Coal quality impacts and gas quality control in oxy-fuel technology for carbon capture and storage – cost impacts and coal value

A report detailing estimates of cost impacts related to coal quality under a research contract to the University of Newcastle, Australia supported by the Xstrata Coal Low Emissions Research

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Table of Contents

COSTS AND UNIT SIZE

Unless otherwise stated, costs, such as LCOE, are presented in 2009 $AUS to allow direct comparison with results from an often-quoted EPRI report. The cost for oxy-fuel units considered in this report is based on units with a (sent out) electricity output of 550 MWe.

ACRONYMS AND ABBREVIATIONS

ACI	Activated carbon injection
AH	Air heater
AL	Air Liquid
APCD	Air pollution control device(s)
ASU	Air separation unit
BAHX	Brazed aluminum heat exchangers
BFW	Boiler feed water
B&W	Babcock & Wilcox
BOP	Balance of plant
CFB	Circulating fluidized bed
CFD	Computational fluid dynamics
COAL	Coal handling system
COP	Callide oxy-fuel project
CPU	CO_2 purification unit
CCS	Carbon capture and storage
DCCPS	Direct contact cooler/polishing scrubber
DOE	Department of Energy in the United States
DRET	Department of Resources, Energy and Tourism
EERC	The Energy & Environmental Research Center
EPA	Environmental Protection Agency
EPRI	Electric Power Research Institute
ESP	Electric static precipitator
FD	Forced draft
FDF	Forced draft fan
FF	Fabric filter

FGC	Flue gas conditioning
FGD	Flue gas desulfurization
GCCSI	Global Carbon Capture and Storage Institute
HgRS	Mercury removal system
HHV	High heating value
ID	Induced draft
IDF	Induced draft fans
IDLH	Immediately Dangerous to Life or Health
IGCC	Integrated gasification combined cycle
ITM	Ion transport membrane
LCOE	Levelised cost of electricity
LHV	Low heating value
NETL	National Energy Technology Laboratory
NGCC	Natural gas combined cycle
OCC1	The 1st Oxy-fuel Combustion Conference
O&M	Operation & maintenance
PC	Pulverized coal
pf	Pulverized fuel
PJFF	Pulse jet fabric filter
PM	Particulate matter
RFG	Recycled flue gas
SC	Super critical
SCR	Selective catalytic reduction
SDA	Spray dryer adsorbent
SNCR	Selective non-catalytic reduction
STG	Steam Turbine Generator
TS&M	Transportation storage & maintenance
UN	United Nation
USC	Ultra super critical
VPSA	Vacuum pressurized swing adsorption
WFGD	Wet flue gas desulfurization

1 ABSTRACT

Costs associated with cleaning of gases from the release of coal sulfur and mercury in oxyfuel combustion for CCS are estimated – these having previously been determined to have greatest coal quality impacts on the technology.

In order to quantify the impact of coal sulphur content on oxyfuel technology for CCS, flowsheets produced by IHI and B&W were used for the basis of plant design, and were compared with those produced by EPRI[1] (Electric Power Research Institute) in 2010 and NETL[2] (National Energy Technology Laboratory) in 2008. S&L models were used to cost the gas cleaning equipment, with the balance of plant costs based on the reports by EPRI and NETL. The results indicate that coal sulphur content will have a significant impact on plant costs. A cost increase of $7/MWH was obtained with an increase in the sulphur content of the feed coal of 3.3 %. Even at low sulphur contents, the cleaning technology selected will have a large impact on costs, with NaOH based scrubbing used at the Callide Oxyfuel Project found to be expensive compared to scrubbers operating with lime, and less expensive sodium options (eg Trona, a naturally occurring sodium carbonate mineral) being recommended. WFGDs, that use limestone, were found to be the cheapest to operate but slightly more expensive to build. Experimental results suggest fabric filters may collect a significant proportion of sulphur prior to scrubbing by other equipment.

Costs of removing mercury gaseous species from oxyfuel combustion to protect the aluminium cold box in the CPU were also considered in this study. Both an atmospheric and pressurized activated carbon packed bed was costed, based on the data presented by Stiegel[3]. The capital costs associated with the removal of the mercury in a pressurized bed is significantly cheaper than an atmospheric bed due to the reduced size. Replacement of the activated carbon is required prior to the bed reaching breakthrough due to the collection by the bed of moisture and other contaminants. Therefore the reduced bed size also reduces variable O&M costs. However, overall the costs associated with the removal of mercury are minimal compared to costs associated with the cleaning of sulphur.

2 EXECUTIVE SUMMARY

Approximately forty percent of the electricity generation in the world today is from coal. It is also forecasted that the use of coal in the developing countries for power generation is expected to increase largely in coming years despite the increasing greenhouse gas emissions and associated global warming concerns. Oxyfuel combustion is one of the most promising technical options for CO_2 capture from coal-fired power generation and thus reduce the greenhouse gas emissions. In an oxy-fuel plant, flue gas impurities concentrations in the flue gas are expected to increase by around three times due to replacement of N_2 by recycled CO_2-rich flue gas. Therefore, impurities removal becomes essential to avoid any corrosion problems in the furnace as well as back end processes. Also impurities removal is required for obtaining the required CO_2 quality for transport and storage. However, the cost associated with flue gas cleaning is imposing large energy penalties and increase the electricity generation cost to a greater extent.

The flue gas cleaning in oxyfuel plants is not only for air emission control. It also differs from the methods used in other advanced technologies such as post- and pre-combustion capture. In the post- and pre-combustion capture adsorbents or membranes are used for separating the CO_2 from the main flue gas stream. The CO_2 capture from oxyfuel combustion employs a flue gas cleaning step to remove undesirable impurities and makes the process more efficient and durable. The major impurities required to be removed in oxyfuel are: SOx, NOx and Hg. The flue gas cleaning for CO_2 capture from oxyfuel combustion could principally rely on conventional flue gas cleaning technologies for SOx, NOx and Hg removal. SOx is expected to be removed by flue gas desulphurization units (FGDs). Although NOx emissions from oxyfuel will be relatively less compare to air-firing, it is still expected to be removed in the compression step. Hg is expected to be removed by activated carbon. Also, remaining Hg can be effectively removed in the compression circuit along with nitrogen as mercury nitrate or mercury nitrite.

In Australia currently coal based thermal power stations do not have FGDs installed. Therefore, costs associated with SOx removal from different types of FGDs have been established in this report. Also mercury removal from flue gas is essential before liquefying CO_2 in a cold box to avoid risk associated with the mercury attack on aluminum heat exchanger plates. It is still unclear if Callide should install an atmospheric carbon bed within the gas cleaning circuit or a pressurized carbon bed in the back end after the CPU. The costs for the carbon bed for atmospheric and pressure conditions have therefore also been estimated.

A comprehensive literature review was carried out in Section 3. The costs associated with sulphur and mercury removal in the oxy-fuel process were determined. B&W have published several flowsheets, each suitable for coals with different sulphur contents. However, no cost estimates on the impact of sulphur removal for these different flowsheets are available in the existing literature. To date, only two detailed reports have been published on the cost of an oxy-fuel thermal power plant. These were published by EPRI (Electric Power Research Institute) in 2010 and NETL (National Energy Technology Laboratory) in 2008. Their summary and comparisons have been made in Section 4. Section 5 describes the different flowsheet options suggested by IHI and B&W with varying coal quality and flue gas cleaning schemes. Section 6 of the report defines the methodology obtained for quantifying the coal quality impact and gas quality control in oxyfuel plants. Detailed process simulations were carried out to obtain gas flow rates and composition at various locations in oxyfuel plants using ASPEN Plus v 7.3 and described in Section 7. As described earlier, SO_x concentrations during oxyfuel combustion are generally higher compared to conventional air firing. The higher SO_x concentrations, particularly SO_3 in combination with high concentrations of water in the recycled stream, increase the acid dew point temperature of the system, thereby increasing allowable flue gas temperatures and reducing the thermal efficiency of a power plant. However, SOx can be captured by fly ash in fabric filters which can reduce the acid dew point temperature of flue gas and thus improve the efficiency. To investigate this in detail, experiments were carried out on a 20 kW furnace at IFK Institute of Combustion and Power Plant Technology, Universitat Stuttgart, Germany. The efficiency improvement by SOx capture in fabric filter has been estimated in Section 8. Costing codes developed using S&L models have been detailed in Section 9. Results are discussed in Section 10 where total plant costs including capital costs ($/kW), fixed operating and maintenance costs ($/kWh/yr) and variable operating and maintenance costs ($/MWh) have been calculated and compared with the published NETL and EPRI reports. The levelised cost of electricity (LCOE) in $/MWh was also calculated and compared for all cases. Also, costs associated with sulphur and mercury removal was obtained for different published flowsheets with varying coal sulphur. Finally, conclusions derived from the study are given in Section 11.

3 INTRODUCTION

In 2010, approximately 30 Gigatonnes of CO_2 was emitted by man worldwide. The energy sector (electricity and heat generation) contributed 41% of these emissions; the largest emitter of any sector. Of these approximately 71% came from combustion of coal and peat[4]. Carbon dioxide is known to be one of the key drivers of global warming and the sheer scale of emissions makes this an extremely difficult problem to resolve. Alternate methods for producing power that are not dependent on coal are developing but are expensive and unlikely to be cost effective in the near future. In the meantime, a technique that uses coal but avoids the release of CO_2 will be required.

Three different technologies are currently being demonstrated for capturing the CO_2 from the combustion of coal to produce electricity:

- Production of syngas in an oxygen fired gasifer. The syngas is a mixture of H_2, CO and CO_2 which produces a moist CO_2 product when combusted,
- Amine scrubbing of the combustion gases from an air fired p.f. furnace to produce a relatively clean flue gas and a separate CO_2 stream, and
- Removal of the nitrogen from the air before combustion in the p.f. furnace occurs to produce a flue gas stream rich in CO_2.

The product liquid from each technology can be sequestered underground. This last option is termed “oxy-fuel combustion” and is suitable for new and retrofitted plant designs, unlike the first option and appears more cost effective, at this stage, compared to the second option[5]. This technology also offers other potential benefits, such as lower overall NOx emissions, and lower flue gas volumes requiring smaller cleaning equipment.

To limit the combustion temperature, the flue gas is recycled through the combustion process which, if there is no sulfur removed from the recycle, results in a concentration of sulphur species in the flue gas that is 2.5 to 4 times higher than the concentration in air combustion[6, 7], which would result in higher concentrations of SO_3, an associated elevation (together with higher water vapor) of the acid dew point, and therefore necessitating higher flue gas temperature without removal in operation and efficiency loss. Desulfurization of the flue gas is required to avoid sulphur associated corrosion when combusting all but the lowest sulphur content coals, and depending on the flue gas sulphur concentration, different scrubbing equipment is required to remove the sulphur. Similarly, the concentration of mercury and other volatile trace elements will increase in the recycled flue gas[8]. Mercury attacks the aluminum cold box and therefore must be removed prior to this stage, which is usually undertaken with an activated carbon bed.

This is the third milestone report for the project “Coal quality impacts and gas quality control in oxy-fuel technology for CCS” supported by Xstrata Coal Low Emissions Research and Development Corporation Pty Limited . The first milestone report[9] provided published flowsheets for oxy-fuel technology for the power plant and CO_2 processing units (CPUs), which determine the capital and operating costs as well as influence the efficiency penalty of the technology. The unit operations in the CPU are primarily determined by the coal sulfur and mercury levels. Different flowsheets produced by gas vendors - Air Liquide, Air Products, LINDE and Praxair –detail integrated gas cleaning and compression schemes, removing SOx, NOx and Hg (all as liquids) as well as Ar, O_2 and N_2 gases during compression. The mechanisms by which the impurities in coal determine gas quality were detailed, for coal sulfur, for coal mercury and for NOx formed from the coal nitrogen.

The second milestone report[10] detailed experimental work carried out at a 20 kW electrically heated once-through furnace at the Institute of Combustion and Power Plant Technology (IFK) of the University of Stuttgart, Germany. In this pulverized coal combustion rig, studies in air and simulated oxy-fuel environments were performed with the same three Australian coals which were previously tested in the Aioi furnace of IHI under a CCSD project[8], allowing a comparison of results. Measurements of sulphur, nitrogen and mercury gas products from combustion in air and oxy-fuel environments were made, with conditions representing different extents of gas cleaning of the recycled flue gas identified in the first milestone report[9].

The flowsheets[9] and gas quality[10] from the two reports are used as inputs to determine costs in the current report.

3.1 OXY-COMBUSTION

Oxy-fuel combustion is the process of burning a fuel using pure oxygen by eliminating nitrogen from air. Historically, the primary use of oxy-fuel combustion has been in welding and cutting of metals, especially steel[11]. There is a growing interest of utilizing this concept for electricity generation with an aim of generating CO_2 enriched flue gas followed by its cleaning, compression and sequestration.

In an oxy-fuel process, oxygen at greater than 95% purity and recycled flue gas are used for fuel combustion, producing a gas that is enriched in CO_2 and water[12]. Recycled flue gas is mainly used to control the flame temperature. It also carries fuel into the boiler and ensures adequate convective heat transfer to all boiler areas. The process schematic is shown in Figure 1.

The oxygen is separated from air using an Air Separation Unit (ASU) and mixed with the recycle stream from the boiler. The flue gas generated this way is enriched in CO_2 which then goes to purification and compression stages. CO_2 and H_2O also have higher thermal capacities than nitrogen, which leads to a higher heat transfer in the convective section of the boiler. The amount of gas passing through the boiler is lower, and heat transfer is increased in the radiative section of the boiler, resulting in a lower heat transfer in the convective section of the boiler and a lower gas temperature at the furnace exit.

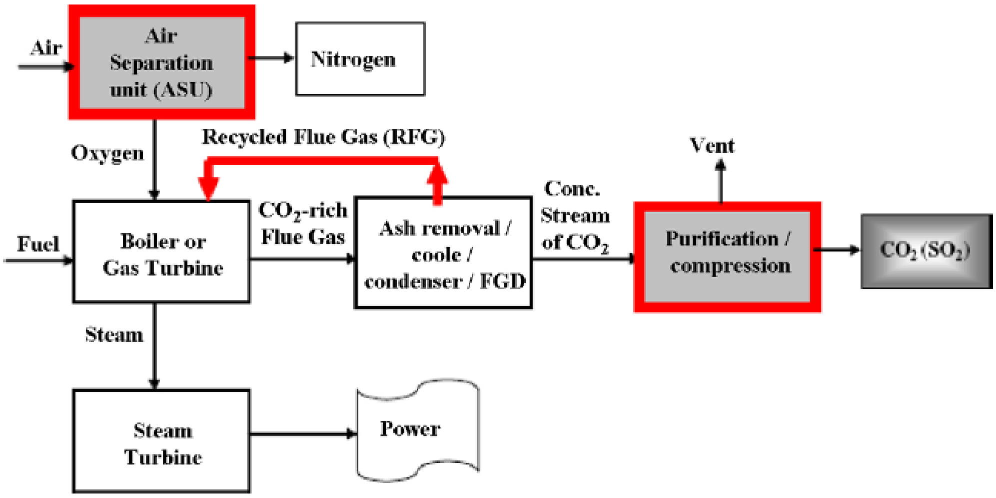


Figure 1: Schematic of the oxy-fuel process[13]

Disadvantages of oxy-fuel

1. Oxygen production is required which is a highly energy intensive process.
2. The technology is not yet operated at large scale. However, 30 MWe scale demonstrations are ongoing and a 150 MWe scale demonstration project (i.e. FutureGen 2.0) has been announced.
3. As the concentration of impurities in the flue gas increase by around three to five times in oxy-fuel, cleaning steps which adds to the auxiliary costs may be needed.
4. Australia does not have SO_x, NO_x and mercury emissions regulations in place. Therefore, currently no measures are used to limit SO_x, NO_x and mercury emissions at existing power plants other than low-NOx burners. For example, no FGDs are installed and so no operating experience with FGDs exists within Australia.
5. The overall process efficiency is reduced by the addition of oxygen production and flue gas cleaning steps. This results in a higher electricity generation price than the conventional air fired systems.

Advantages of oxy-fuel

1. Oxy-fuel provides a zero-emission or reduced emission platform.
2. The technology is already known and has been employed in several industries other than power generation

3. Oxy-fuel combustion reduces the NO_X formation/emission in the furnace.

4. The heat transfer characteristics of the boiler are improved.

5. The oxy-fuel process can be retrofitted to existing coal fired thermal power plants.

6. Designs for new plants will work at higher temperatures which will reduce the footprint size and improve the efficiency of power generation.

7. Concentrations of impurities in the flue gas increase by around three times but the efficiency for impurity removal is expected to increase such that overall lower emissions occur. Comparison of flue gas cleaning in oxy-fuel and conventional air fired systems using ASPEN Plus process simulations shows that recycling enhances the flue gas cleaning performance of fuel borne impurities such as SO_2, SO_3, Hg and dust (fly ash), as described in Figure 2.

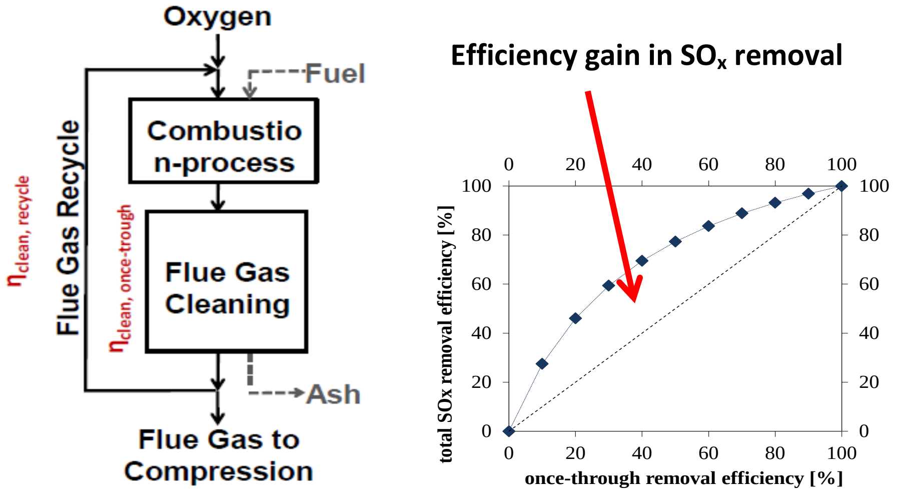


Figure 2: The improved efficiency of SOx removal predicted by an ASPEN Plus process simulations of a once through boiler and flue gas cleaning with flue gas cleaning and the oxy-fuel process with flue gas recycling.

3.2 RESEARCH DEVELOPMENT STATUS

3.2.1 OXYFUEL COMBUSTION

Full oxy-fuel combustion and oxygen enhanced combustion techniques have been used in a number of industrial applications such as glass and cement manufacture and in blast furnaces for iron making. The benefit of oxy-fuel in these cases includes efficiency improvement and NOx reduction. However, the application of oxy-fuel for power generation is relatively new. As shown in Figure 3, until recently the development of the process was undertaken at pilot scale. Early pilot scale studies considered the oxygen requirements to complete a retrofit of an operating air fired furnace. Burner performance and variation in emissions during oxy-fuel compared to air combustion were also investigated. These studies are discussed in the review by Buhre[13].

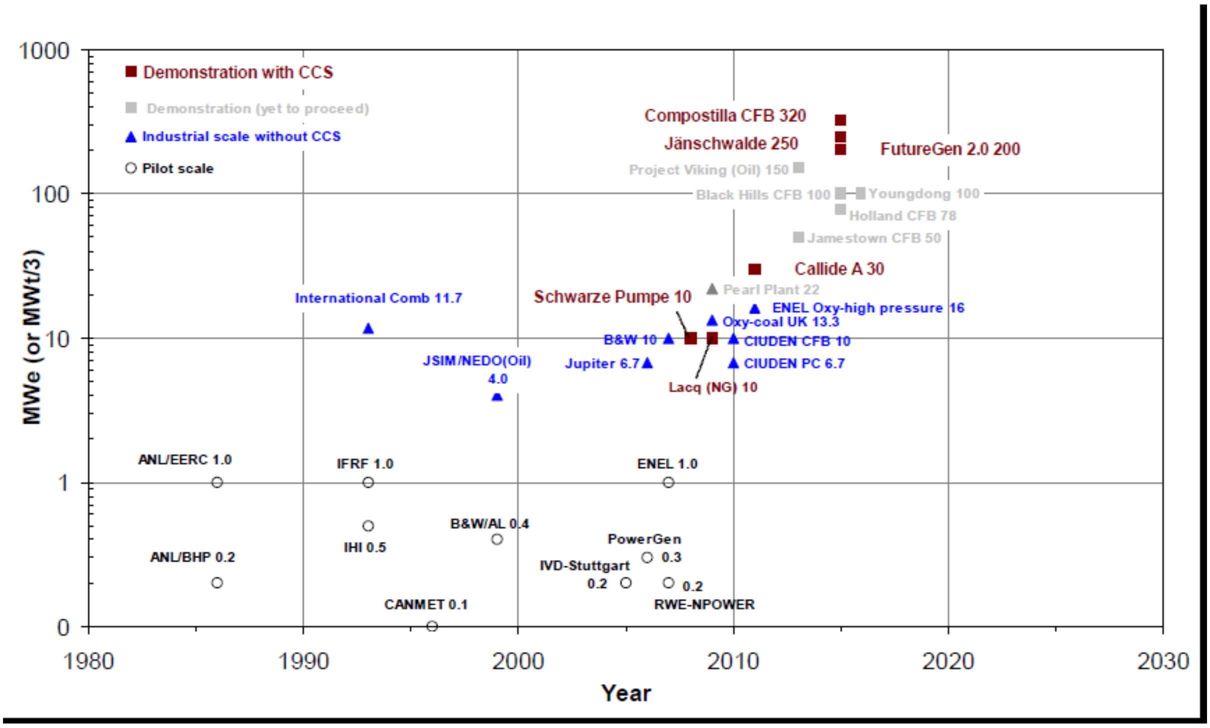


Figure 3: Development of the scale of the oxy-fuel combustion process. (Plants that produce electricity are included in the figure according to their electrical output: MWe. For plants that do not produce electricity, the total plant size MWt is divided by 3).

Further investigation of the oxy-fuel process has been completed more recently on Industrial scale plants, most without sequestration of the cleaned and compressed CO_2 gas. Vattenfall 30 MWt Scwarze Pumpe plant was commissioned in 2008[14]. The plant combusted a black coal or a pre-dried brown coal, with studies focused on combustion issues and gas treatment of SOx, NOx, H_2O and ash. Political difficulties associated with gaining approval for sequestering CO_2 have caused a halt to the planned development of the full scale oxy-fuel plant at Janschwalde.

The CIUDEN test facility comprises two combustion options: a 20 MWt p.f. boiler and a 30 MWt CFB[15]. The plant also has gas treatment and compression and a 3 km CO_2 transportation rig. It aims to validate each phase of the process, from fuel preparation to CO_2 purification, producing a CO_2 stream ready for transport and storage, then produce data for scaling-up the PC, CFB, FGD and compression and purification. A full scale 300 MWe demonstration oxy-fuel CFB plant, Compostilla, has been cancelled [16].

The Callide Oxy-fuel demonstration project, at Callide in Australia[17], is a retrofit of a p.f. boiler, with electricity generation during oxy-fuel firing. Oxygen is produced by 2 x 330 tonne air separation unit at 98% purity and 40% recirculation of the flue gas is completed. Compression is completed but to date no sequestration has been completed, though sequestration tests in the Otway basin are planned.

Sequestration has been demonstrated from an Oxy-fuel plant by Total LACQ CCS plant[18] in France. It comprises a retrofitted 30 MW_t air fired boiler, built in 1957, with gas compression and has been sequestering CO_2 produced since 2010. The CO_2 is sequestered at Rousse, 27 km from the plant in a depleted natural gas reservoir. Project aims are to validate the process, from combustion to sequestration, and collect data required to build a 200 MW thermal oxy-fuel plant.

FutureGen 2.0 is expected to be the first worlds commercial scale near zero emission power plant[19]. This oxy-fuel plant will be a retrofit of the Meredosia power plant in Illinois, USA. The plant design is discussed later in Section 5. CO_2 transportation via a pipeline to a sequester site in north eastern Morgan County, Illlinois 4000 feet below the surface into sandstone is planned. With a total cost of $US 1.65 billion, construction is expected to start in 2014.

Due to the efficiency penalty of 7-10%[20] imposed on the power generation process by the air separation unit and the CO_2 compression, wide spread full scale use of this technology is unlikely until a significant cost on carbon is legislated. Overcoming the public's concern about risks associated with sequestration will also be a major hurdle.

3.2.2 SOX IN OXY-FUEL

Sulphur concentration in the boiler of the oxy-fuel plant is significantly higher than the concentration during air combustion if it is not removed from the recycle. Measurements completed IHI tests furnace at Aioi[21], shown in Figure 4, show that the SO_2 concentration can be three times higher in oxy-fuel. Tan[22] found the SO_2 concentration could be up to 4 times higher than air combustion, depending on the position of the measurement and the fuel. However, the total amount released per kg of coal combusted, or MJ of energy released, is lower. This is because the total volume of the gas in the boiler is reduced with the exclusion of the N_2 from the oxidant (air) which dilutes the concentration of other gases in the flue gas during air combustion.

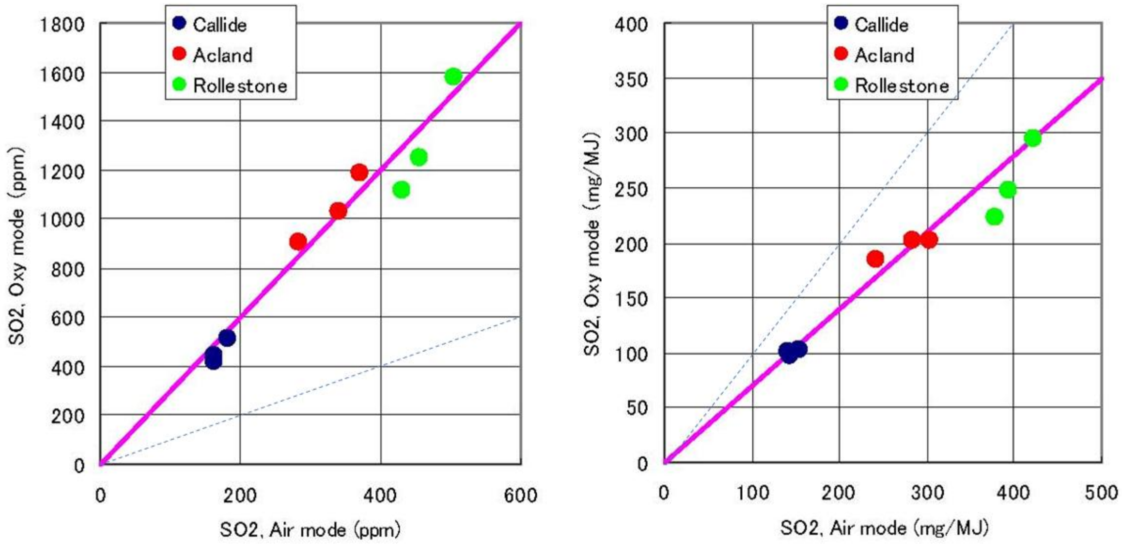


Figure 4: The increased concentration of SO_2 in the flue gas and the decreased amount of SO_2 produced during oxy-fuel combustion, as measured by Suko et al.[21]

In oxy-fuel combustion, the oxygen concentration in the oxidizer stream is generally much higher than in air–fuel combustion, due to the recycle of combustion products to provide similar flame temperatures or heat flux profiles as compared to conventional air-fired conditions[23]. The increased amount of oxygen and SO_2 in oxy-fuel combustion has the potential of affecting the degree of oxidation of SO_2 to SO_3. There are few experimental studies reported in the literature that concludes that under oxy-fuel firing, the SO_3 concentration is higher than air-fired conditions[23, 24]. However, in few studies it has been observed that SO_2 mass emissions (mg/MJ) are lower in recycled oxy-fuel due to SO_3 formation and subsequent sulfur retention. Croiset and Thambimuthu[25] found the conversion of SO_2 to SO_3 decreased from 91% in air combustion to 64% in oxyfuel combustion and Kiga[26] tried to close the mass balance around an oxy-fuel test rig, finding that SOx emissions during oxyfuel combustion were reduced compared to air combustion with increased capture of SOx by ash.

Even though the conversion to SO_3 decreases and the collection by ash increases, the enormous increase in concentration of SO_2 in oxyfuel combustion means that only coals with the very lowest sulphur contents can be used in oxy-fuel without removing SOx from the gas stream. Increasing the SO_3 concentration in the flue gas increases the acid dew point. SO_3 and moisture in the gas form H_2SO_4 which condenses when the temperature falls below the acid dew point. Increasing the acid dew point means either increased corrosion will occur or the flue gas temperature must be increased to avoid corrosion, reducing the efficiency of the plant. Sulphur also poisons the activated carbon used to collect mercury and so must be removed from the gas stream before the flue gas is allowed to pass through the activated carbon bed. Different equipment is available for the removal of sulphur from the flue gas. These are discussed in Section 3.4.

3.2.3 MERCURY IN OXY-FUEL

For oxy-fuel combustion, only limited information on the Hg emission behaviour is available. The Hg speciation and capture in oxy-fuel systems can be impacted by a considerable increase of the H_2O, HCl, SO_x, and Hg concentrations in the flue gas, because of the exclusion of diluting N_2 from the combustion and a change in the flue gas NOx concentrations[27]. Accordingly, oxy-fuel combustion in pilot scale showed a doubling or tripling of Hg concentrations in the flue gas compared to air firing[28]. Changes in the HCl, SOx, and Hg concentrations can potentially improve Hg oxidation but may also decrease Hg capture and depend upon the kind and extent of recycle gas cleaning in an oxy-fuel system. A system with full cleaning of the recycled flue gas should exhibit only minor differences to an air system, while in an uncleaned oxy-fuel recycle system, considerable differences in the Hg oxidation and capture are possible[27]. Changes in the ash loadings in oxy-fuel combustion with increased oxidant O_2 and changes in the temperature profile in the oxy-fuel combustion process because of changed thermal properties of the oxy-fuel flue gas can also affect Hg oxidation and capture. Moreover, the increased residence time of flue gases that are partially recycled and therefore pass flue gas cleaning equipment several times can improve the Hg emission behaviour[29].

3.3 SCIENTIFIC AND TECHNICAL CHALLENGES AND FLUE GAS CLEANING IN OXY-FUEL

Similar to the other advanced low emission clean coal technologies, oxy-fuel also faces several challenges. One of these challenges is to lower the energy penalty associated with oxygen production, which ranges from 7-10%[6]. There are several emerging technologies such as ion transport membrane (ITM), chemical looping air separation and TDA process utilizing ceramic oxygen carriers which may be suitable for low-cost and large-scale oxygen production. Success would result in a key enabling technology that significantly reduces the energy penalty involved in producing oxygen.

Another important challenge is that current design configurations and materials are unable to operate the process at the high temperature ranges required for oxy-fuel combustion. Despite the fact that oxy-fuel will assist in reducing the plant size and will make emissions capture easier, it has been suggested that its full potential is unlikely to be realized until new high-temperature materials become available for combustors and boilers, such that less flue gas needs to be recycled to the boiler. In this regard, a conceptual study was undertaken by Praxair to evaluate the option of a new build boiler which employed more exotic materials of construction to eliminate the need to moderate combustion temperatures, anticipating that cost savings may accrue from an overall reduction in equipment size and utility consumption[30]. However in the preliminary calculations it was showed that the potential savings were more than offset by the increased capital cost.

A final issue is the need to reduce the total energy consumption for CO_2 purification, separation and compression. In this regard, it is essential to understand the impacts of coal quality and gas quality control on the cost of CO_2 purification, separation and compression which forms the basis of the current report. The current work aims to provide cost estimates for a retrofitted oxy-fuel power plant. The issues associated with retrofitting a current coal-fired power station to an oxy-fuel configuration are mainly oxygen production, flue gas cleaning and CO_2 compression. It has been established that under the oxy-firing mode, the flue gas impurities, such as SO_x and Hg, are expected to increase by around three times[6]. SO_x removal becomes essential to avoid any high temperature corrosion problems in the furnace as well as in the back end processes. The oxyfuel process is expected to reduce NOx emissions due to N_2 replacement with recycled flue gas. However, remaining thermal NO_x is expected to be removed during compression. Mercury is expected to be removed mainly in the PM removal system and compression circuit. However, if SO_x is present in the flue gas during compression, it is expected that NO_x and Hg removal may be greatly affected. Therefore, SO_x removal is a critical step in the oxy-fuel process. Moreover, in Australia there are currently no regulations for SO_x emissions due to the abundant availability of low sulphur coal, and therefore coal fired thermal power stations do not have flue gas desulphurization (FGD) units installed. So, understanding the cost for SO_x removal in oxy-fuel combustion with different FGD types is critical to understand the coal quality impacts on the performance of an oxyfuel plant and forms the basis of the current study. Also, literature[31] notes that dedicated mercury removal unit operations such as carbon beds have been tested and further installed in the conventional power plants, which can also be used for the oxy-fuel configuration, as mercury is expected to attack the alumina heat exchanger in the CO_2 liquefaction unit (cold box). The Callide oxy-fuel demonstration project considered the use of a carbon bed to remove mercury and so costs estimates for the removal of mercury have also been included in this study.

3.4 SOX REMOVAL TECHNOLOGICAL OPTIONS

A proportion of the sulphur is expected to be removed in-situ in the electrostatic precipitator (ESP) or fabric filter unit operations with fly ash. However, a large proportion will still be present in the flue gas. Several dedicated commercial technologies are available for sulphur removal:

1. Wet flue gas desulphurization (WFGD)
2. Spray dry absorber (SDA)
3. Circulating dry scrubber (CDS)
4. Wet Caustic scrubber (WSC)

3.4.1 SO_2 REMOVAL

The WFGD was developed in the 1940s. With the introduction of the U.S. clean air act in 1970, WFGD was commercialized and installed in U.S. coal-fired power stations. The SDA system technology was also developed simultaneously, but was mainly used for low sulphur fuel. By the late 1980s, another technology, the CDS was introduced. Over the past 15 years, these dry technologies have been applied to medium and higher sulphur fuels. During this time a modified version of the WFGD process with the ability to make gypsum as by-product became available and quickly became the technology of choice. Wet caustic based scrubbing has also been used for controlling SO_2 emissions from Fluid Catalytic Cracking (FCC) Units and other air emission sources within refineries. This scrubbing technology is well proven in providing the flexibility to handle the added capacity that comes with FCC unit expansions and in providing uninterrupted operation/performance. A brief description of each of these technologies is provided as below.

3.4.1.1 Wet flue gas desulphurization (WFGD)32, 33,33

WFGD systems remove sulphur dioxide (SO_2) from flue gas by providing intimate contact between the gas and a slurry of finely grounded limestone. The slurry absorbs the SO_2 from the flue gas to form a mixture of calcium sulphite and calcium sulphate (gypsum). Limestone for the wet limestone FGD process is usually received as gravel. Preparation of the limestone slurry involves grinding the limestone extremely finely in a horizontal ball mill containing water. The slurry is pumped through banks of spray nozzles to create fine droplets to facilitate intimate and uniform contact with an updraft of flue gas.

After absorbing the sulphur dioxide, the slurry collects in the bottom of the absorber in a reaction tank, where it is aerated to oxidize the bisulphite ions to sulphates. Oxidized slurry is then recirculated to the spray headers. A portion of the slurry is withdrawn to remove the precipitated gypsum. Typically, the slurry is dewatered in a two stage process involving a hydro-cyclone and a vacuum filter system to produce a gypsum cake for disposal or sale. Water removed from the slurry is returned to the process. A portion of this water is removed from the system as wastewater to limit accumulation of chloride salts and other undesirable constituents introduced with the coal.

