
				G (kg/s)	Tbl (°C)	P (bar)
1		to preheater 1	s	26.970	186.42	2.20
2		to deaerator	s	23.142	275.26	5.05
3		to preheater 3	s	19.578	366.81	10.46
4		to preheater 4	s	24.489	455.07	19.42
5		to preheater 5	s	33.657	544.09	34.23
6		to preheater 6	s	36.466	339.12	55.51
7		to preheater 7	s	44.336	388.81	79.48
8		to preheater 8	s	36.380	430.78	105.26

Low temperature heat available from intercooled compressors (more than 100 MW_{th}) is used to preheat water to almost 90°C. Subsequently, water preheating

is mainly attained by rigeneration, by means of bleedings from the turbines. Eight rigenerative preheaters (a surface heat exchanger, then a deaerator and six other surface heat exchangers) are used.
However, the load on the middle preheaters is reduced by heat recovery (about 24 MW_{th}) from flue gas in a gas cooler.

5.1.2 RECYCLES

In case 1, 49.6% of the flue gas exiting the furnace are used as secondary wet recycle, whereas 19.5% of the total flue gas (corresponding to 42.1% of the stream before the splitting point) constitute the primary stream for coal drying. Both the streams contribute to the moderation of the temperature in the furnace.
The mass balance of case 6 is almost identical to the base case.
In case 2, only a dry stream recycle is instead considered. Its mass flow accounts for 68.3% of the total mass flow at boiler outlet. It allows coal drying and acts like a ballast in the furnace.
A single wet recycle is used for case 3 and case 5. Respectively, 42.4% and 26.6% of the flue gas are recirculated to attain an appropriate O_2 concentration in the oxidant. The lower recycle rate in case 5 is lower because of the more sizeable amounts of oxygen injected by the coal feed.
In case 4, a very hot wet recycle steam (34.9% of the flue gas) is used for coal drying and a wet-hot recycle (19.24% of the remaining mass flow) is required for temperature mitigation.
More details about mass balance in cases other than the base one are available, but not reported, even because many values and related considerations are similar to those in the base case.

5.1.3 COMPOSITION OF THE CO_2-RICH STREAM

The gas composition at the CPU inlet and the electric output excluding the CPU for all the analysed cases is reported in the following table.

Table 16. Estimated gas composition at CPU inlet and electric output if CPU is not included for the considered cases.

Composition (% mol.)	without FGD					with FGD
	South African black coal			Loy Yang brown coal		SA bl. Coal
	PC		CFB	PC	CFB	PC
	wet sec. recycle	all-dry recycle				
	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Ar	3.1	3.1	3.2	2.8	3.0	3.1
CO_2	80.5	80.8	83.9	79.6	81.2	80.6
H_2O	4.1	4.1	4.1	4.1	4.1	4.1
N_2	8.7	8.6	5.1	5.1	4.8	8.7
O_2	3.3	3.0	3.7	5.3	6.9	3.3
SO_2	0.4	0.4	0.0	3.1	0.0	0.2
Total	100.0	100.0	100.0	100.0	100.0	100.0
Flowrate (kg/s)	225.5	228.4	217.8	274.0	260.4	224.8
Gross power output (MW)	970.6	936.0	973.0	910.6	909.8	968.2
Net power output (MW)	746.8	716.8	752.3	663.5	649.8	739.0

Vapour content is constant as the conditions for the condensation process are set.

CO_2 is obviously the main component and its concentration is about 80% vol. for all the plants. O_2 concentration is slightly higher for brown coal feed. The concentration at boiler outlet, but at that point H_2O concent is higher for lignite-fired plants.

Oxygen concentration is imposed at the boiler exit, but the concentration at the CPU inlet section changes consistently. In particular, it is high in the cases, for which there is no dry recycle. It is even higher for the lignite-fired plants. These two conditions increase the H_2O presence in the boiler. Since water is condensed before the CPU, O_2 concentration of the flue entering the CPU is higher for those configurations.

SO_2 concentration is quite low for all the cases, with the exception of case 4, whose fuel sulphur content is higher if water and H_2O-generating components are negleted. The lowest SO_2 concentration is observed in the CFB plants, for which in-furnace removal is performed. Importantly, flue gas desulphurization on the small primary recycle stream (in case 6) is able to cut SO_2 concentration by about 50%.

Mass flow rate entering the CPU is largely dependant on the fuel and is higher for lignite-fired plants. Another influencing parameter is efficiency: the higher it it the lower fuel consumption and flue gas flow are.

Table 17. Composition and mass flow of the gas exiting the CPU and net electric output (taking into account CPU consumption).

Composition (% mol.)	without FGD					with FGD
	South African black coal			Loy Yang brown coal		SA bl. Coal
	PC		CFB	PC	CFB	PC
	wet sec. recycle	all-dry recycle				
	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Ar	0.8	0.9	1.0	0.8	0.9	0.8
CO_2	96.5	96.5	96.9	93.0	96.3	96.7
H_2O	0.0	0.0	0.0	0.0	0.0	0.0
N_2	1.3	1.3	0.9	0.8	0.7	1.3
O_2	0.9	0.9	1.2	1.6	2.1	1.0
SO_2	0.5	0.5	0.0	3.8	0.0	0.3
Total	100.0	100.0	100.0	100.0	100.0	100.0
Flowrate (kg/s)	188.5	191.8	190.7	232.4	221.0	188.2
CO_2 captured (%)	93.6	93.8	95.5	94.7	94.2	93.6
Gross power output	970.6	936.0	973.0	910.6	909.8	968.2
Net power output	746.8	716.8	752.3	663.5	649.8	739.0

As shown in the table above, at the exit of the purification unit, the cumulative concentration of oxygen, nitrogen and argon is lower than 4%, that is a quite accepted quality requirement for this major impurities. On top of it, the CO_2 purity is about 96.5% as default, except for the lignite-fired PC plant, for which the SO_2 concentration is quite high and for which SO_2 removal could be required.

SO_2 concentration is low especially in the CFB plants, for which in-furnace SO_2 removal is performed.

Most of the nitrogen and large shares of argon and oxygen are removed from the CO_2-rich stream in the compression and purification unit. The related energy consumption is slightly more than 400 kJ/kg of pure CO_2 produced. However, efficiency penalty related to the CPU is higher for the lignite-fired plants, for which the flue gas mass flow is higher and the power output (without considering the CPU demand) lower.

Before the purification unit the quality requirements about non-condenables are not met, so the CPU is fundamental to reach the goal with acceptable efficiency

penalties (lower than 4.5%), while mantaining high CO_2 capture rates (higher than 93%). The purification process also contributes to the achievement of low costs of CO_2 avoided, because it reduces the demands for the compression of uncondensables and keeps the CO_2 emissions low. However, optimum concentrations for capture cost minimization have not been investigated.
The following table reports the specific CO_2 emissions for all the investigated cases.

Table 18. LHV efficiency, specific CO_2 emissions and net output of the considered plants.

	without FGD					with FGD
	South African black coal			Loy Yang brown coal		SA bl. Coal
	PC		CFB	PC	CFB	PC
	wet sec. recycle	all-dry recycle				
	Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Efficiency	36.87	34.83	37.42	30.81	33.65	36.51
Specific CO_2 emissions (kg/MWh)	60.41	61.30	41.69	59.99	59.63	60.56
Net output (MW)	746.85	716.75	752.28	663.50	649.79	739.00

It can be easily noticed that CO_2 specific emissions are much lower than in traditional coal-fired plants, whose emissions are higher than 600 kg/MWh, even for the most efficient plants. All the plants vent about 60 kg/MWh, with the exception of case 3, which releases only 41.7 kg/MWh, mainly because the gas entering the CPU are more easy to treat: SO_2, O_2 and N_2 concentrations are low, making the purification easier and allowing a high capture rate.

5.2 ENERGY BALANCE

In this section, data from simulations of the actual process, including consumption in each system and thermal efficiency, are reported and commented.

5.2.1 BASE CASE

The base case is quite similar to the PC with all-dry recycle and to the case with FGD on the primary recycle stream. On top of it, it is not too different from the most studied design scheme for oxyfuel coal-fired plants, because it is the natural evolution of conventional air-fired plants. Therefore it is interesting to commented the related results.
In the following table, the energy balance is reported.

Table 19. Resuming table of the energy balance of the base case.

	Power (MW)	% LHV
Thermal input	2025.6	
Mechanical power at blades	985.4	48.65
Electric power at generator	970.6	47.92
ASU	-118.4	-5.85
CPU	-77.5	-3.75
Condenser auxiliaries Consumption	-10.8	-0.53
Compressors	-10.5	-0.52
Pulverization and coal Handling	-3.8	-0.19
Ash handling	-2.1	-0.10
Miscellaneous	-0.6	-0.03
Net electric	746.8	36.95

As expected, the majority of the losses are related to the air separation process and to the compression and purification section, that are typical of oxyfuel power plants and cause a significant efficiency penalty as compared to conventional coal-fired plants.
The CO_2-rich stream (96.5% vol. purity) is compressed to 150 bar before transportation and storage, that is about 19500 tonnes of pure CO_2 per day. The calculated CPU consumption is 413.2 kJ/kg of pure CO_2 entering the CPU, that makes available 640 kJth/kg of pure CO_2. This is consistent with the assumptions: 522 kJth/kg of pure CO_2, or 101.5 MW_{th} were considered in the GS simulation as LT heat from the CPU. 181.87 kg/s of CO_2 are available for storage, representing the 93.55% of the CO_2 total production.
The ASU provides about 14200 tonnes of pure oxygen per day, that can be supplied by two advanced large scale cryogenic ASUs. The LT heat from the ASUs is 28.8 MW_{th}, or 40 kJ/kg of air entering the units.

Table 20. Comparison of the estimated efficiency for base case against the ones reported by other studies on oxyfuel plants.

	This work (PC, black coal)	Mitsui Babcock (base case)	Pelliccia (advanced large, wet recycle)	DOE (case 6)	Davison 2007	ENCAP	MIT
Net output (MW_e)	746.8	532.0	742.3	550.0	532.0	472.0	500.0
Year	2009	2005	2003	2007	2007	2004	2007
LHV efficiency (%)	37.0	35.4	38.5	33.0	35.4	36.0	29.3

The table above shows that the results of the simulations are in line with available literature, although expected efficiency in case of black coal feed is in the slighly higher than the one assumed by many other studies, but some of them have an additional penalty related to the presence of the FGD.
Several other reasons can explain this difference. Some of them are:

- some of the studies consider smaller plants, so efficiency penalties can be foreseen in the performance of some components (for example for the turbomachines).
- the plants simulated here are quite complex. A large share of the available heat is recovered and used within the plant, reducing the efficiency penalty.
- the simulated plants are equipped with optimised steam cycle and turbomachinery.
- some of the reference studies were performed a few years ago and they do not take into account the most recent technological developments.

