Impact of Impurities on CO$_2$ Capture, Transport and Storage

Report Number PH4/32
Date August 2004

IMPACT OF IMPURITIES ON CO_2 CAPTURE, TRANSPORT AND STORAGE

Background to the Study

The IEA GHG R&D programme has issued reports on a range of technologies which reduce or eliminate CO_2 emissions from large power plants and major industrial processes by capturing the CO_2 for geological storage. The focus of these studies has been on the cost and quantity of CO_2 captured but, to date, there has been a tacit assumption that geological storage will involve relatively pure CO_2. However there is increasing interest in so called zero emission technologies which involve the capture not only of the CO_2 component of flue gas but other components such as SO_2 and NOx. The main aims of this study were to establish the range of compositions which captured CO_2 could have if co-capture technologies were applied and the impact of these other components on the transmission and storage of CO_2. The implications for capture, transport and storage of CO_2 particularly when mixed with toxic gases are potentially more far reaching than for capture of CO_2 alone. Also success in capturing and storing CO_2 potentially opens up an easy pathway to capture of other gaseous emissions for long term storage. This possibility should be recognised early and the implications fully explored.

The case of full flue gas capture and storage is also evaluated for the sake of completeness although this proves, as expected, to be a rather unattractive option because of the large parasitic energy demand.

As well as establishing the levels of impurities which might be present in captured CO_2 the effects on the transport system are also of interest. This study addresses the principal effects from an engineering and technical viewpoint. The study is restricted to the use of a variety of coals and heavy fuels as these contain the greatest levels of potential contaminants.

Approach adopted

The study is based on evaluation of two types of coal fired plant which have different emission characteristics and opportunities for co-capture. One process is a conventional pulverised coal steam raising power plant and the other is an integrated gasification combined cycle plant (IGCC). Flow schemes and material balances for these basic plants were developed firing West Australian coal from the Drayton coal field. This is the standard coal which the IEAGHG R&D programme specifies for studies on coal fired plants. Emissions standards for these plants were based on those expected to prevail in Western Europe in the coming decade. Process schemes were then drawn up for the same basic plants modified to co-capture CO_2 and other components present in the flue gas. This was done on the basis that these components could be stored along with the CO_2 and so it would not be necessary to take specific measures to reduce their levels in the flue gas stream to meet current emission standards.

The heat and material balances for these base plants and modified co-capturing plants were then developed. The effects on these basic heat and material balances were then estimated when firing a range of different coals, Orimulsion, heavy fuel oil and petroleum coke. This enabled the envelope of possible contaminant levels for all relevant species to be determined thus defining the worst extremes. The effects of these extremes on compression, pipelines and the target storage reservoirs were then assessed. In the case of reservoirs no attempt was made to evaluate detailed geochemical effects. Such details would need a separate study and will vary from case to case. However the overall effects on EOR in particular were addressed.

Contractor was also asked to investigate what other co-capture process technologies might be employed and what the effect of these might be on types and amounts of impurities.
This work was entrusted to a consortium lead by a major engineering contractor SNC Lavalin, based in Calgary working with the Alberta Research Council supported by consultants from Vikor specialising in reservoir studies (particularly EOR) and University of Regina.

Results and Discussion

Range of fuels

The following nine fuels were considered in the study:-

- East Australian Bituminous (IEA GHG standard)
- High Sulphur Bituminous Coal from Eastern US
- Low Sulphur Sub-bituminous Coal from Western US
- High Ash Bituminous Coal from Chandrapur region of India
- Ultra-Low Sulphur "Enviro" Bituminous Coal from Indonesia
- Lignite from Germany
- Orimulsion– a mixture of bitumen and water from Venezuela;
- High Sulphur Heavy Fuel Oil
- High Sulphur Petroleum Coke

The sulphur levels in the selected fuels vary over a wide range as illustrated below and this element is responsible for the largest effects on contaminant species in CO_2 co-capture processes

Sulphur levels in fuels evaluated

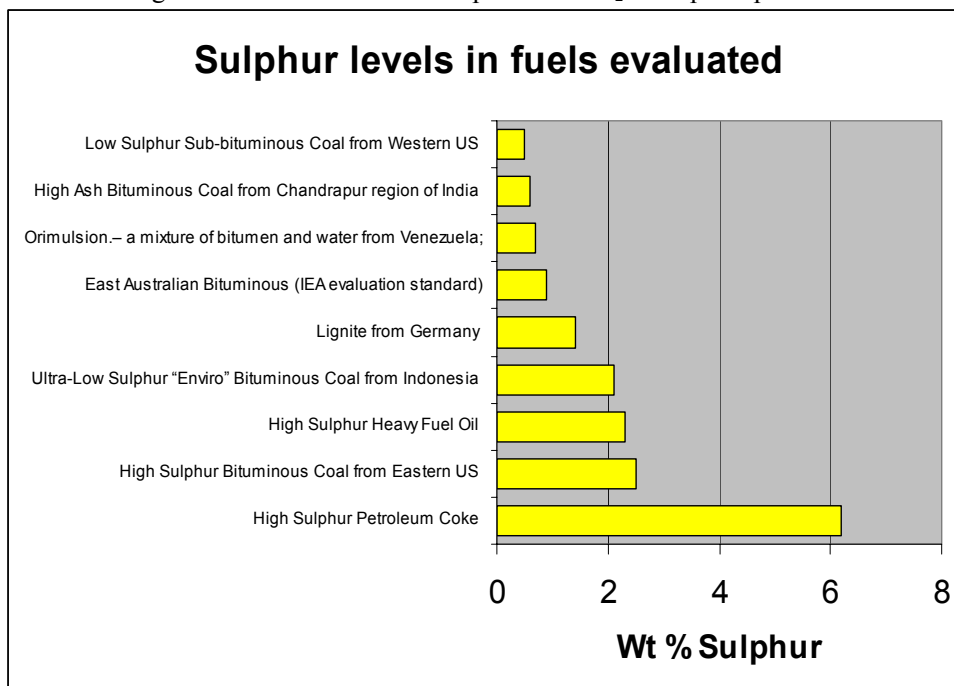| Fuel	
Low Sulphur Sub-bituminous Coal from Western US	
High Ash Bituminous Coal from Chandrapur region of India	
Orimulsion.– a mixture of bitumen and water from Venezuela;	
East Australian Bituminous (IEA evaluation standard)	
Lignite from Germany	
Ultra-Low Sulphur "Enviro" Bituminous Coal from Indonesia	
High Sulphur Heavy Fuel Oil	
High Sulphur Bituminous Coal from Eastern US	
High Sulphur Petroleum Coke	