Process chemistry of WFGD can be defined with the reactions below. The process efficiency is generally stated to be as high as 98% even for high sulphur fuels.

$$CaO + SO_3 + 2\,H_2O \rightarrow CaSO_4{\bullet}2H_2O$$

$$2\,SO_2 + 2\,CaCO_3 + O_2 + 4\,H_2O \rightarrow 2\,CaSO_4{\bullet}2H_2O + 2\,CO_2$$

$$CaCO_3 + SO_3 + 2\,H_2O \rightarrow CaSO_4{\bullet}2H_2O + CO_2$$

WFGD is typically located after the particulate removal system (i.e. baghouse or electrostatic precipitator). In the past, wet limestone FGD systems were applied to coals of all sulphur levels, including low sulphur coals (<1%). However, as the lime spray dryer technology became commercially available, that became the choice for low sulphur coals. Today, WFGD systems are typically installed on medium- to high-sulphur fuels (>2%), where the relatively low cost of reagent can pay back the higher initial capital cost.

Unit size can range from less than 100 MW to more than 1,000 MW, with 400 to 500 MW units common. Alstom Power Environmental, Babcock & Wilcox and Black & Veatch are a few of the leading technology providers.

3.4.1.2 Spray Dry Absorber (SDA)[33, 34]

The development of the spray dryer FGD technology started in 1977. Initially, it was suitable only for low-sulphur coal with a low (70%) SO_2 removal efficiency, but as experience was gained, the technology was shown to be capable of much higher performance. Maximum absorber size has grown from the early 150 MW units to 400 MW today.

In the SDA, the flue gas is treated by mixing the gas stream concurrently with atomized lime slurry droplets. The lime slurry is atomized through rotary cup spray atomizers or through dual fluid nozzles. Some of the water in the spray droplets evaporates, cooling the gas at the inlet from 300°C or higher to 70°C to 80°C, depending on the relationship between the approach to saturation and removal efficiency. The droplets absorb SO_2 from the gas, and the SO_2 reacts with the lime in the slurry. Desulfurized flue gas, along with reaction products, unreacted lime, and the fly ash passes out of the dry scrubber to the fabric filter.

SDA systems are typically located after the air preheater, but before the particulate collector. The process chemistry is described below. SDA process efficiency is generally stated to be 94%.

$$SO_2 + CaO + 1/2\ H_2O \rightarrow CaSO_3 \bullet 1/2\ H_2O$$

$$CaSO_3 + ½O_2 + 2H_2O \rightarrow CaSO_4 \bullet 2H_2O$$

3.4.1.3 Circulating dry scrubber (CDS)[33, 35]

In a CDS system, flue gas is treated by injecting a mixture of dry hydrated lime or lime slurry and recycled by-product (a mixture of reaction products ($CaSO_3$, $CaSO_4$), excess lime ($Ca(OH)_2$), and fly ash). Preparation of hydrated lime requires the use of a lime hydrator. Although hydrated lime can be purchased, converting commercial quicklime into the hydrated lime on-site typically offers a lower cost alternative, particularly for larger units or multiple unit systems. The reagent is fed to the absorber to replenish hydrated lime consumed in the reaction.

The CDS technology is similar to wet and dry FGD processes, but in CDS solids are continuously recycled to the absorber to achieve a high utilization of the reagent. Reactive material also recirculates within the absorber, achieving a high retention time. It is this circulation that produces the CDS high removal efficiency of 98%. Banks of venturis increase the flue gas velocity at the entrance to the fluidized bed reactor. Water is injected in the scrubber near the outlet of the venturis to humidify the flue gas and promote the absorption of SO_2 from the flue gas and the reaction of SO_2 with the lime. For air firing, humidification is controlled to maintain a flue gas temperature of typically 70°C or approximately 16°C above the adiabatic saturation temperature of the flue gas. For oxy-firing the temperature is 87°C. Hydrated lime absorbs SO_2 from the gas and forms calcium sulphite and calcium sulphate. The desulfurized flue gas, along with reaction products ($CaSO_3$, $CaSO_4$), excess lime ($Ca(OH)_2$), and fly ash passes out of the fluidized bed to the fabric filter.

The process chemistry can be described as below.

$$SO_2 + Ca(OH)_2 \rightarrow CaSO_3\bullet 1/2\ H_2O + ½\ H_2O$$

$$CaSO_3\bullet 1/2\ H_2O + ½O_2 \rightarrow CaSO_4\bullet 2H_2O$$

3.4.1.4 Wet Caustic Scrubber (WCS)[36]

In a WCS system, flue gas is treated by caustic solution. The caustic solution with (typically) a 50% concentration is prepared in water, ensuring the process uses large amounts of water. Caustic reacts with SO_2 to form sodium sulphate and bisulphate which is soluble in water, generating acidic waste which needs to be neutralized before disposal. The process plant operates at higher pH, between 4 and 5.5[37]. The process has high efficiency >97% and is generally flexible with SO_2 inlet feed rate.

The process chemistry can be described as below.

$$SO_2 + NaOH \rightarrow HSO_3^- + Na^+$$

$$HSO_3^- + Na^+ + NaOH \rightarrow SO_3^{2-} + 2Na^+ + H_2O \rightarrow Na_2SO_3 + H_2O$$

3.4.2 SO_3 REMOVAL IN PARTICULATE COLLECTION IN OXY-FUEL

Higher partial pressure of SO_2 and O_2 are expected to increase the SO_3 formation in the flue gas during the oxy-fuel process. Higher SO_3 formation will increase the acid dew point temperature by at least 30°C which will reduce the heat recovery in the air-preheater and thus reduce the overall plant efficiency. SO_3 is expected to be captured in the fabric filter if the coal ash is alkaline. Experiments were conducted at the 20 kW once through rig at IFK, Germany[27]. The results are discussed in Section 8. It was found that 40-65% SO_3 was captured in fly ash across the fabric filter and with such capture the efficiency can be improved by 0.3-0.5%.

3.5 HG REMOVAL TECHNOLOGICAL OPTIONS

Depending on the flue gas conditions and type of coal, Hg may be present in the flue gas as elemental mercury vapor (Hg^0), as an oxidized mercury species (Hg^{2+}), and as particulate - bound mercury (Hg^p). Elemental Hg, released into the exhaust gas, can then be oxidized to Hg^{2+} via homogeneous and heterogeneous oxidation reactions. Among all the Hg species, Hg^0 is difficult to capture due to its insolubility in water, lower bonding energy on the sorbent surface, high volatility and chemical inertness. $Hg^{0,}$ not Hg^{2+}, attacks the aluminium heat exchanger in the CO_2 compression unit.

Different in-situ impurity control systems such as fabric filters (FF), FGD units and sorbent injection can be applied to decrease Hg emissions. Hg removal efficiency can vary significantly with different coals making it difficult to find a consistent Hg removal technology for different types of coal burned in boilers.

The interaction of gaseous Hg with particulates causes small amounts of Hg to be adsorbed on the particle surface and captured by particulate control devices. High capture rates of Hg in particulate control systems occur for bituminous and sub-bituminous coals due to their high ash content and the high chlorine present in the coal. Sorbent injection methods can also be included in particulate control systems to remove Hg[38]. Cold-side ESP and hot-side ESP units will capture some mercury associated with the particulate matter but have been shown to be much less effective when compared to those with FF for both bituminous and sub-bituminous coals because there is less contact between gaseous Hg and fly ash in ESP units[38].

Although FGD systems are built for sulphur removal, wet FGD systems are capable of capturing the water-soluble form of Hg, i.e. Hg^{2+}, as a co-benefit. It has been established that up to approximately 90% of Hg^{2+} can be removed on limestone-based wet FGD systems[39]. In a wet FGD system SO_2 is contacted with limestone-based slurry and through forced oxidation, gypsum is generated. The hydrated gypsum is found to capture water soluble oxidized mercury. When calcined for gypsum production, the Hg bound to the hydrated gypsum can be emitted to the environment, causing environmental concern and possible contamination.

Chlorine and bromine[40] present in the flue gas oxidise the Hg present. Oxidised mercury will absorb onto the surface of carbon present in the flue gas, or onto activated carbon provided to remove Hg. Activated carbon injection[41] is one of the most common Hg remediation processes present in power plants. Depending on the location of injection, two different processes[42]; i.e. powdered activated carbon injection (PACI) or fixed-bed granular activated carbon (FGAC); can be used to capture Hg. In PACI, activated carbon is passed through the flue gas stream and removed in FFs or ESP, whereas in FGAC, an adsorption unit is placed downstream of the PM removal system or FGD unit to remove Hg as a final treatment process. FGAC is generally less favourable for mercury control in coal-fired power plants due to the pressure drop across the packed-bed. But in the case of PACI, the footprint for the PM removal system increases with the increased PM load. The adsorption capacity of PACI is also highly depend on temperature and the SO_3, chlorine and bromine concentration in the flue gas stream. Standard PACI works well at approximately 140°C. Above 180°C the capacity decreases dramatically due to the breakage of the powdered activated carbon (PAC) structure. Moreover, if the sulphur content in the flue gas is high, the adsorption capacity of PAC will decrease due to the competition of high levels of SO_3 and Hg for the active sites on the activated carbon; therefore, this process is generally effective for Hg capture on low sulphur bituminous coal applications. PACI is effective in decreasing Hg^{2+} concentrations in flue gas.

Reduced Hg oxidation in the flue gas occurs when the coal concentrations of chlorine and bromine are low. In some situations, the chlorine and bromine concentration inside the flue gas is not adequate for Hg oxidation. Hence, for these situations chemically-treated PAC (i.e. chlorinated or brominated) has been developed to overcome the limitations of PAC.

4 CAPITAL COST COMPARISON OF PUBLISHED EPRI[1] AND NETL[2] REPORTS

A comprehensive literature review was carried out to identify the cost associated with sulphur and mercury removal in the oxy-fuel process. B&W published several flowsheets, each suitable for coals with different sulphur contents. To date, only two detailed reports have been published on the cost of an oxy-fuel thermal power plant. These were published by EPRI (Electric Power Research Institute) in 2010 and NETL (National Energy Technology Laboratory) in 2008. B&W provided the costs for these reports. Their summaries follow.

4.1 NETL REPORT SUMMARY

The objective of the NETL report was to establish a cost and performance baseline for the oxy-combustion process that can be used as a basis for comparison with past and future studies. This study comprises twelve plant configurations, or flowsheets. Mass and energy balances were performed using Aspen Plus software, followed by development of major equipment specifications, with capital and operating cost estimates. For all the spreadsheets considered the coal feed rate (Illinois No. 6 bituminous coal) was adjusted as necessary to maintain a nominal net plant output of 550 MW.

The technical and cost related information for the oxy-combustion boiler and gas cleaning equipment was provided by Babcock & Wilcox. Air Liquide provided the information related to the cryogenic distillation process for the production of oxygen and CO_2 purification and compression processes. The balance of plant designs, costs and cost rollups were competed by Research and Development Solutions LLC (RDS). Technical guidance was provided by the Department of Energy's National Energy Technology Laboratory (NETL).

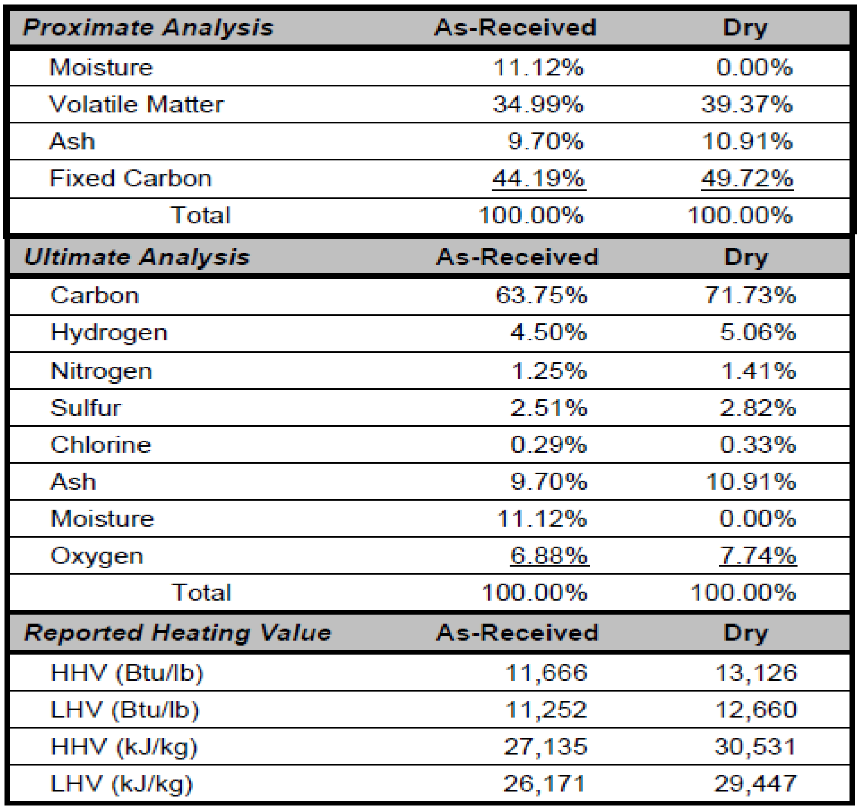
Of the twelve cases examined, four conventional air-based combustion designs (with and without CO_2 control) were included for reference, six oxy-combustion designs with O_2 provided by a cryogenic distillation process and two oxy-combustion designs with O_2 provided by an ion transport membrane (ITM) process were considered. Both supercritical (SC) 240 bar/600°C/620°C (3,500 psi/1,110°F/1,150°F) and ultra-supercritical (USC) steam cycles 275 bar/730°C/760°C (4000 psi/1350°F/1400°F) were analyzed. Oxygen concentrations of 95, 99 and 100 mol%, producing varying CO_2 purities, with less than 0.015% by volume H_2O were considered.

The total plant costs (TPCs) including equipment, materials, construction labor, home office expenses, process contingencies, and project contingencies were obtained. Moreover, Levelized cost of electricity (LCOE) generation was calculated for a period of 20 years.

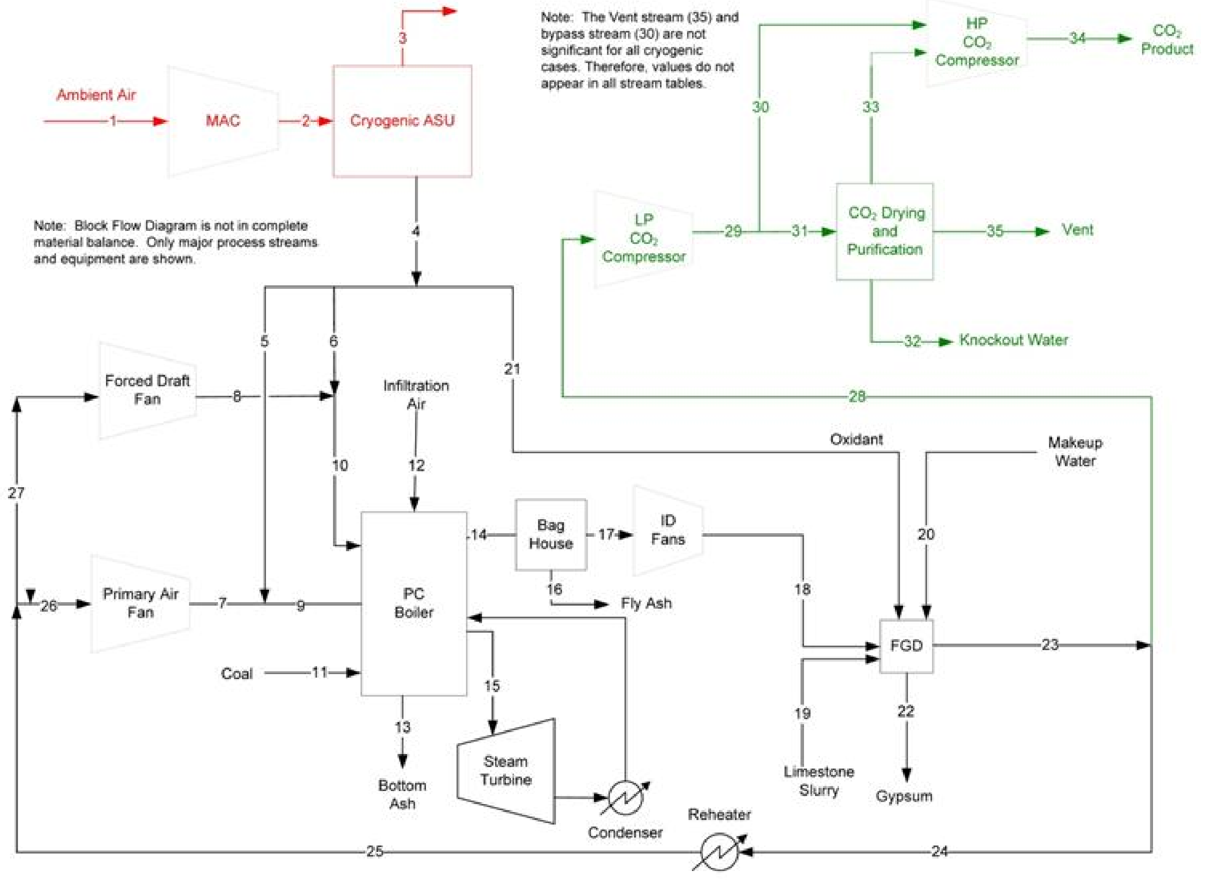
NETLs case 5a, a wall fired PC boiler with flue gas recirculation utilizing 99 mol% oxygen from a cryogenic ASU was selected by this current study for comparison. This design uses a WFGD system for SOx removal due to NETLs selection of a high sulphur coal, Illinios No 6. It's specifications are shown in Table 1. The flowsheet for 'case 5a' is shown in Figure 5. In NETLs sensitivity analysis a low sulphur (<1%) case was investigated where the WFGD system was eliminated at the expense of a condensing heat exchanger, capable of handling the volume of flue gas and higher acidic condensate. By eliminating the WFGD, an 8.03% reduction was achieved in the electricity generation price for the CCS unit, indicating sulphur removal unit operations contribute significantly to the overall energy penalty in advanced low emission clean coal technologies.

However, no other sulphur removal unit operations such as SDA, DCCPS or CDS have been studied or compared in the NETL report, which creates the basis of the current study. In the current report, different flowsheets published by IHI and B&W have been studied to estimate the cost associated with sulphur removal in oxy-fuel combustion.

Table 1: The proximate and ultimate analysis for Illinois number 6 coal, used by NETL to assess the cost of Oxy-fuel combustion.[2]



Proximate Analysis	**As-Received**	**Dry**
Moisture	11.12%	0.00%
Volatile Matter	34.99%	39.37%
Ash	9.70%	10.91%
Fixed Carbon	44.19%	49.72%
Total	100.00%	100.00%
Ultimate Analysis	**As-Received**	**Dry**
Carbon	63.75%	71.73%
Hydrogen	4.50%	5.06%
Nitrogen	1.25%	1.41%
Sulfur	2.51%	2.82%
Chlorine	0.29%	0.33%
Ash	9.70%	10.91%
Moisture	11.12%	0.00%
Oxygen	6.88%	7.74%
Total	100.00%	100.00%
Reported Heating Value	**As-Received**	**Dry**
HHV (Btu/lb)	11,666	13,126
LHV (Btu/lb)	11,252	12,660
HHV (kJ/kg)	27,135	30,531
LHV (kJ/kg)	26,171	29,447



Note: The Vent stream (35) and bypass stream (30) are not significant for all cryogenic cases. Therefore, values do not appear in all stream tables.

Note: Block Flow Diagram is not in complete material balance. Only major process streams and equipment are shown.

Figure 5: NETLs 5a flowsheet for oxy-fuel combustion[2].

4.2 EPRI REPORT SUMMARY

The EPRI report was commissioned by the Australian Department of Resources, Energy and Tourism (DRET) to identify costs and diffusion barriers to the deployment of low-emission technologies and to undertake assessment of the costs of various options for different electricity generation technologies out to 2030. The objective of the work was to establish an up-to-date cost and performance database agreed by Australian stakeholders as supportable in the Australian context. The report also provided a levelised cost analysis of a basket of technologies in 2015 and 2030, providing an agreed basis for comparing globally available power generation technologies and costs.

EPRI evaluated a specific list of technologies focusing on twelve key central station technologies of current and future interest to Australia such as Integrated Gasification Combined Cycle (IGCC); Pulverised Coal (PC); Combined Cycle Gas Turbine (CCCT); Open Cycle Gas Turbine (OCGT); Solar Thermal; Solar Photovoltaic; Wind; Tidal/Wave; Geothermal; Nuclear; Hydroelectric; and Biomass. For each of the above technologies, appropriate plant size, configuration and design approaches were selected and evaluated to develop performance, emissions and cost information, to determine the preferred choices for the future Australian power generation mix.

For each selected configuration, heat and material balances and emissions data were developed based on the selected fuels and capacities. Capital costs and operating and maintenance cost estimates were developed based on US, Gulf Coast rates for equipment, materials and labour. These costs were then adjusted to Australian values based on adjustment factors developed by EPRI's subcontractor by liaising with their Australian and US offices.

Post-, Pre- and Oxy-fuel combustion options were studied in the EPRI report. The oxy-fuel flowsheet provided by EPRI is shown in Figure 6, and used a supercritical Rankin steam cycle. But due to the number of technologies considered in the EPRI report, very little detailed information was provided for the processes or the resulting costs. For example, the type of FGD system for SO_2 removal used in the oxy-fuel combustion flowsheet was not provided. For example, a wet FGD would be a high cost option for a low sulfur coal. Neither were the fixed and variable O & M costs for each unit operation provided. Only total operating and maintenance costs for the entire plant were presented, though the capital costs were provided for each unit operation.

EPRI considered two Australian coals for their study: Latrobe Valley Brown Coal and Hunter Valley Black Coal with low sulphur. Their characteristics are tabulated in Table 2. The costs of oxy-fuel combustion using Hunter Valley black coal was used for comparison in this study.

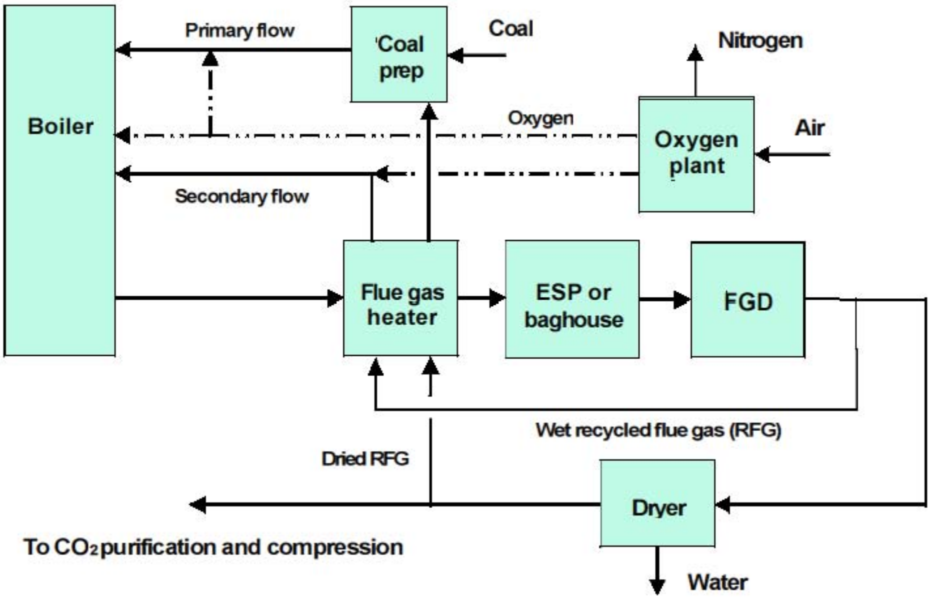
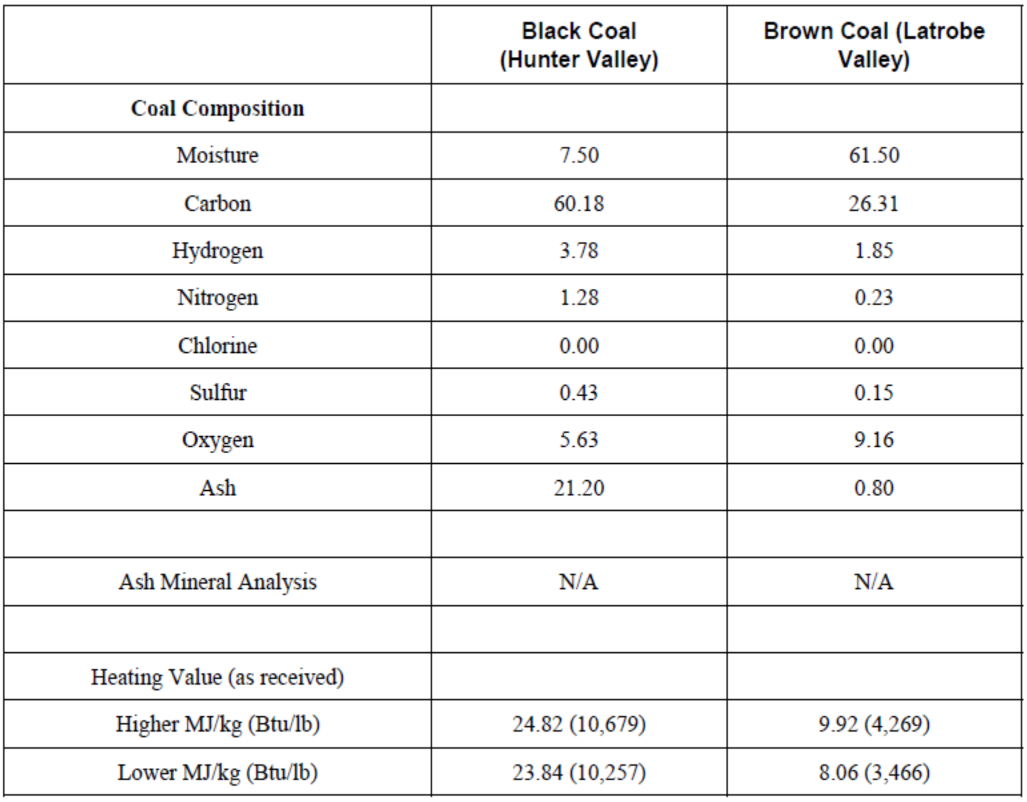


Figure 6: EPRI oxy-fuel flowsheet.[1]

Table 2: Coal analyses used by EPRI for their cost estimation of an oxy-fuel plant.[1]



	Black Coal (Hunter Valley)	Brown Coal (Latrobe Valley)
Coal Composition		
Moisture	7.50	61.50
Carbon	60.18	26.31
Hydrogen	3.78	1.85
Nitrogen	1.28	0.23
Chlorine	0.00	0.00
Sulfur	0.43	0.15
Oxygen	5.63	9.16
Ash	21.20	0.80
Ash Mineral Analysis	N/A	N/A
Heating Value (as received)		
Higher MJ/kg (Btu/lb)	24.82 (10,679)	9.92 (4,269)
Lower MJ/kg (Btu/lb)	23.84 (10,257)	8.06 (3,466)

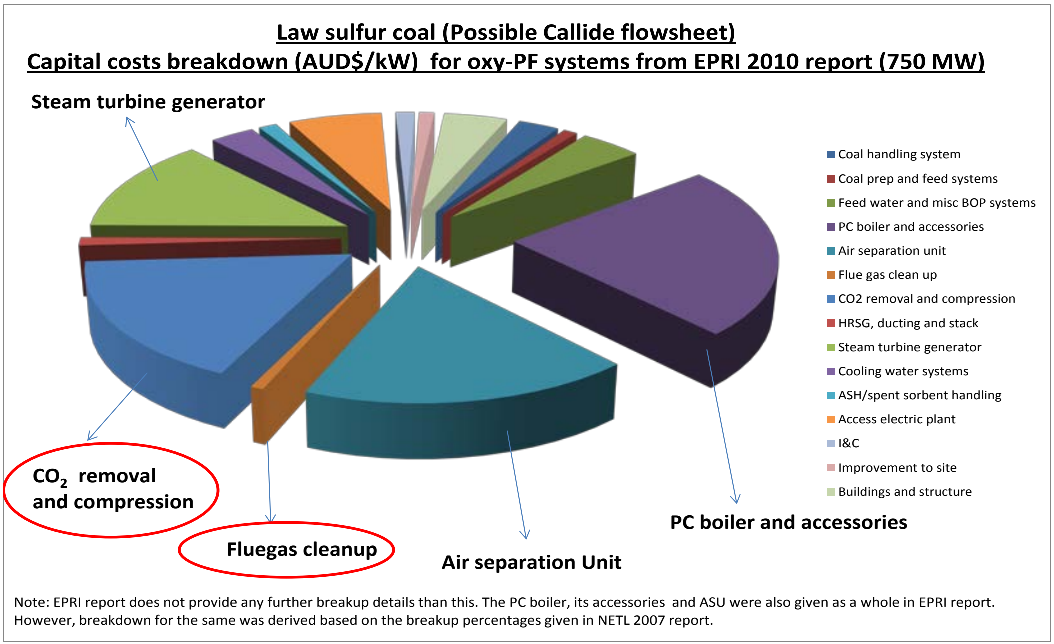
4.3 COMPARISON OF OXY-FUEL CAPITAL COSTS BY NETL AND EPRI

Capital costs for oxy-fuel unit operations produced by NETL and EPRI are presented in Figures 7 and 8. The assumptions, limitations and methodology for the comparison are highlighted in detail further in Appendix 1. The estimates produced by NETL and EPRI for flue gas clean up and CO_2 removal and compression costs are significantly different. This could be due to following reasons:

1. EPRI calculations for low sulphur coal may have overestimated the costs for the CO_2 compression and purification in Australian conditions, or

2. EPRIs costs for FGD may be included in CO_2 purification and compression costs.

The costs for the flue gas cleaning equipment such as ESP/PJFF, wet FGD /SDA and CO_2 purification and compression cannot be obtained directly from any of these reports. Therefore, costing codes have been developed in this work for such unit operations and described in Section 9.



Note: EPRI report does not provide any further breakup details than this. The PC boiler, its accessories and ASU were also given as a whole in EPRI report. However, breakdown for the same was derived based on the breakup percentages given in NETL 2007 report.

Figure 7: The breakdown of capital costs for oxy-fuel combustion provided by EPRI.[1]

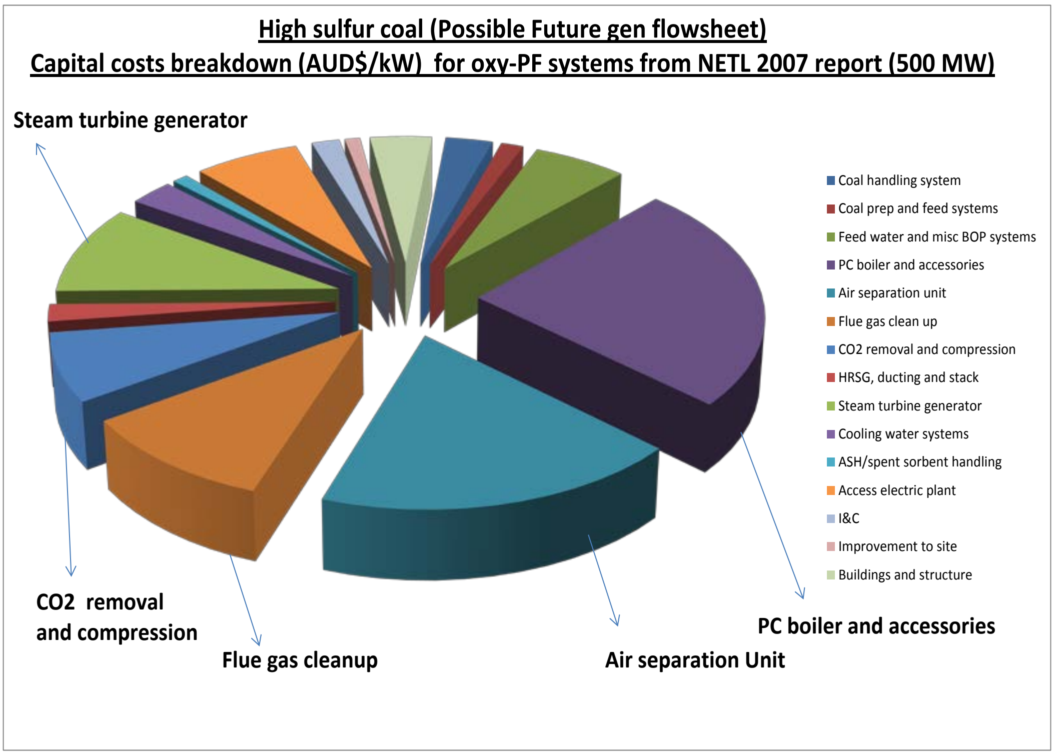


Figure 8: The breakdown of capital costs for oxy-fuel combustion provided by NETL.[2]

5 PLANT DESIGNS (FLOWSHEETS) FOR SO_X AND HG REMOVAL IN OXY-FUEL.

As discussed earlier in Section 3, some SO_x in the flue gas may be removed by the PM handling systems such as Electrostatic precipitators (ESP) or Fabric filters (FF). However, the majority of the SO_2 is expected to be removed by either FGD systems or sorbent injection[43]. In retrofit conditions, the fly ash load on an ESP or FF may increase above its design value when sorbent injection is used. Also, handling more solids in ESPs or FFs has been found to be more challenging. Therefore, FGD systems are the preferred option for removing SO_2 from flue gas depending on the amount of sulphur present in the coal. As described in Section 3, the FGD systems include SDA, WFGD and CDS. A wet/direct contact cooler polishing scrubber (DCCPS) using a caustic solution can also considered for SO_2 removal in a power plant operating with low sulphur coal. Compression can also remove the remaining SO_2 as sulphuric acid in the condensate. However, SO_2 removal in the compression unit can affect the NO_x removal performance; therefore, it is generally recommended to remove SO_2 before the flue gas enters the compression circuit.

SO_3 in the flue gas hydrates to form H_2SO_4 vapour. As the flue gas temperature cools, the H_2SO_4 vapour condenses to form a sub-micron aerosol (acid mist). SO_3 levels in oxy-fuel flue gas are expected to increase by approximately three times compared to air combustion levels and therefore the air-preheater may need to be operated at higher temperatures due to the higher acid dew point. Due to the small size of these aerosol particles (less than 0.5 µm), the acid mist is difficult to capture. However, 20-40% capture rates have been observed as the flue gas pass through the final boiler heat traps (air heaters) and in the air pollution control equipment (ESP/FF/FGD systems).

During combustion, the mercury (Hg) in coal is vaporized and converted to elemental mercury (Hg^0) vapour in the high temperature regions of coal-fired boilers. As the flue gas is cooled, a series of complex reactions begins to convert Hg^0 to ionic mercury (Hg^{2+}) compounds. The presence of chlorine gas-phase equilibrium favours the formation of mercury chloride ($HgCl_2$) at flue gas cleaning temperatures. However, Hg^0 oxidation reactions are kinetically limited and, as a result, Hg enters the flue gas cleaning device(s) as a mixture of Hg^0 and Hg^{2+}. The mercury speciation will have considerable influence on selection of mercury control approaches as mentioned earlier. Some removal of mercury may be achieved via controls used to remove particulate matter (PM), sulphur dioxide (SO_2), and nitrogen oxides (NOx). Other dedicated techniques are activated carbon beds (atmospheric or pressurized), activated carbon injection (ACI), gold and other rare metal coated sorbents, bromination of coal and use of bromine gas. For dedicated systems, the efficiency of mercury removal will be largely affected by SOx present in the flue gas. The atmospheric activated carbon bed can only achieve mercury removal efficiencies of 50-70%. The pressurised mercury bed can achieve up to 90% efficiency, however, the risks associated with fire at pressurised conditions due to hot spot generation in a fixed carbon bed reduce its integration possibilities in oxy-fuel flue gas cleaning systems. Recently gold and other rare metal coated sorbents that can be used in pressurised conditions without any risk of fire have been suggested and are claimed to achieve mercury removal of up to 99%. The use of bromine gas for mercury removal is also in the research phase.