5.2.2 OTHER CASES

Among the other things, GS simulations provide the expected energy balances for the maximum size plants. The resulting efficiencies are reported in the following table, as they are very important to evaluate the design of the plant and to perform the economic analysis.

Table 21. Resuming table of the most important parameters describing the simulated plant.

		without FGD					with FGD
		South African black coal			Loy Yang brown coal		SA bl. Coal
		PC		CFB	PC	CFB	PC
		wet sec. recycle	all-dry recycle				
		Case 1	Case 2	Case 3	Case 4	Case 5	Case 6
Net power output	MW_e	746.8	716.8	752.3	663.5	649.8	739.0
Gross Electric Power	MW_e	970.6	936.0	973.0	910.6	909.8	968.2
LHV Efficiency	%	36.9	34.8	37.3	30.8	33.7	36.5
Pure O2 production	tpd	14223.2	14414.1	14324.4	15372.3	16348.3	14213.8

The highest expected efficiency is the one obtained in the plant without FGD in which a CFB boiler is fed by black coal, whereas the lowest one is associated with the PC plant using brown coal as fuel, where the drying process involves a significant deterioration of the quality of the cycle.
The advantage of using a CFB boiler is slight when using a black coal (similar to the one previously reported in literature [16]), but is more evident if the fuel is brown coal, thanks to the different requirements for drying.
It is noteworthy to highlight that the plants using black coal are clearly more efficient, although the decrease is not so sizeable. Therefore an economic analysis could provide interesting results for this cheaper fuel.
Cases 1, 2 and 6 share the majority of the plant design and adopt black coal for combustion in a PC boiler. The results of the simulations are similar too,

although case 2 and case 6 show an efficiency penalty related, respectively, to the less efficient recycle strategy and to flue gas desulfurization.
As said previously, the gross power changes among the different cases, as the binding parameter is the discharge velocity at the exit of the last turbine stage and the different sections have different amounts of steam bleedings.
Another interesting parameter that is worth to be commented is the oxygen production. In fact, it is necessary for large plants to set more units in parallel to supply the requested mass flow. Currently, the production of the largest designable cryogenic ASU is 7000 tonnes per day, although this limit can be raised to 10000 tonne/day in the future [12]. That means that at least two trains are needed to supply the requested oxygen mass flow. The number of necessary units also depends on the availability and the safety of the plant and on the feasibility of storing the oxygen.
A single study [43] provided an estimate of the efficiency of an PC oxyfuel plant in case of a brown coal feed. The expected efficiency is higher in that study. Reasonably, at least part of that that difference can be brought back to the better quality of the coal (52% moisture vs. 62% in this study).
A recent study [46] evaluated the performance of an oxyfuel CFB fed by black coal with a 512 MW_e net output. It estimated a 36.6% efficiency, that is slightly lower than the one expected in this study, but substantially in line with it. The difference could be attributable to the inferior size of the plant.
Compared to other the proposed solutions being studied for CO_2 capture, the expected efficiency for an oxyfuel plant is slightly higher than the one of a post-capture plant, but usually lower than the one associated with pre-combustion system, although some studies expect similar efficiency [47].

5.3 DIRECT MECHANICAL DRIVE OF COMPRESSORS

Compressors of the air separation units and of the CPU cause a significant efficiency penalty, respectively higher than 5% and 3% in all the analysed configurations. Those compressors are driven by electric motors.
This loss could be reduced, if direct mechanical drive is considered for those compressors. Indeed, the possibility of direct mechanical drive through a joint shaft for compressors in ASUs and CPU has been investigated. In particular, two options are available: compressors can be driven directly by the main steam turbines or by a LP turbine driven by extracted steam. But the main steam turbine shaft cannot be used for this purpose, since it would cause problems both in the compressor units, such as surge at start up and in the shaft itself because of too large thermal motions [43].
Moreover, due to the issues of startup time requirement and upstream location of ASU, steam driven air compressors for ASU are not very interesting.

If extracted steam is adopted instead, it would be taken from the IP/LP crossover or a nearby extraction point. This solution would complicate the steam cycle, causing an increase in power consumption, and start-up procedures [11]. On top of it, the isoentropic efficiency of the large main steam turbines for power generation is higher than the efficiency of smaller turbines.
On the other side, direct mechanical drive would reduce the power losses associated with generator, transformer, motor, frequency converters, and gear loss. Moreover, the low pressure end steam flow would be reduced, increasing the steam turbine efficiency and/or the power output. Savings from steam driven compressors could be up to 10% as compared with those driven by motors, that would result in a 0.4% efficiency gain.
According to another study on oxyfuel power plants [53], more than 10% net saving on CPU consumption (corresponding to 0.5% difference in LHV efficiency) would be obtained for CO_2 compression with four stages, if steam driven CO_2 compressors were applied. The corresponding end steam flow through main steam turbine would be reduced. For CO_2 compressors driven by extracted steam, this is not a problem as they are located downstream of boiler with much less startup time.
Another study estimates [43] instead that direct mechanical drive of both compressors of the ASU and of the CPU would cause a significant power saving (about 0.7% of the net output, or 6.5% of the total ASU and CPU consumption).

5.4 REFERENCE PLANTS

The selected reference plants are advanced USC plants fed by South African bituminous coal. The power output is not set, but the height of the blades turbine last stage and the discharge velocity (255 m/s) are selected, in accordance with the approach adopted for oxyfuel plants.
Two plants have been considered to meet the requirements of the European/U.S. and Australian scenario, respectively. The former simulated plant is equipped with pollution control devices for the abatement of NO_x and SO_x, whereas those compounds are not controlled in Australia, nor is a change in policy foreseen about this issue, so those components are absent in the reference plant for the Australian context.
The procedure to calculate the cost of electricity is the same adopted for oxyfuel plants, although a few different assumptions have been made, as stated previously.
As compared to the oxyfuel combustion configuration, the net output increases not only because the ASU and the CPU are absent, but also because more steam is produced and used in the high- and intermediate-pressure turbine stages to produce electric power. Indeed, the amount of steam flowing through the last

turbine stage is equivalent to oxyfuel cases (because of the assumptions), but more significant mass flows are bled for rigeneration.
In particular, ten preheaters (3 surface heat exchangers, then a deaerator and other 6 surface heat exchangers) have been taken into account.

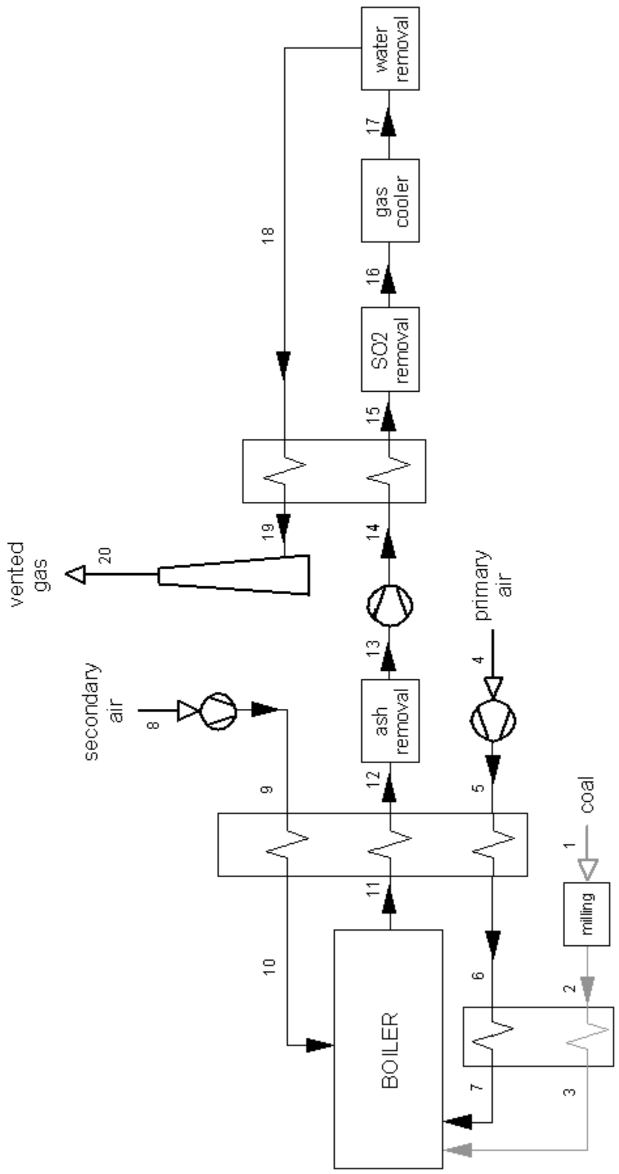

Figure 30. Schematic of the gas side of the reference air-fired plant.

Table 22. Main thermodynamic features at several points of the air-fired reference plant (gas side).

N	gas/ fuel/ liq.	G (kg/s)	T (°C)	P (bar)	W (kg/kmol)
1	f	88.545	15.0	2.00	16.556
2	f	88.545	41.3	2.00	16.556
3	f	88.545	45.1	2.00	16.556
4	g	186.02	15.0	1.01	28.851
5	g	186.02	21.6	1.08	28.851
6	g	186.02	180.0	1.07	28.851
7	g	186.02	175.8	1.07	28.851
8	g	744.08	15.0	1.01	28.851
9	g	744.08	18.7	1.05	28.851
10	g	744.08	300.0	1.04	28.851
11	g	1030.8	350.0	0.99	30.065
12	g	1030.8	130.0	0.96	30.065
13	g	1028.9	128.9	0.96	29.853
14	g	1028.9	140.7	1.05	29.853
15	g	1028.9	97.3	1.04	29.853
16	g	1027.4	97.3	1.04	29.830
17	g	1049.6	47.9	1.02	29.422
18	g	1048.2	47.9	1.02	29.447
19	g	1048.2	90.0	1.01	29.447

Flue gas exiting the boiler (at 350°C) is cooled in a heat exchanger and used to preheat the primary and secondary air. The former one is heated to 180°C and therefore used to dry the coal, whereas the latter one is preheated to 300°C and injected into the boiler.
The cooled flue gas stream at 130°C reach the ash removal section and is then compressed and cooled to less than 100°C. After SO_2 removal, the flue gas stream is cooled to less than 50°C. At that temperature, water is condensed and removed. Before being vented, the flue gas is reheated to 90°C.

Table 23. Composition of the mass flows at several points of the plant (gas side).