Wt % Sulphur

Chart 1 Maximum sulphur levels

Coal contains many other trace elements, mostly heavy metals, some of which can contaminate flue gases in the form of the element or volatile compounds. Nine metals were considered to have the potential to contaminate co-captured CO_2. The figure below shows the maximum levels of these elements present in the selected fuels.
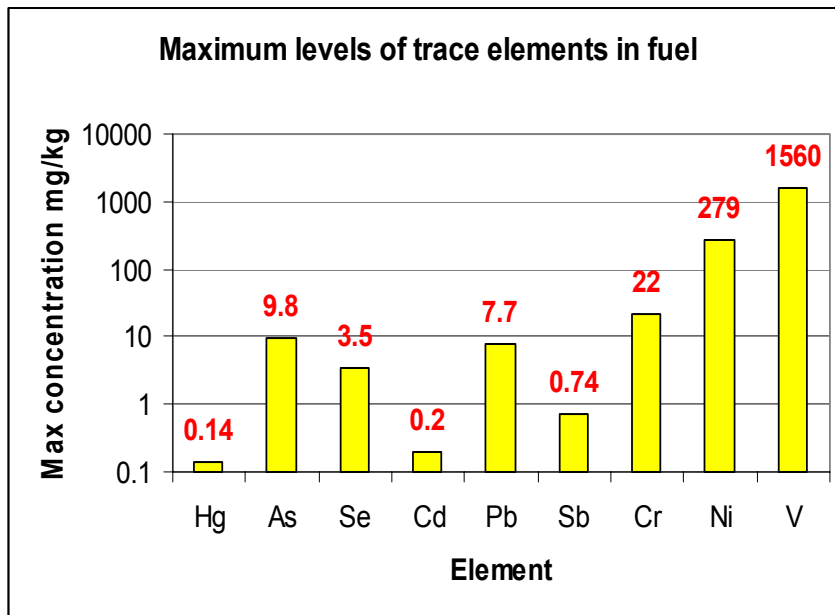
Chart 2 Maximum trace element levels

Baseline power generation processes

The baseline conventional pulverised fuel process included low NOx burners, Selective Catalytic Reduction (SCR) for NOx reduction, particulate removal with bag filters and flue gas desulphurisation (FGD). No mercury reduction step was included.

The baseline IGCC process was based on the Chevron Texaco wet feed gasifier. This leads to a plant with a lower cost of electricity than achieved using dry feed gasifiers (e.g. Shell). This choice was made so as not to overstate the relative economics of the co-capture process. It should be noted however that some low calorific value coals are not suitable for the wet feed process but this is not expected to affect conclusions about impurity levels. Sulphur is removed from the raw gas by amine absorption followed by conversion to elemental sulphur in a Claus plant.

In all cases the costs of CO_2 compression and 100km of CO_2 transportation pipeline were included but no allowance was made for injection well costs.

Capture processes in the baseline plant

For conventional pulverised coal plant amine scrubbing of the flue gases was considered to be the preferred process. Low NOx burners and SCR were retained in order to meet NOx specifications for the inlet of the amine absorber. In order to meet the SO_2 specification at the absorber inlet the FGD has to meet higher removal targets and additional capital allowances for this was made. It is noted that there is some uncertainty as to whether conventional FGD can meet the SO_2 reduction target of around 20ppm, although other recent work suggests that current wet limestone scrubbing technology can reach 10ppm.

The base case IGCC process with capture was taken as that described in the recent IEAGHG report PH4/19. A sour gas shift reactor is used to treat the raw syngas after which H_2S and CO_2 are removed *separately* in a Selexol® unit. Employing sour shift catalyst avoids the need to cool and remove H_2S prior to the shift reaction. The H_2S is then converted to sulphur in a Claus plant.

Selected co-capture processes

For the pulverised coal steam plant conventional FGD was omitted and replaced with a solvent based SO_2 recovery process in order to meet the more stringent SO_2 requirements upstream of the amine

absorber. (Cansolv® was assumed for the evaluation). This process recovers pure SO_2 which is then blended back into the captured CO_2. It is noted that, while regarded as the most suitable process combination, processes like Cansolv® have not yet been proven on a large commercial scale. For co-capture low NOx burners and SCR were still retained in order to meet NOx specs for the inlet of the CO_2 absorber as NOx is expected to pass through the Cansolv® step.

The optimum process for application to IGCC was determined to be a sour shift conversion followed by a Selexol® physical absorption unit designed to capture CO_2 and H_2S together. This saves installation of a sulphur recovery (Claus) plant and simplifies the acid gas absorption unit.

The cost of using these systems is described in a later section.

Other co-capture processes

Full flue gas capture
The total capture and storage of all of the flue gas stream from a conventional pulverised coal steam plant was evaluated in order to establish the efficiency and costs. The parasitic load for compression reduces overall efficiency to very low levels even when the injection site is close by. Furthermore the compressors required are beyond the size limits of current practice and manufacturers have concerns as to the deleterious effects of residual particulates on the compressors. Chart 3 illustrates the costs of CO_2 avoidance using full flue gas capture and storage. The efficiency for the base case, where CO_2 injection pressure is 70 bar, drops to 16.5% because of the large parasitic

Chart 3 Cost of full flue gas capture v injection pressure

power losses. Efficiencies for all other fuels are of similar order. The CO_2 avoidance cost at this pressure is approximately \$130/ton and power would cost US 15¢ /kWh
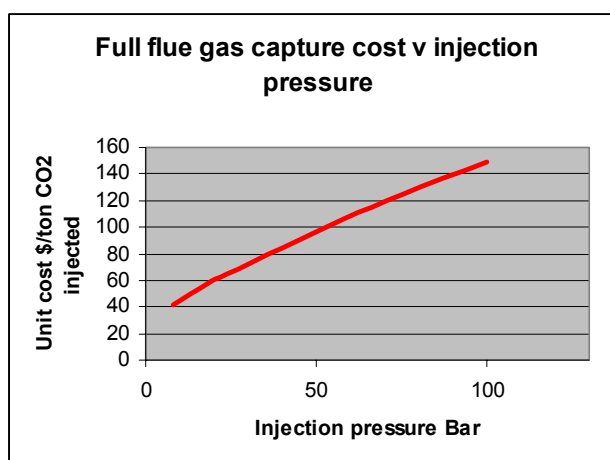
Oxy-combustion and other processes
The most competitive alternative to the baseline co-capture processes was found to be oxy combustion. Accordingly estimates were made of the composition for co-capture from such a plant. The incondensable gases O_2, N_2 and Ar are presumed to be removed from the captured CO_2 in this process leaving SO_2 and NOx as the main impurities. A more detailed study of oxyfuel processes is in preparation by the programme.

Several other processes were considered, such as gas separation membranes and cryogenics but these did not appear to have much chance of being economically viable. Use of gas separation membranes to produce a partially enriched CO_2 stream was analysed in some depth as this defines another potential range of composition between fluegas and concentrated CO_2. However parasitic power consumption was found to dominate in the same way as it does for full flue gas capture. CO_2 avoidance costs were in excess of \$100-110/ton CO_2 and power costs US 10-12¢ /kWh depending on the selected fuel.

Impurities likely to be present in co-captured CO_2.

The most important impurity expected in co-captured CO_2 was sulphur either in the form of:

- H_2S -from IGCC plants which use pre-combustion capture
- SO_2 -from conventional steam plants which use post combustion capture.

Other significant impurities expected are :

- NOx
- Hydrogen – only from IGCC
- CO - only from IGCC

Other contaminants likely to be present are:

- Nitrogen
- Oxygen
- Argon

These will be in small quantities for the base case processes where they enter as a result of physical absorption in the scrubbing processes. As noted above they are presumed to be removed for the most part in oxycombustion processes. The exception is membrane processes which would produce enriched flue gas with around 50% nitrogen and 1-2% oxygen.

Several trace elements in coal are emitted in small quantities in stack gases of conventional plants. Of these the most significant at present is Mercury for which tightening emission regulations are expected. When co-capture is applied only three of the trace elements are expected to appear in the CO_2 stream:-

- Mercury
- Arsenic
- Selenium

It is not possible to calculate the retention of trace elements in a systematic way as their behaviour in the relevant processes is not fully understood but rough estimates have been made.

Maximum levels of contaminants.

The following charts show the maximum levels of each impurity which could be expected when operating the worst combination of fuel type and process. As such they represent the maximum levels for which approval is likely to be sought for single projects. These maximum levels cannot and will not occur all at the same time but do define the extremes of the composition envelope. Compositions for each fuel in each of the processes considered are given in the main report.