Looking to the different removal unit operations and removal mechanisms for SOx and Hg in oxy-fuel plants, several flowsheet options[9] have been suggested by IHI and B&W. They are described in the following subsections.

5.1 SOX REMOVAL

5.1.1 CALLIDE FLOWSHEET 1 (C1) - IHI PLANT DESIGN FOR LOW SULPHUR COAL

The IHI design for low sulphur coal is given in Figure 10. This design is for a plant retrofit, i.e. changing an existing air-fired plant with no CO_2 recovery to an oxy-fuel plant. The suggested retrofit flowsheet has been attempted in Australia at Callide oxy-fuel demonstration project (30 MW_e). This is the first power plant to evaluate the retrofit option, unique from other oxy-fuel demonstration projects around the world. Currently, there are no regulations for SOx and NOx emissions in Australia. Therefore, no SOx and NOx removal unit operations are installed in existing coal fired power stations. As Australia does not have experience with SOx removal unit operations, the IHI design attempts to avoid the need for dedicated SOx removal unit operations. Therefore, this flowsheet is suitable for low sulphur coal (<0.5%).

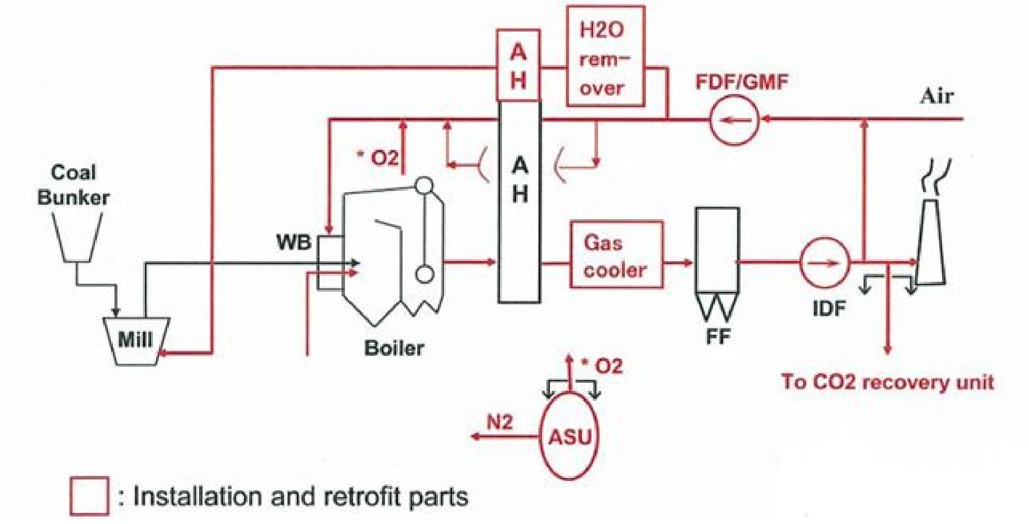


Figure 9: Schematic of the IHI plant design for low sulphur coal, C1.

In the Callide oxyfuel retrofit, the boiler remains un-modified. The boiler flue gas flow rate and temperature change significantly, and so a gas cooler was proposed to cool the gas to the air combustion outlet temperature of 150°C, avoiding excessive temperatures at the fabric filter. Addition of the gas cooler increases the draft head for the fans and so the ID fans and FD fans were modified or replaced. Approximately 70% of the flue gas is recycled to the furnace, divided into a primary gas stream for the pulverizing mills and a secondary gas stream to the windbox. No sulphur cleaning of the recycle streams occur. This recycling of the gas triples the sulphur content within the boiler compared to air combustion, but when burning low sulphur coal this is equivalent to the combustion of bituminous coals with moderate sulphur contents during air combustion. The moisture content of the recycle flue gas also increases compared to air firing conditions, requiring dewatering and reheating of the primary gas stream to ensure it remains above saturation temperature.

Air Liquide designed the CO_2 processing unit (CPU) for the Callide oxyfuel demonstration plant. SO_2 is removed in an atmospheric caustic scrubber (scrubber 1) using caustic solution before compression of the CO_2 rich gas, shown in Figure 10.

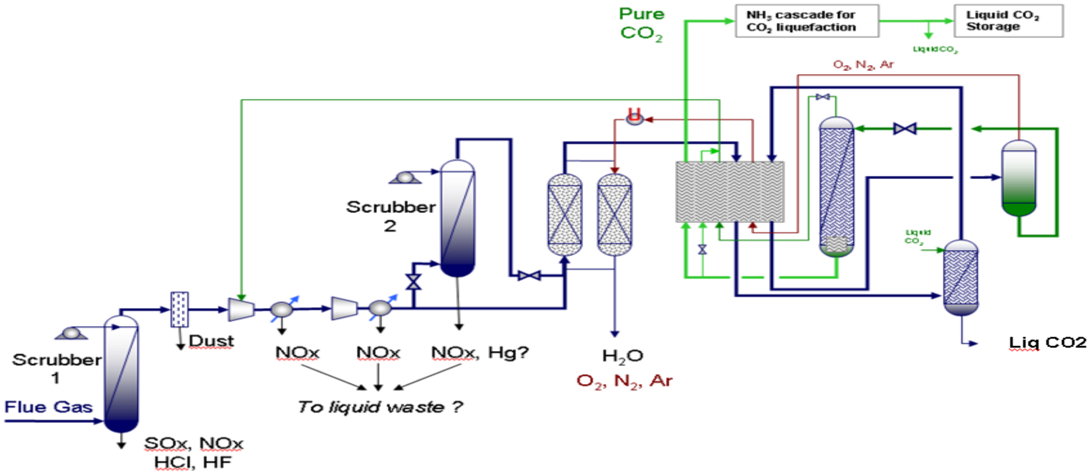


Figure 10: The CO_2 processing unit for Callide oxy-fuel demonstration project

5.1.2 BABCOCK AND WILCOX PLANT DESIGN FOR LOW SULPHUR COAL (<1%S)

B&W have designed several flowsheets for the combustion of coals with varying sulphur levels. Figure 11 shows the process schematic of B&W design for low sulphur coal, referred to as FL1 in this report. Following the recycle heater, the flue gas is split into the secondary recycle, the primary recycle and gas for the CPU. Due to the low sulphur (0.85 wt. % d.b.) level in the design coal, no cleaning of the recycle streams is needed. The design employs cold-recycle of the primary gas stream for coal transport from mill to the burners. Warm-recycle of the secondary gas stream is used; the secondary recycle temperature is decreased to 150^{o}C before passing through a fabric filter (FF) and forced draft fan which returns the flow to the recycle heater for reheating and then to the windbox. The remaining flue gas passes through the SDA, followed by a fabric filter and a polishing sodium-based direct contact cooler scrubber (DCCPS) to remove SO_2 to very low concentrations (<10 ppmv) before the gas enters the compression circuit in the CPU. Any small amount of remaining SO_2 that enters the CPU is condensed.

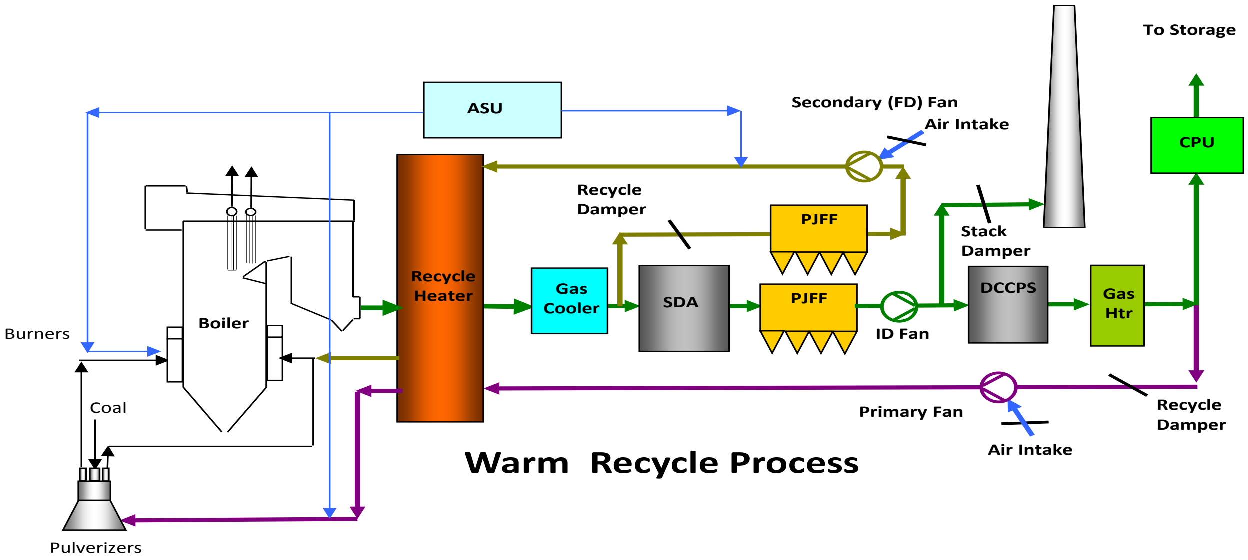


Figure 11: Schematic of B&W plant design for low sulphur coal, FL1.

In addition to the significant NO_x reduction produced by oxy-combustion, the combustion system design incorporates provisions to reduce NO_x formation in the burner zone. NO_x produced in the process passes into the CPU where it is removed during the compression process with only a very small amount remaining in the non-condensable gaseous vent stream to atmosphere. Mercury is removed in both the SDA-FF and the polishing scrubber prior to entering the CPU. The remainder is removed within the CPU process. Particulates are mainly removed from both the secondary recycle and the main flue gas streams by high efficiency fabric filters.

Since there is no SO_2 or moisture removals in the warm recycled secondary recycle steam, the moisture and SO_2 levels in the boiler are high, about the same as experienced in an air-fired boiler burning a moderate sulphur content bituminous coal. The plant heat rate is generally improved by returning warmer recycle gas to the boiler and using some of the heat in the steam cycle.

5.1.3 BABCOCK AND WILCOX PLANT DESIGNS FOR MEDIUM SULPHUR COAL

Two B&W designs are reported in the literature for medium sulphur (~1.5% d.b.) coals. The main difference between the two is the FGD system employed for SO_2 removal. Figure 12 shows the first design, FL2, uses a SDA as the FGD system and cool recycle rather than the warm recycle for low sulphur coal. The SDA keeps the SO_2 and HCl concentrations in the boiler about the same as those during air firing, minimising the high temperature corrosion risk. After the recycle heater, the flue gas is cooled in a gas cooler to 150-180°C before it enters the SDA. A 30°C difference between inlet and outlet temperature is maintained in the SDA. The pulse jet fabric filter (PJFF) installed after the SDA is used to capture fly ash from flue gas, though some fly ash is removed in the SDA. A secondary recycle stream is withdrawn after the PJFF and the remaining stream goes to the DCCPS, whose primary function is to remove moisture from the flue gas, but it also further reduces SO_2. After slightly reheating the flue gas leaving the DCCPS, about half of the flow goes to the CPU for compression, and the remainder is sent to the primary fans. This primary recycle stream is heated in the recycle heater before becoming the coal drying and conveying medium in the pulveriser.

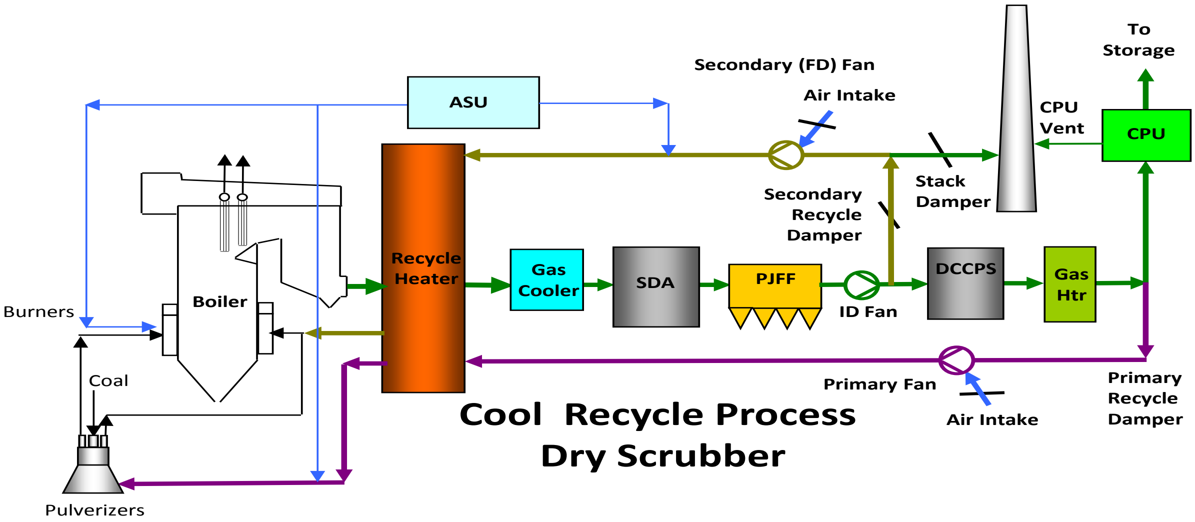


Figure 12: Schematic of B&W plant design 1 for medium sulphur coal, FL2.

The second B&W design for medium sulphur coals, FL3, shown in Figure 13, uses a WFGD which operates at 50-60°C. The flue gas at the outlet of WFGD is saturated with water vapour and therefore requires reheating before being recycled in the secondary recycle stream to avoid low temperature corrosion in the pipeline. As the WFGD operates at lower temperatures, the PJFF can be installed before it to remove fly ash particles before the stream enters the gas cooler. The remaining flue gas proceeds to the DCCPS. After re-heating, nearly half of the flue gas is sent as the primary recycle to the furnace via the mill and the remainder is directed to the CPU. Dry sorbent injection has been suggested upstream of the PJFF to remove SO_3.

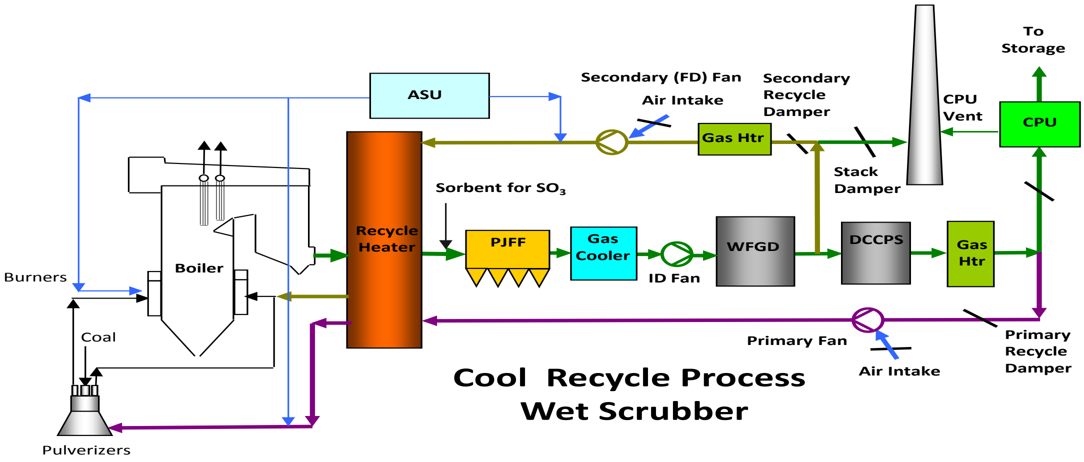


Figure 13: Schematic of B&W plant design 2 for medium sulphur coal, FL3.

5.1.4 BABCOCK AND WILCOX PLANT DESIGN FOR HIGH SULPHUR COAL (>3%S)

The B&W design for high (~3.2 % d.b.) sulphur coal, FL4, shown in Figure 14 also employs cold recycle. This design was earlier proposed for the FutureGen 2.0 oxy-fuel demonstration program. It uses the same unit operations for SO_2 removal as suggested in design 2, FL3, for medium sulphur coal, however, the secondary recycle stream is withdrawn after the DCCPS because the SO_2 concentrations after WFGD when high sulphur coals are used will be too high for recycle to the boiler. After reheating the flue gas stream exiting the DCCPS, the stream is spilt into three streams: 1, the flue gas for compression, 2, the primary recycle and 3, the secondary recycle.

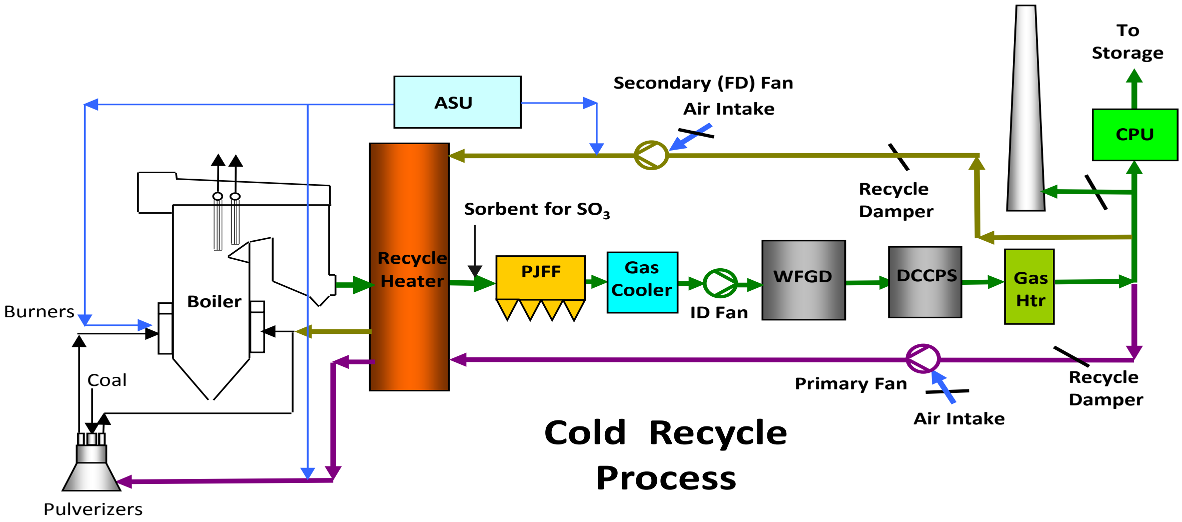


Figure 14: Schematic of B&W plant design for high sulphur coal, FL4.

5.1.5 MODIFIED FUTUREGEN 2.0 PLANT DESIGN BY BABCOCK AND WILCOX (2%S)

Figure 15 shows the schematic of the modified FutureGen 2.0 flowsheet designed for 2% sulphur coal. It uses a circulating dry scrubber (CDS) developed by B&W. This CDS technology is ideal for smaller units that are firing medium to high sulphur coals. B&W state that compared to wet FGD systems, the CDS system's capital costs are lower with potential for higher SO_3 reduction and lower particulate matter (PM) emissions due to the integration of B&W's PGG pulse jet fabric filter. HCl, HF, mercury, heavy metals, dioxins and furans concentrations are also reduce. Like the spray dryer absorber (SDA), the CDS creates a dry solid byproduct and does not require a wastewater treatment facility. But the SDA removal capacity is limited by the higher moisture flue gas. In addition, the CDS is ideal for plant locations with limited space because of its smaller footprint. The B&W FutureGen 2.0 design is a 168 MW plant and scaleup of the CDS may be an issue, or larger plants may require multiple units.

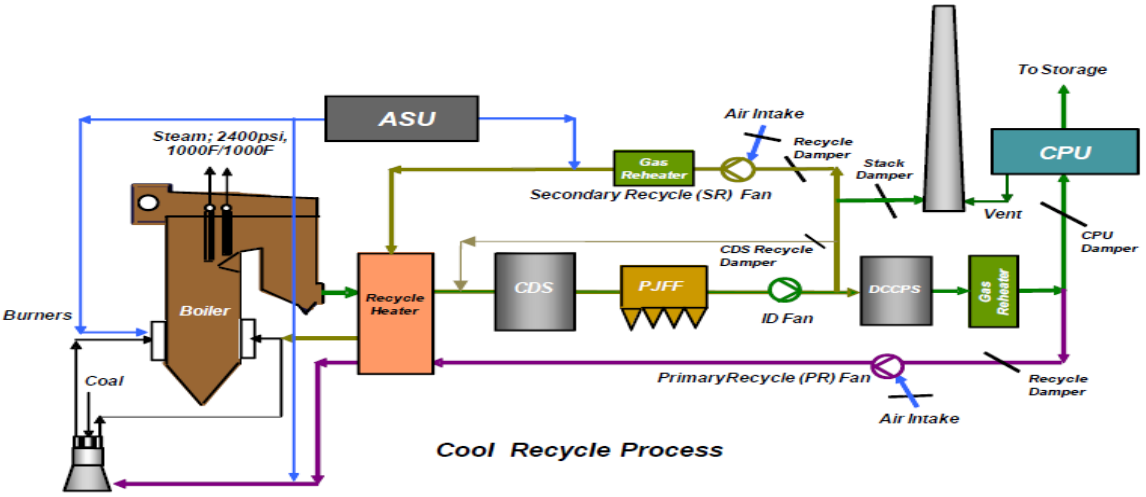


Figure 15: Schematic of B&W FutureGen 2.0 plant design for 2% Sulphur coal.

5.2 MERCURY REMOVAL

5.2.1 IHI DESIGN

Controls used to remove particulate matter (PM), sulphur dioxide (SO_2), and nitrogen oxides (NO_x) will also remove some mercury. Literature indicates that some oxidized mercury may be removed in the ESP/FF or FGD plant, elemental mercury is largely unaffected[44]. Unless sufficient chlorine is present in the coal or an SCR device is used, the total amount of mercury remaining in the flue gas before CO_2 compression may be significant. Research into the speciation of mercury in Australian power plants has shown that about 58% of the mercury emitted in the flue gas is in the elemental form [45]. This mercury will be oxidized during compression and removed as mercury nitrate[46].

An initial plan for the Callide oxyfuel project, designed by IHI, was to include a dedicated Hg removal system, along with indirect Hg removal. Atmospheric or pressurized activated carbon beds were considered. However, no dedicated Hg removal system was installed due to the low mercury content of the Callide coal. In trial runs, no mercury was found to be present after the compression circuit. Mercury is known to attack aluminium heat exchangers. Air Liquide has proposed a method for mitigating Hg corrosion risks using brazed an aluminum heat exchanger (BAHX) as the coldbox[47].

5.2.2 B&W DESIGN

No information is provided for Hg removal in the existing literature for B&W designs. However, mercury is expected to be removed in ESP/FF, FGDs and the compression circuit.

6 REPORT METHODOLOGY AND BASIS FOR COST EVALUATION

Cost estimates have been calculated and compared for SOx removal for each flowsheet. For mercury removal units, such as atmospheric or pressurized carbon beds, the costs have been obtained from literature. Recent references, such as NETL 2008 report[2], EPRI 2010 report [1] and S&L 2011 reports [48-50] were used for cost calculations. However, this report has made several assumptions and therefore accuracy of the data should be considered around +/-30%. Capital costs ($/kW), fixed O&M ($/kW/year) and variable O&M ($/MWh) costs have been calculated for all the cases described above.

6.1 SO_X REMOVAL COST ESTIMATION

A simple methodology for obtaining sulphur removal cost estimates is presented in Figure 16.

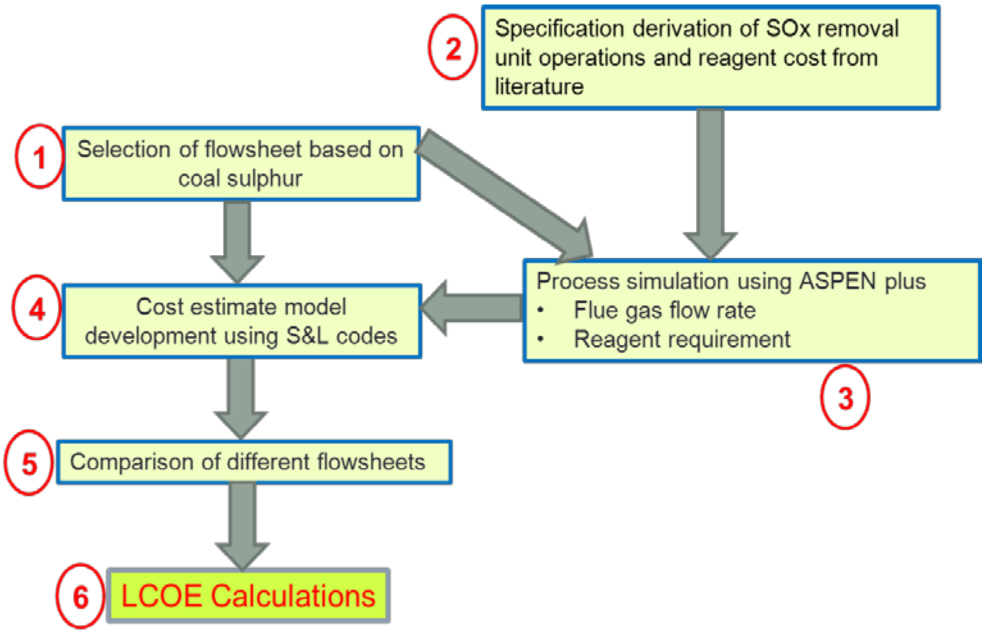


Figure 16: Methodology for obtaining sulphur removal cost estimates

The steps performed to determine the costs associated with the sulphur removal were:

Step 1: Select flowsheets based on coal sulphur

The flowsheets (i.e. C1, FL1, FL2, FL3 and FL4) discussed in Section 5.1 were selected for the current study. Callide coal was selected as the low sulphur coal. Hypothetical medium and high sulphur coals were created by altering the oxygen content in this Callide coal. The proximate and ultimate analyses of low, medium and high sulphur coals are shown in Table 3.

Table 3: Proximate and ultimate analyses of the coals used to determine costs of the sulphur removal plants of the flowsheets considered.

	Details	Unit	Low S Coal	Medium S Coal	High S Coal
Proximate Analysis	Moisture	Wt%, a.d.	3.88	3.88	3.88
	FC	Wt%, d.b.	43.3	43.3	43.3
	VM	Wt%, d.b.	24.24	24.24	24.24
	ASH	Wt%, d.b.	32.46	32.46	32.46
Ultimate Analysis	Carbon	Wt%, d.b.	49.83	49.83	49.83
	Hydrogen	Wt%, d.b.	2.91	2.91	2.91
	Nitrogen	Wt%, d.b.	0.71	0.71	0.71
	Chlorine	Wt%, d.b.	0.00	0.00	0.00
	Sulfur	Wt%, d.b.	0.22	1.50	3.50
	Oxygen	Wt%, d.b.	13.87	12.59	10.59

In the following sections, flowsheets will be named as per nomenclature given in Table 4.

Table 4: Flowsheet nomenclature

Flowsheet	#	Coal type
IHI Design (Callide flowsheet)	C1	Low S Coal
B&W Design 1	FL1	Low S Coal
B&W Design 2	FL2	Medium S Coal
B&W Design 3	FL3	Medium S Coal
B&W Design 4	FL4	High S Coal
B&W Design 5	FL5	Medium S Coal

Step 2: SOx removal unit specifications and reagent costs from literature

Various sulphur removal unit operations have been employed in the studied flowsheets. Each unit operation uses different reagents for the sulphur removal. For example, SDA uses lime while WFGD employs limestone. Caustic solution is used in the DCCPS. Their reaction chemistry (described in Section 3.4), efficiency and reagent costs have been obtained from the literature and tabulated as follows.

Table 5: Specifications of SOx removal unit operation

SO_2 removal unit operation	SO_2 removal efficiency (%)	Stoichiometry (mol/mol)	Slurry concentration (wt%)
SDA	94	Ca:S = 1.4	20
WFGD	98	Ca:S = 1.03	30
DCCPS	95	Na:S = 2	50

The costs for the raw material or reagent, labour rate, electricity rate, waste disposal rate etc. have been assumed as below.

Table 6: Reagent costs

Details	Unit	Value
Limestone cost [48]	\$/ton	15
Lime cost [49]	\$/ton	95
Caustic cost (100%)[51]	\$/ton	940

Step 3: Process simulations

Thermodynamic process simulations were completed using ASPEN Plus v7.3. Several unit operations such as RStioc, RYield and RGibbs reactors have been used to simulate furnace and sulphur removal unit operations for oxyfuel process. The details of the process simulation are given in Section 7. The objective of the process simulations is to identify flue gas flow rates, SO_2 concentrations in the flue gas and reagent requirements.

Step 4: Cost estimate model development

For cost estimates, a detailed literature review was completed. NETL and EPRI reports have been studied and compared in Section 4. However, the cost associated with sulphur removal can't be identified from the published reports. Therefore, individual costing codes developed by S&L for different SO_x removal unit operations such as SDA[49], WFGD[48], DCCPS[49] and PJFF[50] were used in the current study. The process parameters obtained from process simulations such as flue gas flow rates, SO_2 concentrations and reagent requirements have been used to estimate capital, fixed and variable O&M costs. The details for S&L codes have been given in Section 9. The capital costs include the costs for base equipment, reagent preparation and handling equipment, waste handling equipment and balance of plant costs such as ID fan, boosters, piping, ducting etc. The fixed O&M costs take into account the costs for the labour, maintenance and administration. The variable O&M costs comprise only the reagent costs.

Step 5: Comparison of different flowsheets

The costs associated with each flowsheet considered in this report have been compared and discussed in Section 10 along with the flowsheets outlined in the NETL and EPRI reports.

Step 6: Levelised cost of electricity (*LCOE*) calculations

LCOE has been calculated using Equation 1 and discussed in Section 10.

$$LCOE = \frac{\sum_{t=1}^{n} \frac{C + M_t + F_t}{(1+r)^t}}{\sum_{t=1}^{n} \frac{E_t}{(1+r)^t}}$$

Equation 1

Where
LCOE = Average lifetime levelised electricity generation cost
I_t = Investment expenditure in the year t
M_t = Operations and maintenance expenditure in the year t
Ft = Fuel expenditure in the year t
Et = Electricity generation in the year t
r = discount rate
n = Amortisation period

6.2 HG REMOVAL COST ESTIMATION

Mercury removal varies significantly depending on the technology used and the coal combusted. Coal mercury is converted to gaseous mercury (Hg^0) in the combustion flame and is partially oxidised by the combustion gases as the gases cool. The cooling rate of the gases influence the amount of oxidised mercury formed and the presence of chlorine and nitrogen oxides increases the conversion of elemental mercury to mercury II species. It also appears that some fly ash species may increase the oxidation of mercury. Oxidised mercury can be removed with the fly ash on fabric filters and in wet scrubbers without incurring additional costs but the elemental mercury will not. If low levels of oxidation occur increased oxidation must be obtained by addition of bromine or additional chlorine or activated carbon or gold or silver plates must be used to collect the elemental mercury.

Table 7 shows the proportion of mercury removed reported for each type of gas cleaning equipment, but these removal rates are very dependent on the coal used within each classification and can only be used as a general guide. These results are for a conventional power plant. With the lower NOx content in oxyfuel combustion, the conversion of mercury would be expected to decrease slightly.

Table 7: The reported mercury removal of gas cleaning equipment

Control equipment	**Percent of mercury removed (Bituminous / Sub bituminous / Lignite)**
Cold side electrostatic precipitators (ESP)	27[52,53],28/13/8[54], 36/9/1[55]
Fabric Filters (FF)	58[53], 90/72/NT[54], 90/75/NT[55]
ESP + Wet Scrubber	49[53], 69/29/44[54], 81/10/48[55]
FF + Spray dryer	0-99[53], 98/19/NT[54], 98/25/2[55]
FF + Wet Scrubbers	88[53], 98/NT/NT[54], 98/NT/NT[55]
Selective catalytic reduction (SCR) +ESP	1-87[55], 8/69/NT[54]
SCR + ESP + Wet scrubbers	44-90[55], 85/NT/NT[54]

The Callide flowsheet, C1, utilises a NaOH scrubber and fabric filters. Capture of mercury in WFGDs occurs because the oxidised mercury is soluble in water. Therefore capture in NaOH scrubbers would be expected to be similar to WFGDs. Table 7 suggests that the Callide flowsheet C1 could achieve a reasonable removal of mercury without requiring other technologies to increase the mercury capture. However, if further mercury removal was required, collection of the mercury on activated carbon would probably be installed as it is the most common technology that is effective in the removal of mercury. At temperatures from as low as 38°C to above 70°C the concentration of mercury can be reduced to 0.01 µg/Nm^3 by activated carbon.

Injection of the activated carbon into the gas steam allows intimate contact between the carbon and the gas and uses the fabric filters already present in the flowsheet to collect the carbon laden with mercury. The EPA lists costs for a 90% removal of mercury by carbon injection as between $US5,000 and $US28,000/lb of mercury removed, while the US Department of Energy (DOE) quotes $US25,000 to $US70,000/lb of mercury removed, increasing the cost of power generation by 0.1-0.8c/kWh[53, 56] in 2003 US dollars. Unfortunately, collection of the carbon with the fly ash can make the flyash unsalable. If the flyash is then classed as a hazardous material due to the mercury content, the disposal costs can be significant. Neither the loss of income from the sale of flyash or the disposal costs are included in the figures listed above.

Alternately installation of a packed bed of activated carbon can also be used to remove mercury from the gas stream. Parsons Infrastructure and Technology Group Inc. for US DOE/NETL considered the cost of removing mercury using a packed bed of 4 mm diameter activated carbon pellets in the 250 MWe Tampa Electric IGCC Demonstration Project on a gas stream of approximately 40°C and 26 atmospheres pressure[3]. Pellets were selected because the pressure drop through the bed was approximately 20% less than the pressure drop experienced through the alternative bed consisting of 0.4 mm carbon particles. Replacement of the carbon was listed to occur every 18 months, not because the carbon is spent but due to the pressure drop across the bed and the buildup of water and other contaminants within the bed. The cost of a bed operating at atmospheric pressure is noted to be an order of magnitude more expensive due the increased size of the equipment and the requirement for more activated carbon. Table 8 shows the costs, in 2002, presented in the Parsons Infrastructure and Technology Group Inc. study.