% vol.	Ar	CO_2	H_2O	N_2	O_2	SO_2	Ash
1							
2	fuel: see coal composition						
3							
4	0.920	0.030	1.034	77.282	20.733	0.000	0.000
5	0.920	0.030	1.034	77.282	20.733	0.000	0.000
6	0.920	0.030	1.034	77.282	20.733	0.000	0.000
7	0.920	0.030	1.034	77.282	20.733	0.000	0.000
8	0.920	0.030	1.034	77.282	20.733	0.000	0.000
9	0.920	0.030	1.034	77.282	20.733	0.000	0.000
10	0.920	0.030	1.034	77.282	20.733	0.000	0.000
11	0.870	14.983	7.863	73.215	2.487	0.074	0.507
12	0.870	14.983	7.863	73.215	2.487	0.074	0.507
13	0.875	15.059	7.904	73.588	2.500	0.074	0.000
14	0.875	14.905	7.833	73.626	2.687	0.074	0.000
15	0.875	14.905	7.833	73.626	2.687	0.074	0.000
16	0.875	14.905	7.833	73.626	2.687	0.074	0.000
17	0.876	14.915	7.838	73.675	2.689	0.007	0.000
18	0.845	14.400	10.805	71.131	2.596	0.007	0.000
19	0.847	14.431	10.828	71.285	2.601	0.007	0.000
20	0.847	14.431	10.828	71.285	2.601	0.007	0.000

Obviously, gas composition in the air-fired case is very different from the oxyfuel configuration. Indeed, nitrogen is the main constituent of the streams. O_2 concentration is instead similar, whereas H_2O and CO_2 concentration are much lower. More specifically, CO_2 concentration is about 15%, for which only post-combustion capture solutions can be used.
Moreover, SO_2 concentration is definitely lower, because no recycle is performed and the adoption of the FGD is required in the general scenario.

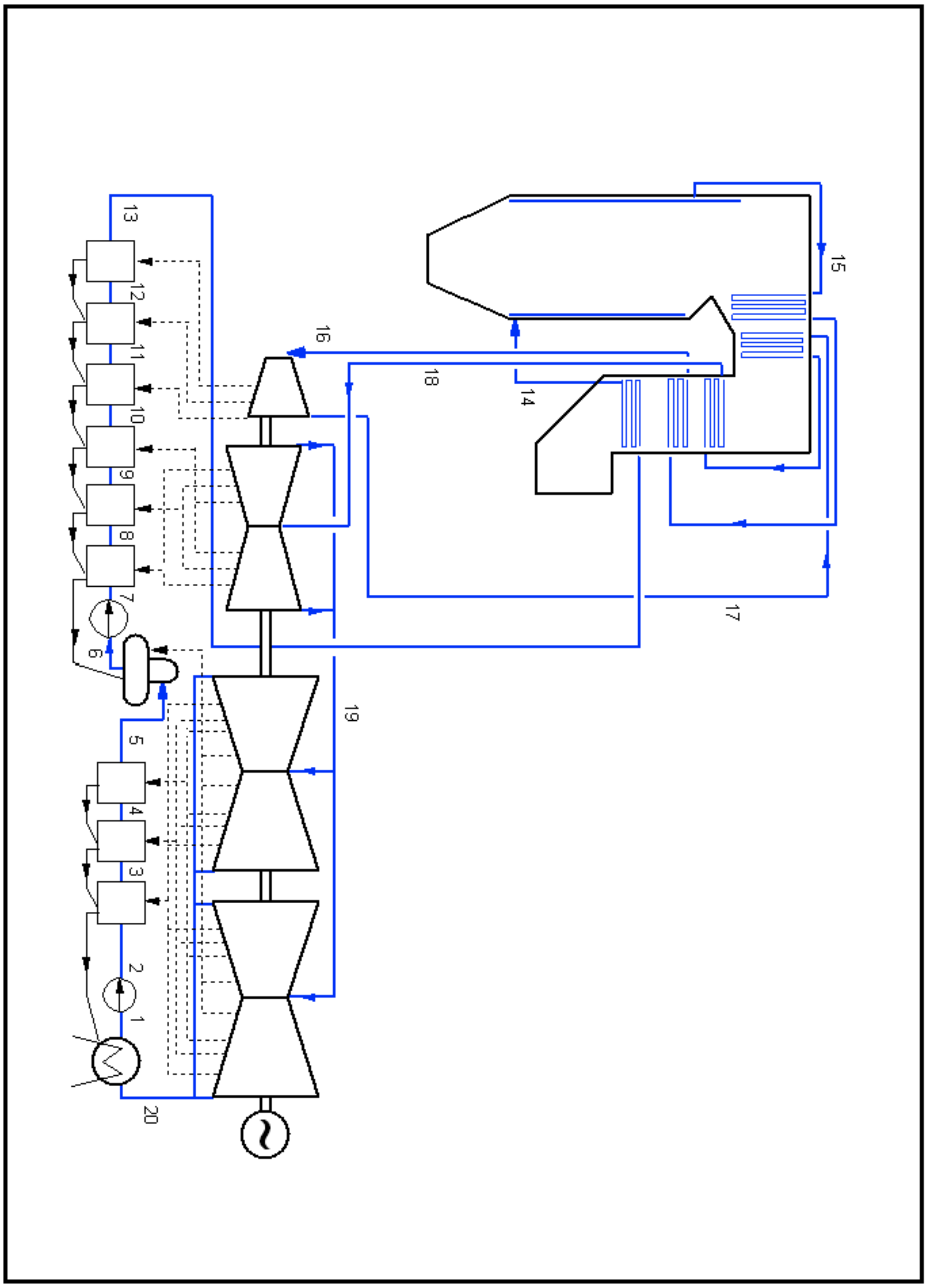

Figure 31. Schematic of the steam-water side of the reference coal-fired USC plant.

Table 24. Main thermodynamic features at several points of the reference air-fired plant (steam-water side).

N	Component preceding point N	water/ steam	G (kg/s)	T (°C)	P (bar)
1	condenser	w	581.52	29.0	0.04
2	LP pump	w	581.52	29.03	9.70
3	preheater 1	w	581.52	64.55	8.70
4	preheater 2	w	581.52	92.34	7.70
5	preheater 3	w	581.52	120.15	6.70
6	deaerator	w	850.20	149.53	4.70
7	preheater 4	w	850.20	154.66	347.00
8	preheater 5	w	850.20	180.68	346.00
9	preheater 6	w	850.20	211.68	345.00
10	preheater 7	w	850.20	244.02	344.00
11	preheater 8	w	850.20	268.07	343.00
12	preheater 9	w	850.20	293.65	342.00
13	preheater 10	w	850.20	314.97	341.00
14	economiser	w	850.20	360.0	337.00
15	evaporator	s	850.20	400.0	317.00
16	SH	s	850.20	598.0	279.00
17	HP turbine	s	750.26	338.7	55.51
18	RH	s	705.17	608.0	50.00
19	IP turbine	s	547.55		6.20
20	LP turbine	x=0.8379	434.89	29.0	0.04
Spillamenti					
			G (kg/s)	Tbl (°C)	P (bar)
1	to preheater 1		32.776	68.72	0.29
2	to preheater 2		26.925	103.90	0.89
3	to preheater 3		28.038	185.81	2.20
4	To deaerator		26.140	274.81	5.05
5	to preheater 5		27.562	366.43	10.46
6	to preheater 6		34.395	454.76	19.42
7	to preheater 7		38.391	543.94	34.23
8	to preheater 8		42.257	338.72	55.51
9	to preheater 9		51.177	388.47	79.48
10	to preheater 10		42.023	430.49	105.26

The only additional required input for the economic analysis is the cost of SCR equipment. So the values proposed by the EPA [54], for which the reported specific capital cost for a 300 MW_e plant is 98 USD/kW, have been selected to estimate the related capital expenses.

The resulting specific total plant cost for the general scenario is 1667 USD/kW, in line with recent literature values, and the expected efficiency obtained from simulations is about 44.7%.
If the assumptions reported previously are accepted, the calculated cost of electricity is 49.5 USD/MWh and the expected CO_2 specific emissions are 785.2 kg/MWh_e.
For the Australian setting reference plant, higher total plant costs (2367 AUD/kW) and efficiency (about 45%) are taken into account, so the COE is 47.7 AUD/MWh and the CO_2 emissions are 767.8 kg/MWh_e.
These calculated values can be used to calculate the cost of CO_2 avoided.

5.5 ECONOMIC ANALYSIS

In this section, the results of the economic analysis performed on the data from the simulations of oxyfuel plants are reported and compared to the values available in literature.

5.5.1 CAPITAL EXPENSES

Capital expenses form the major share of the cost of electricity, even for large scale plants, where the scale economies reduce their role.
The figure below shows that CAPEXes consistenly increase in comparison to the air-fired case, because most of the components increase their size (including key components such as boiler and steam turbine) in proportion to the net output and because new expensive items – the air separation unit and the compression and purification unit – are required.

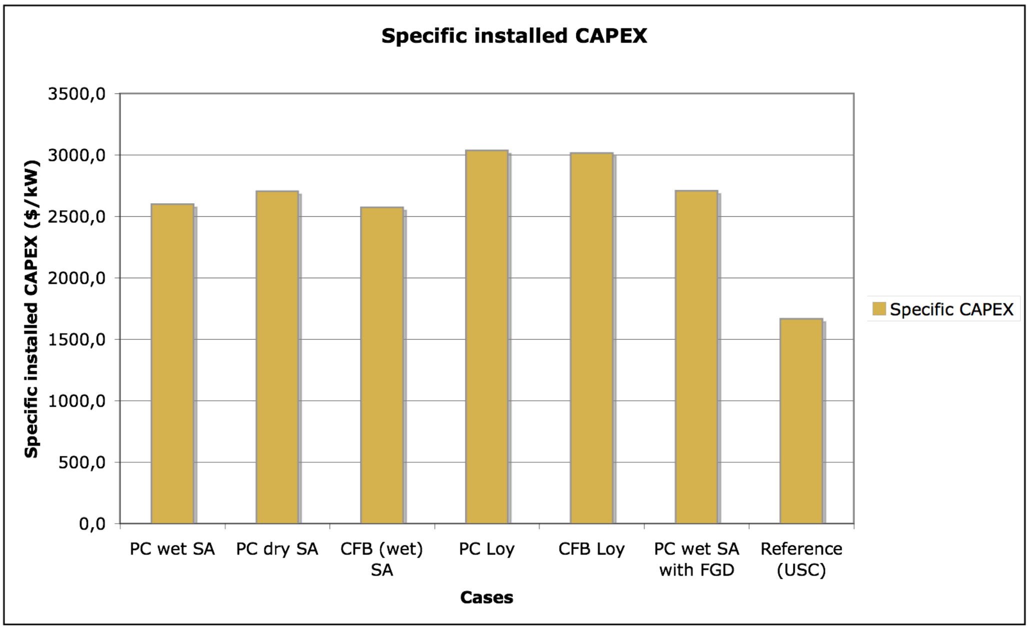

Figure 32. Comparison of the expected installed CAPEX for different configurations at maximum size.

Investment costs are even significantly higher if a high-moisture coal is used and/or if the plant is equipped with pollution control devices. Moreover, the all-dry recycle solution (case 2) appears to be less convenient than the solution with both wet and dry recycles. From the calculation performed in this work, no significant difference between the CFB configuration with in-furnace SO_x capture and the analogous PC one without SO_x removal has been instead observed.