If CO_2 capture and storage were to be carried out on a large scale it is likely that common collection and transport systems would be developed. CO_2 from many different sources might then be mixed in the transport systems and suitable composition specification envelopes would be required.
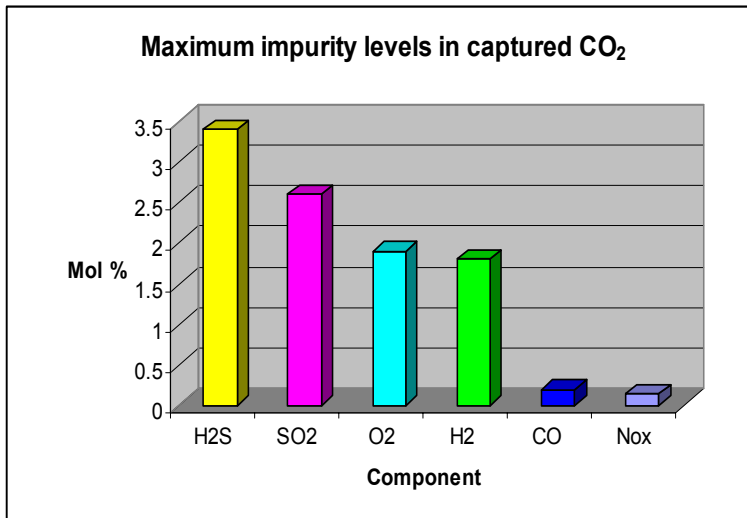
Note that the maximum level of oxygen shown in this chart is that associated with the membrane process which only partially concentrates the captured CO2. In all other processes the level is much lower
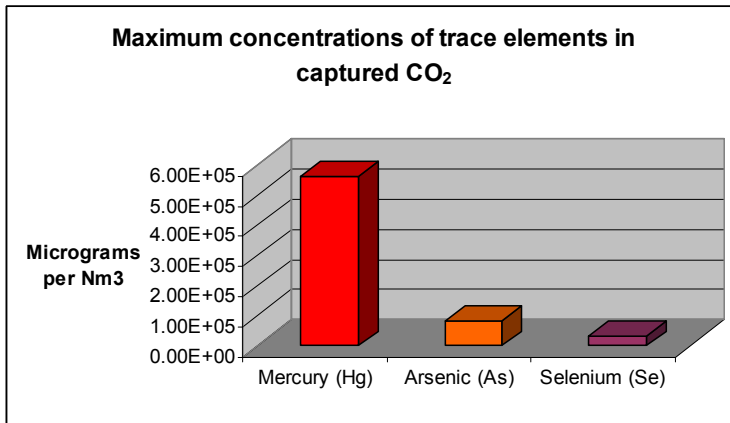
Chart 4 Maximum levels of impurity in co-captured CO$_2$

Chart 5 Maximum levels of trace elements in co-captured CO$_2$

One important potential conflict has been identified, that between H$_2$S and SO$_2$ containing streams. These result from processes in which CO$_2$ is captured in "reducing" or "oxidising" environments respectively. If mixed these components could undergo the Claus reaction to form sulphur which might cause equipment blockages or plug reservoirs. At ambient temperatures in the gas phase the reaction rate is negligible. However surfaces in equipment or reservoirs might catalyse the reaction and further research is recommended. Otherwise there would seem to be no undesirable consequences of mixing co-captured streams apart from well documented effects on toxicity and general corrosion.

Mercury is the only trace element expected to occur in significant concentrations in co-captured CO$_2$. However the levels are well below those which would cause any concerns in pipeline transmission and geologic storage.

Effects on transmission systems

The main effect of co-capture on the transmission system is the introduction of the highly toxic compounds H$_2$S and/or SO$_2$. At levels of 1 or 2 % in pure CO$_2$ both these compounds dominate toxicity of leaks. This is illustrated by the fact that pure CO$_2$ is rendered harmless after dilution 20 times the amount of air whereas to reach the same effect with 1% of these gases present would require 500 to 1000 times dilution. When these compounds are present operators will require personal monitors and processing and compression facilities will need a toxic leak detection system. Pipelines will need state of the art leak detection and to be sectionalised with emergency shutdown valves placed so as to limit the volume of gas leaking from any one section according to the proximity to populated areas. The pipeline

operator will have to set up emergency evacuation programmes which would be required to some extent with pure CO_2 lines but would have to be much more extensive when H_2S or SO_2 are present. It was estimated that for the basecase 100km transmission line costs would be \$1-2million higher as a result.

There are no serious additional corrosion issues as long as the co-captured gases are adequately dried. For higher levels of H_2S and also when Hydrogen is present pipeline steel with limited hardness may have to be selected.

Effects on subsurface storage.

No significant effects on the integrity of geologic storage are expected. Detailed evaluation of possible effects on reservoir rock and cap rock would require extensive testing to confirm this and would need to be reservoir specific. Such evaluation is beyond the scope of this study.

Where storage is accompanied by enhanced production either of gas or oil use of co-captured CO_2 does raise some issues. One is the injection of sour gas into a sweet[1] reservoir thus rendering the reservoir sour. Depending on how sour the reservoir becomes there will be effects on the production equipment. For example extra safety precautions or changes to materials of construction might be required. The industry has extensive experience in handling sour operations and has the capacity to make appropriate technical and financial judgements on this issue.

Another key effect on enhanced oil recovery operations is the change to minimum miscibility pressure (MMP) when gases other than CO_2 are present but only in schemes which rely on the miscibility effect. One theoretical correlation included in the main report shows that SO_2 significantly and H_2S somewhat reduces MMP. Maximum levels of SO_2 could thus reduce MMP by as much as 15% and the H_2S by about 5% thus reducing compression requirements. Lighter gases such as nitrogen or oxygen increase MMP but up to levels of 5% these effects are expected to be small.

If used for Enhanced coal bed methane recovery (ECMB) contamination of either H_2S or SO_2 are likely to have significant effects on the design of production wells and surface production equipment against corrosion after breakthrough because such coal seams do not normally contain these gases. From a theoretical viewpoint both H_2S and SO_2 are expected to preferentially absorb on coal in the same way as CO_2. Experiments would be needed to quantify and confirm this.

The economics of subsurface storage will be adversely affected by the presence of lighter contaminants such as nitrogen or argon simply because these will occupy pore space thus reducing that available for CO_2

Cost advantages of co-capture.