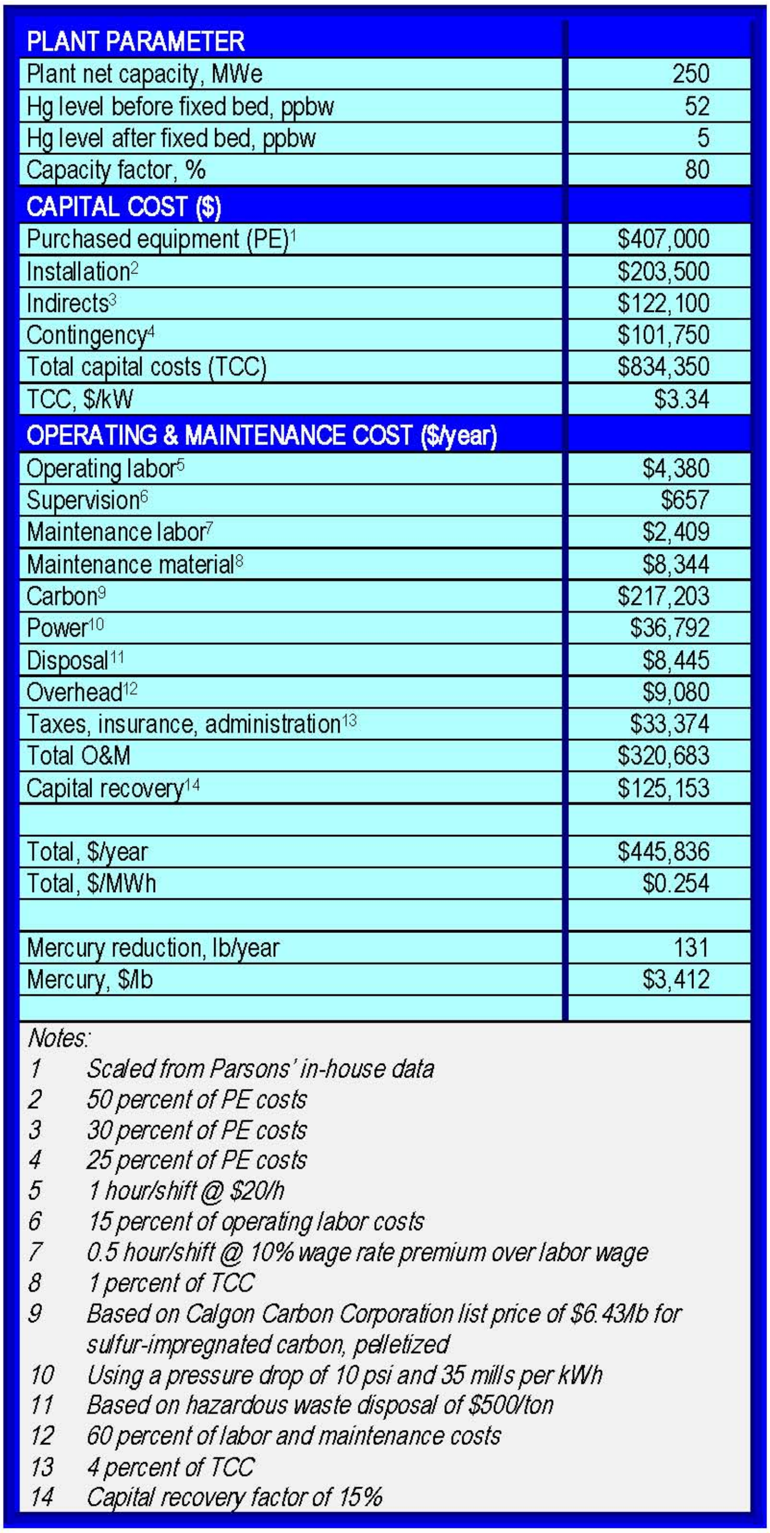
Other forms of carbon, such as cloth, straight and corrugated plates and honeycomb structure can also be installed within the flue ducts to capture mercury but the costs associated with these materials are dependent on the manufacture process and can vary significantly.

To ascertain the impact of mercury removal from the fluegas on the costs of oxyfuel in this study, costs for flowsheet C1 were calculated without any activated carbon as noted above, and with the activated carbon in a packed bed. Two positions are possible for the packed bed:

1. after the caustic scrubber and before compression, or
2. after compression, which will reduce the size of the bed required.

The costs for the activated carbon bed determined here were based on the costs presented by Parsons Infrastructure and Technology Group Inc. in Table 8. These costs were initially changed to 2009 Australian dollar values, assuming an 0.85US$ exchange rate and a 24% per annum inflation rate. These values were then adjusted based on the flowrates determined for C1 in Section 7 and pressures expected during compression. The costs determined are presented in Section 10.4

Table 8: The costs determined for a pressurised packed bed of activated carbon for mercury removal in the 250 MWe Tampa Electric IGCC Demonstration Project in 2002.[3]



PLANT PARAMETER	
Plant net capacity, MWe	250
Hg level before fixed bed, ppbw	52
Hg level after fixed bed, ppbw	5
Capacity factor, %	80
CAPITAL COST ($)	
Purchased equipment (PE)[1]	$407,000
Installation[2]	$203,500
Indirects[3]	$122,100
Contingency[4]	$101,750
Total capital costs (TCC)	$834,350
TCC, $/kW	$3.34
OPERATING & MAINTENANCE COST ($/year)	
Operating labor[5]	$4,380
Supervision[6]	$657
Maintenance labor[7]	$2,409
Maintenance material[8]	$8,344
Carbon[9]	$217,203
Power[10]	$36,792
Disposal[11]	$8,445
Overhead[12]	$9,080
Taxes, insurance, administration[13]	$33,374
Total O&M	$320,683
Capital recovery[14]	$125,153
Total, $/year	$445,836
Total, $/MWh	$0.254
Mercury reduction, lb/year	131
Mercury, $/lb	$3,412

Notes:

1 Scaled from Parsons' in-house data
2 50 percent of PE costs
3 30 percent of PE costs
4 25 percent of PE costs
5 1 hour/shift @ $20/h
6 15 percent of operating labor costs
7 0.5 hour/shift @ 10% wage rate premium over labor wage
8 1 percent of TCC
9 Based on Calgon Carbon Corporation list price of $6.43/lb for sulfur-impregnated carbon, pelletized
10 Using a pressure drop of 10 psi and 35 mills per kWh
11 Based on hazardous waste disposal of $500/ton
12 60 percent of labor and maintenance costs
13 4 percent of TCC
14 Capital recovery factor of 15%

6.3 GENERAL ASSUMPTIONS

The following assumptions have been made:

1. The basis for the calculations is an oxy-fuel power plant with 550 MWe power output assuming overall 30% efficiency.
2. The numbers of operators for an entire day for SDA, WFGD, CPU and PJFF has been assumed to be 12, 12, 18 and 1 respectively, while no additional operators are required for the carbon bed.
3. The labour costs have been assumed to be $60/hr.
4. In the ASPEN Plus models, it is assumed that no SO_2 is removed in the FF and condensers.
5. It is assumed that 25% of Hg is removed in FF.
6. The plant working days is assumed to be 320 days per year.
7. Gypsum is a by-product of the FGD. Incomes from its sale or costs for disposal are not considered in the current calculations.
8. The cost for neutralizing the condensate is not considered.
9. The costs for auxiliary power and water for FGDs are not considered for all flowsheets as they contribute to ~1-5% of the total operating costs[57].
10. A currency conversion factor of 0.85 USD = 1AUD as per 2010 price was used.
11. Instalment costs on capital investment have been calculated for 25 years without considering any interest rate for simplicity.

7 ASPEN PLUS PROCESS SIMULATION MODEL DEVELOPMENT

Process simulations for different oxyfuel flowsheets were made using ASPEN Plus v7.3. The process simulation model for Callide power plant-IHI design is shown in Figure 17. The major unit operations such as the combustion in furnace (incl. steam generator), air pre-heater (APH1 and APH 2), fabric filter (FF) and direct contact cooler polishing scrubber (DCCPS) are highlighted with red blocks in the figure. Remaining flowsheets are presented in Appendix B.

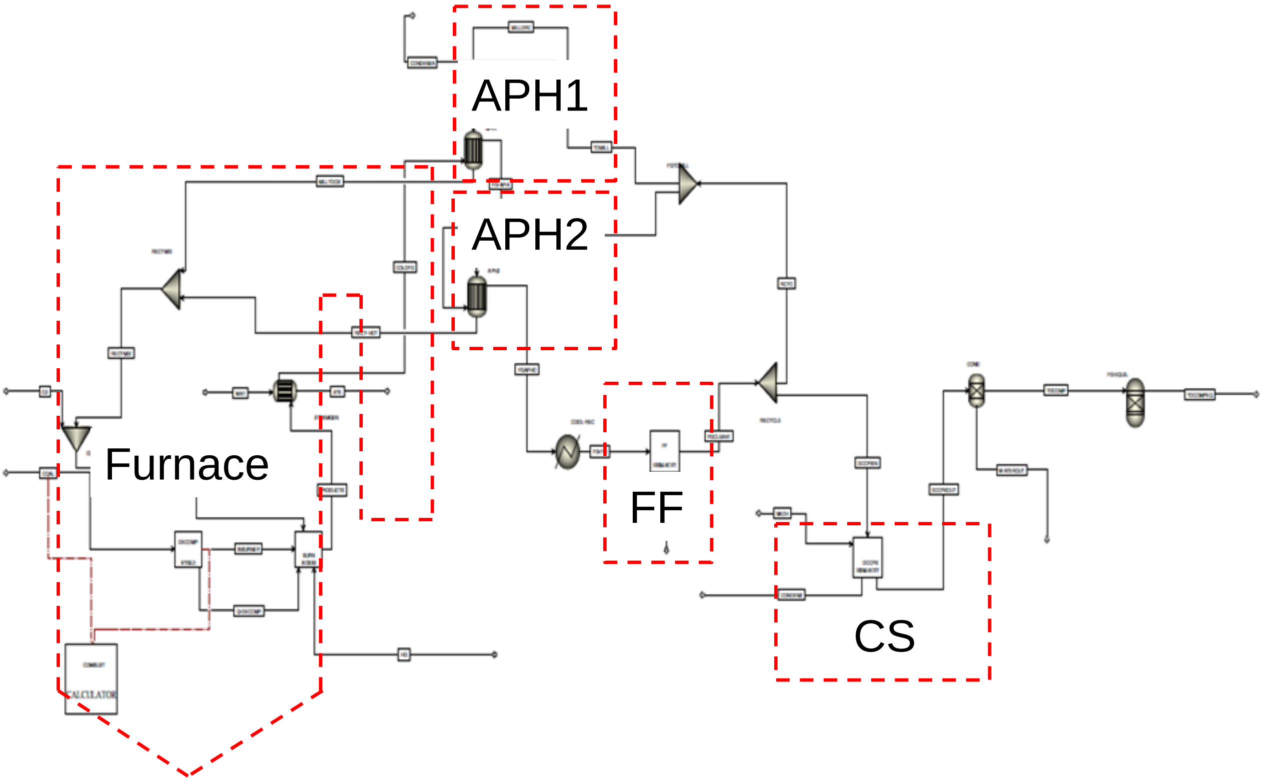


Figure 17: The ASPEN Plus V7.3 process simulation models.

Table 9 provides information on the reactor type, process conditions and purpose of the different unit operations used in the oxyfuel process. The modeling of a combustion process was conducted using RYield and RGibbs models (Aspen plus, 2010). Since coal is a non-conventional component according to the definition of Aspen Plus, it shall be decomposed into constituent elements by the RYield block before it is sent to the RGibbs block. The downstream unit operations include air pre-heater, fabric filter and FGDs. The ideal property method was applied which uses both Raoult's and Henry's Law.

Table 9: Specification and description of unit operation blocks

Unit operation	Reactor type	Conditions	Purpose
Decomposition	RYield	P= 1 bar, T = 75^{o}C	Decomposition of coal to conventional components
Burn	RGibbs	P =1 bar	Combustion
Combustion	Calculator		Calculate the combustion products and properties
Steam Generator	Heat exchanger	HSOT =350^{o}C	To generate steam
APH1	Heat exchanger	CSOT = 130^{o}C	To pre-heat the recycle stream
APH2	Heat exchanger	CSOT = 200^{o}C	
FF	FabFl	Eff. = 99.9%	To remove solids (fly ash)
SDA	RStoic	P = 1 bar, T = 150^{o}C Eff. = 94%	To remove SOx
WFGD	RStoic	P = 1 bar, T = 60^{o}C Eff. = 95%	
DCCPS	RStioc	P= 1 bar, T= 50 ^{o}C, Eff. = 95%	

The reaction scheme, type of chemical reactions and its values used in the process simulations are highlighted in Table 10. All the reaction types used in ASPEN Plus are fractional conversion. The conversion values are defined based on the efficiency described for each unit operation in the existing literature.

Table 10: Specifications of the reactions in different FGD unit operation blocks

Unit operation	Reaction scheme	Type	Value
DCCPS	$O_2 + 4\ NaOH + 2\ SO_2 \rightarrow 2\ Na_2SO_4 + 2\ H_2O$	Fractional conversion	0.95
SDA	$2\ SO_2 + 2\ CaO + O_2 + 4\ H_2O \rightarrow 2\ CaSO_4{*}2H_2O$	Fractional conversion	0.94
	$CaO + SO_3 + 2\ H_2O \rightarrow CaSO_4{*}2H_2O$		
WFGD	$2\ SO_2 + 2\ CaCO_3 + O_2 + 4\ H_2O \rightarrow 2\ CaSO_4{*}2H_2O + 2\ CO_2$	Fractional conversion	0.98
	$CaCO_3 + SO_3 + 2\ H_2O \rightarrow CaSO_4{*}2H_2O + CO_2$		

The main objective of the process simulations was to obtain inlet flue gas flow rates to the SOx removal unit operations as well as to identify reagent requirements. The process simulation results for the Callide plant design are tabulated in Table 11. The results for the remaining flowsheets are described in Appendix B.

Table 11: ASPEN Plus V7.3 process simulation results for flowsheet C1.

ASPEN PLUS RESULTS								
Stream	**recyc. FG**	**recy. FG partly dried**	**FG filter inlet**	**FG Filter Outlet**	**CS-Inlet stream**	**CS-Outlet stream**	**CS-condensate**	**Dry FG to compression**
	RCYC	RECYMIX	FG-FI-I	FGCLEAN1	DCCPSIN	DCCPSOUT	CONDENS	TOCOMP
Substream: MIXED								
Mole Frac								
O2	0.03115194	0.03196459	0.03115194	0.03115194	3.12E-02	0.03513113	3.59E-05	0.03991814
CO2	0.71542742	0.73391611	0.71542742	0.71542742	0.71542742	0.82138302	0.00752245	0.93330565
SO2	0.00119108	0.00122219	0.00119108	0.00119108	0.00119108	8.85E-06	1.03E-06	1.01E-05
H2O	0.23173889	0.21187175	0.23173889	0.23173889	0.23173889	0.11992066	0.98284126	0
CACO3	0	0	0	0	0	0	0	0
CA++	0	0	0	0	0	0	0	0
CO3--	0	0	0	0	0	0	0	0
H+	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0	0
OH-	0	0	0	0	0	0	0	0
SO3	7.99E-07	8.19E-07	7.99E-07	7.99E-07	7.99E-07	8.19E-07	6.31E-07	9.31E-07
H2SO3	0	0	0	0	0	0	0	0
H2SO4	0	0	0	0	0	0	0	0
HCO3-	0	0	0	0	0	0	0	0
SO4--	0	0	0	0	0	0	0	0
HSO4-	0	0	0	0	0	0	0	0
SO3--	0	0	0	0	0	0	0	0
HSO3-	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0
CASO4*2W	0	0	0	0	0	0	0	0
CAO	0	0	0	0	0	0	0	0
CA(OH)+	0	0	0	0	0	0	0	0
CASO3	0	0	0	0	0	0	0	0
CASO4	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0	0
N2	4.30E-03	4.41E-03	4.30E-03	4.30E-03	4.30E-03	0.00494084	4.19E-06	0.00561408
NO	3.31E-04	3.40E-04	3.31E-04	3.31E-04	3.31E-04	0.00038107	1.41E-07	0.000433
S	2.67E-09	2.74E-09	2.67E-09	2.67E-09	2.67E-09	2.10E-14	1.96E-08	0
H2	0.00101705	0.00104361	0.00101705	0.00101705	0.00101705	0.00116934	8.21E-08	0.00132868
CL2	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0.00E+00	0	0	0
C	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0	0
CO	1.48E-02	1.52E-02	1.48E-02	1.48E-02	1.48E-02	1.71E-02	1.29E-05	0.01938944
HG	1.84E-09	1.89E-09	2.46E-09	1.84E-09	1.84E-09	2.12E-09	0	2.41E-09
NO2	0	0	0	0	0	0	0	0
NAOH	0	0	0	0	0	0	0.00091251	0
NA2SO4	0	0	0	0	0.00E+00	1.25E-46	0.00866893	0
Total Flow kmol/hr	0.13168739	0.12833563	0.18626222	0.18626222	0.05457483	0.04746664	0.0074491	0.04177441
Total Flow kg/hr	4.90957196	4.84853199	6.94423193	6.94423191	2.03465995	1.90171651	0.14381857	1.79916937
Total Flow l/min	76.967143	74.2678695	107.790009	108.864417	31.8972743	20.9661201	0.00254853	18.4518491
Temperature K	423.15	423.15	423.15	423.15	423.15	322.975931	323.190391	322.975931
Pressure atm	0.99013077	1	1	0.99013077	0.99013077	1	1.00E+00	1
Vapor Frac	1	1	1	1	1	1	0	1
Liquid Frac	0	0	0	0	0	0	1	0
Solid Frac	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0	0
Enthalpy cal/mol	-79965.993	-80565.525	-79965.993	-79965.993	-79965.993	-84353.676	-70150.113	-88005.045
Enthalpy cal/gm	-2144.8943	-2132.4861	-2144.8943	-2144.8943	-2144.8943	-2105.4589	-3633.435	-2043.3646
Enthalpy cal/sec	-2925.1425	-2872.0631	-4137.401	-4137.401	-1212.2585	-1112.2183	-1.45E+02	-1021.2109
Entropy cal/mol-K	**3.117847**	**3.30163902**	**3.09815093**	**3.117847**	**3.117847**	**1.64690582**	**-36.957578**	2.39968124
Entropy cal/gm-K	0.0836287	0.08739096	0.0831004	0.0836287	0.0836287	0.04110659	-1.91E+00	0.05571753
Density mol/cc	2.85E-05	2.88E-05	2.88E-05	2.85E-05	2.85E-05	3.77E-05	4.87E-02	3.77E-05
Density gm/cc	0.00106313	0.00108807	0.00107372	0.00106313	0.00106313	0.00151173	0.94053263	0.0016251

The inlet SO_2 rate (Kg/KJ) and reagent requirement for SOx removal unit obtained from the ASPEN calculations are tabulated in Table 12 for all studied flowsheets. These parameters are used to estimate capital costs and fixed and variable O&M costs in Section 10.

Table 12: Inlet SO_2 for sulphur removal unit operations in each flowsheet calculated using ASPEN Plus V7.3

Flowsheet	Sulphur in Coal (wt%)	SOx removal unit operation	SO_2 rate (Kg/KJ)	Reagent type	Reagent requirement (T/T of coal)
C1	0.22	CS	2.31217E-07	NaOH	0.01087513
FL1	0.22	SDA	2.31952E-07	Lime	0.00510675
		DCCPS	1.39171E-08	NaOH	0.00027512
FL2	1.5	SDA	1.64135E-06	Lime	0.0352996
		DCCPS	5.71188E-08	NaOH	0.00127002
FL3	1.5	WFGD	1.6146E-06	Limestone	0.04635238
		DCCPS	1.7486E-08	NaOH	0.00034219
FL4	3.5	WFGD	3.7337E-06	Limestone	0.10815563
		DCCPS	7.46737E-08	NaOH	0.0016296

8 EXPERIMENTAL OBSERVATIONS ON SOX AND HG

As discussed earlier, the SO_x concentrations during oxy-fuel combustion are generally higher compared to conventional air firing. The higher SO_x concentrations, particularly SO_3 in combination with high concentrations of water in the recycled stream, increase the acid dew point temperature of the system, thereby increasing allowable flue gas temperatures and reducing the thermal efficiency of a power plant.

To investigate these impacts in detail, experiments were carried out on a 20 kW once-through combustion rig at The Institute of Combustion and Power Plant Technology (IFK) at The University of Stuttgart, firing coals in both air and oxyfiring. A novel methodology to investigate oxy-fuel process configurations was employed, for which “impurities” were injected with the oxidant gas into the once-through reactor to simulate different extents of oxy-fuel recycle gas treatment. Three Australian coals (A, B and C) were studied, shown in Table 13, which were previously tested in the IHI’s furnace at Aioi. The detailed results can be found elsewhere[10].

As shown in Table 14, different gas compositions were investigated: Along with combustion in air, oxy-fuel combustion with 6 different concentrations of H_2O, NO, SO_2 and Hg^0 and one “clean oxy-fuel” configuration (labelled by OC) simulating removal of 100% of all impurities was completed. Oxy-fuel experiments with impurities included simulations of a 20% H_2O removal rate and a constant NO concentration (approx. 1000 ppm). SO_2 removal rates of approx. 50%, 20% and 0% and Hg removal rates of approx. 80%, 50% and 0% were investigated.

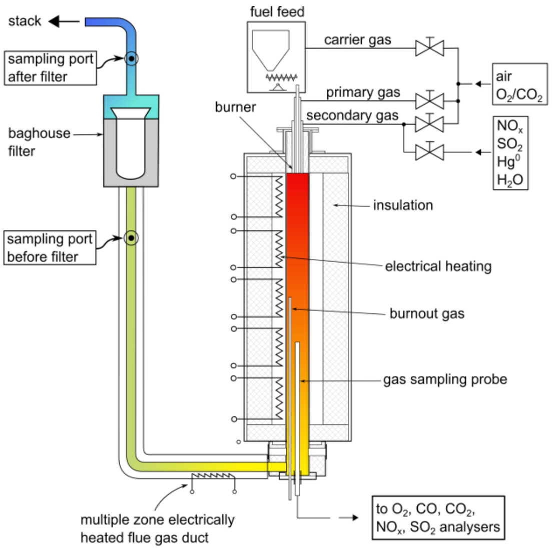
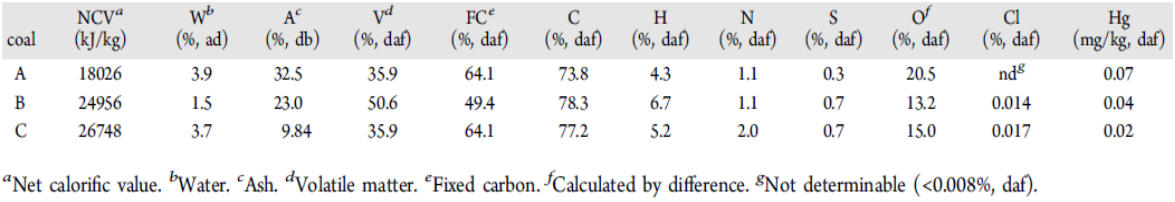


Figure 18: Schematic of the experimental combustion rig used for air and oxy-fuel investigations at IFK.

The experiments are indexed by a alpha-numerical code enabling an unambiguous identification of experimental conditions, e.g. the index C-O2S5H indicates an experiment with coal C (index: C) under oxy-fuel conditions (index: O) with simulated removal of 20% of the SO_2 (index: 2S) and 50% of the Hg (index: 5H) from the recycled gas. Coals A, B and C (shown in Table 13) were tested under air-firing, clean oxy-fuel and O2S5H conditions and coal A was investigated with 5 additional oxy-fuel settings. Slight deviations from the desired removal rates can be observed, but experiments with different coals and similar removal rates are considered comparable.

Table 13: Net Calorific Value (NCV), Proximate, Elemental, Cl, and Hg Analyses of the Investigated Australian Coals. (Percentages on a Weight Basis)



coal	NCV[a] (kJ/kg)	W[b] (%, ad)	A[c] (%, db)	V[d] (%, daf)	FC[e] (%, daf)	C (%, daf)	H (%, daf)	N (%, daf)	S (%, daf)	O[f] (%, daf)	Cl (%, daf)	Hg (mg/kg, daf)
A	18026	3.9	32.5	35.9	64.1	73.8	4.3	1.1	0.3	20.5	nd[g]	0.07
B	24956	1.5	23.0	50.6	49.4	78.3	6.7	1.1	0.7	13.2	0.014	0.04
C	26748	3.7	9.84	35.9	64.1	77.2	5.2	2.0	0.7	15.0	0.017	0.02

[a]Net calorific value. [b]Water. [c]Ash. [d]Volatile matter. [e]Fixed carbon. [f]Calculated by difference. [g]Not determinable (<0.008%, daf).

Table 14: List of all experimental settings and simulated removal rate.

Index	Coal	Experimental settings				
		air/oxy	Simulated capture rates [%]			Simulated NO conc. in rec. flue gas [ppm, dry]
			H_2O	SO_2	Hg	
A-A	Coal A	Air	-	-	-	-
A-OC		Oxy28 – clean	100	100	100	-
A-O2S5H		Oxy28	23	23	44	962
A-O5S5H		Oxy28	23	52	44	962
A-O2S8H		Oxy28	23	23	79	962
A-O0S5H		Oxy28	23	4	44	962
A-O2S0H		Oxy28	23	23	7	962
A-O0S0H		Oxy28	23	4	7	962
B-A	Coal B	Air	-	-	-	-
B-OC		Oxy28 – clean	100	100	100	-
B-O2S5H		Oxy28	24	24	51	949
C-A	Coal C	Air	-	-	-	-
C-OC		Oxy28 – clean	100	100	100	-
C-O2S5H		Oxy28	23	23	62	959

SO_x emissions and conversion of SO_2 to SO_3 were measured and compared to existing literature. Capture of SO_x in a baghouse filter has also been evaluated. In Figures 19 and 20 the mean concentrations of SO_3 measured before and after the experimental rig's baghouse filter are shown. The error bars in the figures represent the range between measured minimum and maximum SO_3 concentrations. As expected, in all experiments SO_3 is captured on the baghouse filter of the combustion rig and concentrations after the filter are considerably lower than those before the filter. For coals A and C barely any SO_3 was detected under air firing conditions and coal C generated only measurable SO_3 concentrations in the oxy-fuel setting with SO_2 injection.

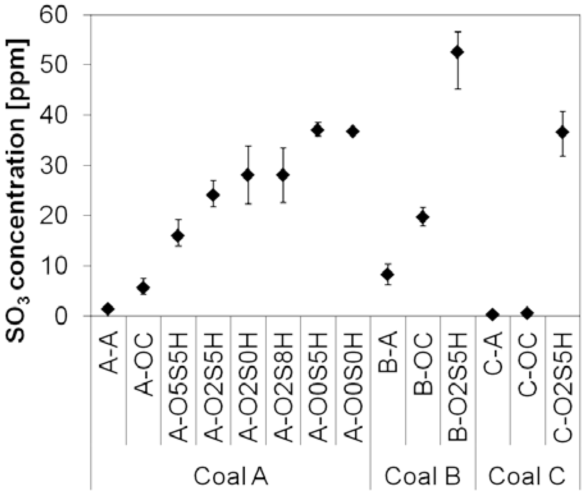


Figure 19: SO_3 concentrations measured before the baghouse filter (error bars indicate measured min. and max. concentrations).

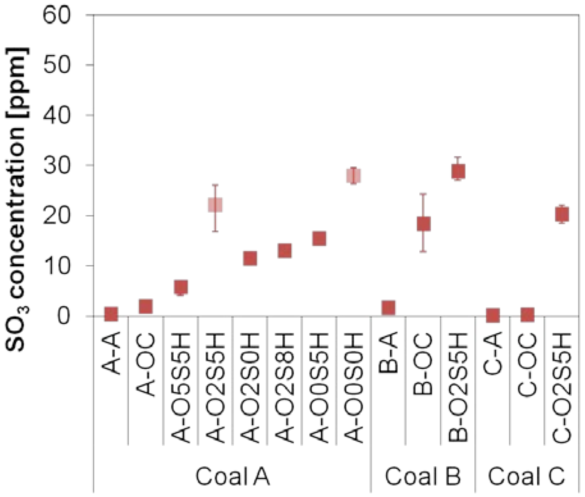


Figure 20: SO_3 concentrations measured after the baghouse filter (error bars indicate measured min. and max. concentrations)

Based on measured SO_3 concentrations, the H_2SO_4 dew point temperatures were calculated according to ZareNehzhad's correlation[58] for coal A. In Figure 21, differences between the acid dew point of the experiments O5S, O2S and OC and the experiment without SO_2 removal (O0S) are presented. At low acid dew points, the oxidant pre-heater can be safely operated at lower temperatures, which can increase a plant's efficiency. Based on these acid dew points, the flue gas composition and the NCV of coal A, equivalent efficiency gains were calculated for oxy-fuel combustion without SO_2 removal, according to Equation 2.

$$\Delta\eta = \frac{n_{FG,O0S}c_{p,FG,O0S}T_{H2SO4,O0S} - n_{FG,l}c_{p,FG,l}T_{H2SO4,l}}{NCV}$$

Equation 2

Where $n_{FG,O0S}$ is molar flow rate of flue gas at zero SO2 removal, $c_{p,FG,O0S}$ is specific heat of the flue gas at zero SO_2 removal, $T_{H2SO4,O0S}$ is the acid dew point temperature at zero sulphur removal, $n_{FG,l}$ is the molar flow rate of the flue gas for reference case, $c_{p,FG,l}$ is the specific heat of the flue gas for reference case and $T_{H2SO4,l}$ is the acid dew point temperature of the flue gas for reference case.

A minor efficiency benefit of 0.02% appears to be provided from partial recycle gas desulfurization of 20% and a considerable benefit of 0.14% for 50% desulfurization. The clean oxy-fuel configuration has a relatively high efficiency benefit potential of 0.47%, due to the combined effect of low SO_3 and low H_2O concentrations. It should be noted that the efficiency benefit of the clean oxy-fuel configuration is of limited significance, since a complete H_2O and SO_2 removal from recycled flue gas would require cooling of the gas to temperatures far below the H_2SO_4 dew point. With such improved process efficiency with SO_3 capture in the bag filter by fly ash, the impact on capital costs and operating and maintenance costs is further calculated and described in Section 10.

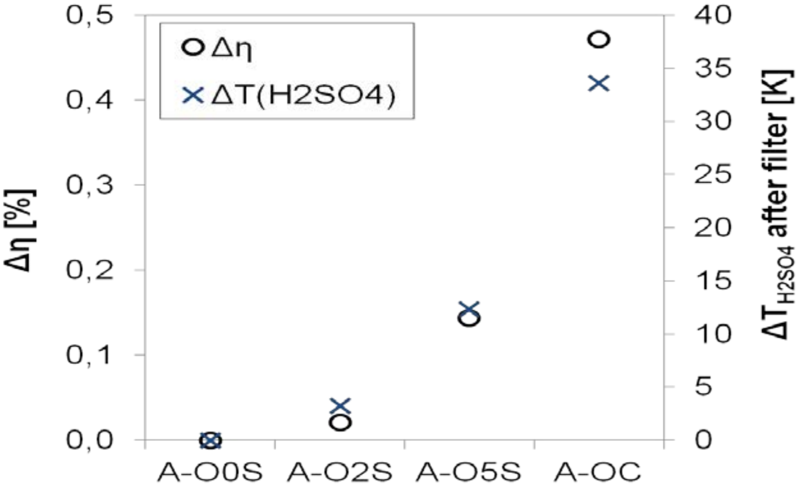


Figure 21: Differences in acid dew point temperatures and respective process efficiencies for coal A oxy-clean (OC) and 20% (O2S) and 50% (O5S) SO_2 removal settings, compared to an uncleaned recycle configuration (O0S).

Capture of elemental and ionic forms of Hg in a baghouse filter has also been investigated. Measured Hg concentrations increased when switching from air to oxy-fuel operation for all investigated coals and oxy-fuel settings, even when no Hg^0 is injected to the oxidant gas (not shown). Hg capture by fly ash in the baghouse filter has been found to reduce the Hg emissions considerably, as shown in Figure 22. Reduction rates of between 10 and 50% were observed, depending on the coals and experimental settings. The results are described more in detail elsewhere[27]. Overall, this will also reduce the capital and operating costs associated with the activated carbon bed technology included in the current report. This is further discussed in the Section 10.

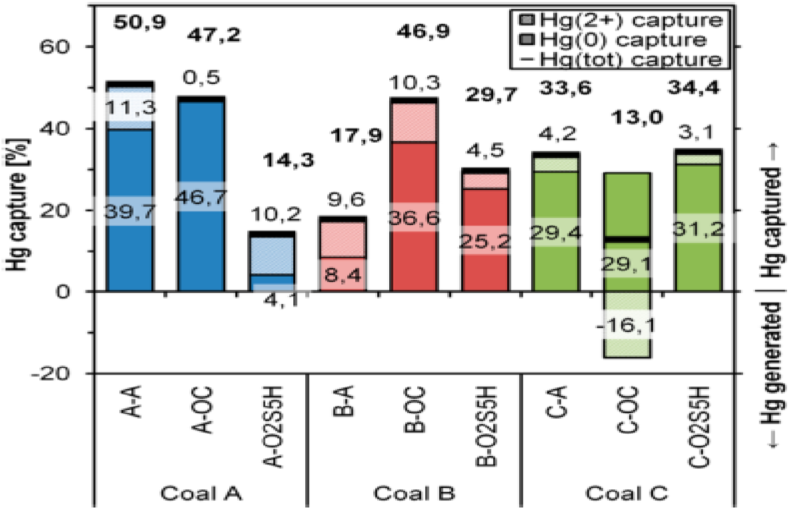


Figure 22: The mercury capture measured during experiments at IFK.

For all process simulations, SO_3 and Hg capture rates were established in the fabric filter based on IFK data.

9 S&L COSTING MODELS

Sargent and Lundy L.L.C prepared a series of reports for the National Lime Association[59], Perrin Quarles Associates Inc.[48, 49, 60] and Systems Research and Applications Corporation[50]. In these reports a series of models are developed that predict the capital cost and fixed and variable operating and maintenance costs of a spray drier absorber flue gas desulfuriser (SDA FGD)[49], a wet flue gas desulfuriser (WFGD)[48], and a bag house containing fabric filters[50] for particulate control. The models provide costs in 2009 dollars that do not include escalation.

Each of these models were used in this study to determine the impact of coal sulphur content on the capital and operating costs of selected oxyfuel plant designs and compared with the published costs of oxyfuel plants by EPRI[1] and NETL[2]. Using mass balances completed for combustion and flue gas desulfurisation, discussed in Section 7, each unit could be sized and then costed by the models described below. Symbols listed in all the equations within this section are listed in Table 15.

Table 15: The symbols used in calculation of the SDA FGD, WFGD and Fabric Filter capital costs.

Symbol	Variable	Units	Comments
A	Gross plant size	MW	550MW plant used, impact of plant size considered in Appendix D (Sensitivity analysis)
B	Retrofit factor		Value of 1 used
D	SO_2 rate	lb/MMBtu	
E	Bag house air to cloth ratio		4.0 used
F	Coal factor		Bituminous=1, Semi-bituminous=1.05 Lignite=1.07, 1.05 used in this study.
G	Gross heat rate/1000	Btu/kWh	
T	Operating labour rate	\$/hr	\$60/hr used, includes all benefits.
K	Lime, limestone or NaOH rate	ton/hr	Calculated by Aspen modelling, Chapter 6.
L	Particulate removal cage cost	\$/cage	\$30 per cage used
N	Particulate removal bag cost	\$/bag	\$80 per bag used
P	Lime, limestone or NaOH cost	\$/ton	Lime cost of \$95/ton used Limestone cost of \$15/ton used NaOH cost of \$940/ton (100%) used

9.1 SPRAY DRIER ABSORBER FLUE GAS DESULFURISER (SDA FGD)

SDA FGD models described by Sargent and Lundy should not be applied to fuels with more than 3 lb SO_2/MMBTU (1.29g/MJ). To determine the costs associated with the SDA FGD used in flowsheets FL1 and FL2, the following models were used:

$$\boldsymbol{SDA\ FGD\ capital\ costs}(\$) = absorber\ costs + reagent\ prep\ unit\ \&\ waste\ handling + unit\ cost + balance\ of\ plant\ costs \qquad \textbf{Equation 3}$$

Balance of plant costs includes ID or booster fans, piping, ductwork, electrical etc.