In summary, costs of installation are slighly lower or higher than 2000 $/kW, depending on the specific configuration. Therefore, total plant cost is instead in the 2500-3100 $/kW$_e$ net range.

As shown in the figure below, representing the role of the considered items in the determination of CAPEX for the base case, the most important systems for an oxyfuel plant are in order: ASUs (more than 20% of the total), boiler (almost 20%) and steam turbine (almost 15%).

The importance of the ASU is associated with an increase in uncertainty about the total CAPEX: it is a component that is still under development, especially for oxyfuel applications. The design of boiler and steam turbine are instead based on well-established experience. Improvements in the investment cost for the ASUs can be expected, if the technology improves for the conventional cryogenic solution or if the OTMs to be a viable way to produce O_2-rich streams.

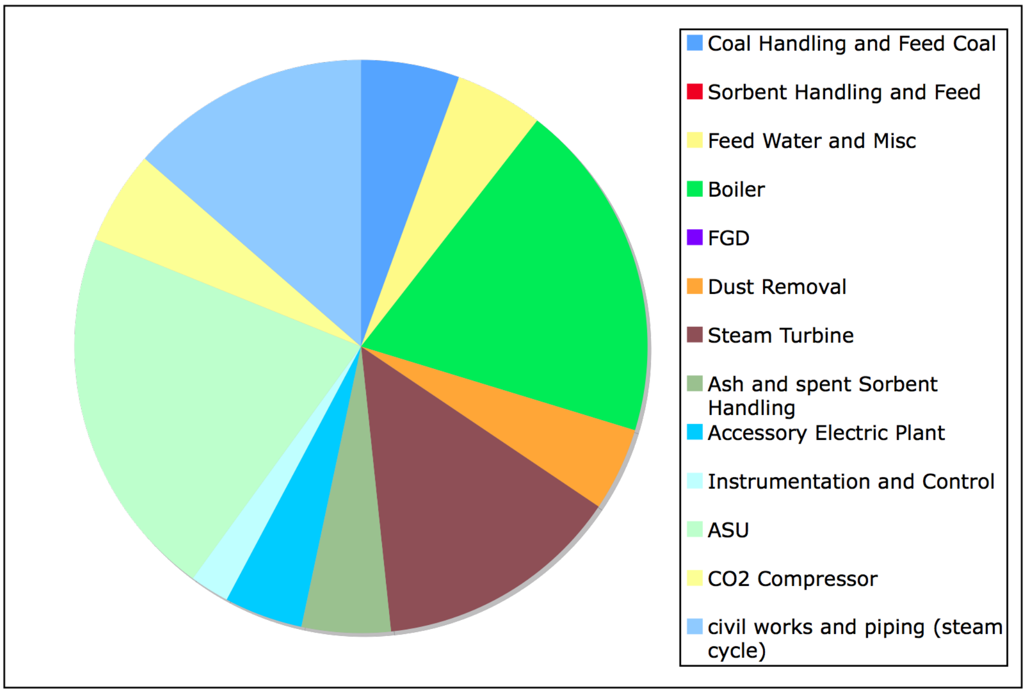

Figure 33. Composition of total CAPEX in the base case.

The breakdown of capital expenses for the base case is approximately the same observed for case 2 and case 6.
Even when case 3 is analysed, minor changes can be detected. Slight decreases in the costs associated to the boiler and to the coal feed and handling system and the addition of costs for sorbent handling (1.3% of the total), partially offset by increases in the expenses for the devices for ash handling, are the most important modifications, which are linked to the different boiler design and to in-furnace capture.
For the simulated plants fed by lignite from the Lao Yang coal mine, sizeable changes can be noticed. Costs for coal preparation system increase dramtically (especially in the PC case). On the other side, expenses for ash handling decrease relevantly, because of the lower ash content in the fuel. The differences between PC and CFB results for lignite-fired plants are exactly the same detected for plants using black coal.
The comparison of the total CAPEX against other studies has been performed for the base case, whose configuration is similar ro the ones considered in other studies (with the exemption of the study by Chalmers University [43], similar to case 4), after adjusting the available values to take into account the year in which the study had been published and the difference in size. The former adjustment is done by using the CEPCI indexes and the plant scaling factor.

The following table compares the adjusted expected total capital costs from this and other studies.

Table 25. Comparison of calculated total CAPEX for base case against literature values.

Estimated total specific costs (in 2009 USD/kW$_e$) for an 750 MW plant	This work	DOE/NETL Case 6	Mitsui Babcock(IEA Greenhouse Gas Research and Development Programme)	Pelliccia
FGD	no	yes	no	yes
Coal Handling and Feed Coal	130.2	100.2		141.2
Sorbent Handling and Feed				29.2
Feed Water and Misc	127.3	131.2		117.3
Boiler	498.4		635.8	417.1
FGD				179.6
Dust Removal	110.2			98.3
Steam Turbine	361.6	223.4	392.1	307.9
Ash and spent Sorbent Handling	119.8	22.2		98.7
Accessory Electric Plant	113.6	150.8		126.9
Instrumentation and Control	59.8	41.0		66.8
ASU	589.1		508.2	585.0
CO_2 Compressor	134.2	174.5	196.8	106.3
Civil works and piping (BOP)	353.4	179.0	524.8	341.3
Boiler (wet recycle) including ASU		1111.8		
Flue gas cleanup		224.3		
Coal and ash handling			152.4	
Total	2597.6	2358.4	2410.1	2615.6

For [16], in 2005, for both PF and CFB hardcoalfiring cases, the specific investment cost has been slightly below 2,000 €/kW$_e$ (assuming 1 € = 1.20 $) for oxycombustion power plants.
A single available study, performed in 2005, provides the estimate of investment

costs for a lignite-fired plant [43] (1890-2015 USD/kW_e net), that is slightly lower than the correspondent one obtained for case 4, after adjustment to present USD/kWnet.
Importantly, estimated capital expenses for a plant of this size are in line with previous works discussing this subject, even though estimates for single cathegories are quite different, mainly because of different accounting strategies and different estimates for both installed costs and start-up costs. Nevertheless, although relevant uncertainties exist over the real value, the range that is considered to be reasonable by the major experts is not too wide.
In summary, the total capital expenses calculated in this work are generally slightly higher than in other studies or similar, mainly because higher set-up incremental costs are used. In fact, the ratio total expenses/installed costs have been conservatively set to 1.33, that is higher than for the other studies, in which it ranges between 1.17 and 1.27.
The differences among the values assigned for each component or system are instead mainly ascribable to a different strategy for the splitting up of costs. Anyway, this does not impact the results, as only the total value has impact on the economic analysis.
These aspects are in case suitable for a deeper investigation. In particular, the values attributed to the main systems that require modifications to be used in an oxyfuel plant (especially ASU and boiler) should be verified.
The uncertainties about the capital expenses necessary for new-built plants make the comparison hard, but it is likely that oxyfuel plants are more expensive than pre-combustion systems and slightly more expensive than (or comparable to) post-combustion plants, if a comparison is made with literature values.

5.5.2 COST OF ELECTRICITY

In this paragraph, the estimated COEs are reported and commented.
The figure below shows a comparison among the expected COEs for maximum size plants with different configurations.

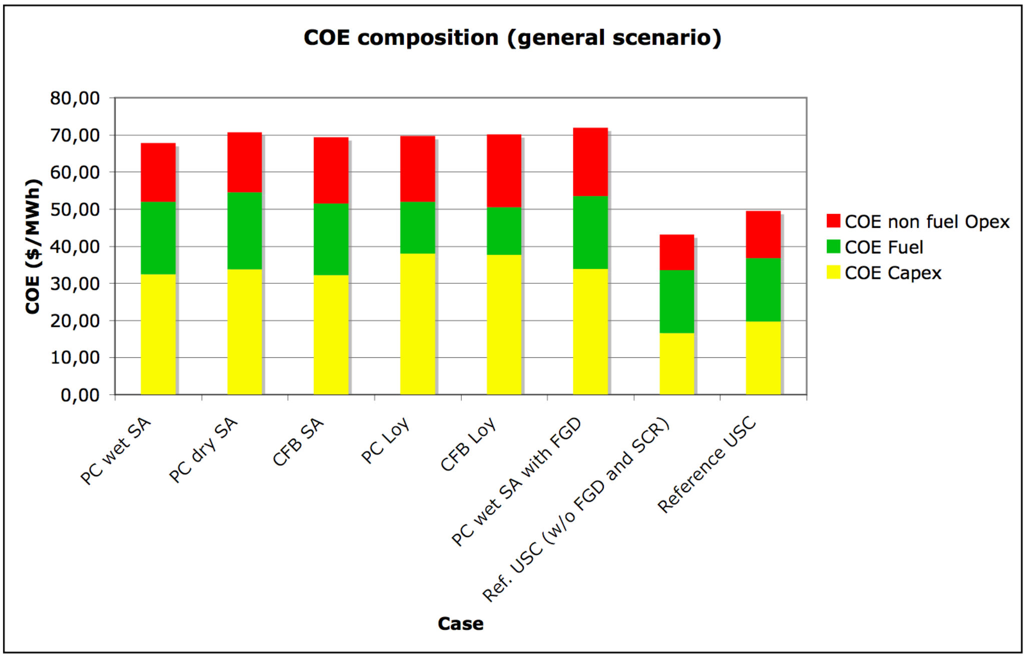

Figure 34. Cost of electricity for maximum size simulated plant (general scenario).

Other than investment costs, it is evident that fuel consumption represents an important share of the cost of electricity, that is very important, even because fuel price is quite volatile. Specific fuel consumption increases significantly with respect to the air-fired case (20-25% higher), because of the decrease in efficiency.

It is more significant if the fuel is black coal, that is much more expensive than brown coal. In fact, the decrease in efficiency just eases that effect.

The fuel total cost is obviously related to the expected thermal efficiency, once the fuel price is defined. Therefore the considerations about the differences among the cases and the comparison against literature would be identical to the one reported previously with respect to thermodynamic results.

O&M costs constitute another important share of the COE and are composed by a fixed and a variable constituent, that can be traced back to the CAPEX and to the efficiency, respectively. So they are a mixed component and both CAPEX and efficiency have impact on them.

It is interesting to notice that the minimum total expected COE among oxyfuel considered configurations is attained by the base plant, whose design is object of the largest number of studies. However, cases 3, 4 and 5 are associated with a similar estimated economic performance. With respect to the lignite-fired plants, the effect of higher investment costs is offset by lower fuel expenses, because low-rank fuel is significantly cheaper than black coal.

The PC case with all-dry recycle is more expensive than the one with wet secondary recycle, suggesting that this choice should be considered only if the problems related to the ESP and to the recirculation fan, that are working conditions, and to the corrosion in the low-temperature heat exchangers and in the furnace are not surmountable.
Moreover, the effect of the addition of a small FGD (that captures SO_2 from the primary recycle stream) is noticeable. If sulphur abatement is necessary, the CFB solutions would deserve particular attention, since in-furnace removal is cheap and effective.