For conventional steam power plant with amine capture of CO_2 there appears to be some scope for cost reduction with co-capture particularly for high sulphur fuels. This relies on the fact that FGD has to be designed to reduce SO_2 to the low levels which are not harmful to the amine absorption system. This step (for which CANSOLV® is a leading contender) can, for little extra cost, perform the entire FGD step, easily meet the specification and deliver a pure stream of SO_2 for blending back into the CO_2. The saving varies with sulphur content and ranges as follows:

Savings from co-capture – conventional steam boiler plant – injection site 100km away

Fuel	Standard Coal	6% sulphur Petroleum coke
Power cost reduction US ¢/kWh	US 0.3 ¢/kWh	US 1.6 ¢/kWh
CO_2 avoidance cost reduction \$/ton	6 \$/ton	16 \$/ton

---

[1] "Sweet" and "sour" are oil industry terms referring to whether or not a reservoir contains significant amounts of the acid gas hydrogen sulphide

There are also significant savings to be made when co-capture is practised in combination with IGCC because it is possible to use a physical solvent process able to extract H_2S and CO_2. together. This saves on the costs of sulphur recovery and sulphur handling plant. However plant electrical efficiency and hence saving is reduced due to loss of the fuel value of co-captured H_2S and the increased co-absorption of hydrogen in the Selexol® solvent

Savings from co-capture IGCC – injection site 100km away

Fuel	Standard Coal	>3% sulphur fuels
Power cost reduction US ¢/kWh	US 0.3 ¢/kWh	US 0.7-0.8 ¢/kWh
CO_2 avoidance cost reduction $/ton	4 $/ton	8-12 $/ton

The relative magnitude of the reduction in CO_2 avoidance costs is illustrated in charts 6 and 7
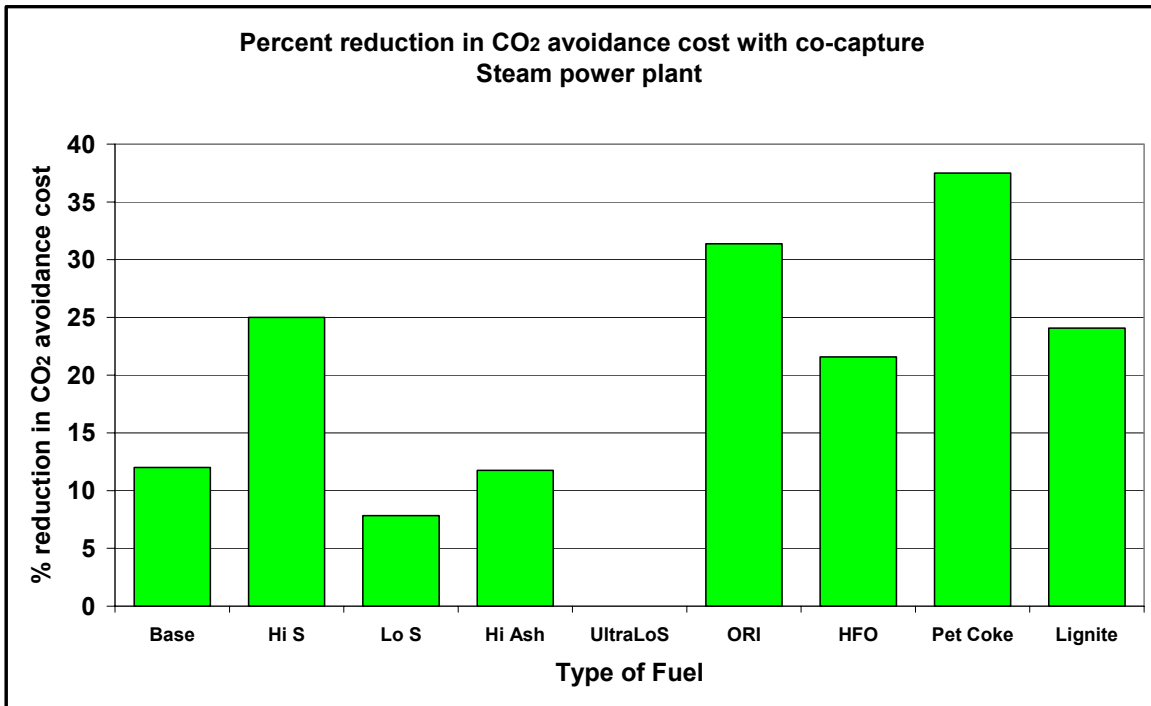

Chart 6 Reduction in CO₂ avoidance costs due to co-capture- Steam power plant

Percent reduction in CO₂ avoidance cost with co-capture IGCC power plant
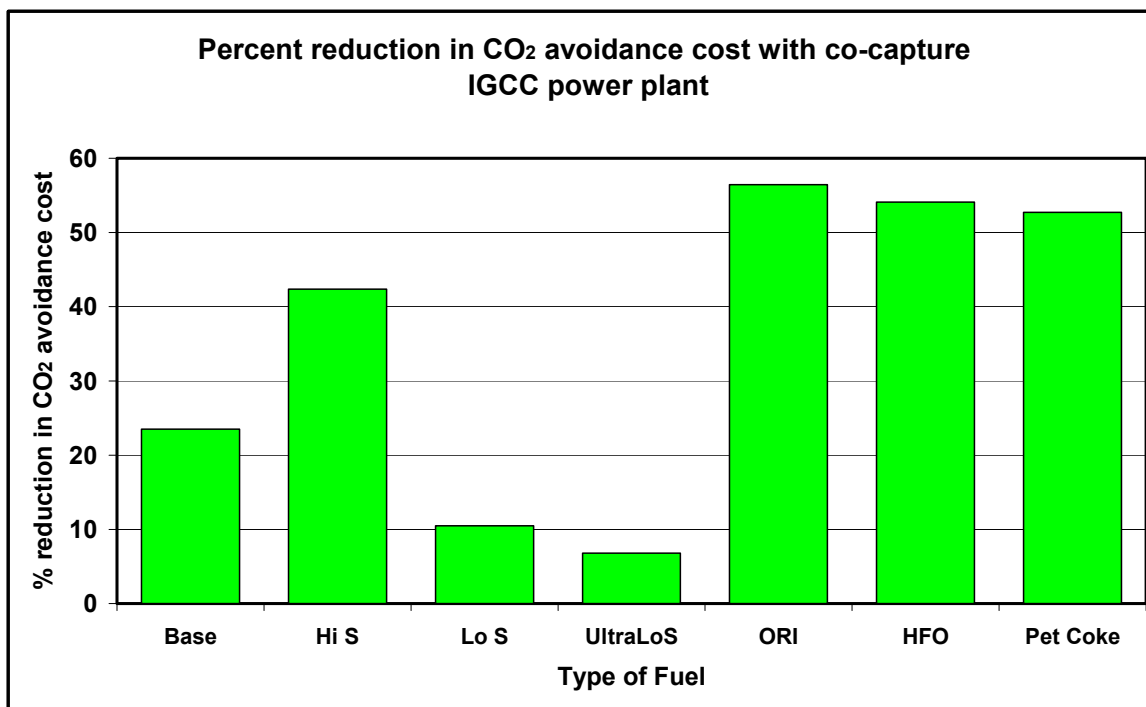

Chart 7 Reduction in CO₂ avoidance costs due to co-capture - IGCC power plant

Expert Reviewers' Comments

Reviewers felt that more detailed and precise information should be provided about the behaviour of impurities in the capture and compressions steps. Detailed simulations of this nature are outside the scope of this report, for trace elements of this type prediction is difficult. Reviewers felt also that more information could be provided about the interaction of impurities in saline aquifers. Again this is a highly specialised area and for specific reservoir rock is likely to require experimentation. The draft report as reviewed was compiled from contributions from several organisations and improvements to the integration and presentation of the information were suggested. Some useful technical points were raised, for example the possible importance of the odour, which some trace contaminants exhibit, on the acceptability of pipeline transport. It was also suggested that more should be said about the effect of co-capture on simplification of CO_2 purification in oxy-combustion processes. As the scope did not cover detailed simulation in this specialised area this is not covered in detail in this report. An IEAGHG study on oxy-combustion is in preparation and will provide more insight into this issue.

Major Conclusions

There are small but significant cost savings through use of co-capture processes and slight simplifications to the overall power generation process. The potential for savings rises, as might be expected, for higher sulphur fuels.