$Absorber\ costs\ (\$) = 98000.A$, if the gross plant size, A, is >600MW or

$= 566000(A^{0.716})B(FG)^{0.6}\left(\frac{D}{4}\right)^{0.01}$, if A is ≤600MW **Equation 4**

$$reagent\ prep\ unit\ \&\ waste\ handling\ unit\ cost\ (\$) = 48700.A,$$

if A is >600MW, else

$$= 300000(A^{0.716})B(DG)^{0.2} \qquad \textbf{Equation 5}$$

$$Balance\ of\ plant\ costs\ (\$) = 129900.A, \text{ if } A \text{ is >600MW, else}$$
$$= 300000(A^{0.716})B(FG)^{0.4} \qquad \textbf{Equation 6}$$

$$\boldsymbol{SDA\ FGD\ Fixed\ O\&M\ costs}\ (\$) \\ = operating\ labour\ costs \\ + maintenance\ material\ \&\ labour\ costs \\ + administrative\ labour\ costs \qquad \textbf{Equation 7}$$

$$Fixed\ O\&M\ operating\ labour\ costs\ (\$) = \#\ of\ operators.2080\frac{T}{1000A} \qquad \textbf{Equation 8}$$

For SDA FGD the number of operators was assumed to be 12; 3 per shift.

$$Fixed\ O\&M\ maintenance\ material\ \&\ labour\ costs \\ = \frac{0.015(SDA\ FGD\ capital\ costs)}{1000.B.A} \qquad \textbf{Equation 9}$$

$$Fixed\ O\&M\ Administrative\ labour\ costs\ (\$) \\ = 0.03(Fixed\ O\&M\ operating\ labour\ costs + 0.4 \\ \times Fixed\ O\&M\ maintenance\ material\ \&labour\ costs) \qquad \textbf{Equation 10}$$

Variable Operating and maintenance costs were limited to the cost of lime. No waste disposal, auxiliary power or makeup water costs were included:

$$Variable\ O\&M\ costs\ (\$) = cost\ of\ lime = K\frac{P}{A} \qquad \textbf{Equation 11}$$

9.2 WET FLUE GAS DESULFURISER (WET FGD, WFGD)

The cost of the WFGD in flowsheets FL3 and FL4, the cost of the caustic scrubber in flowsheet C1 and the cost of the direct contact cooler/ polishing scrubber (DCCPS) of flowsheets FL1, FL2 , FL3 and FL4 were determined using the Wet FGD models outlined below.

$$\boldsymbol{Wet\ FGD\ capital\ costs}\ (\$) \\ = absorber\ costs + reagent\ prep\ costs + waste\ handling \\ + costs + balance\ of\ plant\ costs \qquad \textbf{Equation 12}$$

Balance of plant costs includes ID or booster fans, new wet chimney, piping, ductwork and minor waste water treatment.

$$Absorber\ costs\ (\$) = 550000(A^{0.716})B(FG)^{0.6}\left(\frac{D}{2}\right)^{0.02}$$ **Equation 13**

$$reagent\ prep\ costs\ (\$) = 190000(A^{0.716})B(DG)^{0.3}$$ **Equation 14**

$$Waste\ handling\ costs\ (\$) = 1000000(A^{0.716})B(DG)^{0.45}$$ **Equation 15**

$$Balance\ of\ plant\ costs\ (\$) = 1010000(A^{0.716})B(FG)^{0.4}$$ **Equation 16**

$$\boldsymbol{Wet\ FGD\ Fixed\ O\&M\ costs}\ (\$) = operating\ labour\ costs + maintenance\ material\ \&\ labour\ costs + administrative\ labour\ costs$$ **Equation 17**

$$Fixed\ O\&M\ operating\ labour\ costs\ (\$) = \#\ of\ operators.2080\frac{T}{1000A}$$ **Equation 18**

If the plant size (A) is greater than 500 MW, sixteen (16) operators are required for the Wet FGD, however if the plant size is less than 500 MW, only twelve (12) operators are required. For the DCCPS unit operations it was assumed no additional operators would be required as each plant also had additional operators for the SDA FGD or WFGD unit operations. The caustic scrubber of flowsheet C1 was assumed to require twelve operators for any size plant considered.

$$Fixed\ O\&M\ maintenance\ material\ \&\ labour\ costs = \frac{0.015(Wet\ FGD\ capital\ costs)}{1000.B.A}$$ **Equation 19**

$$Fixed\ O\&M\ Administrative\ labour\ costs\ (\$) = 0.03(Fixed\ O\&M\ operating\ labour\ costs + 0.4 \times Fixed\ O\&M\ maintenance\ material\ \&labour\ costs)$$ **Equation 20**

Wet FGD Variable Operating and maintenance costs were limited to the cost of limestone or NaOH for the DCCP's in flowsheets FL1, FL2, FL3 and FL4 and the caustic scrubber in flowsheet C1. No waste disposal, auxiliary power, makeup water or waste water treatment costs were included:

$$Variable\ O\&M\ costs(\$) = cost\ of\ limestone = K\frac{P}{A}$$

Equation 21

9.3 PARTICULATES REMOVAL

The cost of the baghouses required in each flowsheet was calculated using the models set out below. Flowsheet FL1 required two baghouses.

$$\boldsymbol{Baghouse\ capital\ costs}\ (\$) = E.B.(G^{0.81})$$

Equation 22

E represents a constant based on the bags air to cloth ratio. If the required air to cloth ratio is 6.0, E = 422, however if the air to cloth ratio required is 4.0, then E becomes 476. In this study, an air to cloth ratio of 4.0 was assumed to be required for all flowsheets.

Capital costs include the cost of equipment, installation, buildings, foundations and electrical installations required. It also includes a cost for difficulty in retrofitting to an established plant. Equipment costs include the costs of ID or booster fans, piping and ductwork.

$$\begin{aligned}\boldsymbol{Baghouse\ Fixed\ O\&M\ costs}\ (\$) \\ &= operating\ labour\ costs \\ &+ maintenance\ material\ \&\ labour\ costs \\ &+ administrative\ labour\ costs\end{aligned}$$

Equation 23

$Fixed\ O\&M\ operating\ labour\ costs\ (\$) = 0$, no additional operators are required to operate the baghouse

Equation 24

$$\begin{aligned}Fixed\ O\&M\ maintenance\ material\ \&\ labour\ costs\ (\$) \\ &= \frac{0.005(Baghouse\ capital\ costs)}{1000.B.A}\end{aligned}$$

Equation 25

$$\begin{aligned}Fixed\ O\&M\ Administrative\ labour\ costs\ (\$) \\ &= 0.03(0.4 \\ &\times Fixed\ O\&M\ maintenance\ material\ \&labour\ costs)\end{aligned}$$

Equation 26

$$\boldsymbol{Baghouse\ variable\ O\&M\ costs}\ (\$) = 0.005\left(\frac{N}{3}+\frac{L}{9}\right)$$

Equation 27

N and L represent the cost of the bags and cages required in the baghouse, respectively. Values of $80 per bag and $30 per cage were used in this study. No costs associated with ash disposal have been included.

Costs calculated by these models are discussed in Section 10.

10 RESULTS AND DISCUSSION

10.1 METHODOLOGY FOR DETERMINING PLANT COSTS, 550 MWE

To obtain an indication of the total expected costs for each of the flowsheets considered, without designing units and approaching suppliers for costs, the cost estimates produced by EPRI[1] and NETL[2] in their studies were used for the comparison. Each study looked at the costs associated with an oxy-fuel plant. EPRI's plant was a 750 MW with a CO_2 capture rate of 85-90%, which is compressed to 160 bar (16 MPa) after sulphur and NO_x is removed. NETL's plant was a 550 MW with CO_2 compressed to 2215 psia (15.7 MPa), 95°F after gas cleaning. After the costs provided by the two reports were stripped of any Engineering, Management, Process and project contingencies, adjusted to 2009 Australian dollars and determined on a cost per kW basis (Refer to Section 3), plant costs were assigned to selected flowsheets. Sulphur is a key feature of the costs. Increasing the coal sulphur content affects the overall design of the fluegas cleaning, and therefore the cost of the unit operations. The sorbent preparation and handling costs and the waste handling and disposal costs are also significantly affected by the amount of sulphur that needs to be treated. Though EPRI's plant was designed to burn a low sulphur coal, and NETL's plant was designed to burn a high sulphur coal, the total fluegas cleaning, purification and compression costs were very similar from the two reports. But EPRI provided a very low value for fluegas cleaning and a large value for purification and compression. This suggested that EPRI had included some of the fluegas cleaning in the purification section. The fluegas cleaning costs were removed from the estimates and replaced with the costs provided by the S&L models determined in Section 8 for each of the flowsheets C1-4, FL1-4, and the cost of purification and compression provided by NETL was used for all flowsheets, while other aspects that are not significantly affected by S, such as the boiler, air separation unit and the cooling water circuit, the costs of EPRI or NETL were used. Costs for sorbent handling and disposal were not included in the S&L model calculations and so sorbent handling from the NETL and EPRI were included here.

10.1.1 FLOWSHEETS FOR LOW SULPHUR COALS

EPRI costs were based on oxy-fuel combustion for a black Hunter Valley coal with a sulphur content of 0.43%, as listed in Table 16. The design coal for flowsheets C1, 2, 3 and FL1 have a sulphur content of 0.217% d.b., and as such should be easily handled by the equipment in the EPRI report. As the coal used in Flowsheet C1, 2, 3 and FL1 has a lower net calorific value, the amount of coal required to produce the same amount of energy will be larger, and the coal handling costs listed by EPRI were increased proportionally. EPRI have combined the ash handling costs with the sorbent handling costs. Flowsheet C1 will have higher ash handling because its coal has a significantly larger amount of ash, however, it contains approximately half the sulphur and so the sorbent handling costs will be halved approximately. Therefore no change was made to the ash and sorbent handling costs.

EPRI do not specifically state how the sulphur is removed from the gas stream. WFGD is expected to be used.

Table 16: Comparison of the design coals for each flowsheet considered in this report and the EPRI, NETL and estimated Futuregen2 design coals.

	Flowsheet C1,2,3, FL1	Flowsheet FL2, FL3	Flowsheet FL4	EPRI P3-2 EPRI	NETL	FutureGen 2
Design coal	Low S Qld Black coal	Med S	High S	Hunter Valley Black Coal	Illinois No. 6	Blend of Illinois No 6 & PRB coal*
Moisture	3.88 %, a.d.	3.88 %, a.d.	3.88 %, a.d.	7.5	11.12% a.r.	
FC	43.3 %, d.b.	43.3 %, d.b.	43.3 %, d.b.		49.72% d.b.	
VM	24.24 %, d.b.	24.24 %, d.b.	24.24 %, d.b.		39.37 %d.b.	
Ash	32.46 %, d.b.	32.46 %, d.b.	32.46 %, d.b.	21.2	10.91% d.b.	10.34% d.b.*
Carbon	49.83 %, d.b.	49.83 %, d.b.	49.83 %, d.b.	60.18	71.73% d.b.	
Hydrogen	2.91 %, d.b.	2.91 %, d.b.	2.91 %, d.b.	3.78	5.07 d.b.	
Nitrogen	0.71 %, d.b.	0.71 %, d.b.	0.71 %, d.b.	1.28	1.41% d.b.	
Chlorine	0 %, d.b.	0 %, d.b.	0 %, d.b.	0	0.33% d.b.	
Sulfur	0.217 %, d.b.	1.5 %, d.b.	3.5 % d.b.	0.43	2.825 d.b.	2 % d.b.
Oxygen	13.87 %, d.b.	12.587 %, d.b.	12.587 %, d.b.	5.63	7.74% d.b.	
Net Calorific Value	18026 kJ/kg	18026 kJ/kg	18026 kJ/kg	23840 kJ/kg	26171 kJ/kg	24091 kJ/kg*

Note: EPRI no basis given. *FutureGen has not released the name of the Powder River Basin coal and so the ash content and net calorific value used in the calculations have been estimated from a typical PRB coal.

10.1.2 FLOWSHEETS FOR HIGH SULPHUR COALS

The NETL design coal was Illinois No. 6, a bituminous coal with 2.825% d.b. sulphur. It has lower ash content than the design coal for FL4, allowing higher fixed carbon and volatile contents which results in a higher calorific value. The design coal for FL4 has 3.5% d.b. sulphur, which will significantly increase the cost of gas cleaning. To complete the comparison, the cost of coal prep and feeding systems was increased proportionally with the increased amount of coal required to attain the same energy. As the ash content of the coal is double that of NETL's design coal and the sulphur content is 24% greater than the NETL coal, the cost of ash handling and spent sorbent handling was assumed to be double that of the NETL plant.

Futurgen 2 is expected to use a blend of Illinois No. 6 coal and a Powder river Basin coal. The Powder River Basin coal has not been named at this date, so a "typical" Powder River

Basin coal was selected to provide specifications for the design and costings outlined here, as shown in Table 16. The FutureGen 2 blend has a sulphur content of 2% d.b., a calorific value between the EPRI and NETL coals and ash content similar to the NETL coal. The costs of the FutureGen 2 plant are based on the higher sulphur (2.8% d.b.) NETL plant costs, with unit operations in the fluegas cleaning circuit costed using S&L models (See Chapter 8), as per the other flowsheets. Gas flowrates and sulphur loadings were based on the gas flows calculated for FL2, but adjusted for FutureGen 2's coal specifications. Costs for FutureGen2's circulating dry scrubbing system (CDS) are not available for a 550 MW plant. Several units may be required. Therefore the S&L SDA model was used to provide a basic capital cost which was multiplied by 1.4 to give an estimate of the more complex CDS. The cost of the coal handling, preparation and feeding systems were adjusted slightly to account for the slightly lower calorific value of the coal blend.

10.2 TOTAL PLANT COSTS

10.2.1 CAPITAL COSTS

Table 17 shows the plant capital costs for flowsheets C1 and FL4 and FutureGen 2 compared with the EPRI and NETL cost estimates. Even though the capital costs per kilowatt for the EPRI and NETL costs are very similar, the costs associated with increasing sulphur content appear to increase the capital costs per kilowatt, as shown by flowsheets C1, FL1, FutureGen 2 and FL4.

The S&L model suggests the sulphur content of the coal significantly impacts on the cost of the plant, with a 23% increase in total plant cost when the coal is changed from a low sulphur coal to a high sulphur coal and the flowsheet changes accordingly (C1 to FL4). The flue gas cleanup up increases from 7.5% of the plant costs for flowsheet C1 to 22.3% for FL4. However, the NETL cost, based on a high sulphur coal is not significantly different to the cost predicted by flowsheet FL1, which is a low sulphur coal plant. The difference in the costs between FL4 and the NETL plant is due to the significant difference in cost of flue gas cleanup in the two plants, and the much higher ash content, the higher sulphur content and the markedly lower calorific value of the coal used in FL4.

Table 17: Capital cost estimates for unit operations for flowsheets C1, FL1 and FL4, and FutureGen 2.

	C1	FL1	EPRI	FutureGen 2	FL4	NETL	Costs for C1, FL1, FL4 & FutureGen calculated from:
	2009 AUD/kW	2009 AUD/kW	2009 AUD/kW	2009 AUD/kW	2009 AUD/kW	2009 AUD/kW	
Coal handling system	129.8	129.8	104.3	127.9	154.4	117.8	EPRI/NETL costs changed based on coal CV
Coal prep and feed systems	44.7	44.7	36.0	60.8	73.4	56.0	EPRI/NETL costs changed based on coal CV
Feed water and misc BOP systems	159.2	159.2	159.2	233.8	233.8	233.8	EPRI/NETL costs
PC boiler and accessories	865.3	865.3	865.3	958.9	958.9	958.9	EPRI/NETL costs
Air separation unit	652.8	652.8	652.8	717.4	717.4	717.4	EPRI/NETL costs
Flue gas clean up	252.3	490	620.7	727	974.8	399.4	S&L models
Purification and Compression	287.5	287.5	25.8	287.5	287.5	287.5	NETL cost
HRSG, ducting and stack	40.4	40.4	40.4	72.3	72.3	72.3	EPRI/NETL costs
Steam turbine generator	475.7	475.7	475.7	374.8	374.8	374.8	EPRI/NETL costs
Cooling water systems	114.5	114.5	114.5	110	110	110	EPRI/NETL costs
ASH/spent sorbent handling	46.3	46.3	46.3	39.6	79.3	39.6	NETL costs changed based on coal ash & S
Access electric plant	238.1	238.1	238.1	261.8	261.8	261.8	EPRI/NETL costs
I&C	48.5	48.5	48.5	69.5	69.5	69.5	EPRI/NETL costs
Total	**3355**	**3593**	**3427**	**4042**	**4368**	**3699**	

10.2.2 FIXED OPERATING AND MAINTENANCE COSTS

Calculation of the fixed O & M cost was not described in detail in the EPRI report. A basic labour rate of $AUS40/hr was used with an associated labour burden of 30%. Administrative and support labour was calculated as 25% of the operation and maintenance labour. However the number of operators or basis for calculating the maintenance costs were not provided. A total value of $60.1/kWh was presented, but no breakdown was given. Therefore only the costs provided by NETL can be compared with the costs predicted using S&L's models, however NETL's costs were adjusted to use the same base labour rate as the other calculations, shown in Table 18. NETL assumed a total of 14 operators per shift, the number also assumed for flowsheets FL3 and FL4, while flowsheets C1 and FL1 and Futuregen2 were expected to require 13 operators per shift. Maintenance costs were calculated assuming they are 1.5% of the installed capital costs of the plant[61], as per the S&L model. This is significantly more than the cost provided by NETL ($17.9/kWyr).

Table 18: The Fixed operating and maintenance costs produced by NETL and costs calculated using S&L models for flowsheets C1, FL1, FutureGen 2 and FL4

		NETL	**NETL Adjusted**	**C1**	**FL1**	**FutureGen 2**	**FL4**
Labour rate	$/hr	43	60	60	60	60	60
Operators per shift		14	14	13	13	13	14
Operator labour	M$/yr	5.261	6.989	6.490	6.490	6.490	6.989
Operator labour	$/kWyr	9.6	12.7	11.8	11.8	11.8	12.7
Maintenance cost	$/kWyr	17.9	55.5	50.3	53.9	60.6	65.5
Admin. and support labour	$/kWyr	6.9	1.05	0.96	1.00	1.08	1.17
Total	**$/kWyr**	**34.3**	**78.4**	**71.4**	**75.5**	**73.5**	**90.0**

The cost of administrative and support staff calculated using the S&L model appears very low and is much smaller than the cost provided by NETL. The S&L model is:

$$\text{Additional administrative labour costs} = 0.03 \times \left(\text{Operating labour costs} + 0.4 \ \text{Maintenance, materials and additional labour} \right),$$

which also results in a value much lower than the method often employed of 15% of the total operating and maintenance costs[61].

As the number of operators required for each of the flowsheets is fairly consistent, the main differences in the costs are associated with the maintenance costs, which are based on the capital costs of the plants.

10.2.3 VARIABLE OPERATING AND MAINTENANCE COSTS

Again EPRI provided a single value for variable operating and maintenance costs, with no breakdown provided; $AU9.1/MWh. NETL provided more details with each unit process, including costs for consumables and the corresponding amount required per annum. NETL included the cost for water and water treatment chemicals as well as catalyst costs for the selective catalytic reduction of NOx and ash disposal costs. However, they did not include the costs of fabric filter bags and cages. To allow comparison, the NETL costs were converted to 2009 dollars and the costs of each consumable changed to match the prices used in this study, as shown in Table 19. The cost of fabric filter bags and cages, as calculated by the S&L model, was included in the NETL costs. As the coals used for flowsheets C1, FL1-4 have very high ash contents, the disposal costs were included for all the flowsheets. The water and water treatment chemicals were also included but are common for all sheets. The cost of disposal of gypsum was not considered by either study and the cost of coal has not been included.

FutureGen 2 and NETL appear the cheapest to operate, due mainly to the lower cost of disposing fly and bottom ash due to the lower ash content of the coals. NETL does not use the expensive caustic soda but does have a SCR catalyst which adds to its costs. Without disposing ash, FL1 has the least expensive variable operating costs.

Table 19: The variable operating and maintenance costs for NETL and flowsheets C1, FL1 and FL4

Consumable	**Cost**	**NETL adjusted**	**C1**	**FL1**	**FutureGen 2**	**FL4**
		\$/MWh	\$/MWh	\$/MWh	\$/MWh	\$/MWh
Water	\$1.03/gallon	1.05	1.05	1.05	1.05	1.05
WT chem	\$0.16/t	0.81	0.81	0.81	0.81	0.81
Limestone	\$15/t	1.16				0.97
Lime	\$95/t			0.29	0.87	
NAOH	\$940/t		6.13	0.16	0.30	0.92
SCR Catalyst	\$5500/t	0.40				
Fabric filter bags and cages		0.15	0.15	0.30	0.15	0.15
Subtotal		3.57	8.15	2.61	3.19	3.90
Fly and bottom ash disposal	\$25/t	1.86	4.25	4.25	1.06	4.25
Total		**5.43**	**12.25**	**6.56**	**4.24**	**8.01**

10.2.4 COST OF COAL

As each plant produces 550 MW of power, and the cost of each tonne of coal was assumed to be the same for each plant irrespective of the coal quality (\$60/tonne), the total cost of coal per annum for each plant is directly proportional to the net calorific value of the coals selected and the efficiency assumed for each plant. Flowsheets C1, FL1, FutureGen 2 and FL4 all assume similar efficiencies of 30%, while NETL assumes a lower efficiency of 28.8%. For the plants with the same efficiency, the higher the calorific value, the lower the total amount of coal required and the higher cost per kW produced, as shown in Table 20.

Table 20: The cost associated with coal for each flowsheet

	NETL adjusted	C1	FL1	FutureGen 2	FL4
Coal Net Calorific Value	26171 kJ/kg	18026 kJ/kg	18026 kJ/kg	24091 kJ/kg*	18026 kJ/kg
Coal cost	\$28.16/MWh	\$32.73/MWh	\$32.73/MWh	\$24.50/MWh	\$32.73/MWh

10.2.5 LEVELISED COST OF ELECTRICITY (LCOE)

The cost of electricity associated with each plant design was considered for a 40 year operating period by determining the average lifetime levelised cost of electricity generation (LCOE), as shown in Equation 28;

$$LCOE = \frac{\sum_{t=1}^{n} \frac{C + M_t + F_t}{(1+r)^t}}{\sum_{t=1}^{n} \frac{E_t}{(1+r)^t}} \qquad \textbf{Equation 28}$$

Where:

C is the capital costs associated with construction,

M_t is the operational and maintenance costs, both fixed and variable, in the year t,

F_t is the cost of fuel in the year t,

E_t is the electricity generation in the year t,

r is the discount rate and

N is the life of the system.

The cost of each plant was determined over 40 years with a discount rate of 3.5%, assuming the plant operates 24 hours per day for 320 days per year, and is shown in Table 21 and Figure 23.

Table 21: The levelised cost of electricity generation in the flowsheets considered with a plant life of 40 years and a discount rate of 3.5%.

	NETL	C1	FL1	FutureGen 2	FL4
LCOE \$/MWh	64.40	72.96	69.14	68.70	76.81

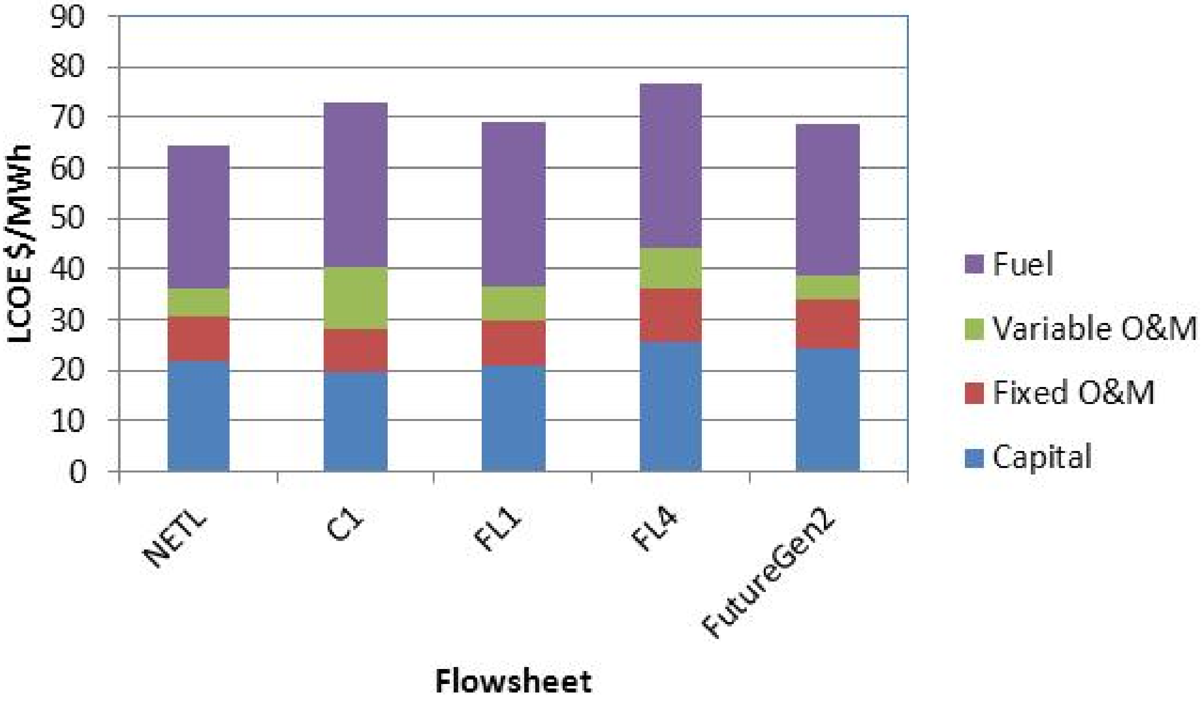


Figure 23: The LCOE for each flowsheet showing the impact of cost type.

The NETL and FutureGen 2 flowsheets have the lowest cost of electricity production due primarily to the higher calorific value and lower ash contents of the selected coals, while flowsheet FL1 has the next cheapest costs. FL1 has slightly higher variable operating and maintenance costs and a higher fuel rate. If an equivalent amount of fuel was used in both plant designs, these costs would be very similar. C1 is the next expensive plant due mainly to the high NaOH use, making the variable operating and maintenance costs almost double that of FL1. FL4 is most expensive in producing electricity due to higher capital and fixed operating and maintenance costs due to the higher sulphur content in the fuel and the lower CV resulting in higher amounts of fuel.

10.3 COSTS ASSOCIATED WITH S REMOVAL

The significant differences between all the flowsheets considered is the treatment of the recycle stream and cleaning of the flue gas for sulphur removal, and in the case of flowsheets C2 and C3, mercury removal. These differences are dependent on the coal sulphur content. Flowsheets C1 and FL1 are designed for the combustion of low sulphur coals, typical of Australian coals and where FGD's are not required to meet SOx emission regulations. SOx must be removed from the flue gas before compression and sequestration but due to the relatively low sulphur content of the coal and the resulting SOx content in the combustion gases, the recycle stream needs only to have particulates removed. SOx removal from the recycle stream is not required. Flowsheet C1 uses a caustic scrubber to remove the SOx from the flue gas, while FL1 uses a SDA with a DCCPS.

Flowsheets FL2 and FL3 represent oxy-fuel combustion plants where a medium sulphur coal (1.5% d.b. S) is combusted and the sulphur must be removed from the flue gas prior to the recycle back to the boiler to avoid excessive corrosion. FL2 uses a SDA prior to the recycle and a DCCPS to complete the final polishing of the flue gas. FL3 uses a WFGD prior to recycle and also uses a DCCPS to meet the low flue gas SOx concentration requirements (10ppm).

Flowsheet FL4 is designed to burn a high sulphur coal (3.5% d.b. S). All of the combustion gases are treated to remove SOx using a WFGD and DCCPS prior to any recycle and compression.

Table 22 shows the SOx concentrations calculated in Chapter 6 at the inlet of the main absorber in each flowsheet, indicating that flowsheet FL4 has significantly more sulphur to remove before compression and sequestration.

Table 22: SOx concentration in the combustion gases prior to absorbers.

	C1 **at caustic scrubber inlet**	**FL1** **at SDA inlet**	**FL2** **at SDA inlet**	**FL3** **at WFGD inlet**	**FL4** **at WFGD inlet**
SOx ppm	1192	806	2446	2376	5377
SOx g/MJ	0.23	0.23	1.64	1.61	3.73

Removing sulphur from the gas stream therefore would be expected to have a significant impact on the cost associated with both the construction of the plant and the operating and maintenance costs, depending on the coal sulphur content and the plant design, i.e. the equipment used to remove SOx from the gas stream and the treatment required of the recycled stream. To gain a better understanding of this the cost of the gas cleaning associated with SOx removal was determined using the Sargent & Lundy models described in Section 8. A boiler size of 550 MW was selected and the mass balance from the boiler to compressor inlet was calculated, as explained in Section 6 for each flowsheet outlined in Section 4. Costs of the boiler, ASU, gas cooler, gasheater, compressor and stack were not calculated as these items are common to all the flowsheets. Only items associated with sulphur removal were sized and costed.

10.3.1 CAPITAL COSTS

Capital costs for sulphur removal in each flowsheet are shown in Figure 24.

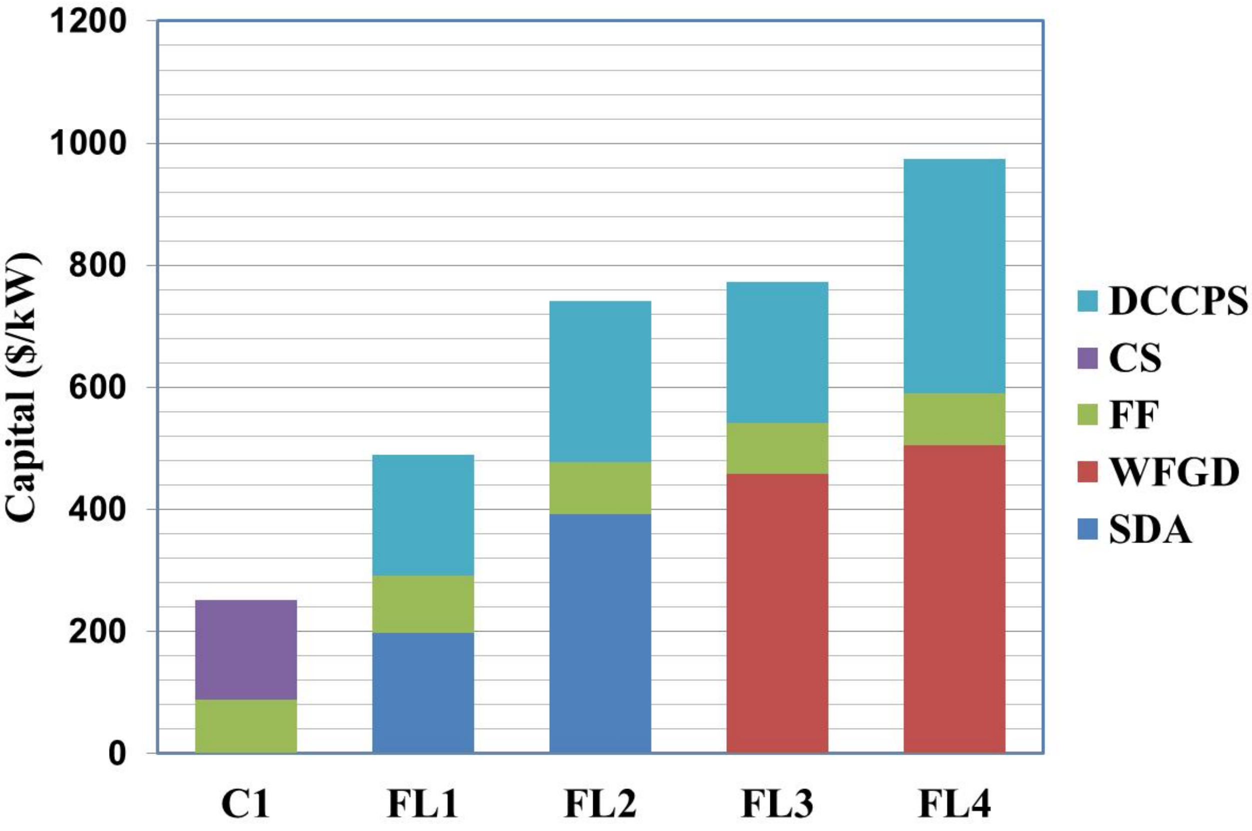


Figure 24: The capital costs/kW for the SOx removal equipment in each flowsheet considered.

The capital costs associated with using two unit operations, as in FL1, is approximately double that of using one, C1. As the amount of combustion gas that needs to be treated increases, the size of the units and the capital costs increase accordingly. Recycling the combustion gases without removing the SOx, as in flowsheets C1 and FL1 allows the caustic scrubber in C1 and the SDA in FL1 to be smaller, reducing the costs. Similarly, as the sulphur content of the gas stream increases, as in FL3 and FL4, the residence times in the WFGD increases, increasing the size of the scrubbers and increasing the capital costs. In FL4, the recycle stream is taken after the DCCPS, not before as in FL3, requiring a much larger DCCPS, as shown in the costs of the DCCPS in these two flowsheets. The sulphur content of the coal combusted in flowsheets FL2 and FL3 are the same, but the capital costs for the SDA, as shown in FL2 is less than the WFGD in FL3 but the DCCPS in FL2 is slightly larger making the total capital costs only slightly less than FL3.

The capital costs of the fabric filters (FF) are similar for all flowsheets. FL1 fabric filter costs is slightly higher because it has two smaller units, with total gas flows of 0.1 and 0.08 kmol/kg fuel, instead of one large unit in the other flowsheets with total gas flows nearing 0.2 kmol/kg of fuel, but the difference is small.

10.3.2 FIXED OPERATING AND MAINTENANCE COSTS

Maintenance on the unit operations is predicted by the models to be proportional to the size of each item. That is the larger the unit, the more it will cost to build and the more it will cost to maintain. Therefore the fixed operating and maintenance costs are determined by the number of operators and the capital costs of the unit operations selected. As shown in Figure 25, the unit operations associated with SOx removal in flowsheet C1 have the lowest total fixed O&M costs. This is because the overall capital for the plant is relatively low and 12 extra operators were required.