Table 26. Resume of costs of electricity as reported in literature.

	COE with capture (USD/MWh)	COE without capture (USD/MWh)	Δ COE (USD/MWh)	Currency	Cost year
This work (PC, black coal)	67.9	49.5	18.4	USD	2009
MitsuiBabcock (base case)	72.8	49.8	23.0	USD (and EUR=1.2 USD)	2005
Pelliccia (advanced large. wet recycle)	67.4	41.3	26.1	EUR (=1 USD)	2003
DOE (case 6)	95.9	64.3	31.6	USD	2007
Davison 2007	77.6			EUR=1.23 USD	2007
MIT	76.9			USD	2007
IPCC SRCCS 2005	73 (62-86)	46 (43-52)	27.0	USD	2005
Andersson (Chalmers University, fuel: lignite)	61.2	42.1	19.1	USD	2005
This work (PC, brown coal)	70.1	49.5	20.4	USD	2009

The obtained cost estimates are generally lower than the values available in literature, especially if the particular currency rate and the pricing year are taken into account when they are compared to values from old studies.
The reason for this seeming abnormality partially depends on the adopted procedure for the calculation of the cost of electricity. For example, many economic analyses use the capital charge rate method (or others) to calculate the share of COE attributable to CAPEX and include the effect of taxation, that is

neglected in this analysis. Moreover, different assumptions are adopted about cost of fuel, fixed and variable O&M expenses.
These differences make a comparison with those studies not fully sensible. However, the approach in this study provides reasonable estimate of the real cost of electricity and favours a comparison against current and future prices and an assessment of real future potential of CCS technologies.
The results for the Australian case are quite different than the ones for the general case, as shown in the figure below. These changes are due to three major reasons:

- coal for domestic use is definitely cheap. This causes a concurrent fall in the impact of it in the cost of electricity.
- variable O&M costs are also historically lower in Australia.
- on the other hand the capital expenses are usually remarkably higher in Australia, as previously described. Thus the role of the CAPEX in the formation of the COE increases in this scenario.

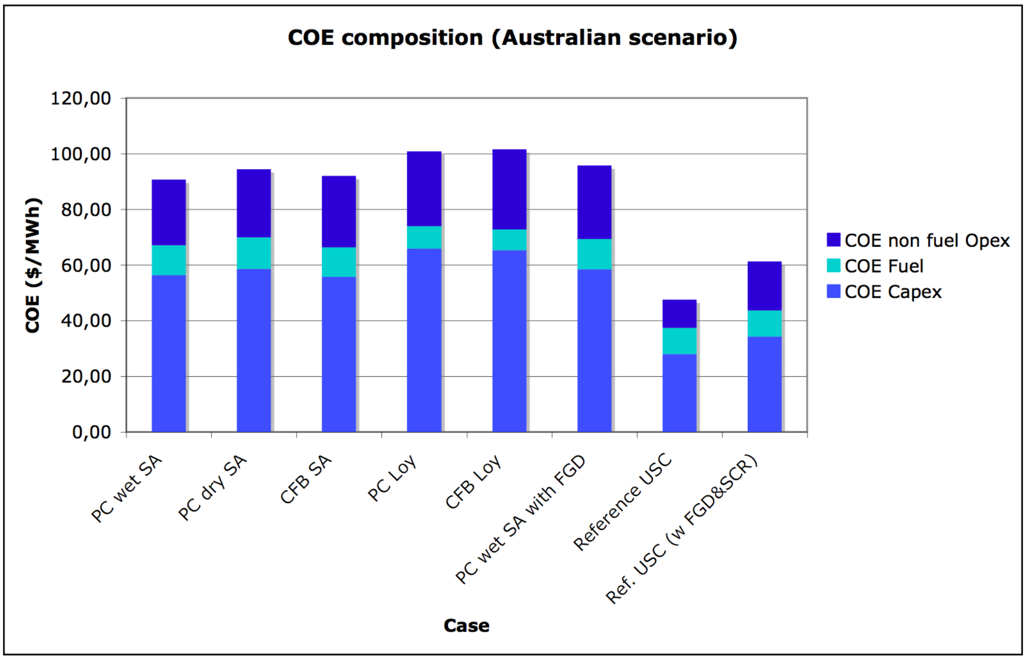

Figure 35. Cost of electricity for simulated maximum size plants (Australian scenario).

The magnitude of the total cost of electricity is eventually similar to the general case, because the increase in capital and fixed O&M expenses is offset by the decrease in variable costs. Actually the costs are slighly higher.

However, the main problem is the increase of the CAPEX, that makes the investment more risky. However, on the other hand, the economic results are less influenced by the volatility of the prices of coal.
Total COE is definitely lower for the black coal-fed plants, as they are characterized by lower specific capital costs, with a minor advantage for the PC configurations.
It is likely that an advanced plant would have a worse economic performance than a more conservative and simple plant in the Australian setting. In particular, the cost of electricity would probably decrease, if the investment costs decrease, in spite of the lower net efficiency.
This last feature is very useful in the context of a development technology. Indeed, for first-of-a-kind plants, it is not realistic to expect high efficiency. At the same time, the control of the capital expenses can be a priority.
For this reason, Australia is a suitable place to build and test pilot oxyfuel plants, beacuse the total project would be less risky and subject to the volatility of the price of fuel than anywhere else and the cost of avoided CO_2 is not prohibitive.
Other reasons that encourage the undertaking of this venture in Australia are:

- the chance to perform enhanced oil recovery (EOR) or enhanced gas recovery (EGR), that would give an economic advantage that could mitigate the project costs.
- the presence of a well-estabilished research structure with competence in capture and storage technologies.
- the strategic interests of several stake-holders (for example coal suppliers and government) in developing technologies able to reduce the emissions of greenhouse gas in coal-fired plant [55]. The effect of an international legislation that aims at a reduction of these emissions would result in a massive damage for the economics of conventional coal-fired plants. In order to be able to mantain the present levels of coal export, reduce the impact on the cost of electricity in the domestic market and to avoid an excessive increase of the cost of electricity on the domestic market, the development of CCS technologies with a low economic impact is fundamental. On top of it, the market of new “clean coal” power plants is potentially very wide and attractive.

On the other side, the adoption of oxyfuel technology in the long term would be more problematic. This is because in Australia, desulphurisation and denitrification are not required and the COE is lower than elsewhere. So the penalty in terms of COE is higher. In other terms, the prospective adoption of this technology would provide an expensive environmental improvement, that is not requested at the moment.
The previous considerations about the all-dry recycle configuration still remain valid: it is economically disadvantaged even in the Australian setting.

Economically, the best configurations are, in order, the PC plant with wet secondary recycle (base case) and the CFB one fed by black coal, whereas the lignite-fired plants do not appear to be convenient.

5.5.3 COST OF CO_2 AVOIDED

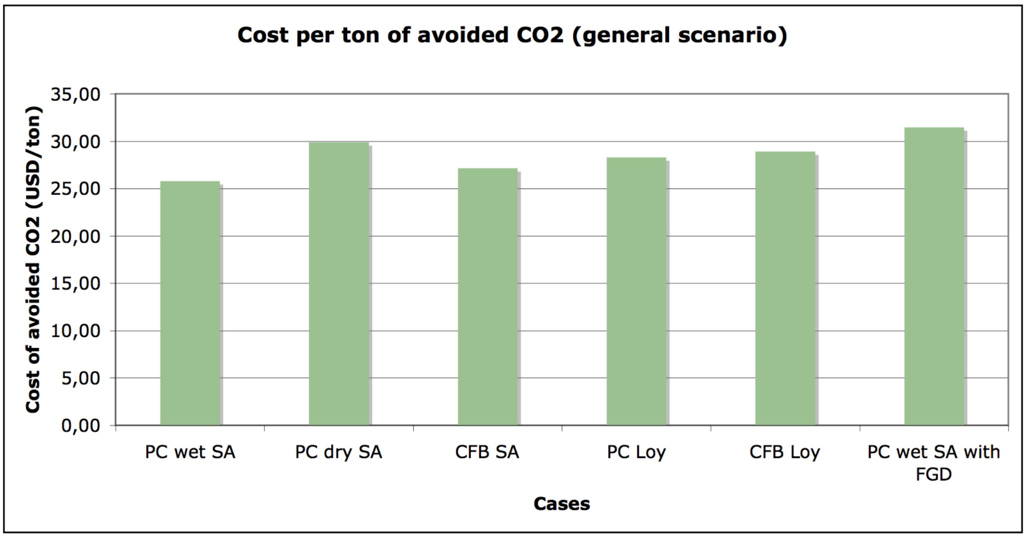

Figure 36. Estimated cost for CO_2 avoided (transport and storage costs not included) for the general scenario.

In the general case, the cost of avoided CO_2 is 26-31 USD/ton (excluding the expenses for transport and storage), that represents an interesting economic performance. In particular, the plants without FGD have a favourable economic performance, with a cost for capture of about 26 USD/ton or slighly higher. Nevertheless, the technical feasibility of some of those solutions still has to be proven.
Again, it should be noticed that the adoption of the FGD treating the small primary recycle stream has a sizeable impact on the economic performance of the plant.

Table 27. Resume of several estimates of the cost of CO_2 avoided in literature, compared to the results from this work.

	Cost of avoided CO_2 (USD/ton)	Currency	Cost year
This work (PC, black coal)	25.0	USD	2009
This work (PC, brown coal)	27.47	USD	2009
MitsuiBabcock (base case)	35.0	USD (and EUR=1.2 USD)	2005
Pelliccia (advanced large, wet recycle)	40.12 (storage included)	EUR (=1 USD)	2003
DOE (case 6)	37.1	USD	2007
Andersson (Chalmers University, fuel: lignite)	22.3	USD	2005
Davison 2007	36.0	EUR=1.23 USD	2007
ENCAP	25 (18-38)	EUR	2004
MIT	40.4	USD	2007
IPCC SRCCS 2005	41 (29-51)	USD	2005

The previous table shows that the results from this work are in the low part of the range of expected costs of avoided emissions in literature. The reasons have already been explained previously, with respect to the cost of electricity.
However, a cost comparison against other studies is not very reasonable, since they differ in many ways: operating and economic assumptions, procedure for cost estimation and plant configuration.