Co-capture would introduce H_2S and/or SO_2 into captured CO_2 at levels which could have serious consequences for the approval of transport pipelines and storage sites. In contrast to the transport and storage of CO_2 the presence of these components introduces a higher level of risk because of their toxicity. Also small leaks would have environmental consequences because of the low smell threshold of these compounds. From the engineering point of view these risks can be countered at a small extra cost through additional safety features. However co-capture needs to be considered in the context that extensive transport networks crossing populated areas might be necessary for large scale application of CCS technology. It could be significantly more difficult for countries or regions to make the decision to

go down a path leading to widespread transport of toxic gas. It is difficult to predict to what extent such plans would be accepted.

Co capture and storage is not expected to have significant effects on the geologic storage process. However the introduction of sulphur compounds into sweet reservoirs of any type has potential downsides. For any enhanced recovery applications the souring of the reservoir will have implications for the design and operation of the surface facilities. For other reservoirs the burden of proof of integrity of the storage site will increase because of the higher toxicity of the material to be stored.

Recommendations

Include an analysis of the option to co-capture other components and make clear decisions on the subject, whenever planning new CO_2 transport and geologic storage systems.

Extend the scope of research into geochemical interactions between CO_2 and reservoir rocks to include effects of co-captured components, particularly H_2S and SO_2.

Specifiy and select steel materials of construction for pipelines, fittings and wells destined for captured CO_2 service after a careful analysis of the extent to which co-captured components H_2S and/or Hydrogen may be present at some future date. Check what the extra cost will be and consider using the better materials especially if the extra costs are insignificant. Steels which have the necessary hardness and impact resistance are often available at the similar prices.

Cover the consequences and desirability of co-capture in all activities which aim to adopt CCS as one of the alternatives for reduction of GHG emissions to atmosphere, unless it has become clear that co-capture is no longer an option.

Include co-captured components in ongoing work on the how CO_2 is treated as a waste in applicable regulations, laws and treaties.

NOTICE to READER

This document contains the expression of the professional opinion of SNC-Lavalin Inc. ("SLI") as to the matters set out herein, using its professional judgment and reasonable care. It is to be read in the context of the agreement dated May 21, 2003 (the "Agreement") between SLI and IEA Environmental Projects Ltd. Greenhouse Gas R&D Programme (the "Client"), and the methodology, procedures and techniques used, SLI's assumptions, and the circumstances and constrains under which its mandate was performed. This document is written solely for the purpose stated in the Agreement, and for the sole and exclusive benefit of the Client, whose remedies are limited to those set out in the Agreement. This document is meant to be read as a whole, and sections or parts thereof should thus not be read or relied upon out of context.

SLI has, in preparing the document, followed methodology and procedures, and exercised due care consistent with the intended level of accuracy, using its professional judgment and reasonable care, and is thus of the opinion that there is a high probability that the outcome will fall within the specified error margin. However, no warranty should be implied as to the accuracy of estimates. Unless expressly stated otherwise, assumptions, data and information supplied by, or gathered from other sources (including the Client, other consultants, testing laboratories and equipment suppliers, etc.) upon which SLI's opinion is based, has not been verified by SLI, and SLI makes no representation as to its accuracy, and disclaims all liability with respect thereto.

SLI disclaims any liability to the Client and to third parties in respect of the publication, reference, quoting, or distribution of this report or any of its contents to and reliance thereon by any third party.

EXECUTIVE SUMMARY

SNC-Lavalin Inc. (SLI) was awarded a study by the International Energy Agency Greenhouse Gas R&D Programme to evaluate the benefits and issues associated with the combined capture and transport of all components of flue gas for geological sequestration. Previous studies have focused on CO_2 only, and it is thought that there could be advantages to capturing other components such as SOx and NOx along with the CO_2. It is widely reported that capture of CO_2 from large stationary sources may be a way to achieve the deep reductions in CO_2 emissions specified under the Kyoto Protocol.

The mandate of the study was as follows:

❑ Set up base cases for conventional boilers and gasification systems;
❑ Identify the impurities that could be captured along with CO_2 when firing a range of coals and other fuels;
❑ Identify the most promising co-capture alternatives and the relevant impurities in each case;
❑ Identify compression, transport, and storage issues.

SLI evaluated the following base case scenarios:

❑ Supercritical boiler with no capture of CO_2, and with capture and injection of the entire flue gas stream;
❑ Integrated Gasification Combined Cycle (IGCC) plant with no capture, and with conventional CO_2 capture.

Our analysis was based on a 750 MW output for each system before capture. A Texaco slurry-fed gasifier was used for the IGCC cases. For the impact on storage options, we focused on Enhanced Oil Recovery (EOR), Enhanced Coalbed Methane recovery (ECBM), and storage in depleted oil and gas reservoirs and deep saline aquifers. Six coals were evaluated, as well as a heavy fuel oil, high sulphur Pet Coke and Orimulsion®. Table 1 summarizes the study results.

Overall some promising co-capture options were identified. For conventional boilers, the co-capture of SO_2 along with CO_2 could offer cost savings compared to capture of CO_2 only, due to the elimination of costly sulphur reduction equipment. For gasification systems, our report confirms that the co-capture of H_2S and CO_2 appears promising. Co-capture of sulphur species could be particularly promising for high-sulphur fuels. In general, trace elements are unlikely to be present in any of the co-captured streams at significant levels, due to condensing and removal in successive compression stages.

Further to the above, SLI makes the following conclusions on the benefits and issues of co-capture of other flue gas components in addition to CO_2:

Capture of Entire Flue Gas Stream

❑ Capture of the entire flue gas stream is not practical and results in significantly higher capture costs and parasitic power losses. The captured flue gas would not be acceptable for EOR, and would not be practical for storage in depleted oil and gas reservoirs. One potential application could be ECBM projects close to the capture source;

Co-Capture of SO_2 from Conventional Boilers

❑ There could be an opportunity to co-capture SO_2 using either a Cansolv® or similar system, or via oxygen-enriched combustion ("Oxy-fuel"). Either option results in a CO_2 stream with 0.5%-3.0% SO_2. Savings for a 2% sulphur coal compared to conventional CO_2 capture using either alternative, could be on the order of $13-25/t CO_2 avoided and $0.010/kWh.

❑ Relative savings for either alternative increase as the sulphur content of the fuel increases. Savings with the 6% sulphur Petroleum Coke were estimated to be $20-32/t CO_2 avoided and $0.016-0.023/kWh.

❑ Both alternatives are technically feasible but neither have been demonstrated at the scale required for a 750 MW power plant. Oxy-fuel has been demonstrated at a pilot-scale only;

❑ The above streams could be acceptable for EOR in certain reservoirs, and would likely be acceptable for ECBM. There are concerns over conversion of SO_2 to sulphur in reservoirs containing H_2S. These streams must be dehydrated well below the dew point of water prior to transportation to minimize corrosion;

Co-Capture of H_2S from Gasification Systems

❑ There could be an opportunity for co-capture of H_2S during gasification via a sour shift and recovery of H_2S and CO_2 using Selexol® or similar process. This results in a CO_2 stream with 0.5-3% H_2S and 2% hydrogen. A disadvantage is that the energy value of the H_2S and hydrogen is lost;

❑ Savings relative to "CO_2-only" capture were estimated to be $4/t CO_2 avoided and $0.003/kWh for the base coal. Savings appear to increase as the sulphur content of the fuel increases. Savings with the 6% sulphur Pet Coke were estimated to be $12.5/t CO_2 avoided and $0.007/kWh;

❑ The above streams would be acceptable for EOR in reservoirs that already contain H_2S. There are significant concerns over injection of H_2S into a "sweet" EOR reservoir, or for use of a stream containing H_2S for ECBM;

❑ There are serious but not insurmountable safety concerns regarding transport of capture streams containing H_2S, as it is extremely toxic. Gases containing higher levels of H_2S are transported near populated areas in North America, but there could be public resistance in areas not used to the practice, particularly in the EU;

❑ We would expect to see similar safety concerns raised for transport of capture streams containing SO_2, which is also toxic;

Co-Capture Using Membranes

❑ Opportunities for co-capture using gas separation membranes do not appear promising. Estimated capture costs for this alternative were about \$120/t on average for the fuels studied, based on an inlet pressure of 18 bar. The results are very sensitive to the inlet pressure. A capture stream of about 50% CO_2 and 45% nitrogen with 1-2% oxygen and 1-2% SO_2 could be produced. This stream would not be acceptable for EOR but could be acceptable for ECBM;

❑ Capture of CO_2 from coal-fired boilers using gas separation membranes may have significant technical concerns with fouling from sub-micron fly ash particles.