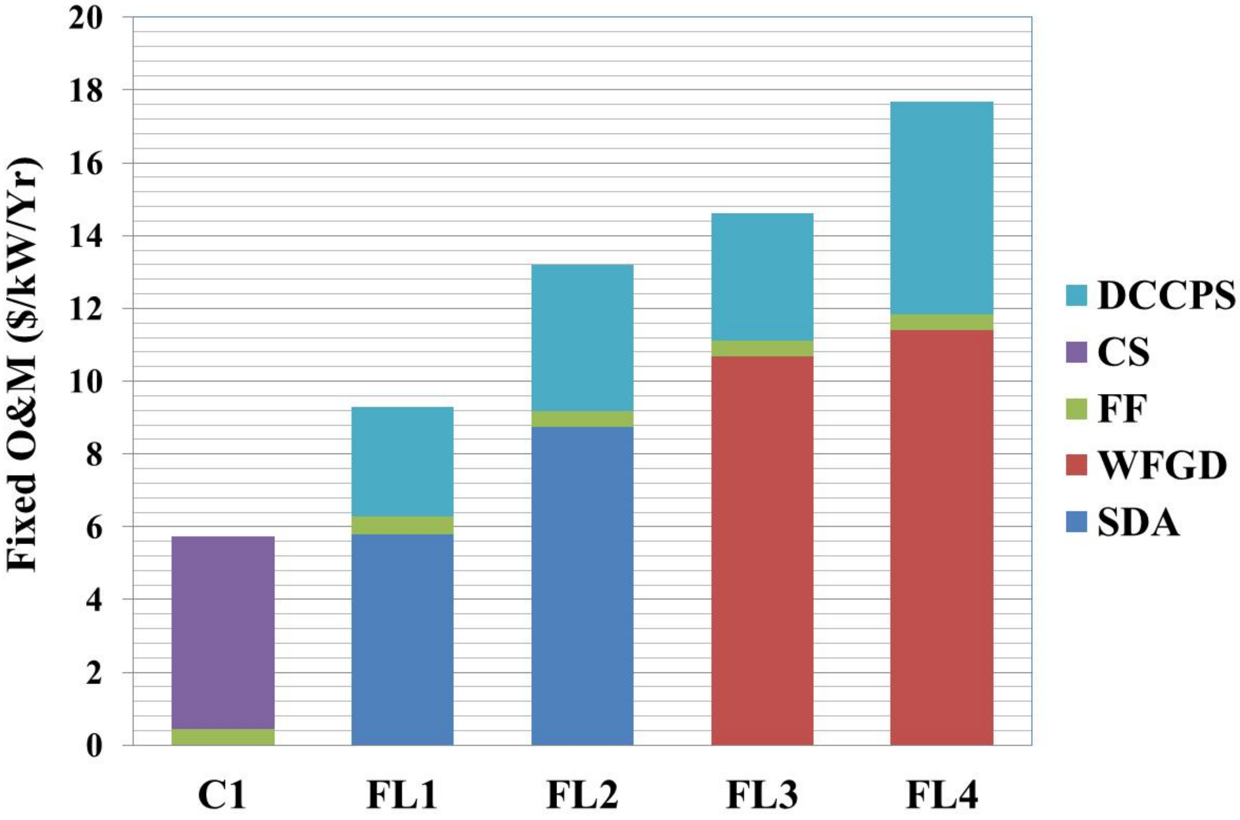


Figure 25: The Fixed Operating and maintenance costs for the SOx removal equipment in the flowsheets considered.

Fixed operating and maintenance costs of the WFGD's is significantly higher because the capital costs of these units are approximately double the costs of the caustic scrubber, due to the higher gas flows and significantly higher amount of SOx required to be cleaned, and the additional 16 personnel required to operate these units.

According to the Sargent & Lundy models WFGD's are more expensive than SDA's and so the maintenance costs are also slightly higher. In addition, the SDA's require 12 operators, not 16. However, the DCCPS required after the WFGD in FL3 is smaller than that required after the SDA in FL2 because the WFGD is more efficient, assumed to be able to remove 98%

of the SOx, lowering the SOx concentration to 49 ppm, whereas the SDA can remove 94%, producing 140 ppm SOx.

10.3.3 REAGENT COSTS

In this study the variable operating and maintenance costs for the absorbers were limited solely to the cost of reagent materials. As the cost of NaOH used in this study is very high compared to the cost of lime and limestone, shown in Table 23, the caustic scrubber (CS) in C1 and the DCCPSs in FL1-4 are expensive to operate. All sulphur removal in flowsheet C1 is completed using NaOH, making this flowsheet the most expensive in terms of variable operating and maintenance costs, shown in Figure 26.

Table 23: Costs of reagents used in this study.

	NaOH	**Lime**	**Limestone**
Cost	\$940/tonne (100%)	\$95/tonne	\$15/tonne

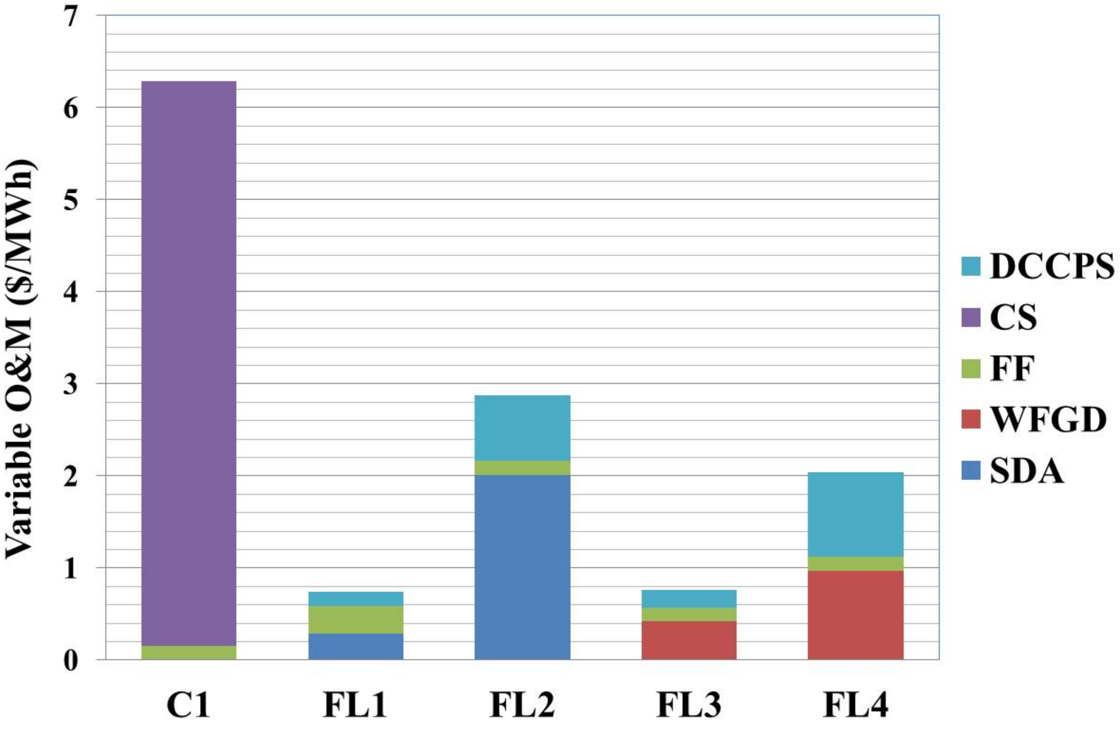


Figure 26: The variable operating and maintenance costs for the SOx removal equipment in the flowsheets considered.

The DCCPs units also use NaOH as the feed reagent but they are a finishing polisher, removing the last small amount of sulphur in the gas stream. Compared to the caustic scrubber in C1, which is treating an inlet gas stream containing 1192 ppm SOx, the DCCPS have a much lower amount of SOx to remove, as shown in Table 24.

Table 24: The inlet SOx concentration and total gas flows entering the DCCPS units in each flowsheet.

	FL1	FL2	FL3	FL4
SOx ppm	48	140	49	110
DCCPS inlet flow tonnes/hr	910	1247	1175	2262

Though burning the same coal as flowsheet C1, flowsheet FL1 has drastically lower reagent costs. The cost of lime used in FL1 is approximately one tenth of the cost of the NaOH used in C1. In addition, approximately twice the amount of reagent is used in the NaOH scrubber in C1 compared to the SDA in FL1 due the molar requirement of the chemical reactions involved. The caustic scrubber requires 2 moles of NaOH (2x40g) per mole of SO_2 absorbed:

$$2\mathrm{NaOH} + \mathrm{SO_2} \Rightarrow \mathrm{Na_2SO_3} + \mathrm{H_2O}$$

While the SDA uses one mole of CaO (56g) for every mole of SO_2:

$$\mathrm{SO_2} + \mathrm{CaO} + 1/2\ \mathrm{H_2O} \Rightarrow \mathrm{CaSO_3} \bullet 1/2\ \mathrm{H_2O}$$

The efficiency of these two units was assumed to be 95% for the caustic scrubber and 94% for the SDA.

The cost of NaOH would need to drop to below $350/tonne to bring the reagent costs of C1 to below the costs incurred by FL2, a flowsheet burning coal with 1.5% sulphur, shown in Figure 27. An even greater drop, to $46/tonne, would be required to bring the reagent costs below FL1 which is designed to burn the same coal as C1.

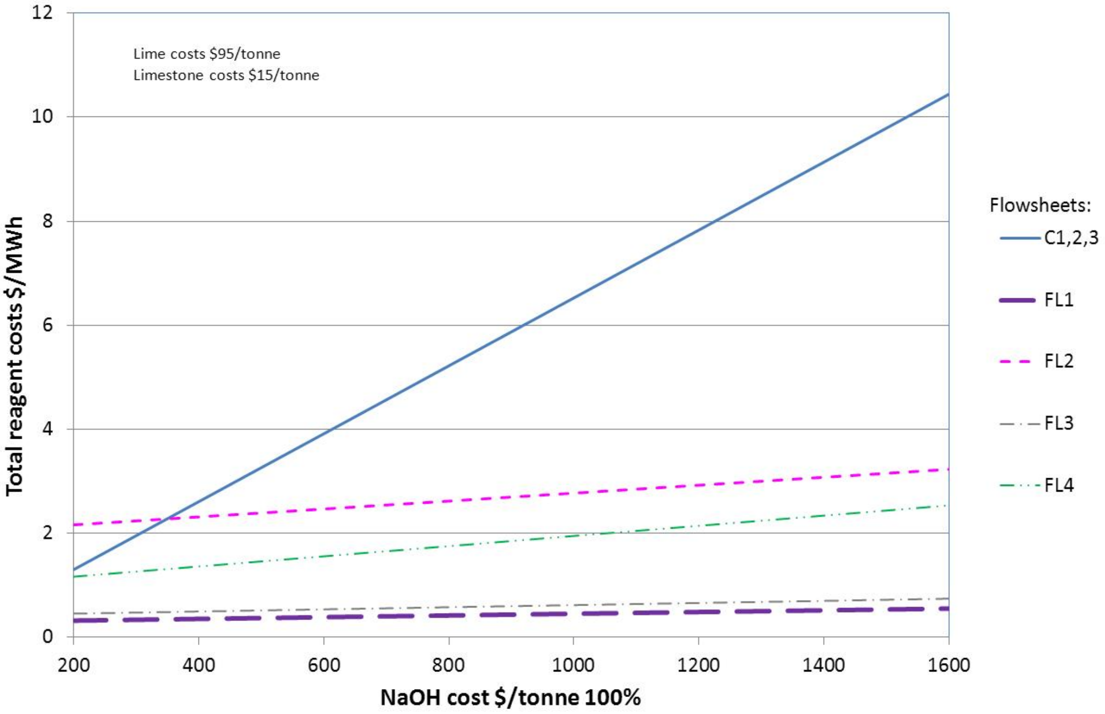


Figure 27: The reagent cost for all flowsheets, with changing cost of NaOH.

Similarly, the cost of lime must increase enormously for the cost of reagent in FL1 to come close to that of C1, and limestone would have to quadruple in price before the reagent costs of FL4, a flowsheet burning coal with 16 times the amount of sulphur, approaches the costs of C1, as shown in Figure 28.

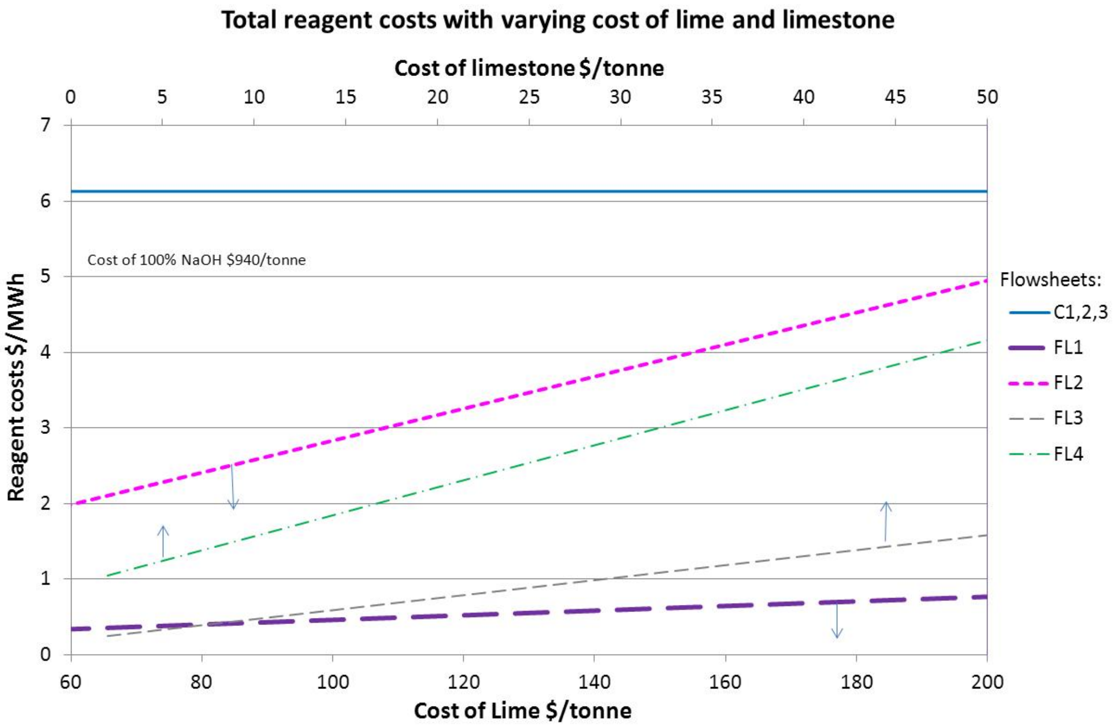


Figure 28: The impact of changing lime and limestone prices on reagent costs, with comparison to C1.

WFGD's are the cheapest to operate, due to the low cost of limestone and the assumed higher efficiency of 98%. FL2 and FL3 burn the same coal but the cost of reagent for the WFGD is \$0.42/MWh compared to \$2/MWh for the SDA while the concentrations of SO_2 in the inlet stream of these two units are virtually the same.

During the calculation of the mass balance, it was assumed no SO_2 was collected by the fabric filters in all the flowsheets. Experiments completed at IFK Stuttgart[58] suggest up to 50% capture of SOx by the filters can occur. The amount collected is dependent on the gas SOx concentration in the gas stream, the gas volume passing through the filters and the ash composition. The surface area of the filters will also have an impact. Therefore the costs associated with SOx collection in absorbers discussed here will be greater than what will be found in practice.

The cost of replacement fabric filter bags and cages are consistent throughout the flowsheets. FL1 has higher fabric filter maintenance costs because there are two baghouses in this flowsheet.

The above costs do not include the cost of the coal as the cost of the coal was assumed to be the same for each flowsheet. However, a coal with 3.5% sulphur (flowsheet FL4) would be expected to be cheaper than a coal with 0.22% sulphur (flowsheets C1 & FL1). Therefore how much cheaper would a high sulphur coal need to be to make it economic to use, requiring a flowsheet that can treat it? Figure 29 shows the costs incurred in sulphur removal for every tonne of coal combusted for each flowsheet considered, with the cost of capital assumed to be distributed across 25 years. The cheapest cost is produced by the flowsheet FL1 which burns the low sulphur coal (0.22% d.b. sulphur). This is \$4/tonne of coal cheaper than FL3, which is designed to combust coal with 1.5% d.b. sulphur, and \$9/tonne of coal cheaper than FL4 which is designed to combust coal with 3.5% d.b. sulphur. FL4 will only be the cheapest flowsheet if coal with high sulphur content can be purchased at least \$9/tonne cheaper than coal with much lower sulphur contents.

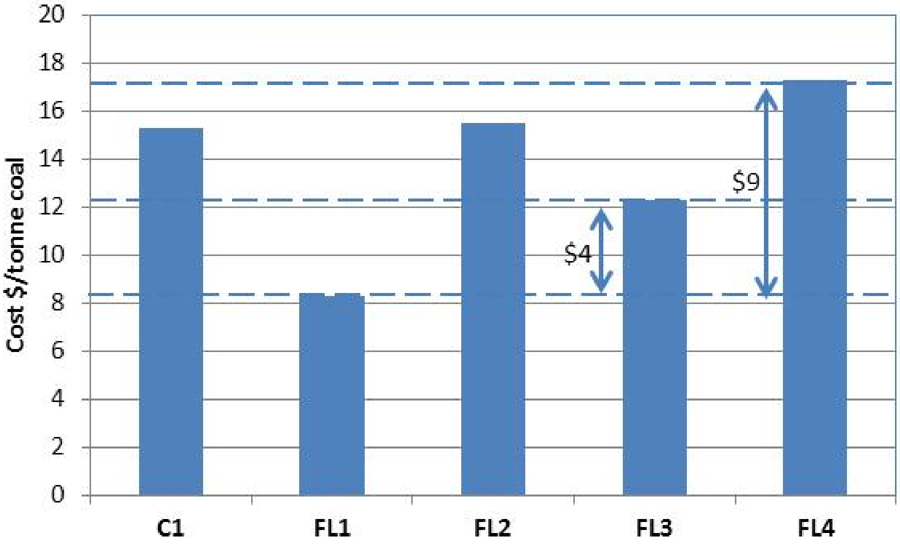


Figure 29: The net cost of sulphur removal per tonne of coal used for each of the flowsheets.

10.4 COSTS ASSOCIATED WITH HG REMOVAL

Power plants in Australia do not remove mercury from the fluegas, due mainly to the low mercury content of coal burnt and remote location of power plants. However in oxyfuel combustion, concern over mercury attack of the aluminium heat exchanger in the compression phase may make mercury removal necessary. The coal used as the basis for the calculations completed in this study has a mercury content of 43 ppb (43 ng per kilogram of coal). This represents a low mercury content coal. Cost of removing mercury from the flue gas in flowsheet C1 were determined based on the costs presented by Stiegel[3]. Two options were considered:

1. an atmospheric packed bed of activated carbon situated after the NOAH scrubber and the subsequent dryer or
2. a pressurised activated carbon bed after compression before the cold box.

Cost for the activated carbon was assumed to be $1600/tonne, while its disposal was assumed to be $500/tonne, due to its mercury content it is assumed to be contaminated waste. As the atmospheric activated carbon bed accommodates a much larger volumetric gas flowrate, it requires a significantly larger capital outlay, as shown in Table 25. As the amount of carbon required is not a function of the mercury to be collected but the volumetric flow rate of the gas, the variable operating and maintenance costs, which includes the cost of the carbon and its disposal costs, is larger for the atmospheric bed. However these costs are small overall as shown by the change in costs for the entire plant when the mercury removal is included, as shown in Table 26.

Table 25: The costs associated with mercury removal

Costs		Atmospheric activated carbon packed bed	Pressurised (24 atm) activated carbon packed bed
Capital	\$/kW	7.06	2.26
Fixed O&M	\$/kW/y	0.13	0.06
Variable O&M	\$/MWh	0.0124	0.0031

Table 26: Total plant costs including an activated carbon packed bed

Costs		C1, with no Hg removal	C1 with atmospheric packed bed	C1 with pressurised packed bed
Capital	\$/kW	3354.99	3362.05	3357.25
Fixed O&M	\$/kWy	63.08	63.21	63.14
Variable O&M	\$/MWh	12.4006	12.4131	12.4037
LCOE	c/kWh	7.3106	7.3168	7.3130

The cost of removing mercury when the flue gas is compressed, i.e. the pressurized activated carbon bed, is significantly smaller than the costs associated with the removal of mercury at atmospheric pressure, or after the sulphur scrubber. These differences would undoubtedly direct construction of a pressurized bed, if mercury removal is required. The concentrations of mercury predicted to be present in the compression stage are very low due to the use of a coal with a low mercury content. Under these conditions, operators may elect to accept a slightly shortened life of the cold box. However, the cost of removing mercury from the flue gas to avoid corrosion of the aluminium cold box is minimal compared to the cost of removing sulphur from the flue gas. Additional capital costs associated with mercury removal are 2.94% the capital costs of sulphur removing equipment when an atmospheric carbon bed is used, and 0.009% when a pressurized bed is used. Similarly the fixed operating and maintenance costs of the flue gas cleaning circuit are increased by between 1 and 2% by the addition of a packed bed and the variable operating and maintenance costs are increased by between 0.05 and 0.2%.

10.5 OTHER SOX IMPACTS ON COSTS

For each of the flowsheets considered above, the sulphur removal from the recycle and fluegas streams were fixed, depending on the flowsheet design and the unit operations selected. Flowsheets C1 and FL1 have no cleaning of the sulphur from the recycle stream, except for collection of sulphur with particulates in the fabric filter, assumed in these calculations to be zero. These cases are specific to low sulphur coal, typically the flowsheets to be used when combusting most Australian coals. However experimental studies, discussed in Section 8 have indicated that allowing a greater portion of the sulphur to remain in the recycle gas stream increases the acid dew point. Ensuring the gas temperature of the recycle stream is maintained above the acid dewpoint can decrease the plant efficiency, shown in Figure 21. Compared to no desulphurisation of the recycle stream, a reduction in sulphur by 50% could improve the plant efficiency by 0.14%. This could be achieved by an increased collection of the sulphur in the fabric filters, reducing the sulphur content of the feed coal, or cleaning sulphur from the recycle stream prior to recycle, as in flowsheets designed for higher sulphur coals. Increasing the plant efficiency means less coal is required to produce the equivalent amount of power, and an overall reduction in the amount of sulphur to be removed, which may decrease the size of the sulphur cleaning equipment required. Cleaning the entire flue gas prior to the CPU would require additional unit operations and so were not considered further. However improvements in efficiency resulting from the collection of sulphur species by the fabric filters not considered in the previous calculations can deliver some cost savings.

A sensitivity analysis of the costs associated with the sulphur removal equipment was completed to assess the impact of this change in efficiency on costs for flowsheets C1 and FL1. Only these two flowsheets were considered because other flowsheets already remove sulphur from the recycle flue gas stream, being designed for medium and high sulphur coals. Increasing the plant efficiency decreases the capital cost associated with the removal of sulphur as the amount of gas to be treated and the total amount of sulphur to be removed decreases slightly. For both flowsheets C1 and FL1 the reduction in capital costs associated with an increase in plant efficiency of 0.05% (a 25-30% collection of sulphur by fabric filters) is approximately 0.22%, as shown in

Figure 30, resulting in a saving of $135 million and $273 million in the capital costs of the sulphur cleaning equipment for flowsheets C1 and F1 respectively.

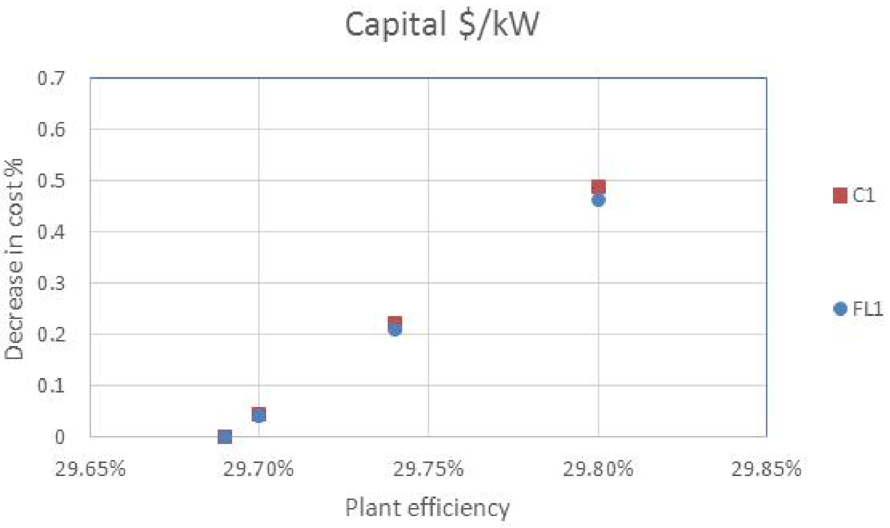


Figure 30: The percentage change in capital cost associated with sulphur removal unit operations with improvement in plant efficiency due to increased acid dew point.

In these calculations the Fixed Operating and Maintenance costs are based on the capital costs of the plant and the number of operators required. No reduction in the total labour was assumed. With an improvement in efficiency of 0.1% associated with removal of sulphur in the fabric filters, the reduction in variable operating and maintenance costs for C1 is approximately $7,300 per annum and $16,000 per annum for FL1.

The variable operating and maintenance costs are solely based on the amount and cost per tonne of the sulphur absorbent and the cost of replacement bags and cages in the fabric filters. A reduction in the total amount of sulphur fed with the coal feed resulting from a reduced coal feed rate due to increased efficiency reduces the amount of the absorbent required. The reduction in cost is directly applicable to the cost of the absorbent, in the case of C1: NaOH in the caustic scrubber, and for FL1: lime in the SDA and NaOH in the DCCP. An increase in efficiency of 0.1% due to a decrease in the acid dew point reduces the cost of absorbent by $107,000 per annum for flowsheet C1 and by $8,000 per annum for FL1. The absorbent cost for C1 is significantly greater than that for FL1.

11 CONCLUSIONS

A comprehensive literature review was carried out to identify the impact of coal quality and gas quality control on the oxyfuel process. So far only two detailed reports (i.e. NETL and EPRI) have been found to be published indicating the costs of an oxyfuel plant. The objective of the NETL report was to establish a cost and performance baseline for the oxy-combustion process that can be used as a basis for comparison with past and future studies. NETL used high sulphur Illinois 6 coal and a WFGD for SOx removal. The EPRI report was commissioned by the Australian Department of Resources, Energy and Tourism (DRET). In contrast to the NETL report, EPRI used low sulphur Australian coals such as Latrobe Valley Brown Coal and

Hunter Valley Black Coal. The EPRI report did not mention which SOx removal unit operation was employed for sulphur removal. Also, other sulphur removal unit operations such as SDA, DCCPS or CDS have not been studied or compared in either the NETL or EPRI reports. IHI (C1) and B&W (FL1, FL2, FL3 and FL4) have published several flowsheets highlighting flue gas cleaning in oxyfuel plants with varying sulphur and mercury removal. However, no cost estimates have been found on these flowsheets in the existing literature.

In this report, the impact of sulphur and mercury levels in coal and their removal on the overall costs of the oxyfuel process has been estimated. Process simulations based on published IHI and B&W flowsheets have been carried out in ASPEN Plus v 7.3. The costing codes were developed using S&L models. Total plant costs including capital costs ($/kW), fixed operating and maintenance costs ($/kWh/yr) and variable operating and maintenance costs ($/MWh) have been calculated and compared with the published NETL and EPRI reports. The levelised cost of electricity (LCOE) in $/MWh was also calculated and compared for all cases. Also, costs associated with sulphur and mercury removal was obtained for different published flowsheets with varying coal suphur.

For low sulphur coal, it was estimated that though capital investment for the C1 (IHI-Callide design) configuration was found to be the lowest (~50%), due to the higher price of the caustic reagent currently used in the caustic scrubber, the LCOE of the C1 was higher (~5%) compared to the FL1 flowsheet. A less expensive form of sodium (eg trona) would reduce this cost. It was also estimated that an increase in the sulphur content of coal increases the capital, fixed and variable operating and maintenance costs for sulphur removal to a great extent. The capital costs of WFGD units are found to be the highest. In contrast, variable operating and maintenance costs of the WFGD units were found to be the lowest compared to SDA and a caustic scrubber due to the lower raw material costs for limestone and higher operational efficiency of WFGDs. The cost associated with mercury removal using carbon bed was found to be less significant (<3%) compared to sulphur removal.

It was also observed that SO_3 and mercury removal in fabric filters can improve the efficiency of an oxyfuel plant. The cost benefits of such capture in fabric filters have been estimated in the report.

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APPENDICES

APPENDIX A: CAPITAL COST COMPARISON OF EPRI AND NETL REPORTS, ASSUMPTIONS, LIMITATIONS AND METHODOLOGY

The capital costs estimates of two Oxy-fuel plants provided in two separate reports have been compared. The reports were:

- “Pulverized Coal Oxycombustion Power Plants”, DOE/NETL report 2007/1291, 2008. (NETL report)
- “Australian Electricity Generation Technology Costs – Reference Case 2010”, EPRI, Palo Alto, CA and Commonwealth of Australia: November 2009. (EPRI Report)

The NETL flow sheet, shown in Figure 5 (Body of report), is designed for a Illinois No. 6 coal, containing 2.82% (d.b.) sulphur, while the EPRI plant, Figure 6 (Body of report), is designed to be used with a low sulphur coal (0.43% (no basis given, appears to be as received)) from the Hunter Valley.

It is difficult to determine all the assumptions used in the development of the assessments. Those that have been found in each report are listed in Table 27. As shown in the table, very little detail is provided in the EPRI report. Many of the unit operations used in each study are listed in **Table 28**.

Table 27: The design parameters for equipment costs in the NETL report and the EPRI report

	NETL Report, case 5a	**EPRI Report**
Plant size	550 MW	750 MW
Efficiency	29.3% HHV	30.1% HHV sent out
Boiler	Babcock and Wilcox supercritical once through spiral wound Benson boiler.	"Single reheat supercritical plant"
Steam cycle	3500 psig 1110°F/1150°F	267 bar/596°C/596°C
ASU	Conventional cryogenic distillation producing 99 mol% oxygen	
Product gas CO_2	87.35 mol% CO_2, 0.015% (by volume) water, compressed to supercritical fluid at 2215 psia, 95°F, 17215 TPD	CO_2 capture rate of 85-90% compressed to 16 MPa (2321 psi), 626193 kg/hr
NOx	B&W DRB-4Z Low NOx burner and over fire air on front and rear walls to 0.07 lb/10^6 BTU.	
SOx	FGD to 0.1 lb/10^6 BTU.	CO_2 product < 100 ppmv total S
Hg removal	(included in FGD) 90% removal assumed. Design coal is high in Cl so Hg will be oxidized (85-90%) with additional oxidation occuring in the air heater and captured in the wet scrubber (with an addition of an unidentified additive to enhance removal) and by unburnt carbon in the baghouse.	
Bag house	2 single stage in line 10 compartment units, 8 m long polyphenylsulfide (PPS) with an intrinsic teflow (PTFE) coating. Design removal efficiency: 99.8%	

Table 28: Unit operations listed in each study

	NETL Report, Case 5a	**EPRI Report**
Coal Handling	Coal is delivered to site in 100 100-tonne rail cars, unloaded by a trestle bottom dumper into 2 receiving hoppers. 4 vibrating feeders, 4 conveyer belts, a magnetic plate separator, surge bin, crusher with tower boiler silos.	Crusher and “associated delivery system”
Steam Generation	Babcock and Wilcox supercritical once through spiral wound Benson boiler- a Carolina (two pass) design with the primary 2 stage superheater and economiser in one pass and a horizontal reheater in the other, Low NOx burners, light oil ignitors/warmup system, soot blowers, air preheaters (Ljungstrom type), coal feeders B&W 89G pulverisers, spray type desuperheater, Primary air fans, Induced draft fans, forced draft fans, overfire air system air heater	Coal pulverisers, burners, waterwall-lined furnace, superheater, reheater, economiser heat transfer surface, soot blowers, Ljungstrom air heaters, forced-draft and induced draft fans.
Turbine-generator	Tandem compound turbine with single flow HP section and double flow IP section Hydrogen cooled generator, DC motor lube oil pumps, main lube oil pumps, condenser.	Steam turbine, power generator plus the main, reheat and extraction steam piping, feedwater heaters, boiler feedwater pumps, condensate pumps, and an air cooled condenser
Water/steam loop	Steam water separator, boiler recirculating pump, piping	Condensate pumps, low pressure feedwater heaters Feedwater pumps De-aerator High pressure feedwater heaters

Table 28 Cont'

	NETL Report, Case 5a	**EPRI Report**
ASU	Conventional cryogenic distillation	No detail given
Flue gas clean-up	SCR, bag house, ID fans, limestone based forced FGD including limestone storage and preparation system, the absorber, cyclone separators, oxidation air compressor, vacuum filtration system, stack	Flue gas reheater, ESP or baghouse, FGD, Dryer
CO_2 purification and compression	Centrifugal compressors, electric motors, intercoolers, 2 radial flow bed adsoption dryers with activated alumina adsorbents	No details given
Ash handling	Fly ash storage silo with wet conditioner and unloading chute, blower, clinker grinder, dewatering hydrobins	No details given

NETL "boiler and ancillary equipment" costs, which include boiler, baghouse, FGD, condensing heat exchanger were provided by Babcock and Wilcox. Costs for the air separation unit (ASU) and CO_2 purification were provided by Air Liquide.

Comparison of plant costs would be meaningless if the plants that are being compared are different in size and efficiency. Therefore this report has endeavoured to compare the two plants on the same basis, by using a cost per kW produced after correcting for plant size and efficiency. Corrections for product quality could not be completed as EPRI, in most cases, did not provide quality data.

The equipment costs for each unit operation were collected from the reports. These costs did not include engineering costs, management costs and project and process contingencies. Including these costs in a comparison on a kW basis would make larger plants more attractive as these costs should not increase significantly with increasing plant size, but equipment costs do. However, the EPRI report did not provide these costs separately and so they were assumed to be 20% of the total capital costs, such that:

EPRI equipment costs=0.8 x EPRI capital costs.

The NETL analysis was completed in 2008 using 2007 US dollars, while the EPRI report was based on 2009 Australian dollars. An average cost increase of 24% per annum was assumed and an exchange rate of \$AUS 1 to \$US 0.85 was used. This was the average exchange rate in 2009. Therefore:

NETL 2009 US equipment costs = NETL 2007 US x 1.24 x 1.24

$$NETL\ 2009\ AUS\ equipment\ costs = \frac{NETL\ 2009\ US\ equipment\ costs}{0.85}$$

To compare the equipment costs on a kW basis, the 2009 equipment costs were divided by the plant size:

$$NETL\ 2009\ AUS\ equipment\ costs/kW = \frac{NETL\ 2009\ AUS\ equipment\ costs}{550}$$

$$EPRI\ 2009\ AUS\ equipment\ costs/kW = \frac{EPRI\ 2009\ AUS\ equipment\ costs}{750}$$

As site costs are dependent on the location selected for the plant, and each study selected a different site, the Improvement to site and Building and structure costs were not included in the total plant equipment costs/kW.

The costs are shown in Table 29. Each cost/kW provided by the two sources are not significantly different, falling within the accuracy of each report, both ±30%. The costs of the Flue gas clean up and the CO_2 removal and compression are significantly different. Costs provided by NETL for CO2 purification and compression are half the cost provided by EPRI, while NETL's costs for Flue gas clean-up is an order of magnitude greater than that provided by EPRI. Given the small value for flue gas clean up and the very large value for CO_2 removal and compression in the EPRI report, it appears EPRI may be:

- removing the SOx and NOx in the compression unit, and not in a FGD. This is possible given it is using a low sulphur coal, though the cost of CO_2 removal and compression still appear high, or
- the costs of the FGD has been included in CO_2 removal and compression steps.

Table 29: The Capital costs of the NETL and EPRI studies compared on a $AU/kW basis

								High sulphur coal	Low sulphur coal	
	NETL , Case 5a, 2007 (USD) Total cap costs	NETL 2007 (USD) costs excluding engineering, management, process and project contingencies	EPRI 2009(AUD) includes engineering, management, process and project contingencies	EPRI 2009(AUD) less 20% engineering, management, process and project contingencies	NETL projected 2009 (USD)	NETL 2009 (USD/kW)	NETL 2009 (EQ AUD/kW)	NETL 2009 (EQ AUD/kW) Efficiency Correction	EPRI 2009 (AUD/kW)	Difference (NETL vs. EPRI) %
Power plant size (MW)	550	550	750	750	550	550	550	550	750	
Coal handling system	43,675	34,855	97,797	78,238	53,593	97	115	118	104	13
Coal prep and feed systems	20,721	16,566	33,717	26,974	25,472	46	54	56	36	20
Feed water and misc BOP systems	88,059	69,207	149,221	119,377	106,413	193	228	234	159	75
PC boiler and accessories	389,255	283,808	811,205	648,964	436,383	793	933	959	865	94
Air separation unit	256,188	212,331	611,962	489,569	326,480	594	698	717	653	65
Flue gas clean up	142,374	118,203	24,162	19,330	181,749	330	389	399	26	374
CO2 purification and compression	111,942	85,093	581,883	465,506	130,839	238	280	288	621	-333
HRSG, ducting and stack	26,789	21,406	37,884	30,307	32,914	60	70	72	40	32
Steam turbine generator	135,246	110,917	445,926	356,741	170,546	310	365	375	476	-101
Cooling water systems	40,420	32,549	107,342	85,874	50,047	91	107	110	114	-5
ASH/spent sorbent handling	14,172	11,733	43,428	34,742	18,041	33	39	40	46	-7
Access electric plant	95,226	77,486	223,196	178,557	119,142	217	255	262	238	24
I&C	25,220	20,574	45,438	36,350	31,635	58	68	70	48	21
Improvement to site	14,887	11,297	36,437	29,150	17,370	32	37			
Buildings and structure	56,936	45,417	151,622	121,298	69,833	127	149			
Total	1,461,110	1,151,442	3,401,220	2,720,976	1,770,457	3,219	3,787	3,699	3,427	21

APPENDIX B: ASPEN PROCESS MODEL FLOWSHEETS AND RESULTS

Table 30: ASPEN Plus V7.3 process simulation results for flowsheet FL1.