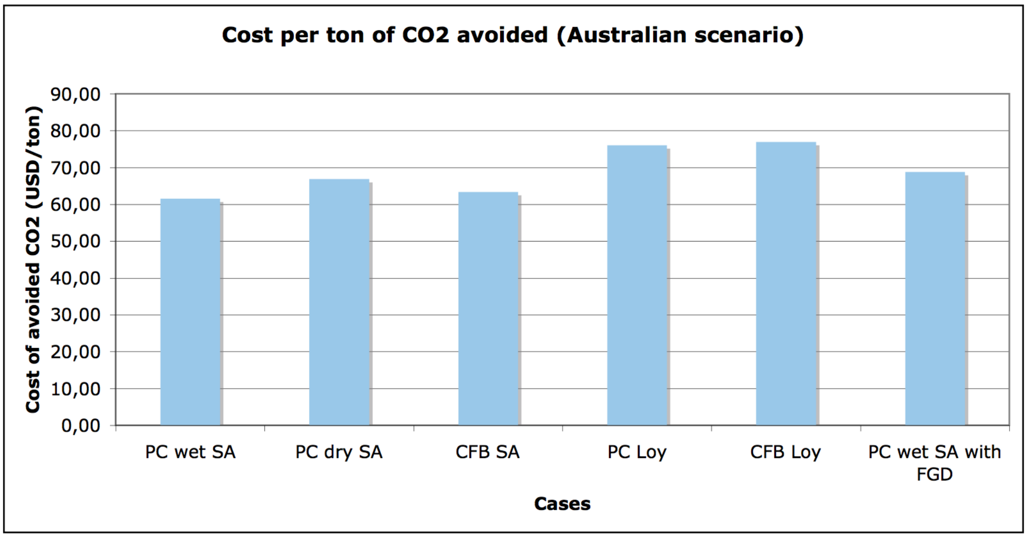

Figure 37. Estimated cost for CO_2 avoided (transport and storage costs not included) for the Australian scenario.

In the Australian setting, the cost of avoided CO_2 is higher, although the COE for the plant with capture is similar, because the reference plant is different. More specifically, the Australian legislation does not require any pollution control device, so the associated COE is lower and the specific CO_2 emissions slightly inferior. Due to this, the impact of the adoption of low-emission technologies in the electric market would be greater and would probably face more resistance.

Several studies compared directly the cost of CO_2 capture by post-combustion and oxyfuel technologies [36, 46, 56]. Other ones just evaluate post-combustion plants with and without CCS [57].
Comparing those results with the ones obtained in this work, it is possible to draw some conclusions.
Currently, carbon capture by means of oxyfuel combustion appears to be competitive with post-combustion technologies and the expected cost for capture are in the same magnitude. Usually, post-combustion plants ensure a lower incremental capital cost, but they are also characterized by a lower efficiency.
This is probably one of the major reasons why the oxyfuel technology has not been tested on a large scale yet, as it implies sizeable investments and, although many suppliers of the components warrant that it is a potentially mature technology, uncertainties still remain on the design and on the operating conditions of the plant, making the investment risky. In connection with this issue, it is worth reminding that the implementation of the oxyfuel technology as a retrofitting is fairly difficult (studies are ongoing anyway [58]), if compared to post-combustion capture, that requires only a downstream treatment of the flue gas, so minor modifications to the original power plant are necessary.

6 SENSITIVITY ANALYSIS

In this work, the effect of a change in the values influencing the economic results has been analysed. In particular, the sensitivity analysis included the following parameters:

- fuel cost
- total CAPEXes
- life time
- real discount rate
- availability

In the following table, the considered variation ranges are reported.

Table 28. Possible ranges of changes for the sensitivity analysis.

	Fuel (USD/ MJ)		Total CAPEX (USD/ kW_e)		Life time (years)		Real discount rate (%)		Availability (%)	
Assumed value	2 (1.2 for lignite)		Depending on the case		25		7		85	
Likeliness of a major change	high		medium		moderate		high		low	
	min	max	min	max	min	max	min	max	min	max
Range	0.5	2.5			15	35	5	20		
Percentual change			-30	105	-40	40	-29	186	-18	12

For the Australian case, the same values (adequately translasted) are generally used. The only exception is fuel cost, for which the range is 0.5-2.5 AUD, lower than for the general case, since the starting current cost is lower.
Some parameters are more volatile than others and are therefore more important for the sensitivity analyses, even because the possible change is wider. Fuel cost and real discount rate are the best examples for that and have a heavy influence on the economic performance of the project.
Another interesting parameter is the total CAPEX, as they have not been exactly determined yet. In particular higher costs are possible (for example if very binding purity requirements are set), but even lower expenses might be observed [59], if currently employed technologies become more effective or new technologies emerge.

Evidently, some parameters are related to others. For example, it is unlikely to see an increase in the fuel price without remarking an associated increase in the price of raw materials (and therefore of CAPEX). On top of it, different design configurations could obviously bring about simultaneous variations in total CAPEX, life time and availability.

It is hard to interprete the results of the sensitivity analysis. Firstly, it should be observed that even the cost of electricity in the reference air-fired case would change if the fuel price and/or the CAPEX change. That effect can be easily estimated. But the effect on other plants (for example non coal-fired plants) cannot be defined, since other parameters influence their COE.

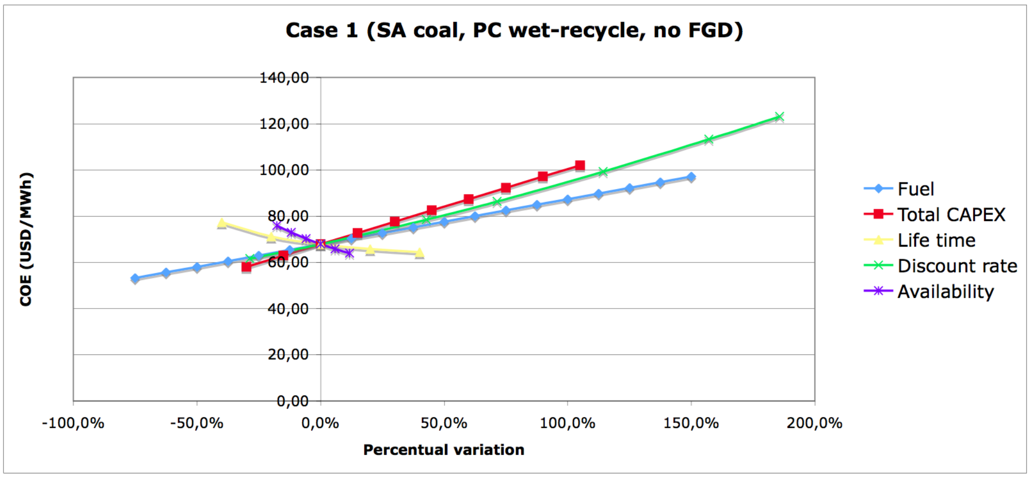

Figure 38. Effect of the changes of some main parameters on the cost of electricity of the base case oxyfuel plant.

In that case, different goals can be set. The most used values are:

- ratio between the costs of electricity in the capture and in the reference plants (associable to the relative increase). This approach is in line with the general DOE guidelines, that set a cost increase lower than 35% as goal. This parameter decreases (and therefore betters) if the fuel cost and/or the CAPEX increase, although the incremental cost raises and the comparison with other technologies becomes less favourable. It could therefore provide misrepresentative impressions on the effect of the changes.
- absolute increase of COE, that supplies a more correct idea of the cost of carbon capture in coal-fired plant, but does not provide indications about the comparison against other technologies.
- The cost of avoided CO_2, for which the reasoning is the same than for the increase in COE.

However, none of the parameters described above is completely satisfactory, so the reported graphs show the effect of the variations on the COE both the plants with capture and the reference plants.

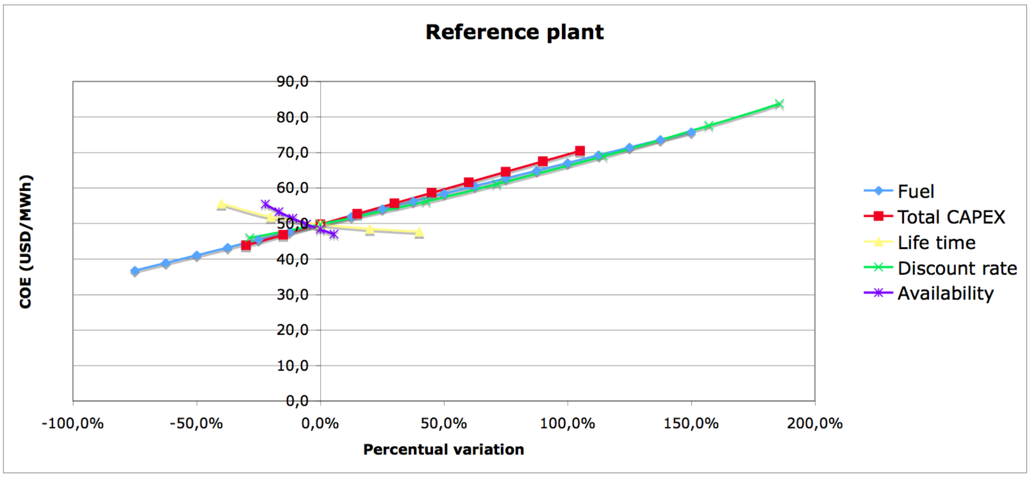

Figure 39. Effect of the changes of some main parameters on the cost of electricity of the reference plant.

Availability and life time appear to have a moderate influence on the definition of the COE of the plant with capture. In fact, the considered range for the availability is quite narrow and the cash flows in the last years of the life time have a minor impact on the cost of electricity.
Naturally, sizeable changes in the cost of coal or in the CAPEX have instead very large influence on the cost of electricity, that changes linearly as those parameters change. The influence of the CAPEX is more evident (especially for plant that are fed by brown coal), as the main share of the COE of oxyfuel plants is ascrivable to the capital costs.
The discount rate has a significant non-linear impact on the COE, that becomes more pronounced as the discount rate increases.
In the Australian setting, the main conclusions formulated for the general case remain valid. Nevertheless, the impact of the changes in the CAPEX and in the discount rate becomes even more conspicuous, since the higher investment costs make the COE more influenced by the capital expenses. At the same time, variations in the cost of electricity associate to changes in the price of the fuel become less influential.
Changes in the COE for reference plant are smaller than for plants with capture in absolute terms, because both the efficiency-related and the CAPEX-related expenses are smaller.

This means that, if the CAPEX or the discount rate (or, at a smaller scale, the cost of the fuel) increase for both conventional and oxyfuel plants, the cost of CO_2 avoided increases.
At the same time, if the cost of coal increases, the ratio between the COE with and without capture decreases. However, this index, which is used as a decisive parameter in the DOE guidelines, becomes misleading as far as heavy variations are taken into account.

7 CONCLUSIONS

Carbon capture by means of oxyfuel combustion represents an interesting option for coal-fired plants in the future. It is indeed a quite reliable technology with an acceptable additional efficiency penalty, in comparison with conventional modern coal-fired plants, although tests on medium and large scale plants are needed to confirm the expected outcomes.

A high purity CO_2 stream can be easily obtained with the addition of a compression and purification unit, which does not affect the CO_2 capture rate too negatively. In particular, the economic performance will depend on the requirements for control of impurities (especially sulphur oxides), that have not been defined yet and have substantial impact on both capital and operating expenses. The position and type of required abatement devices are also important.