Conventional Capture Using FGD-Amines Systems

❑ Previous studies on CO_2 recovery using a conventional FGD-amine system may be side-stepping the difficulty of meeting very low SO_2 levels (20 ppm) using conventional FGD, particularly with high sulphur fuels.

Trace Elements

❑ Trace elements are not expected to be found at significant levels in the co-capture streams reviewed as most metal compounds, including mercury, are removed during particulate removal, or else during cooling, compression, and dehydration, or exit via the rejected gases. The highest level of trace elements may occur in the Oxy-fuel capture stream, as this system has the fewest processing steps;

Other Transportation Issues

❑ It is likely that for a common pipeline network, CO_2 specifications will evolve similar to natural gas distribution networks. This could limit co-capture applications to situations where capture sources are supplying dedicated storage locations;

❑ The regulatory process should not be an impediment to transport activities. We anticipate regulatory bodies will treat SO_2/CO_2 streams similar to H_2S/CO_2 streams, for which there are well established appproval procedures and precedents in North America.

Recommendations:

❑ More detailed evaluations of co-capture of SO_2 using Cansolv® or Oxy-fuel is required to confirm the potential savings relative to conventional FGD-amines systems. Sensitivity of fuel prices to sulphur levels should be included in the analysis;

❑ The impact of SO_2 or H_2S on EOR and other storage options such as deep saline aquifers needs to be studied further;

❑ SO_2 removal capability upstream of amine-based CO_2 recovery systems should be investigated further, particularly for high sulphur fuels;

❑ Development of emerging oxygen production technologies should be monitored, as lowering the cost of oxygen appears to be the biggest challenge in oxy-fuel systems.

Table 1

Summary Results for Each Case
Using IEA Base Coal (C1 - E. Australian Bituminous, 0.86% Sulphur)

	Conventional Boiler Cases						Gasification Cases		
	C	**D**	**E**	**F**	**G**	**H**	**A**	**B**	**B1**
	Flue Gas	**No**	**CO_2 Only**	**CO_2+SO_2**	**CO_2+SO_2**	**Gas Sep.**	**No**	**CO_2 Only**	**CO_2+H_2S**
Parameter	**Capture**	**Capture**	**(Ref.)**	**Cansolv**	**Oxy-fuel**	**Mem.**	**Capture**	**Selexol**	**Selexol**
Gross Power Output, MW	750	761	761	750	750	750	989	973	980
Net Power Output, MW	292	750	551	547	533	421	827	730	742
Total Capital Cost, $ millions	1,350	830	1,120	1,070	1,010	1,510	980	1,090	1,050
Specific Capital Cost, $/kW	4,620	1,110	2,030	1,960	1,890	3,590	1,190	1,490	1,420
Power Cost, $/kWh	0.142	0.040	0.074	0.071	0.066	0.109	0.046	0.057	0.054
Plant Efficiency, %	16.5%	41.8%	30.7%	30.9%	30.1%	23.8%	38.0%	31.5%	32.0%
$/t CO_2 Avoided	128	n/a	50	44	33	118	n/a	17	13
CO_2 Emitted, kg/MWh	-	790	110	110	22	210	830	150	150
Capture Stream CO_2 %	16.0%	n/a	99.9%	99.4%	99.6%	53.1%	n/a	97.4%	97.2%
Primary Impurities	$N_2 + O_2$	n/a	none	SO_2	SO_2	$N_2, O_2,$ SO_2, NOx	n/a	none	$H_2S + H_2$

IEA GREENHOUSE GAS R & D PROGRAMME

IMPACT of IMPURITIES on CO$_2$ CAPTURE, TRANSPORTATION and STORAGE

TABLE OF CONTENTS

LIST of FIGURES

GLOSSARY

ARC	Alberta Research Council Inc.
ASU	Air Separation Unit
BFD	Block Flow Diagram
EOR	Enhanced Oil Recovery (specifically using CO_2)
ECBM	Enhanced Coal Bed Methane recovery (using CO_2 or other gases)
FGD	Flue Gas Desulphurization (generally using limestone or similar)
IEA GHG	IEA Greenhouse Gas R& D Programme
MEA	MonoEthanolAmine (chemical used in CO_2 recovery)
MMP	Minimum Miscibility Pressure
PSA	Pressure Swing Adsorption
SCR	Selective Catalytic Reduction (removal of NOx from flue gas)
TSA	Temperature Swing Adsorption

1.0 INTRODUCTION and STUDY METHODOLOGY

1.1 Introduction

The International Energy Agency Greenhouse Gas R&D Programme (IEA GHG) was established in 1991 to evaluate technologies that could be used to avoid emissions of greenhouses gases, particularly from fossil fuels, and to identify targets for useful research and development into this issue. IEA GHG is an international organization, supported by eleven countries, the European Commission and a number of industrial organizations and companies.

SNC-Lavalin Inc. (SLI) was awarded a study by the International Energy Agency Greenhouse Gas R&D Programme (IEA GHG) to evaluate the benefits and issues associated with the combined capture and transport of all components of flue gas for geological sequestration. Previous studies have focused on the capture and transport of CO_2 only, and it is thought that there could be advantages to co-capturing other components such as SOx, NOx and mercury as well as CO_2. It has been widely reported that sequestration of CO_2 from large stationary sources such as coal-fired power plants may be a potential course to obtaining the deep reductions in CO_2 emissions specified under the Kyoto Protocol.

1.2 Study Mandate

The mandate of the study was as follows:

❑ Set up coal fired power generation base cases for conventional boilers and gasification systems;
❑ Identify the range of impurities that could be captured along with CO_2 for the base cases when firing a range of coals and other fuels;
❑ Review and screen potential co-capture processes for both conventional and gasification systems;
❑ Identify the most likely co-capture scenarios and the range of impurities present in the captured stream for each scenario;
❑ Identify impacts and issues associated with compression and transport;
❑ Identify impacts and issues associated with storage and utilization options.

In evaluating the impact on storage and utilization options, SLI focused on Enhanced Oil Recovery with CO_2 (EOR), Enhanced Coal Bed Methane recovery (ECBM), and storage in either depleted oil and gas reservoirs or deep saline aquifers.

As the scope of the study covered specialized areas and issues, SLI engaged the services of three sub-consultants to augment our own expertise. The following sub-consultants contributed to the study:

❑ Alberta Research Council Inc. (ARC)   Gasification Cases;
❑ Vikor Energy Inc. (Vikor)   Impact on EOR Operations;
❑ Wreford Consulting   Development of Boiler Cases.

1.3   Design Basis and Key Assumptions

Following discussion with IEA GHG, the key parameters and assumptions used to set up the base cases and co-capture options are summarized below. Section 2.3 contains additional discussion of these parameters and how they were determined.