ASPEN PLUS RESULTS										
Stream	**primary recyc. FG (dried)**	**recy. FG rest**	**FG Filter in**	**Filter out**	**SDA flue gas in**	**SDA flue gas out/filter in**	**DCCPS-Inlet stream**	**DCCPS-Outlet stream**	**FG-out after recycle**	**Dry FG to compression**
	PRIMRECY	RESTRECY	RECFFIN	RECFFOUT	SDAIN	OUTFFIN	DCCPSIN	DCCPSOUT	FGOUT	TOCOMP
Substream: MIXED										
Mole Frac										
O2	3.59E-02	0.0317336	0.0317336	0.0317336	0.0317336	3.09E-02	3.09E-02	0.0359483	3.59E-02	0.0408944
CO2	0.817777	0.713789	0.713789	0.713789	0.713789	0.7026356	0.7026356	0.817777	0.817777	0.9302945
SO2	8.9749E-06	0.00080563	0.00080563	0.00080563	0.00080563	4.7582E-05	4.75822E-05	8.97488E-06	8.9749E-06	1.02097E-05
H2O	0.1209482	0.231604	0.231604	0.231604	0.231604	0.2447293	0.2447293	0.1209482	0.1209482	0
CACO3	0	0	0	0	0	0	0	0	0	0
CA++	0	0	0	0	0	0	0	0	0	0
CO3--	0	0	0	0	0	0	0	0	0	0
H+	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
OH-	0	0	0	0	0	0	0	0	0	0
SO3	3.2092E-08	5.2578E-07	5.2578E-07	5.2578E-07	5.2578E-07	3.1054E-08	3.11E-08	3.21E-08	3.21E-08	3.65E-08
H2SO3	0	0	0	0	0	0	0	0	0	0
H2SO4	0	0	0	0	0	0	0	0	0	0
HCO3-	0	0	0	0	0	0	0	0	0	0
SO4--	0	0	0	0	0	0	0	0	0	0
HSO4-	0	0	0	0	0	0	0	0	0	0
SO3--	0	0	0	0	0	0	0	0	0	0
HSO3-	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0
CASO4*2W	0	0	0	0	0	0	0	0	0	0
CAO	0	0	0	0	0	0	0	0	0	0
CA(OH)+	0	0	0	0	0	0	0	0	0	0
CASO3	0	0	0	0	0	0	0	0.00E+00	0	0
CASO4	0.00E+00	0.00E+00	0.00E+00	0	0	0	0.00E+00	0.00E+00	0.00E+00	0
N2	4.92E-03	4.29E-03	4.29E-03	4.29E-03	4.29E-03	4.22E-03	4.22E-03	0.00492151	4.92E-03	0.00559865
NO	3.97E-04	3.46E-04	3.46E-04	0.0003464	0.0003464	0.00034098	3.41E-04	3.97E-04	3.97E-04	0.000452122
S	1.648E-14	2.2238E-09	2.2238E-09	2.2238E-09	2.2238E-09	2.189E-09	2.18903E-09	1.648E-14	1.648E-14	0
H2	0.00127521	0.00111137	0.00111137	0.00111137	0.00111137	0.00109401	0.00109401	0.00127521	0.00127521	0.00145066
CL2	0	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0	0.00E+00	0.00E+00	0.00E+00	0
C	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.000E+00	0.000E+00	0.000E+00	0.000E+00
CO	1.87E-02	1.63E-02	1.63E-02	0.0163196	0.0163196	0.0160646	1.61E-02	1.87E-02	1.87E-02	2.13E-02
HG	**2.1135E-09**	**1.8419E-09**	**2.4559E-09**	**1.8419E-09**	**2.4559E-09**	2.4175E-09	1.81315E-09	2.11348E-09	2.1135E-09	2.40428E-09
NO2	0	0	0	0	0	0	0	0.00E+00	0	0
NAOH	0	0	0	0	0	0	0	0	0	0
NA2SO4	4.04E-48	0	0	0	0	0	0.00E+00	4.0371E-48	4.04E-48	0
Total Flow kmol/hr	0.0229754	0.1053207	0.1053207	0.1053207	0.080945	0.0822299	8.22E-02	0.0705448	0.0475693	0.0418159
Total Flow kg/hr	0.9189462	3.922474	3.922474	3.922474	3.014648	3.034542	3.034542	2.821569	1.902623	1.798973
Total Flow l/min	10.46801	68.83027	60.94913	61.55665	46.84293	47.58649	48.06082	31.1763	21.67338	19.05202
Temperature K	333.15	473.15	423.15	423.15	423.15	423.15	423.15	323.1468	333.15	333.15
Pressure atm	1	0.9901308	1	9.90E-01	1.00E+00	1	0.9901308	1	1	1
Vapor Frac	1	1	1	1	1	1	1	1	1	1
Liquid Frac	0	0	0	0	0	0	0	0	0	0
Solid Frac	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Enthalpy cal/mol	-84026.91	-79333.94	-79816.17	-79816.17	-79816.17	-79467.47	-79467.47	-84116.11	-84026.91	-87680.18
Enthalpy cal/gm	-2100.84	-2130.162	-2143.111	-2143.111	-2143.111	-2153.408	-2153.408	-2103.07	-2100.84	-2038.067
Enthalpy cal/sec	-536.2664	-2320.974	-2335.082	-2.34E+03	-1.79E+03	-1815.168	-1815.168	-1648.322	-1110.307	-1018.453
Entropy cal/mol-K	**1.967999**	**4.234456**	**3.137927**	**3.157623**	**3.137927**	2.986873	3.006569	1.696138	1.967999	2.7404
Entropy cal/gm-K	0.0492038	0.1136976	0.0842551	0.084784	8.43E-02	0.0809382	0.0814719	0.0424068	0.0492038	0.0636987
Density mol/cc	3.6581E-05	2.5503E-05	2.88E-05	2.8516E-05	2.88E-05	2.88E-05	2.85159E-05	3.77129E-05	3.6581E-05	3.65805E-05
Density gm/cc	0.0014631	0.00094979	0.00107261	0.00106202	0.00107261	0.00106282	0.00105233	0.00150839	0.0014631	0.00157374

Table 31: ASPEN Plus V7.3 process simulation results for flowsheet FL2.

ASPEN PLUS RESULTS									
Stream	**primary recyc. FG (dried)**	**recy. FG rest**	**SDA sorbent in**	**SDA flue gas in**	**SDA flue gas out/filter in**	**DCCPS-Inlet stream**	**DCCPS-Outlet stream**	**Fg-out after recycle**	**Dry FG to compression**
	PRIMRECY	RESTRECY	CA-H2O	SDAIN	OUTFFIN	DCCPSIN	DCCPSOUT	FGOUT	TOCOMP
Substream: MIXED									
Mole Frac									
O2	3.51E-02	0.02905734	0	0.03156755	2.91E-02	2.91E-02	0.03507725	0.03507725	0.03990076
CO2	0.82034589	0.67918562	0	0.71100947	0.67918562	0.67918562	0.82034589	0.82034589	0.93315246
SO2	1.01E-05	0.00014007	0	0.00244405	0.00014007	0.00014007	1.01E-05	1.01E-05	1.14E-05
H2O	0.12088761	0.27204714	1	0.2344906	0.27204714	0.27204714	0.12088761	0.12088761	0
CACO3	0	0	0	0	0	0	0	0	0
CA++	0	0	0	0	0	0	0	0	0
CO3--	0	0	0	0	0	0	0	0	0
H+	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0
OH-	0	0	0	0	0	0	0	0	0
SO3	9.80E-08	9.41E-08	0	1.64E-06	9.41E-08	9.41E-08	9.80E-08	9.80E-08	1.11E-07
H2SO3	0	0	0	0	0	0	0	0	0
H2SO4	0	0	0	0	0	0	0	0	0
HCO3-	0	0	0	0	0	0	0	0	0
SO4--	0	0	0	0	0	0	0	0	0
HSO4-	0	0	0	0	0	0	0	0	0
SO3--	0	0	0	0	0	0	0	0	0
HSO3-	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0
CASO4*2W	0	0	0	0	0	0	0	0	0
CAO	0	0	0	0	0	0	0	0	0
CA(OH)+	0	0	0	0	0	0	0	0	0
CASO3	0	0	0	0	0	0	0.00E+00	0	0
CASO4	0.00E+00	0.00E+00	0.00E+00	0	0	0.00E+00	0.00E+00	0	0
N2	4.94E-03	4.08E-03	0	4.27E-03	4.08E-03	4.08E-03	0.00494054	0.00494054	0.00561992
NO	3.86E-04	3.19E-04	0	0.00033419	0.00031923	3.19E-04	3.86E-04	0.0003863	0.00043942
S	3.26E-14	5.31E-09	0	5.56E-09	5.31E-09	5.31E-09	3.26E-14	3.26E-14	0
H2	0.00119591	0.00098823	0	0.00103453	0.00098823	0.00098823	0.00119591	0.00119591	0.00136036
CL2	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0.00E+00	0.00E+00	0	0
C	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.000E+00	0.000E+00	0.000E+00	0
CO	1.72E-02	1.42E-02	0.00E+00	0.01484337	0.014179	1.42E-02	1.72E-02	1.72E-02	0.0195155
HG	2.11E-09	1.74E-09	0	2.43E-09	2.32E-09	1.74E-09	2.11E-09	2.11E-09	2.40E-09
NO2	0	0	0	0	0	0	0.00E+00	0	0
NAOH	0	0	0	0	0	0	0	0	0
NA2SO4	1.09E-47	0	0	0	0	0.00E+00	1.09E-47	1.09E-47	0
Total Flow kmol/hr	0.04728905	0.08301208	0.01036547	0.18880136	0.19764782	1.15E-01	0.09472668	0.04743762	0.041703
Total Flow kg/hr	1.89331772	3.00949749	0.18673684	7.02912379	7.16547023	4.15597272	3.79258394	1.89926622	1.7959554
Total Flow l/min	21.5456612	54.2509061	8.12487495	109.259412	114.37886	67.0009869	41.8618247	21.6133542	19.0005674
Temperature K	333.15	473.15	573.15	423.15	423.15	423.15	323.136818	333.15	333.15
Pressure atm	1	0.99013077	1	1.00E+00	1	0.99013077	1	1	1.00E+00
Vapor Frac	1	1	1	1	1	1	1	1	1
Liquid Frac	0	0	0	0	0	0	0	0	0
Solid Frac	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0
Enthalpy cal/mol	-84223.627	-78322.917	-55480.509	-79799.038	-78801.923	-78801.923	-84312.977	-84223.627	-87901.873
Enthalpy cal/gm	-2103.6383	-2160.41	-3079.6363	-2143.3919	-2173.6226	-2173.6226	-2105.8699	-2103.6383	-2041.1266
Enthalpy cal/sec	-1106.3488	-1806.0413	-159.74488	-4.19E+03	-4326.3968	-2509.3101	-2218.5246	-1109.8248	-1.02E+03
Entropy cal/mol-K	1.91905282	3.74699573	-5.2061151	3.09146733	2.65765798	2.67735405	1.64674816	1.91905282	2.68413716
Entropy cal/gm-K	0.04793183	0.10335477	-0.2889833	8.30E-02	0.07330716	0.07385044	0.04113053	0.04793183	6.23E-02
Density mol/cc	3.66E-05	2.55E-05	2.13E-05	2.88E-05	2.88E-05	2.85E-05	3.77E-05	3.66E-05	3.66E-05
Density gm/cc	0.00146457	0.00092456	0.00038305	0.00107223	0.00104411	0.0010338	0.00150996	0.00146457	0.00157535

Table 32: ASPEN Plus V7.3 process simulation results for flowsheet FL3.

ASPEN PLUS RESULTS											
Stream	**primary recyc. FG (dried)**	**recy. FG rest**	**FG filter inlet**	**FG filter outlet**	**WFGD flue gas in**	**WFGD flue gas out**	**main recycle**	**DCCPS-Inlet stream**	**DCCPS-Outlet stream**	**Fg-out after recycle**	**Dry FG to compression**
	PRIMRECY	RESTRECY	FGAPHO	OUTFFOUT	TOWFGD	FGDOUT	RECMAIN	DCCPSIN	DCCPSOUT	FGOUT	TOCOMP
Substream: MIXED											
Mole Frac											
O2	3.49E-02	3.37E-02	0.03363114	0.03363114	0.03363114	3.37E-02	3.37E-02	3.37E-02	0.03494434	3.49E-02	0.03975371
CO2	0.82076353	0.7902805	0.76035002	0.76035002	0.76035002	0.7902805	0.7902805	0.7902805	0.82076353	0.82076353	0.93372479
SO2	8.83E-06	4.92E-05	0.00237466	0.00237466	0.00237466	4.92E-05	4.92E-05	4.92E-05	8.83E-06	8.83E-06	1.00E-05
H2O	0.12097919	0.1535829	0.18200406	0.18200406	0.18200406	0.1535829	0.1535829	0.1535829	0.12097919	0.12097919	0
CACO3	0	0	0	0	0	0	0	0	0	0	0
CA++	0	0	0	0	0	0	0	0	0	0	0
CO3--	0	0	0	0	0	0	0	0	0	0	0
H+	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
OH-	0	0	0	0	0	0	0	0	0	0	0
SO3	3.40E-08	3.37E-08	1.63E-06	1.63E-06	1.63E-06	3.37E-08	3.37E-08	3.37E-08	3.40E-08	3.40E-08	3.87E-08
H2SO3	0	0	0	0	0	0	0	0	0	0	0
H2SO4	0	0	0	0	0	0	0	0	0	0	0
HCO3-	0	0	0	0	0	0	0	0	0	0	0
SO4--	0	0	0	0	0	0	0	0	0	0	0
HSO4-	0	0	0	0	0	0	0	0	0	0	0
SO3--	0	0	0	0	0	0	0	0	0	0	0
HSO3-	0	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0	0
CASO4*2W	0	0	0	0	0	0	0	0	0	0	0
CAO	0	0	0	0	0	0	0	0	0	0	0
CA(OH)+	0	0	0	0	0	0	0	0	0	0	0
CASO3	0	0	0	0	0	0.00E+00	0	0	0.00E+00	0	0.00E+00
CASO4	0.00E+00	0.00E+00	0.00E+00	0	0	0.00E+00	0	0.00E+00	0.00E+00	0.00E+00	0.00E+00
N2	4.88E-03	4.70E-03	4.53E-03	4.53E-03	4.53E-03	4.70E-03	4.70E-03	4.70E-03	0.00488028	4.88E-03	0.00555195
NO	3.87E-04	3.73E-04	0.00035948	0.00035948	0.00035948	3.73E-04	0.00037266	3.73E-04	3.87E-04	3.87E-04	4.40E-04
S	0	1.62E-13	5.47E-09	5.47E-09	5.47E-09	1.62E-13	1.62E-13	1.62E-13	0	0	0
H2	0.00086632	0.00083385	0.00080436	0.00080436	0.00080436	0.00083385	0.00083385	0.00083385	0.00086632	0.00086632	0.00098555
CL2	0	0	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0.00E+00	0	0.00E+00	0.00E+00	0.00E+00	0.00E+00
C	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.000E+00	0.00E+00	0.000E+00	0.000E+00	0.000E+00	0.000E+00
CO	1.72E-02	1.65E-02	1.59E-02	0.01594326	0.01594326	1.65E-02	0.01652715	1.65E-02	1.72E-02	1.72E-02	1.95E-02
HG	**2.01E-09**	**1.93E-09**	**2.48E-09**	**1.86E-09**	1.86E-09	1.93E-09	1.93E-09	1.93E-09	2.01E-09	2.01E-09	2.28E-09
NO2	0	0	0	0	0	0.00E+00	0	0	0.00E+00	0.00E+00	0.00E+00
NAOH	0	0	0	0	0	0	0	0	0	0	0
NA2SO4	0.00E+00	0	0	0	0	0	0	0.00E+00	0	0.00E+00	0
Total Flow kmol/hr	0.04809312	0.08454221	0.19114919	0.19114919	0.19114919	0.18438869	0.08454221	0.09984647	0.09610372	4.80E-02	0.04220232
Total Flow kg/hr	1.92617189	3.31696542	7.36995788	7.36995786	7.36995786	7.23438478	3.31696542	3.91741936	3.84903918	1.92286729	1.81822942
Total Flow l/min	21.9120069	68.5800095	129.785556	131.046864	94.9331525	82.7185311	37.9264465	44.7920846	42.4722831	21.8744139	19.2280651
Temperature K	333.15	593.15	496.349216	496.349209	363.15	328.02668	328.02668	328.02668	323.151885	333.15	333.15
Pressure atm	1	1	0.99975327	9.90E-01	1.00E+00	1	1.00E+00	1	1	1	1
Vapor Frac	1	1	1	1	1	0.99999995	0.99999995	0.99999995	0.99999835	1	1
Liquid Frac	0	0	0	0	0	4.74E-08	4.74E-08	4.74E-08	1.65E-06	0	0
Solid Frac	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Enthalpy cal/mol	-84268.393	-80709.411	-80702.233	-80702.233	-81982.049	-83319.209	-83319.209	-83319.209	-84357.636	-84268.393	-87955.97
Enthalpy cal/gm	-2104.0333	-2057.1069	-2093.1146	-2093.1146	-2126.3083	-2123.6249	-2123.6249	-2123.6249	-2106.2616	-2104.0333	-2041.5171
Enthalpy cal/sec	-1125.7583	-1895.3757	-4285.0463	-4.29E+03	-4.35E+03	-4267.5333	-1.96E+03	-2310.8693	-2251.9676	-1123.8269	-1031.0962
Entropy cal/mol-K	**1.91340412**	**7.27061192**	**5.18175924**	**5.20096513**	2.18645305	1.4915683	1.4915683	1.4915683	1.64143023	1.91340412	2.67859095
Entropy cal/gm-K	0.04777433	0.18531204	0.13439549	1.35E-01	0.05670842	0.03801682	3.80E-02	0.03801682	0.04098362	0.04777433	0.06217189
Density mol/cc	3.66E-05	2.05E-05	2.45E-05	2.43E-05	3.36E-05	3.72E-05	3.72E-05	3.72E-05	3.77E-05	3.66E-05	3.66E-05
Density gm/cc	0.00146508	0.0008061	0.00094642	0.00093731	0.00129388	0.00145763	0.00145763	0.00145763	0.00151041	0.00146508	0.00157602

Table 33: ASPEN Plus V7.3 process simulation results for flowsheet FL4.

ASPEN PLUS RESULTS										
Stream	**primary recyc. FG (dried)**	**recy. FG rest**	**FG filter inlet**	**FG filter outlet**	**WFGD flue gas in**	**DCCPS-Inlet stream**	**DCCPS-Outlet stream**	**FG-Out after recycle**	**main recycle**	**Dry FG to compression**
	PRIMRECY	RESTRECY	FGAPHO	OUTFFOUT	TOWFGD	DCCPSIN	DCCPSOUT	FGOUT	RECMAIN	TOCOMP
Substream: MIXED										
Mole Frac										
O2	3.30E-02	3.30E-02	0.03385216	0.03385216	0.03385216	3.18E-02	3.30E-02	3.30E-02	3.30E-02	3.75E-02
CO2	0.8268677	0.8268677	0.77653817	0.77653817	0.77653817	0.79603495	0.8268677	0.8268677	0.8268677	0.94066242
SO2	8.84E-06	8.84E-06	0.00537314	0.00537314	0.00537314	0.00010947	8.84E-06	8.84E-06	8.84E-06	1.01E-05
H2O	0.12097296	0.12097296	0.16613193	0.16613193	0.16613193	0.15361559	0.12097296	0.12097296	0.12097296	0
CACO3	0	0	0	0	0	0	0	0	0	0
CA++	0	0	0	0	0	0	0	0	0	0
CO3--	0	0	0	0	0	0	0	0	0	0
H+	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
OH-	0	0	0	0	0	0	0	0	0	0
SO3	8.27E-08	8.27E-08	4.02E-06	4.02E-06	4.02E-06	8.20E-08	8.27E-08	8.27E-08	8.27E-08	9.41E-08
H2SO3	0	0	0	0	0	0	0	0	0	0
H2SO4	0	0	0	0	0	0	0	0	0	0
HCO3-	0	0	0	0	0	0	0	0	0	0
SO4--	0	0	0	0	0	0	0	0	0	0
HSO4-	0	0	0	0	0	0	0	0	0	0
SO3--	0	0	0	0	0	0	0	0	0	0
HSO3-	0	0	0	0	0	0	0	0	0	0
H3O+	0	0	0	0	0	0	0	0	0	0
CASO4*2W	0	0	0	0	0	0	0	0	0	0
CAO	0	0	0	0	0	0	0	0	0	0
CA(OH)+	0	0	0	0	0	0	0	0	0	0
CASO3	0	0	0	0	0	0.00E+00	0.00E+00	0	0.00E+00	0
CASO4	0.00E+00	0.00E+00	0.00E+00	0	0	0.00E+00	0.00E+00	0	0.00E+00	0
N2	4.84E-03	4.84E-03	4.58E-03	4.58E-03	4.58E-03	4.66E-03	4.84E-03	0.00484277	0.00484277	5.51E-03
NO	3.52E-04	3.52E-04	0.00033299	0.00033299	0.00033299	3.39E-04	3.52E-04	0.00035248	3.52E-04	0.00040099
S	0	0	7.19E-09	7.19E-09	7.19E-09	1.93E-13	0	0	0	0
H2	0.00061348	0.00061348	0.00057955	0.00057955	0.00057955	0.00059039	0.00061348	0.00061348	0.00061348	0.00069791
CL2	0	0	0	0	0	0	0	0	0	0
HCL	0	0	0	0	0	0.00E+00	0.00E+00	0	0.00E+00	0
C	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.000E+00	0.00E+00	0.00E+00	0.000E+00	0.00E+00
CO	1.34E-02	1.34E-02	1.26E-02	0.01261292	0.01261292	1.28E-02	0.01335035	0.01335035	1.34E-02	1.52E-02
HG	1.95E-09	1.95E-09	2.46E-09	1.84E-09	1.84E-09	1.88E-09	1.95E-09	1.95E-09	1.95E-09	2.22E-09
NO2	0	0	0	0	0	0.00E+00	0.00E+00	0	0.00E+00	0
NAOH	0	0	0	0	0	0	0	0	0	0
NA2SO4	0.00E+00	0	0	0	0	0	0	0	0.00E+00	0
Total Flow kmol/hr	0.07307357	0.06292786	0.19533717	0.19533717	0.19533717	1.92E-01	0.18453384	4.85E-02	0.06292786	0.04266129
Total Flow kg/hr	2.93370852	2.52638551	7.63545677	7.63545675	7.63545675	7.5410259	7.40854007	1.94844604	2.52638551	1.84267638
Total Flow l/min	33.2935081	49.6066374	139.790508	141.149044	97.0130894	86.0226832	81.5532637	22.1121504	28.6709589	19.437178
Temperature K	333.15	576.417807	523.15	523.149976	363.15	328.032048	323.151839	333.15	333.15	333.15
Pressure atm	1	1	0.99975327	9.90E-01	1.00E+00	1	1.00E+00	1.00E+00	1	1.00E+00
Vapor Frac	1	1	1	1	1	1	0.99999839	1	1	1
Liquid Frac	0	0	0	0	0	0	1.61E-06	0	0	0
Solid Frac	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00	0.00E+00
Enthalpy cal/mol	-84741.234	-82336.129	-81152.443	-81152.443	-82710.55	-83769.431	-84830.611	-84741.234	-84741.234	-88493.669
Enthalpy cal/gm	-2110.7567	-2050.8497	-2076.1153	-2076.1153	-2115.9762	-2130.063	-2112.9829	-2110.7567	-2110.7567	-2048.7885
Enthalpy cal/sec	-1720.0958	-1439.2325	-4403.358	-4.40E+03	-4.49E+03	-4461.9057	-4.35E+03	-1.14E+03	-1481.2737	-1.05E+03
Entropy cal/mol-K	**1.79047435**	**7.16828096**	**5.80830566**	**5.82751123**	2.26499888	1.37417338	1.51809321	1.79047435	1.79047435	2.53868364
Entropy cal/gm-K	0.0445976	0.1785494	0.14859334	1.49E-01	0.05794525	0.03494205	3.78E-02	4.46E-02	0.0445976	5.88E-02
Density mol/cc	3.66E-05	2.11E-05	2.33E-05	2.31E-05	3.36E-05	3.72E-05	3.77E-05	3.66E-05	3.66E-05	3.66E-05
Density gm/cc	0.0014686	0.0008488	0.00091034	0.00090158	0.00131175	0.00146105	0.00151404	0.0014686	0.0014686	0.00158002

APPENDIX C: SENSITIVITY ANALYSIS OF THE PREDICTION OF COSTS ASSOCIATED WITH SOX.

To determine the impact of changing variable costs on the overall cost in each spreadsheet considered, a sensitivity analysis was completed that considered changes in the price of consumables: NaOH, Lime and Limestone. In addition, the cost of labour and coal were considered and the impact of the plant size was also assessed.

As described in Section 8, the basic unit costs of caustic scrubbers, fabric filters, spray drier absorbers (SDA), wet flue gas desulfurisation scrubbers (WFGD), direct contact cooler-polishing scrubber (DCCPS) and fabric filters for the flowsheets considered were determined using S&L models[48-50], based on the mass balances completed in Aspin (Section 6). The unit operations considered are listed in Table 34. Other equipment in the oxy-fuel plants proposed are common to all flowsheets and therefore not considered.

Table 34: The unit operations costed in the flowsheets considered

Flowsheet	Unit operations costed
C1, C2, C3	Caustic Scrubber, fabric filter
FL1, FL2	Spray Drier Absorber (SDA), fabric filters and Direct contact cooler/polishing scrubber (DCCPS)
FL3, FL4	fabric filters, wet flue gas desulfurisation scrubbers (WFGD) and the direct contact cooler/polishing scrubber (DCCPS)

Cost of NaOH

Caustic soda is used in the caustic scrubber in Flowsheets C1, C2 and C3 to remove sulphur from the gas stream before compression (C1) or before the carbon bed is used to remove mercury (Hg) from the gas (C2 and C3). Sulphur poisons the carbon bed and so must be removed beforehand. It is also used in the direct contact cooler-polishing scrubber (DCCPS) in spreadsheets FL1, FL2, FL3 and FL4. In each of these cases the majority of the sulphur is removed prior to the DCCPS, in the spray drier absorber (SDA) in flowsheets FL1 and FL2, which use lime, and in the wet flue gas desulfurisation scrubber (WFGD) for flowsheets FL3 and FL4, which use limestone. The DCCPS are required to treat a much lower concentration of sulphur in the gas stream prior to compression, bringing the SOx concentration below the compression inlet specification of 10ppm.

This analysis provides the cost of all the reagents used in removing sulphur species but assumes the cost of other reagents used to remove sulphur species from the gas stream in FL1, 2, 3 and 4 did not change from the base cast of lime of $95/tonne and limestone of $15/tonne. Variations in the costs of these reagents are considered later.

For most calculations the cost of caustic soda was set at $940/ton (100%), which is a current industrial market price[51]. However, if the cost of caustic changes significantly, the impact would be most felt by the plant reliant on the caustic scrubber for removal of all the sulphur in the gas stream: C1, C2 and C3, as shown in Figure 31.

Though both flowsheets Fl2 and FL3 burn the same coal, the SDA in FL2 removes less of the sulphur species in the gas stream compared to the WFGD in FL3, leaving 3.5 times the sulphur for the DCCPS to remove in FL2. Therefore changes in price of NaOH have a greater impact on the costs associated with this unit operation for FL2. FL1 burns a low sulphur coal and therefore has only a small amount of sulphur species for the polisher to clean, requiring only small amounts of NaOH and therefore very little effect is seen. FL4 uses a high sulphur coal (3.5 wt%) but the majority of the sulphur in the gas stream is removed prior to the DCCPS, leaving only a slightly higher amount compared with FL2, and shown by the slightly higher slope of the line on Figure 31 for FL4 compared to that for FL2. The line for FL2 in Figure 31 starts at a higher cost on the y axis because the cost of lime for the SDA in flowsheet FL2 is greater than the cost of limestone for the WGFD in flowsheet FL4.

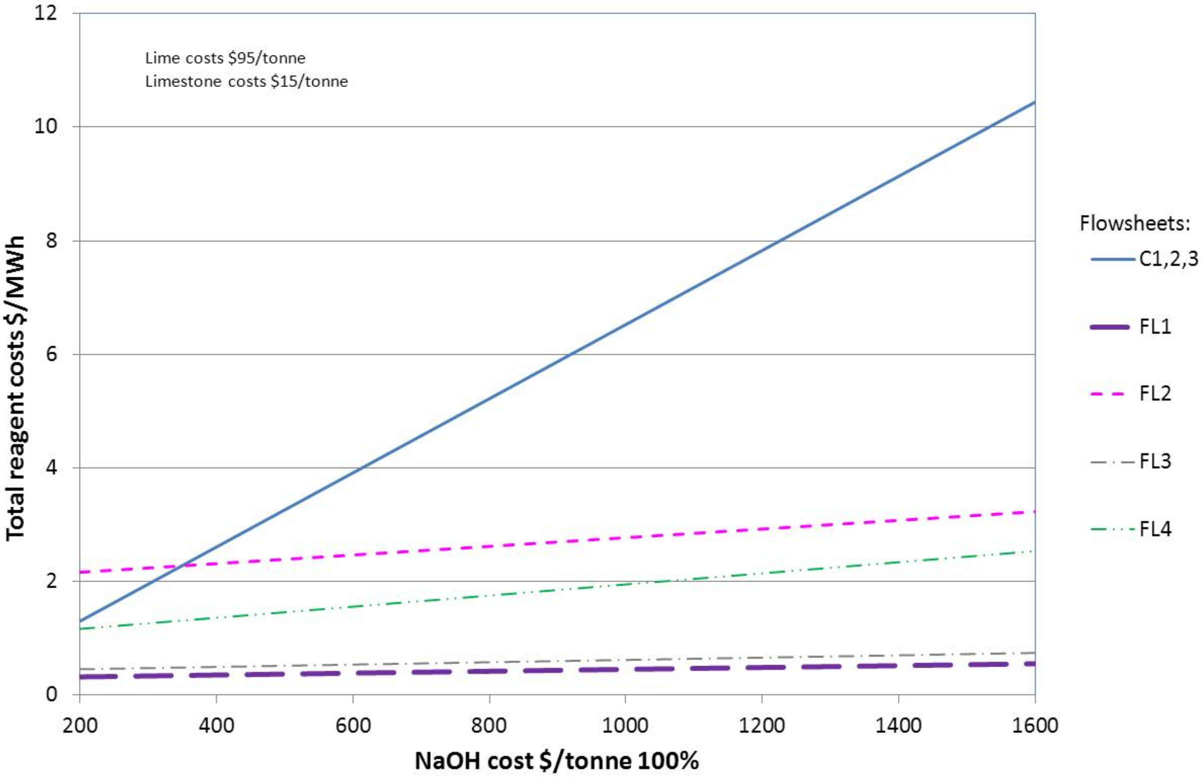

Figure 31: The variation in reagent costs with changing NaOH price

Cost of Lime and Limestone

If the cost of NaOH is constant at $980/tonne (100%), but the cost of lime and limestone is changed the cost of reagent for flowsheets FL2 and FL4 start to approach the costs of flowsheets C1-3, shown in Figure 32. But even with a doubling of the expected price of lime and limestone, the reagent costs for C1-3 is well above that of the other flowsheets. The total reagent costs for flowsheets C1, C2, C3 remains unchanged because these flowsheets only use NaOH and do not use lime or limestone. Of the other flowsheets, FL1, which uses the coal with the lowest sulphur content, has the lowest costs and is least effected by increasing costs. FL2 and FL3 use the same medium sulphur coal, but the reagent costs of FL2 are much higher because it uses the more expensive lime, where FL3 uses limestone. FL4 uses the cheaper limestone, but has a much larger amount of sulphur to remove because it burns the high sulphur containing coal.

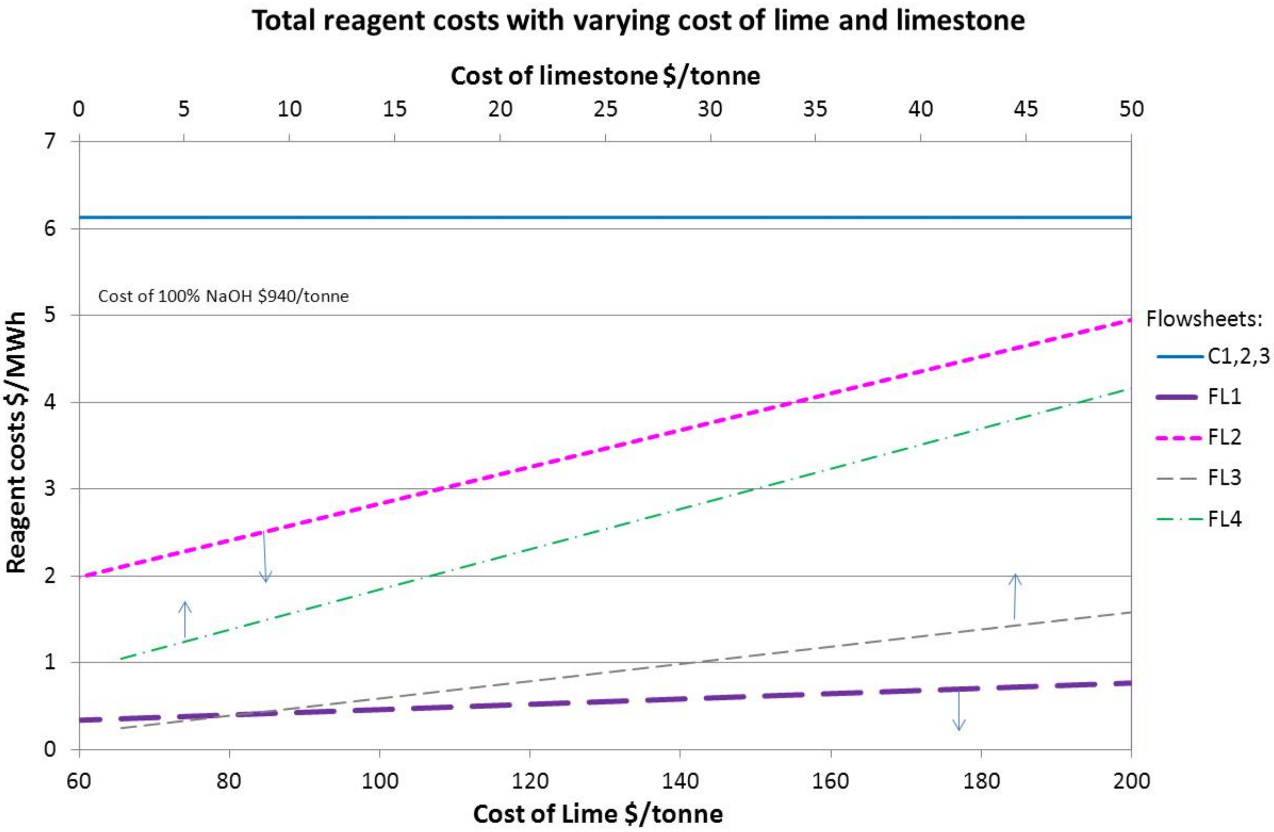

Figure 32: The impact of lime and limestone price on reagent costs for each of the flowsheets considered.