Importantly, oxycoal-fired plants are associated with a favourable environmental profile, especially in reference to NO_x and mercury emissions. On top of it, the flue gas would probably be able to satisfy the quality requirements for transport and storage, if minor treatments are performed on the flow.

Six different plant configurations have been analysed and simulated. The cases differ in SO_x control systems, fuel (bituminous coal and lignite), boiler configuration (PC and CFB) and recycle features. The GS code has been used to simulate the majority of the plant with the exception of the CPU, whose performance has been estimated by means of Aspen Plus. A large amount of data has been obtained and evaluated. Several parameters (pressure, temperature, mass flow, composition, etc.) have been estimated at many points, as well as the performance of the components and systems of the plant.

Then an economic analysis has been performed on the outputs of the simulations

In summary, the performed simulations return efficiencies of 30.8-37.4%. The efficiency decreases s compared to a traditional analogous coal-fired plant is therefore 7-8%.

For the general case, the expected total CAPEX for a new-built oxyfuel plant are 2500-3100 USD/kW, up 55-80% from the reference plant with FGD.

For the Australian case, the total capital costs are instead 4500-5200 AUD/kW, that represent a 90-100% increase from the value for reference Australian case without devices for air pollution control.

For the general scenario, the calculated cost of electricity range from 25.8 to 31.5 USD/MWh. The best results are associated with the base case, but even the CFB plant fed by black coal and the lignite-fired plants have a similar outcome.

Analogous conclusions can be drawn with regard to the Australian setting, with the exception of the economic feasibility of lignite-fired plants, which are in fact not interesting.
The addition of a small FGD has a significant impact on the cost of electricity, suggesting that the effect of SO_x on the plant and on the storage reservoir plays a key role in the assessment of oxyfuel solutions. In particular, strict quality requirements could encourage the adoption of CFB boilers with in-furnace SO_2 capture. Another reason to investigate the option of CFB combustion oxyfuel power plants is that it is characterised by fuel flexibility. Moreover, the configuration with single all-dry recycle seems to be a less interesting option, that should be adopted only if technical hurdles make the adoption of a secondary recycle difficult. Interestingly, the adoption of low rank fuel does not impact largely the economic performance in the general setting, whereas that is not true for the Australian scenario.
Uncertainties exist over the thermodynamic and economic results of the components that must be modified to adapt them to the operating conditions in the oxyfuel plant. The ongoing tests will probably shed light on these issues. Moreover, although it is derived from conventional steam cycle based thermal power stations and the majority of the components rely on well-estabilished technologies, the design of ASUs and CPU is not defined yet.
Significant cost reductions could come from improved or new devices for air separation, such as the OTM ASU with heat integration, and for the abatement of impurities (if this is needed). In the longer term, chemical looping combustion could represent a breakthrough in cutting the costs.
Currently, even the more optimistic expected economic performance does not accomplish the goals set for capture technologies in terms of relative COE increase and cost of avoided CO_2. Nevertheless, capture by oxyfuel combustion is a competitive solution, if compared with post- and pre-combustion capture current technologies.
Australia is a suitable location for a scaled-up plant, thanks to the low domestic coal price and to the presence of active research programs and fit storage sites, although the CAPEX would probably be higher than elsewhere. However, its adoption for large-scale power production would result in a sizeable cost penalty.
On the contrary, the impact on the cost of electricity in Europe and North America would be less relevant, as plants are already equipped with emission control devices, and oxyfuel combustion technology could be a viable solution to limit greenhouse gas emissions. Moreover, it would allow to use cheap fuel, that will be largely available for a long time.

This study has neglected some operability issues, that are important even for the economics of the plant. They include flexibility, controllability, start-up/shutdown characteristics. Studies are currently evaluating them and their

impact on the economic performances [60].
On top of it, the options of retrofitting, repowering or refurbishment have been ignored, since major issues exist about it. Generally, it is believed that the switching to the oxyfuel configuration is viable with modifications of several components. Many studies are exploring the technical and economical feasibility of these solutions [58, 61-63], especially the ones about the Callide A project in Australia [64].

ACRONYMS

ASU	Air Separation Unit
AUD	AUstralian Dollar
BATC	Best Available Commercial Technology
BAT	Best Available Technology
BOP	Balance Of Plant
CAPEX	CAPital EXpense
CIF	Cost Including Freight
CCR	Capital Charge Rate (or Carrying Charge Rate)
CCS	Carbon Capture and Storage
CEPCI	Chemical Engineering Plant Cost Index
CFB	Circulating Fluidised Bed
CO2CRC	CO_2 Cooperative Research Centre
COE	Cost Of Electricity
CPU	CO_2 Processing Unit (or Compression and Purification Unit)
DCF	Discounted Cash Flow
DR	Discount Rate
EGR	Enhanced Gas Recovery
ENCAP	ENhanced CAPture of CO_2
EOR	Enhanced Oil Recovery
EPC	Engineering, Procurement and Construction (costs)
EPCO	Engineering, Procurement, Construction and Owner's (costs)
EPRI	Electric Power Research Institute
ESP	ElectroStatic Precipitator
EU	European Union
EUR	EURo
FG	Flue Gas
FGD	Flue Gases Desulphurisation
FOB	Free-On-Board (costs)
GDP	Gross Domestic Product
GHG	GreenHouse Gas
GS	Gas-Steam
HHV	Higher Heating Value
HP	High Pressure
HTD	HydroThermal Dewatering
IEA	International Energy Agency
IPCC	Intergovernmental Panel on Climate Change
ISO	International Organization for Standardization
ITM	Ion Transport Membrane

LHV	Lower Heating Value
LP	Low Pressure
LT	Low Temperature
MEA	Mono-EthanolAmine
MTE	Mechanical Thermal Expression
NASA	National Aeronautics and Space Administration
NG	Natural Gas
NETL	National Energy Technology Laboratory
NO_x	Nitrogen Oxides
O&M	Operation and Maintenance
OECD	Organisation for Economic Co-operation and Development
OTM	Oxygen Transport Membrane
PC	Pulverised Coal
PCI	Potere Calorifico Inferiore
PF	Pulverised Fuel
PR	Progress Ratio
RFG	Recycled Flue Gas
SA	South African
SCR	Selective Catalytic Reduction
SF	Scaling Factor
SO_x	Sulphur Oxides
SD	Steam Drying
TIC	Total Installed Cost
TPC	Total Plant Cost
tpd	tonnes per day
UE	Unione Europea
UNFCCC	United Nations Framework Convention on Climate Change
US	United States
USC	Ultra Super Critical steam cycle
USD	United States Dollar
WTA	WirbelschichtTrocknungsAnlage (English translation: fluidized-bed drying with internal waste-heat utilization)

SYMBOLS

$Avoided\_Cost$	cost of avoided CO_2 emissions, usually expressed in USD/ton of avoided CO_2.
$C_{\max\_power,no\_FGD}$	total capital costs of the largest-scale plant $(C_{ref\_power})_i$ total capital cost of the component i from a source in literature, associated with a certain size $(S_0)_i$.
$C_{\max\_power,w\_FGD}$	total capital costs of the largest-scale plant (with additional FGD) considered.
$(C_{S_0})_i$	total capital cost of the component i for the largest-scale plant considered.
$c_{\max\_power,no\_FGD}$	total variable costs of the largest-scale plant (without additional FGD) considered.
$c_{\max\_power,no\_FGD}$	total variable costs of the largest-scale plant (with additional FGD) considered.
$CO2_{\max\_power,no\_FGD}$	CO_2 emissions of the largest-scale plant (without additional FGD) considered.
$CO2_{\max\_power,w\_FGD}$	CO_2 emissions of the largest-scale plant (with additional FGD) considered.
$COE_{with\_removal}$	total COE for the plant with CO_2 capture (and with additional FGD, if required) with the requested net output.
$COE_{w/o\_removal}$	total COE for a conventional plant without CO_2 capture, expressed in USD/MWh.
$(COE.CAPEX)_i$	share of the COE related to capital expenses for the component i of the plant (with additional FGD, if required) with the requested net output.
$(COE.CAPEX_{Max\_power})_i$	share of the COE related to capital expenses for the the component i of largest-scale plant (with additional FGD, if required) considered.
$COE.fOPEX$	share of the COE related to total fixed O&M expenses for the plant (with additional FGD, if required) with the requested net output.
$COE.fOPEX_{Max\_power}$	share of the COE related to total fixed O&M expenses for the largest-scale plant (with additional FGD, if required) considered.

$COE.fuel$	share of the COE related to total fuel expenses for the plant (with additional FGD, if required) with the requested net output.
$COE.fuel_{Max\_power}$	share of the COE related to total fuel expenses for the largest-scale plant (with additional FGD, if required) considered.
$COE.vOPEX$	share of the COE related to total variable O&M expenses for the plant (with additional FGD, if required) with the requested net output.
$COE.vOPEX_{Max\_power}$	share of the COE related to total variable O&M expenses for the largest-scale plant (with additional FGD, if required) considered.
$Efficiency$	thermal LHV efficiency of the plant (with additional FGD, if required) with the requested net output.
$Efficiency_{Max\_power}$	thermal LHV efficiency of the largest-scale plant (with additional FGD, if required) considered.
$Emissions_{with\_removal}$	specific CO_2 emissions from the plant with CO_2 capture (and with additional FGD, if required) with the requested net output, expressed in tons/MWh.
$Emissions_{w/o\_removal}$	specific CO_2 emissions from a conventional plant without CO_2 capture, expressed in tons/MWh.
f	scaling exponent for multiple trains.
m	minimum number of largest-size trains needed to reach the size $(S_0)_i$ with the addition of a smaller train.
$Max\_power_{no\_FGD}$	net power output of the largest-scale plant (without additional FGD) considered.
$Max\_power_{w\_FGD}$	net power output of the largest-scale plant (with additional FGD) considered.
n	minimum number of largest-size trains needed to reach the size $(S_0)_i$.
$output$	requested net power output of the plant (with additional FGD, if required).
$output_{Max\_power}$	net power output of the largest-scale plant (with additional FGD, if required) considered.
$(S_0)_i$	size of the component i from a source in literature, associated with a certain total capital cost $(C_{ref\_power})_i$.

$(S_{\max\_power})_i$	size of the component i for the largest-scale plant considered.
SF_i	scaling factor for the component i.

(without additional FGD) considered.

$SOx - SOx_{allowed}$	removed sulphur oxide mass flow.