❑ Nominal 750 MW output;
❑ Super-critical boiler for conventional boiler cases;
❑ Flue gas clean-up for conventional boiler to generally meet current or expected emissions standards for EU countries;
❑ Netherlands coastal location;
❑ Texaco slurry-fed gasifier for IGCC cases.

1.4   Nomenclature and Units

All costs are in US dollars, third quarter 2003 unless otherwise indicated. "Tonnes" and "t" refer to metric tonnes. Values are expressed in metric units except for some benchmarks in Imperial units (e.g. - $/bbl oil price).

The term "co-capture" is used somewhat loosely in the report. For some cases, other components are captured separately and re-combined with CO_2 for transport and storage. For other systems, the impurities are only "co-captured" through lack of selectivity in the CO_2 capture process. Despite this, the term co-capture is used to describe all the systems in the report to simplify the nomenclature.

Unless specifically limited to combustion of coal only, references to "coal-fired boilers" include firing with the other fuels considered also.

Although there are other types of enhanced oil recovery systems, such as steam-assisted or thermal, for the purpose of this report the term "EOR" is used to describe only enhanced oil recovery using primarily CO_2. We also distinguish between "CBM", which is the recovery of methane from coal seams, and "ECBM", which is the injection of CO_2 and other gases to enhance the amount of methane recovered.

Finally, we make no attempt to distinguish between CO_2 "storage" and "sequestration". The report uses the term storage, which is meant to imply the permanent trapping of CO_2 in a geological formation and implies a time frame of the order of many hundreds of years.

2. BASE CASE DEVELOPMENT and FUEL SELECTION

2.1 General

The purpose of the study was to evaluate the benefits and issues associated with the combined capture and geological sequestration of other flue gas components in addition to CO_2. Previous studies have generally focused on the capture and transport of CO_2 only, and it is thought that there could be possible advantages to capturing other components such as SOx, NOx and mercury as well as CO_2.

To carry out the study mandate, SLI developed four base cases for conventional Rankine cycle steam power plants and integrated combined cycle gasification (IGCC) units. Potential alternatives for capturing other impurities in the flue gas along with CO_2 were then evaluated. Once the capture stream composition was determined for the co-capture alternatives for the range of fuels considered, the impact of the impurities on compression, transportation, and storage was then evaluated.

2.2 Selection of Fuels

It is accepted that a wider range of contaminants are present in coals versus other fuels such as natural gas or fuel oil, and hence potentially present in the captured flue gas stream. Further, combustion of coal produces more CO_2 per unit of energy than natural gas or oil, and therefore coal is the primary focus of this study.

The "IEA standard" coal was used as the base case fuel (designated "C1"), a bituminous coal from Eastern Australia, with a sulphur content of 0.86% and LHV of 25.87 MJ/kg. SLI was told that this coal is typical of coals from this region of Australia and that this type of coal has been imported extensively for use in EU power plants.

To get a better idea of the range of impurities that could be expected under the proposed co-capture alternatives, eight other fuels were also selected. Studying other fuels also enabled SLI to determine if there was any level of impurities that could significantly impact the capture alternative selected, and to determine if there were any trends based on the impurity content of each fuel (e.g. sulphur content in fuel vs. CO_2 capture cost).

A summary of the fuels used is shown in Table 2.1. Proximate and ultimate analyses, and trace elements levels typical of each type of fuel, are included in Tables 2.3 and 2.4 at the end of section 2.5.

Table 2.1

Summary of Fuels Used									
Parameter	**C1** **Base**	**C2** **Hi S**	**C3** **Lo S**	**C4** **Hi Ash**	**C7** **UltraLoS**	**ORI** **Orimul.**	**HF2** **HFO**	**PC1** **Pet Coke**	**C8** **Lignite**
Type	Bit.	Bit.	sub-Bit.	Bit.	Bit.	Orimulsion	Fuel Oil	Pet Coke	Lignite
Origin	Australia	USA	USA	India	Indonesia	Venezuela	Canada	Unknown	Germany
Sulfur, wt.	0.9%	2.5%	0.5%	0.6%	0.1%	2.7%	2.3%	6.1%	1.4%
LHV, kJ/kg	25,870	29,668	17,942	14,341	20,923	27,830	39,776	28,344	11,374
$/GJ	$1.50	$1.50	$1.50	$1.50	$1.50	$1.70	$2.25	$0.90	$1.50

Because previous studies have indicated that the level of SOx in the flue gas is a key parameter in the design of monoethanolamine (MEA) based CO_2 recovery systems, a primary focus in fuel selection was the sulphur content of the fuel. Coals at either end of the sulphur scale were included: an eastern US "high sulphur " bituminous coal with 2.5% sulphur and typical of coals used for power generation in the north-eastern US, and an ultra-low sulphur coal from Indonesia with 0.05% sulphur.

A high ash coal from India was chosen to evaluate the impact of ash levels and low LHV. Indian coals will possibly come under more scrutiny as the energy consumption in India increases, and as Western companies look to India to execute CO_2 reduction initiatives as part of Clean Development Mechanisms proposed under the Kyoto protocol.

2.3 Base Case Development

SLI developed the following base case scenarios:

❑ Conventional coal fired boiler with no capture of CO_2 (Case D);
❑ Conventional coal fired boiler with capture and injection of the entire flue gas stream (Case C);
❑ Integrated Gasification Combined Cycle (IGCC) plant with no CO_2 capture (Case A);
❑ IGCC plant with conventional water shift and CO_2 capture using physical absorption (Case B).

A summary of all cases including the co-capture alternatives is shown below.

Table 2.2

```
                    Summary of Cases

Conventional Boilers

    Case              Name                    Goal
      C           Flue Gas Capture      Capture all components
      D           No Capture            Reference
      E           FGD-Amines            Pure CO₂ - reference
      F           Cansolv               Co-capture SO₂
      G           Oxy-fuel              Co-capture SO₂
      H           Gas Sep. Membranes    CO₂ + other components

Gasification Systems

    Case              Name                    Goal
      A           No Capture            Reference
      B           CO₂ Capture           Pure CO₂ - reference
      B1          Co-capture using Selexol    Co-capture H₂S
```

Following discussion with IEA GHG, SLI adopted the following key design parameters and conditions:

❑ Net station output of 750 MW prior to parasitic power demands related to the capture processes and emissions controls;
❑ Netherlands coastal location;
❑ Once-through sea water cooling, 12C° average annual inlet temperature;
❑ Flue gas to meet current or expected emissions standards for EU countries;
❑ Cost of fuel to be $1.50/GJ for coals;
❑ Costs for other fuels were $1.70/GJ for Orimulsion®, $2.25/GJ for Heavy Fuel Oil, and $0.90/GJ for Pet Coke;
❑ Pipeline distance to be 100 km;
❑ CO_2 transported in dense phase at appropriate pressure and temperature;
❑ 70 bar discharge pressure for gas phase cases
❑ No credit for sulphur removal by-products, such as gypsum or marketable pure SO_2.

IEA GHG's "Technical and Financial Assessment Criteria" was used as the primary basis for cost estimating and economic evaluations.

We were told that IEA GHG has used 500 MW as a reference output for previous studies. An output of 750 MW was selected for this study primarily as it is believed to be a more appropriate output for IGCC. An output of 750 MW

apparently corresponds to two state-of-the-art gas turbines that might be specified for a base load IGCC plant.

There are implications for increasing the reference plant output for the conventional boiler cases. Although 750 MW is believed to be in line with a trend towards larger base load units, particularly for supercritical service, there are economies of scale versus 500 MW that should be acknowledged when comparing results from prior studies.