Coal Price

For a 550 MW plant with a coal feed of 300 t/hr, an increase in the cost of coal by $1/tonne increases the cost of power produced by $0.55/MWhr or $M2.3/yr, as shown in Figure 33.

This is by far the greatest cost to the plant and decreasing coal costs by accepting higher sulphur can be well justified if the coal price can be maintained at the lower value.

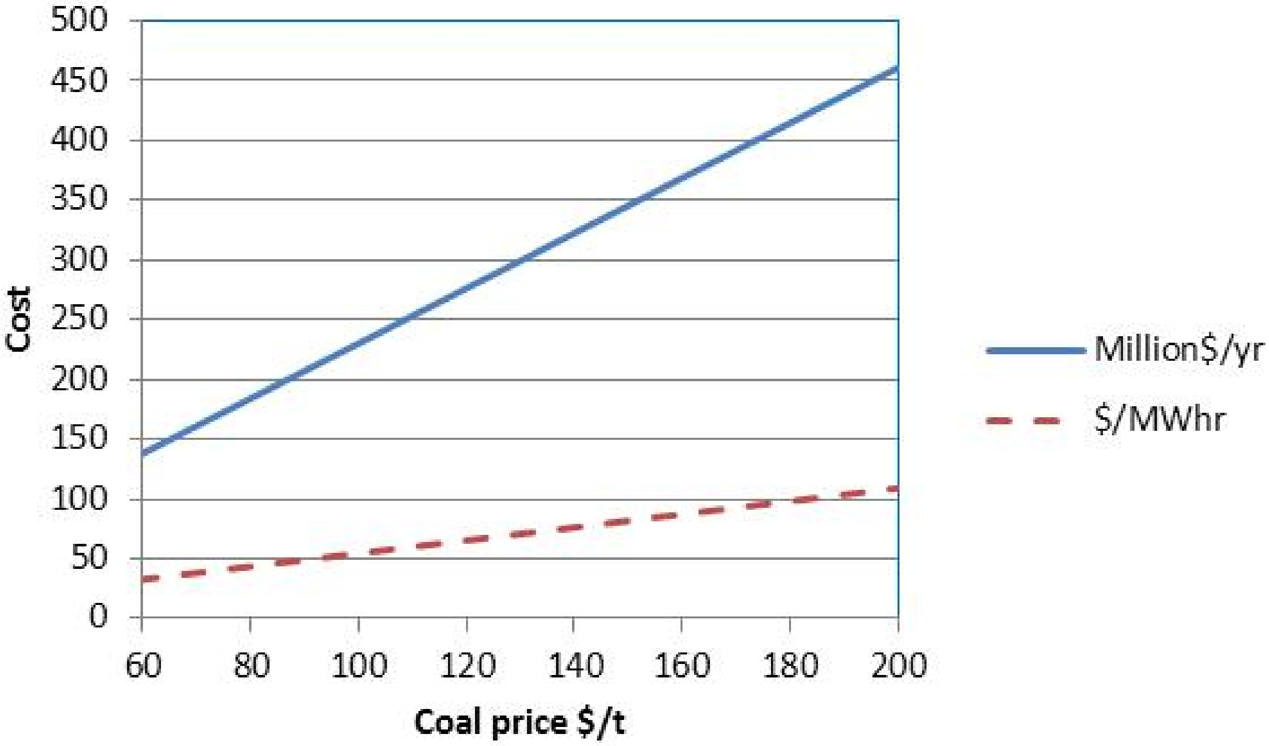

Figure 33: The impact of coal price on costs

Labour costs

As each flowsheet utilises different unit operations, labour costs were expected to impact the costs for each flowsheet differently. The labour costs considered here are solely for the gas cleaning equipment that differ amongst the flowsheets considered, and do not represent the fixed O&M costs for the entire plant. Fixed operating and maintenance costs for each unit operation were calculated as the sum of the operating labour costs, the maintenance, materials and additional labour and the additional administrative labour costs:

Total Fixed O&M costs = *Operating labour costs* + *Maintenance, materials and additional labour costs* + *Additional administrative labour costs*

Each of these costs were based on the models defined by S&L[48-50], and described below:

$$Operating\ labour\ costs = \frac{number\ additional\ operators\ x\ 2080\ x\ labour\ rate}{Gross\ plant\ size\ x\ 1000}$$

SDA and caustic scrubber utilise 12 operators, while the WFGD scrubbers require 16 operators if the plant size is greater than 500 MW and 12 if under this size. The fabric filters and DCCPS were expected to require no more operators than is already utilised in the plant, and so produced no additional cost.

$$Maintenance, materials\ and\ additional\ labour\ costs = \frac{Base\ module\ cost\ x\ N}{Retrofit\ factor\ x\ Gross\ plant\ size\ x\ 1000}$$

For the fabric filters the constant N has a value of 0.005, while for the SDAs, caustic scrubber, WFGDs and DCCPSs the value is 0.015.

$$\begin{array}{l} Additional\ administrative \\ labour\ costs \end{array} = 0.03\ x\ \left(\begin{array}{l} Operating \\ labour\ costs \end{array} +0.4 \begin{array}{l} Maintenance,\ materials \\ and\ additional\ labour \end{array} \right)$$

Therefore, the greatest impact on the Fixed O&M costs is the number of operators and the size of each unit operations. Figure 34 shows the Fixed O&M costs with changing labour rate for a 550 MW plant. The slopes of the lines for each flowchart are determined by the number of operators utilised. As the fabric filters and DCCPSs have been assumed to utilise no extra operators, these costs are associated solely with the selection of a caustic scrubber, SDA or WFGD scrubber.

The intercept or position of the line on the figure is due to the size of each unit operationand the amount of sulphur in that stream, or the size and concentration of the gas stream being treated. The closer to the boiler that the recycle steam occurs, the smaller the costs associated with the unit operations after the recycle stream is removed. The C1 flowsheet is the cheapest because it uses the relatively cheap caustic scrubber and fabric filter. It does not require the DCCPS which increases the costs for the other flowsheets.

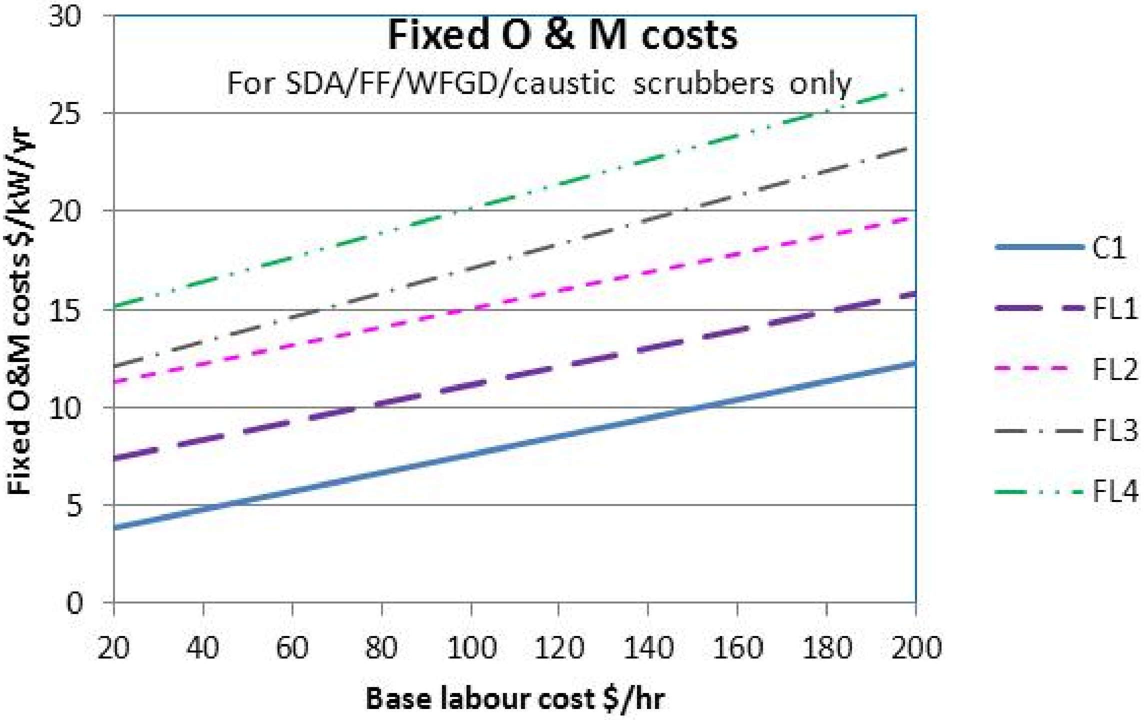

Figure 34: The fixed operating and maintenance costs for a 550 MW plant with changing labour cost for the flowcharts considered.

Plant size

As each of the flowsheets are the same for combustion of the coal to generate power but differ in the treatment of the off gas for sulphur removal and position of the recycle stream, the only costs considered where the costs associated with treatment of the flue gas. All other unit operations, such as the boiler, gas cooler, etc. are common to all of the flow charts and were not pursued for costs for different size plants.

Capital costs

The capital costs associated with the removal of sulphur have been considered here for plants of different size. Again the costs have been based on models produced by S&L[48-50] and for the caustic scrubber, SDAs and WFGD they are the sum of:

- Base absorber island cost
- Base reagent preparation cost
- Base waste handling cost
- Balance of plan cost (ID fan, boosters, piping, ducting elec.)

For the fabric filters, only one model is required for overall costs.

Each part of the models is set out below to identify the key variables.

Base absorber island cost

$$Base\ absorber\ island\ cost = N(A^{0.716})B(F.G)^{0.6}\left(\frac{D}{Z}\right)^{Y}$$

Where *N*, *Z* and *Y* are constants, shown in Table 35 below.

A is the effective plant unit size, equivalent to gas flows for an air fired plant,

B is the retrofit factor (taken to be 1 in this study),

F is the coal factor (1.05 for a semi bituminous coal),

G is the heat rate: BTU/KWh/10000 and

D is the SO_2 rate, lb/MMBTU

Table 35: The model constants for the base absorber island cost

Unit operation	***N***	***Z***	***Y***
SDA	566000	4	0.01
WFGD/caustic scrubber/ DCCPS	550000	2	0.02

However, this equation only applies to the SDA if *A*, the effective plant unit size, equivalent to gas flows for an air fired plant is less than 600 MW. If *A*>600 MW:

$$SDA\ Base\ obsorber\ island\ cost = A.92000$$

The sulphur rate, ie the amount of sulphur in the feed coal, and the effective unit size are the main contributors to the base absorber island cost.

Base reagent preparation costs and waste recycle / handling costs

$$Base\ reagent\ preparation\ cost = N(A^{0.716})B(D.G)^{M}$$

Where, once again, *N* and *M* are constants, shown in **Table 35** below.

A is the effective plant unit size, equivalent to gas flows for an air fired plant,

B is the retrofit factor (taken to be 1 in this study),

G is the heat rate: BTU/KWh/10000,

D is the SO_2 rate, lb/MMBTU.

Table 36: The model constants for the base reagent preparation and waste recycling or handling costs

Unit operation	***N***	***M***
SDA reagent and waste handling costs	300000	0.2
WFGD/ caustic scrubber/ DCCPS Reagent cost	190000	0.3
WFGD/ caustic scrubber/ DCCPS Waste handling cost	100000	0.45

Again, the SDA equation only applies if A<600 MW. If A>600 MW:

$$Base\ reagent\ preparation\ and\ waste\ handling\ cost = A.48700$$

As the coal heat rate is set in these studies, the significant variables are the effective plant unit size, A, and the SO_2 rate, D, which also increases with A.

Balance of plan cost

Balance of plan cost includes ID or booster fans, wet chimney, minor WWT, piping, ductwork and electrical items.

$$Balance\ of\ plan\ cost = N(A^{0.716})B(FG)^{0.4}$$

Where N is a constant as set out in **Table 37**,

A is the effective plant unit size, equivalent to gas flows for an air fired plant,

B is the retrofit factor (taken to be 1 in this study),

F is the coal factor (1.05 for a semi bituminous coal),

G is the heat rate: BTU/KWh/10000.

Table 37: The model constants for the balance of plan cost

Unit operation	***N***
SDA	799000
WFGD/ Caustic scrubber/ DCCPS	1010000

Again, the SDA equation only applies if A<600 MW. If A>600 MW:

$$Balance\ of\ plan\ cost = A.48700$$

The major variable in these calculations is the effective plant size, A.

Fabric filter cost

Sargent and Lundy[50] state that an air to cloth ratio of 4 is required unless the baghouse is installed after other collection devices, such as an ESP, when activated carbon injectant is used for mercury removal, when an air to cloth ratio of 6 is used. Therefore, for this study an air to cloth ratio of 4 was used. Capital costs for the fabric filters include:

- Duct work modifications,
- Foundations,
- Structural steel,
- ID fan modifications or new booster fans, and
- Electrical modifications.

$$Fabric\ Filter\ capital\ cost = 476.B.G^{0.81}$$

Where *B* is the retrofit factor (taken to be 1 in this study) and *G* is the flue gas rate (acfm).

The major variable in this calculation is the flue gas rate, *G*.

Varying the size of the plant changes the cost associated with its construction as shown in Figure 35 and Figure 36, where the capital costs of sulphur removal per kilowatt generally decrease with increasing plant size and the total capital cost for the equipment increases with plant size. The capital cost of the sulphur removal for flowsheet C1 is the cheapest of all the options considered. This is because this flowsheet has only one caustic scrubber and a fabric filter. The other flowsheets have either a SDA or a WFGD as well as a DCCPS and one or two fabric filters. However, the cost per kilowatt for this flowsheet is almost unchanged once the plant size is greater than 300 MW. For the other flowsheets, the cost per kilowatt continues to decrease with increasing plant size. Capital costs increase with increasing sulphur in the coal feed as the size of the unit operations increase to treat the increasing sulphur content of the flue gas, shown by flowsheets 1 to 4. FL1 uses the same coal, with the same sulphur content, as flowsheet C1 but has additional equipment, producing higher costs even though the recycle stream is only treated to remove particulates.

Flowsheets FL2 and FL3 burn the same coal with 1.5% sulphur, but the cost of the handling system for lime and waste from a SDA and the balance of plan costs is slightly cheaper than the handling system for limestone and gypsum and the balance of plan costs for the WFGD. The strange dip in the FL2 curve is due to the change in model when the plant size increases above 600 MW. Flowsheet FL4 is the most expensive in terms of capital costs for the treatment of sulphur because it burns a coal with the highest sulphur content, 3.5 wt% d.b. and it treats the flue gas prior to the recycle streams removal.

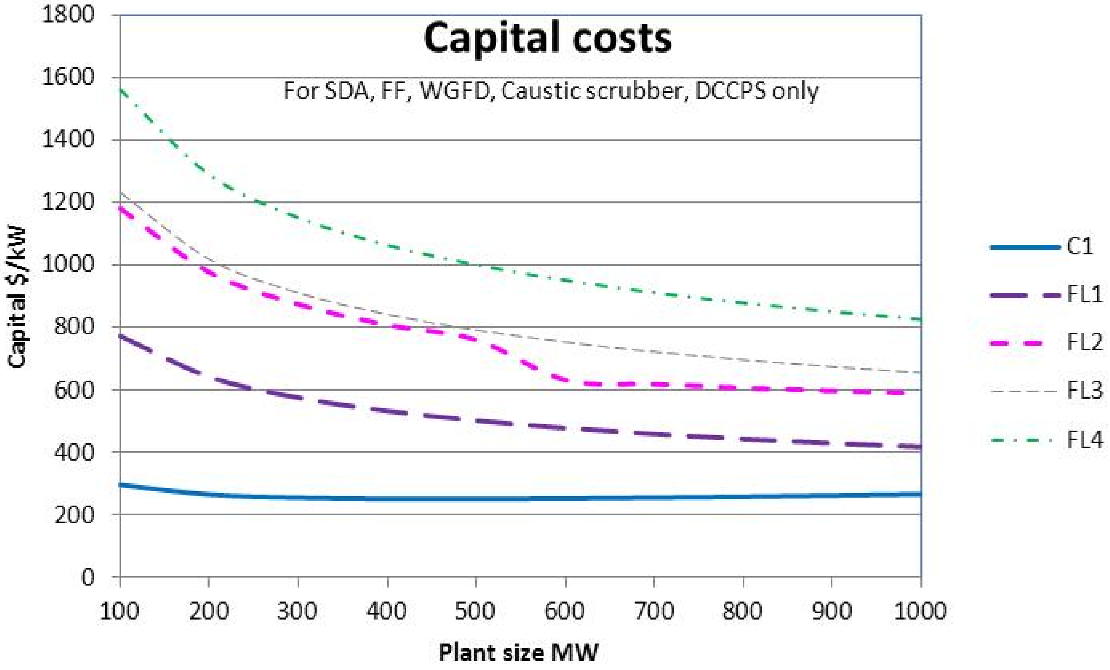

Figure 35: Capital costs per kilowatt for equipment associated with sulphur removal from the flue gas, for various plant sizes.

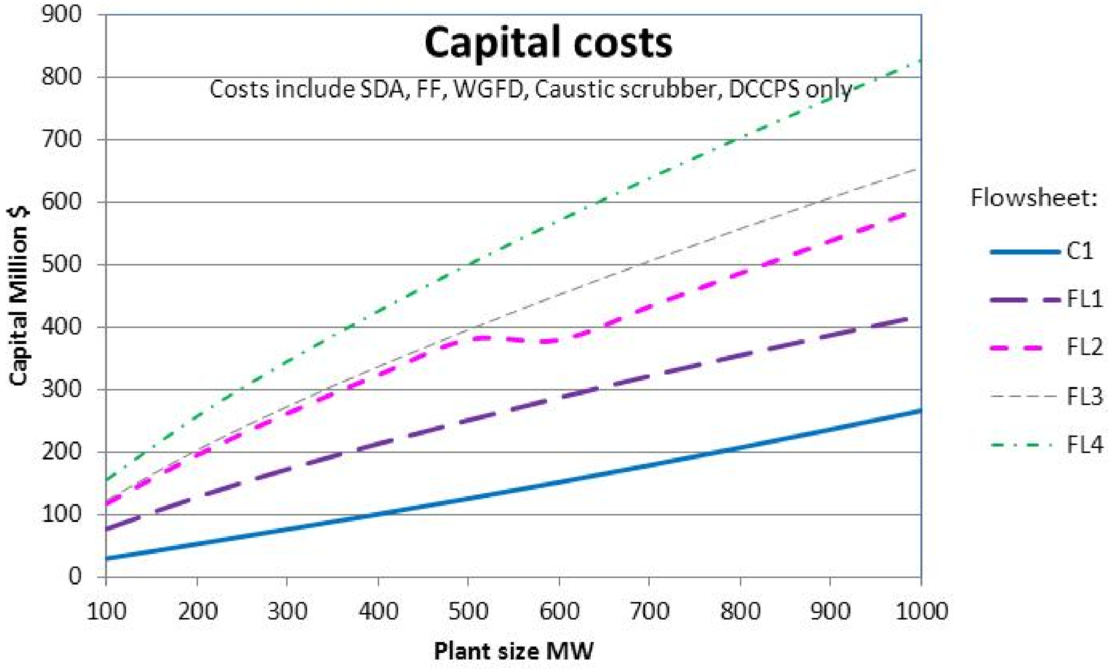

Figure 36: The total capital associated with sulphur removal from the flue gas for various size plants.

Fixed operating and maintenance costs

Calculation of the fixed operating and maintenance costs have been explained in Section 4 of this Appendix.

The impact of plant size on Fixed operating and maintenance costs for the sulphur removal from the flue gas in each of the flowsheets considered is shown in Figure 37 and Figure 38. As the plant size increases the cost of sulphur removal in $/kW/yr decreases, while the actual cost, in millions of dollars per year, increases.

In each of the flowsheets the cost of the fabric filters is very similar. The differences shown in the figures is due to the difference in size of the unit operation removing the sulphur, ie the SDA, WFGD and caustic scrubber, which is determined by the amount of sulphur in the coal. Therefore the flowsheet that uses high sulphur coal, FL4 is the most expensive. Flowsheets FL2 and FL3 both use the medium sulphur coal but FL3's WFGD requires 16 operators, while FL2's SDA only requires 12 for a 550 MW plant considered here. Flowsheets C1 and FL1 both require a low sulphur coal, but FL1's flowsheet has an additional unit operation in the DCCPS, making the flowsheet C1 cheapest in fixed operating and maintenance costs.

Figure 37: Fixed operating and maintenance costs in $/kW/yr for the sulphur removal in each flowsheet.

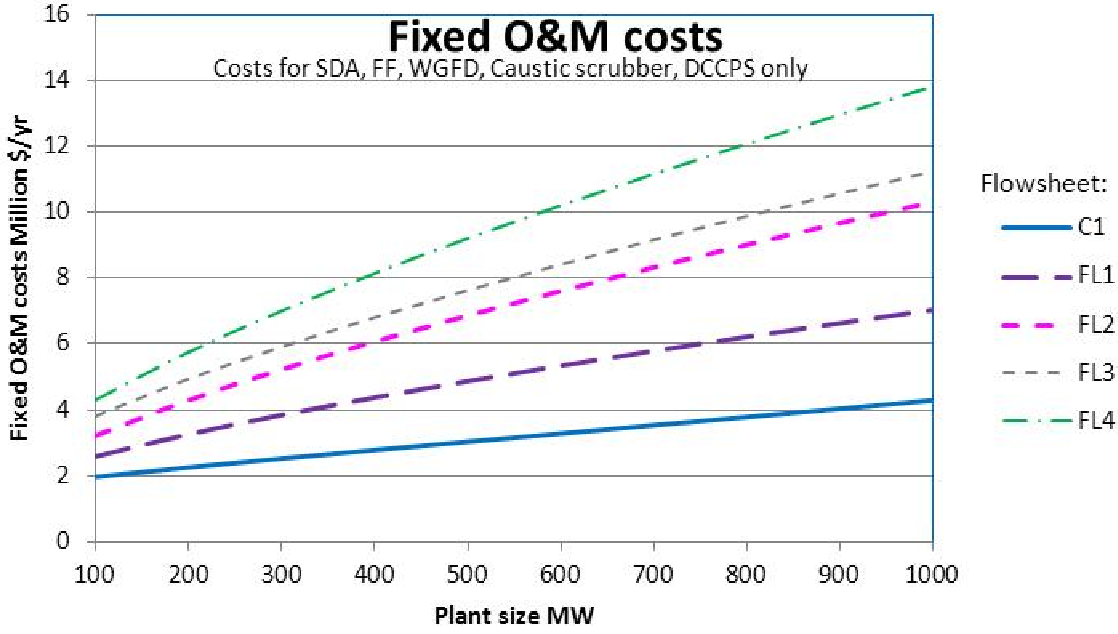

Figure 38: Fixed operating and maintenance costs in millions of dollars per year for the removal of sulphur in each flowsheet considered.

Variable operating and maintenance costs

Only the cost of the main consumable was considered in this analysis. No cost for waste disposal, or income from selling a by-product, such as gypsum, has been included. Neither was the cost of additional auxiliary power, makeup water or waste water treatment included.

The cost of the consumable was determined by the S&L model[48, 49] for SDA, WFGD, Caustic scrubber and DCCPS:

$$Variable\ O\&M\ cost = \frac{K.P}{A}$$

Where *K* is the required rate of caustic, lime or limestone, based on the mass balance outlined in Section 6,

P is the cost of caustic, lime or limestone. For this analysis the cost of caustic was set at $940/tonne (100%), lime was set at $95/tonne and limestone was set at $15/tonne. These were considered market rates at the time of analysis.

As above, *A* is the effective plant unit size, equivalent to gas flows for an air fired plant.

For the fabric filters, the variable operating and maintenance cost is the cost of replacement bags and cages. S&L's[50] model for this cost is:

$$Variable\ operating\ and\ maintenance\ cost = 0.005\left(\frac{K}{3}+\frac{L}{9}\right)$$

Where *K* represents the bag cost, $80/bag, and *L* is the cost of replacement cages, $30/cage.

The cost of the consumables increases with plant size but is constant for each MWh produced. As the cost of caustic is most expensive, flowsheet C1 has the highest costs for the consumables. Flowsheets FL2 and FL4 have higher costs for consumables because they have more sulphur to remove from the higher sulphur coals they consume, compared to FL1 and FL3 respectively. FL2, a flowsheet that utilises a medium sulphur coal, is more expensive than FL4, which uses a high sulphur coal because the cost of limestone, used in the WFGD is much less than that of lime, used in the SDA. FL1 has less sulphur to remove from the gas stream, as it uses lower sulphur coal compared to FL3 but it uses the more expensive lime, whereas FL3 uses limestone, and so the costs are similar.

The cost of the fabric filter consumables was the same for each fabric filters considered.

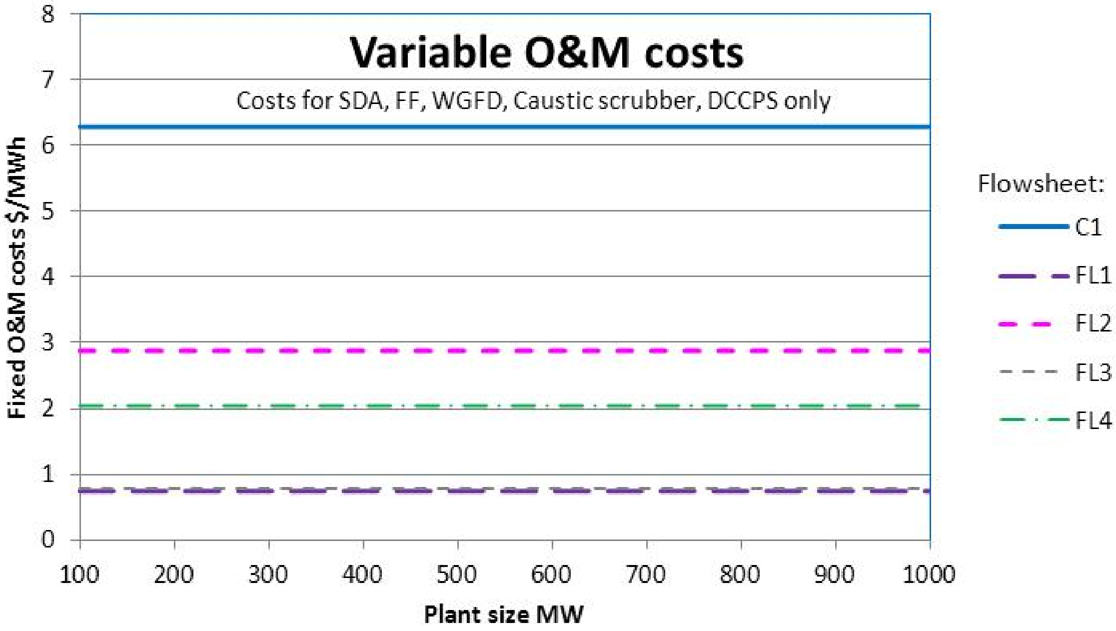

Figure 39: The variable operating and maintenance costs in $/MWh of the sulphur removal units in the flowsheets considered.

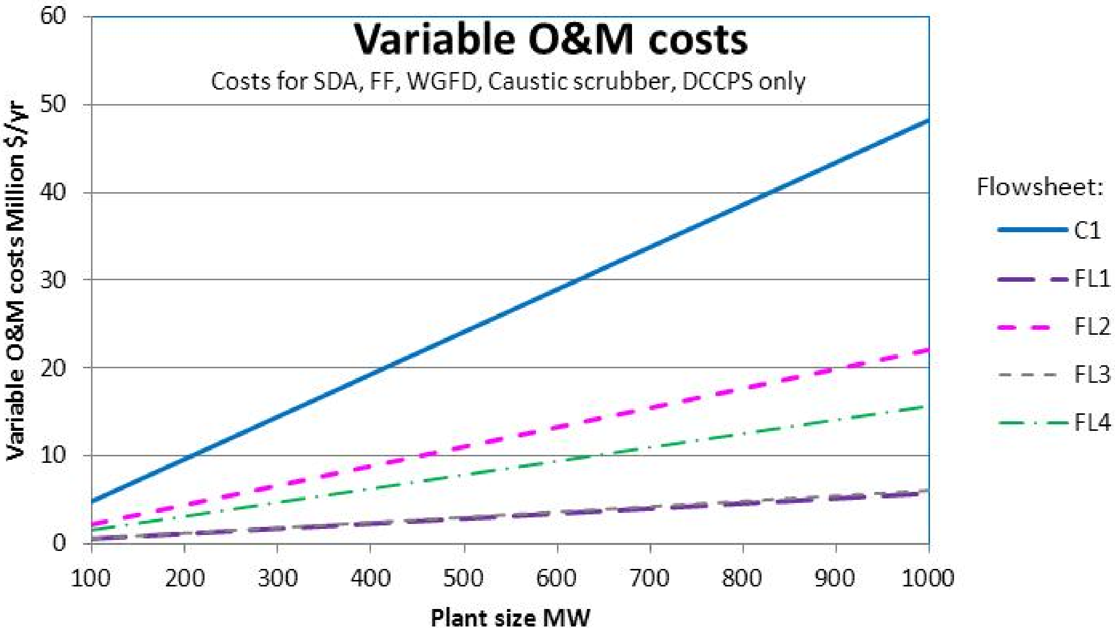

Figure 40: The variable operating and maintenance costs of the sulphur removing unit operations in each of the flowsheets considered, in millions of dollars per year.

Total Cost of Sulphur removal

If instalment costs on the capital invested in the sulphur removal equipment are calculated over 25 years, with no interest, the instalment and operating costs together for each flowsheet would decrease with increasing plant size, as shown in Figure 41. Due to the high cost of caustic soda required in flowsheet C1, this flowsheet is very expensive. Its costs become similar at large plant sizes to flowsheet FL4, which uses very high sulphur coal. Flowsheet FL1 is the most economic at all plant sizes. Given that flowsheets FL2, 3 and 4 are designed to use coals with higher sulphur contents, the cost of coal, not included in the calculations for the cost of sulphur removal, should be cheaper for these flowsheets. For a 550 MW plant, Flowsheet FL2 would require a cost saving of at least $6.75/t of coal to make this flowsheet economic compared to FL1. Flowsheets FL3 and FL4 would require $4.04/t and $9/t of coal respectively to match the costs of FL1 in a 550 MW plant.

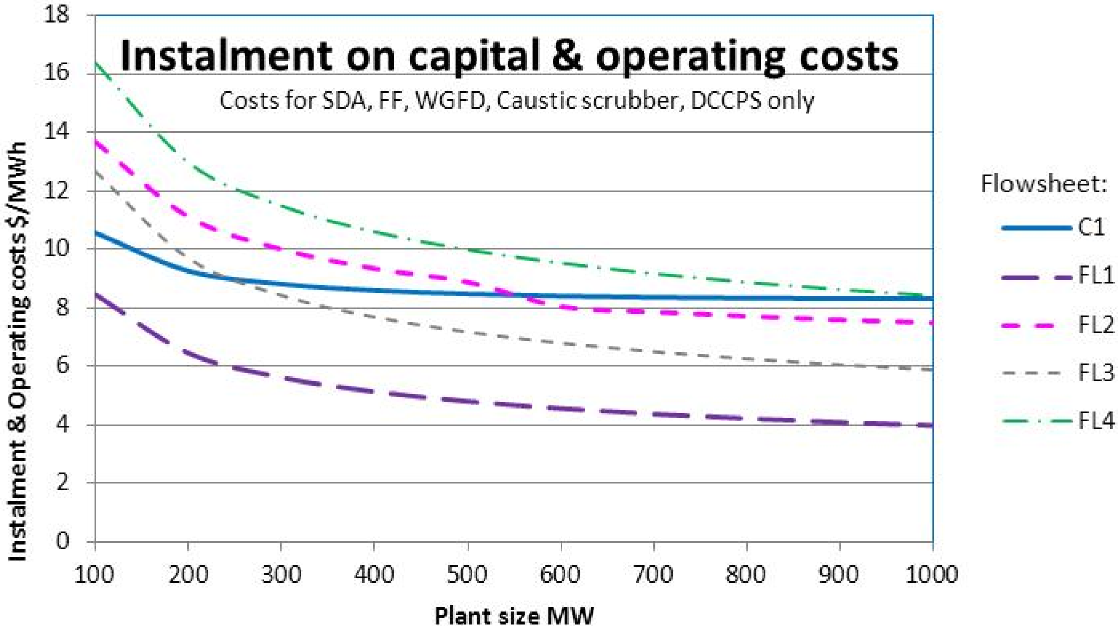

Figure 41: The instalment on capital and operating costs associated with the sulphur removal equipment for the flowsheets considered with changing plant size.

Conclusions

From this analysis it is clear that:

- Flowsheets C1, 2 and 3 are most affected by changes in NaOH costs.
- While large changes in lime cost can significantly affect the reagent costs for flowsheet FL2, and changes in the cost of limestone can impact significantly on the reagent cost for flowsheet FL4, the total cost of reagents for these two flowsheets is still significantly lower than the cost of reagents for flowsheets C1, 2 and 3 because NaOH, used exclusively in flowsheets C1, 2 and 3, is expensive compared to lime, used in FL2, and limestone, used in FL4.
- The cost of coal is the major single cost associated with operating such plants.
- Removing the recycle stream from the gas cleaning circuit, as in flowsheet FL1, reduces all costs for the flowsheet as the size of the unit operations in the gas cleaning circuit are reduced, and the amount of reagent needed in the removal of sulphur is reduced.
- Increasing the amount of sulphur in the coal increases the cost associated with its removal. Flowsheet 4 (the flowsheet designed to treat a high sulphur coal) is significantly more expensive in terms of capital costs and fixed O&M costs compared to flowsheets C1 and FL1 which were both designed to use low sulphur coal. However, Flowsheet C1 is vastly more expensive in variable O&M costs even though it burns low sulphur coal because it uses the very expensive NaOH as the reagent.

- Flowsheets FL1-4’s requirement of a DCCPS to ensure the sulphur concentration of the gas stream is below specification significantly increases the capital cost of gas cleaning in these flowsheets compared to flowsheet C1.

- Although the total capital cost increases with increasing plant size for all flowsheets, the cost per kilowatt decreases for all flowsheets except for C1, which remains approximately constant after 300 MW. The Fixed O&M costs also increase with increasing plant size, but on a cost/kW/yr basis the costs decrease with increasing plant size for all flowsheets. The variable O&M costs on a cost/kW/yr basis do not change with increasing plants size, while the actual costs increase with increasing plant size.

- Flowsheets FL1 and FL3 have similar variable O&M costs even though FL1 uses the more expensive lime compared to FL3’s limestone. Flowsheet C1 has the most expensive variable O&M costs.

In short, the cost of NaOH and flowsheet C1’s total dependence on NaOH makes this flowsheet more expensive than flowsheet FL1. The relatively low cost of limestone makes flowsheet FL3 the least expensive option for medium sulphur coals.