REFERENCES

1. Metz B., a., *IPCC Special Report on Carbon Dioxide Capture and Storage*. 2005.
2. Toftegaard, M.B., et al., *Oxy-fuel combustion of solid fuels.* Progress in Energy and Combustion Science. **In Press, Corrected Proof**.
3. http://www.co2crc.com.au/. 2010; Cooperative Research Centre for Greenhouse Gas Technologies web page].
4. Darde, A., et al., *Air separation and flue gas compression and purification units for oxy-coal combustion systems.* Energy Procedia, 2009. **1**(1): p. 527-534.
5. Scheffknecht, A.K.a.G., *The oxycoal process with cryogenic oxygen supply.* Naturwissenschaften, 2009 **96**(9): p. 9993-1010.
6. Lars Strömberg, a., *Vattenfall's 30 MWth Oxyfuel Pilot Plant Project.* 2007.
7. Strömberg, L., et al., *Update on Vattenfall's 30 MWth oxyfuel pilot plant in Schwarze Pumpe.* Energy Procedia, 2009. **1**(1): p. 581-589.
8. McCauley, K.J., et al., *Commercialization of oxy-coal combustion: Applying results of a large 30MWth pilot project.* Energy Procedia, 2009. **1**(1): p. 439-446.
9. White, V., et al., *Purification of oxyfuel-derived CO2.* Energy Procedia, 2009. **1**(1): p. 399-406.
10. Li, H., J. Yan, and M. Anheden, *Impurity impacts on the purification process in oxy-fuel combustion based CO2 capture and storage system.* Applied Energy, 2009. **86**(2): p. 202-213.
11. D J Dillon , V.W., R J Allam, R A Wall and J Gibbins, *Oxy Combustion Processes for CO2 Capture from Power Plant*. 2005, IEA Greenhouse Gas.
12. Allam, R.J., *Improved oxygen production technologies.* Energy Procedia, 2009. **1**(1): p. 461-470.
13. Davison, J., *Performance and costs of power plants with capture and storage of CO2.* Energy, 2007. **32**(7): p. 1163-1176.
14. Stadler, H., et al., *Oxyfuel coal combustion by efficient integration of oxygen transport membranes.* International Journal of Greenhouse Gas Control. **In Press, Corrected Proof**.
15. Myöhänen, K., et al., *Near Zero CO2 Emissions in Coal Firing with Oxy-Fuel Circulating Fluidized Bed Boiler.* Chemical Engineering & Technology, 2009. **32**(3): p. 355-363.

16. Jäntti T., E., T., Hotta A., Hyppänen T., Nuortimo K., *Circulating Fluidized-Bed Technology - Toward Zero CO2 Emissions*, in *Power Gen Europe 2007*: Madrid, Spain.
17. Koornneef, J., M. Junginger, and A. Faaij, *Development of fluidized bed combustion--An overview of trends, performance and cost.* Progress in Energy and Combustion Science, 2007. **33**(1): p. 19-55.
18. Suraniti, S.L., N.y. Nsakala, and S.L. Darling, *Alstom oxyfuel CFB boilers: A promising option for CO2 capture.* Energy Procedia, 2009. **1**(1): p. 543-548.
19. Hotta, A. *Foster wheeler's solutions for large scale CFB boiler technology: Features and operational performance of Lagisza 460 MWe CFB boiler*. 2009.
20. Yan J., a., *Conceptual Development of Flue Gas Cleaning for CO2 Capture from Coal-fired*
Oxyfuel Combustion Power Plant. 2007.
21. Pipitone G., B.O., *Power generation with CO2 capture: Technology for CO2 purification.* International Journal of Greenhouse Gas Control, 2009. **3**: p. 528-534.
22. Sass, B.M., et al., *Considerations for treating impurities in oxy-combustion flue gas prior to sequestration.* Energy Procedia, 2009. **1**(1): p. 535-542.
23. Stanger, R. and T. Wall, *Sulphur impacts during pulverised coal combustion in oxy-fuel technology for carbon capture and storage.* Progress in Energy and Combustion Science. **In Press, Corrected Proof**.
24. Doctor, R.D. and J.C. Molburg, *High-sulfur Coal Desulfurization for Oxyfuels*, in *7thAnnual Conference on Carbon Capture & Sequestration*. 2008: Pittsburgh, Pennsylvania.
25. Liu, H. and Y. Shao, *Predictions of the impurities in the CO2 stream of an oxy-coal combustion plant.* Applied Energy, 2010. **87**(10): p. 3162-3170.
26. Wheeldon, J., *Review of CO2-Capture Development Activities for Coal-Fired Power*. 2007, EPRI (Electric Power Research Institute): Palo Alto, CA.
27. Croiset, E., K. Thambimuthu, and A. Palmer, *Coal combustion in O-2/CO2 mixtures compared with air.* Canadian Journal of Chemical Engineering, 2000. **78**(2): p. 402-407.
28. DOE/NETL, *Pulverized Coal Oxycombustion Power Plants*. 2007.
29. Wall T., a., *An overview on oxyfuel coal combustion-State of the art research and technology development* CHEMICAL ENGINEERING RESEARCH & DESIGN, 2009. **87 (8A)**: p. 1003-1016.
30. Romano, M., *ADVANCED COAL-FIRED POWER PLANTS WITH CO2 CAPTURE*, in *Dipartimento di Energia*, POLITECNICO DI MILANO.

31. Wheeldon, J. and D. Dillon, *CoalFleet Guideline for Advanced Pulverized Coal Power Plants*. 2007, Electric Power Research Institute (EPRI): Palo Alto, CA.
32. Allardice, D.J. and B.C. Young, *Utilisation of Low Rank Coals.* 2001.
33. Carbo, M.C., et al., *Opportunities for CO2 capture through oxygen conducting membranes at medium-scale oxyfuel coal boilers.* Energy Procedia, 2009. **1**(1): p. 487-494.
34. Ströhle, J., et al., *Simulation of a Chemical Looping Combustion Process for Coal*, in *1st International Oxyfuel Combustion Conference*. 2009: Cottbus.
35. Rubin, E.S., et al., *Use of experience curves to estimate the future cost of power plants with CO2 capture.* International Journal of Greenhouse Gas Control, 2007. **1**(2): p. 188-197.
36. Davison, J. and K. Thambimuthu, *An overview of technologies and costs of carbon dioxide capture in power generation.* Proceedings of the Institution of Mechanical Engineers, Part A: Journal of Power and Energy, 2009. **223**(3): p. 201-212.
37. Pelliccia, G., *Decarbonized Electricity Production from the OxyCombustion of Coal and Heavy Oils*, in *Dipartimento di Energetica*, Politecnico di Milano: Milano.
38. ENERGY, P.D.M.-D.O., *GS: Gas – Steam Cycle Simulation Code*.
39. Schmidt, E., *Properties of Water and Steam in S.I. units*, ed. Springer-Verlag. 1982, Berlin.
40. (editor), G.W.C., *Combustion chemistry*, ed. Springer-Verlag. 1984, New York.
41. Reynolds, W.C., *The Element Potential Method for Chemical Equilibrium Analysis: Implementation in the Interactive Program STANJAN, Version 3";*. 1986: Department of Mechanical Engineering, Stanford University, Stanford, California, USA.
42. Allinson G., N.P., Ho M., Wiley D. and McKee J., *CCS economics methodology and assumptions*. 2006.
43. Andersson, K. and F. Johnsson, *Process evaluation of an 865 MWe lignite fired O2/CO2 power plant.* Energy Conversion and Management, 2006. **47**(18-19): p. 3487-3498.
44. *OECD home page: http://www.oecd.org/home/*. 2010; Interest and exchange rates].
45. OECD, *Projected Costs of Generating Electricity - 2010 Edition*. 2010.
46. Bouillon, P.-A., S. Hennes, and C. Mahieux, *ECO2: Post-combustion or Oxyfuel-A comparison between coal power plants with integrated CO2 capture.* Energy Procedia, 2009. **1**(1): p. 4015-4022.
47. Ekström, C., et al., *Techno-Economic Evaluations and Benchmarking of Pre-combustion CO2 Capture and Oxy-fuel Processes Developed in the European ENCAP Project.* Energy Procedia, 2009. **1**(1): p. 4233-4240.

48. Peters, M.S.a.T., K. D., *Plant design and economics for chemical engineers*, ed. McGraw-Hill. 1980, New York.
49. *Flue gas desulphurisation (FGD) technologies*, D.o.T.a. Industry, Editor. 2000, Department of Trade and Industry (DTI): London.
50. Sharp, G.W., *Update: What's That Scrubber Going to Cost?* Power, 2009. **153**(3): p. 64-66.
51. Srivastava, R.K., *Flue Gas Desulfurization: The State of the Art.* Journal of the Air & Waste Management Association, 2001. **51**: p. 1676-1688.
52. Bendixen, K., *Experiences with Coal fired USC Boilers in Denmark*.
53. Hack, H., et al., *Pathway to Supercritical Flexi-Burn™ CFB Power Plant to Address the Challenge of Climate Change*, in *2009 International Pittsburgh Coal Conference*. 2009: Pittsburgh, PA, USA.
54. Jozewicz, W., *Cost of Selective Catalytic Reduction (SCR) Application for NOx Control on Coal-fired Boilers*. 2002, United States Environmental Protection Agency: Cincinnati, OH.
55. Rai, V., D.G. Victor, and M.C. Thurber, *Carbon capture and storage at scale: Lessons from the growth of analogous energy technologies.* Energy Policy, 2010. **38**(8): p. 4089-4098.
56. Hadjipaschalis, I., G. Kourtis, and A. Poullikkas, *Assessment of oxyfuel power generation technologies.* Renewable and Sustainable Energy Reviews, 2009. **13**(9): p. 2637-2644.
57. Abu-Zahra, M.R.M., et al., *CO2 capture from power plants: Part II. A parametric study of the economical performance based on mono-ethanolamine.* International Journal of Greenhouse Gas Control, 2007. **1**(2): p. 135-142.
58. Xiong, J., et al., *An economic feasibility study of O2/CO2 recycle combustion technology based on existing coal-fired power plants in China.* Fuel, 2009. **88**(6): p. 1135-1142.
59. van den Broek, M., et al., *Effects of technological learning on future cost and performance of power plants with CO2 capture.* Progress in Energy and Combustion Science, 2009. **35**(6): p. 457-480.
60. Alie, C., P.L. Douglas, and J. Davison, *On the operability of power plants with CO2 capture and storage.* Energy Procedia, 2009. **1**(1): p. 1521-1526.
61. Geisbrecht, R. and P. Dipietro, *Evaluating options for US coal fired power plants in the face of uncertainties and greenhouse gas caps: The economics of refurbishing, retrofitting, and repowering.* Energy Procedia, 2009. **1**(1): p. 4347-4354.
62. Singh, D., et al., *Techno-economic study of CO2 capture from an existing coal-fired power plant: MEA scrubbing vs. O2/CO2 recycle combustion.* Energy Conversion and Management, 2003. **44**(19): p. 3073-3091.

63. Simbeck, D.R., *CO2 MITIGATION ECONOMICS FOR EXISTING COAL-FIRED POWER PLANTS*, in *First National Conference on Carbon Sequestration*. 2001: Washington, DC.
64. Palfreyman, D., et al., *TECHNO-ECONOMICS OF OXYGEN-FIRED PF POWER GENERATION WITH CO2 CAPTURE*. 2006, Cooperative research centre for coal in sustainable development (CCSD): Pullenvale, Qld, AUSTRALIA.