SLI is of the opinion that the impact of going to a 500 MW plant as a reference output might add 10 – 15% to the cost of key parameters such as specific capital cost, power cost, and CO_2 capture cost. We would expect these relationships to apply for the other co-capture cases as well.

The impact of meeting "current or expected emissions standards" for flue gas is important as it tends to increase power costs for the conventional boiler base case, and thus tends to decrease the cost penalty associated with CO_2 capture. In SLI's opinion CO_2 capture alternatives should be evaluated with respect to the emissions standards that could be expected to govern new plant construction.

The choice of 100 km as a pipeline distance appears to have only a minor impact on overall capture costs and becomes an important assumption only when the gas phase capture streams are considered, as pipeline pressure drop and diameter become important considerations. For gas phase cases SLI used a pipeline inlet pressure of 70 bar and a nominal terminal pressure of 50 bar.

2.4 Co-capture Alternatives

The following alternatives for capture of other flue gas impurities along with CO_2 from conventional boilers were reviewed:

- ❑ Capture of SO_2 using Cansolv® or similar system prior to amine scrubbing of CO_2, followed by re-combining the SO_2 and CO_2 streams (Case F);
- ❑ Combustion in oxygen-enriched air (Case G);
- ❑ Gas separation membranes (Case H);
- ❑ Physical absorption;
- ❑ Cryogenic separation;
- ❑ Adsorption.

"Conventional" CO_2 recovery by chemical absorption using an amine solution and preceded by conventional wet limestone flue gas desulphurisation (FGD) was also modelled for comparison purposes.

The following gasification co-capture alternatives were reviewed:

- ❑ Co-capture of H_2S and H_2 using sour water shift and Selexol (Case B1);

□ Co-capture of H_2S and H_2 using sour water shift and Rectisol;
□ Gas separation membranes.

2.5 Criteria for Further Study of Co-Capture Alternatives

The following criteria were considered in determining which co-capture alternatives to evaluate for further study:

□ Potential for co-capture of other pollutants – some systems are not amenable to co-capture;
□ Potential for reductions in capture costs relative to the conventional FGD-amines system;
□ Parasitic energy demands;
□ Potential for process simplification;
□ State of development of the technology;
□ Potential for future break-throughs;
□ Potential problems with any of the fuels;
□ Impact on choice of storage option.

Assuming there was a potential to capture other pollutants along with the CO_2, the next primary consideration was the parasitic energy demands. In our opinion this precluded physical absorption from further study for conventional boiler cases due to the minimum pressure required for the system to operate effectively. Parasitic energy demands also form an integral element in the estimation of the capture costs, as the power lost through the capture process must be made up through additional power generation.

Another important consideration was the state of the development of the technology. In general we considered only current technology, and commented on the potential for further improvements to each alternative.

2.6 Estimating Sources and Considerations

The estimates included in this study should be considered "order-of-magnitude" or "screening" level estimates only, with a nominal accuracy range of +50% - 30%. Most costs were developed from previous SLI experience or from published reports. Estimates developed from first principles include an appropriate level of contingency. Estimates developed using bench-mark costs (e.g. - $900kW for installed coal-fired conventional power plants) do not include contingency as the benchmark is based on the final cost of actual plants.

In our opinion the estimates were normalized to the same estimating base, so that despite their wide confidence interval they should be directly comparable.

2.6.1   Capital Costs

Capital costs for the conventional boiler cases were developed using the following benchmarks:

- $900/kW gross output was used for base plant cost for coals and Pet Coke;
- $810/kW was used for the heavy fuel oil plant cost, and $820/kW for the Orimulsion® plant;
- $12.50/kW gross output was used for LoNOx burners;
- $25/kW for SCR systems;
- Particulate removal systems were based on $40/kW for ESPs and $50/kW for fabric filter systems;
- FGD systems were based on $100/kW for medium range sulphur coals.

The above costs were developed from previous SLI studies, in-house data, or publicly available information. In all cases the costs developed for this report are for the purpose of screening alternatives. Costs should only be used as absolute values with caution; their primary purpose is to point to promising directions for further work.

FGD costs were adjusted based on the gas flow, quantity of sulphur removed, and level of SO_2 reduction. FGD capital costs were increased where SO_2 reduction requirements exceeded 95%, as it was judged that the equipment required to meet this level would cost more than a standard system, which are generally based on 90% SO_2 removal.

Compression capital costs were based on $/kW of power required including expected economies of scale for larger units or multiple units where applicable. Cooling and other utilities and site specific costs were included.

Costs for the Cansolv® system were developed from information received from Cansolv® for the base coal (C1) and Pet Coke. These cases "book-ended" the expected sulphur quantity for the fuels, and costs were developed for the other cases based on the gas and sulphur flows for these two cases.

Costs for the amine systems were developed based on information from Mitsubishi Heavy Industries (MHI) for previous SLI studies, and from published information from Fluor, MHI, and ABB Lummus Global. In all cases the published costs were adjusted based on both gas flow through the absorber and CO_2 flow in the regenerator process.

Costs for oxy-fuel systems were developed from published studies and adjusted for differences in flue gas flow and composition. Costs for air separation units were taken from published information on commercially available standard cryogenic oxygen systems.

Costs for the gas separation membranes were developed from previous SLI work on CO_2-methane units. The costs were adjusted to reflect what SLI believes to be a realistic membrane inlet pressure of around 18 bar.

Capital costs for the gas separation membranes have probably the greatest uncertainty of all the costs developed. Most vendors were unable to give what they felt was appropriate costs, given the emerging nature of the technology in this application. Gas separation membranes are currently being used in CO_2-EOR to separate CO_2 from the produced oil and gas stream prior to re-injection. However, suppliers of this equipment were reluctant to speculate on how their equipment would perform under a flue gas scenario and declined to provide costs.

While SLI used our professional judgement to develop what we consider to be realistic costs for the gas separation membranes, it needs to be recognized that technology for this application is not commercially available, and estimates should be treated with caution.

Capital costs for the gasification cases were based on costs reported in IEA Report PH4/19. A total capital cost of $981 million was carried for the base case using the IEA base fuel (coal C1). The following adjustments were then made to the capital costs for other cases and fuels:

❑ Base plant costs were reduced slightly for the Orimulsion® and Heavy Fuel Oil plants, to account for simpler fuel handling and feed systems;
❑ Sulphur recovery unit costs were adjusted according to sulphur throughput;
❑ Acid gas removal and compression costs were adjusted according to the total throughput in each case.

2.6.2 Operating Costs

Operating costs were based on the following:

❑ $1.50/GJ for all coals;
❑ $1.70/GJ for Orimulsion®, $2.25/GJ for Heavy Fuel Oil, and $0.90/GJ for Pet Coke;
❑ 85% availability;
❑ Operating, maintenance, taxes and insurance were estimated at 6% of capital costs in aggregate for the conventional boilers, and 7.5% in aggregate for the gasification cases;
❑ Capital charges and depreciation estimated at 12% per year (based on 10% discount rate and 25-year project life).

Operating costs for the flue gas clean-up systems were estimated using the following:

- ❑ 10% of capital costs for SCR systems;
- ❑ 3-4% of capital costs for particulate removal systems;
- ❑ $300/t SO_2 removed for FGD systems;
- ❑ $100/t SO_2 removed for sulphur recovery units in gasification cases.

FGD operating costs were based on the medium sulphur coals and adjusted for economies of scale for higher sulphur coals.

Operating costs for the various co-capture systems were estimated using the following:

- ❑ Cansolv® operating costs were obtained from discussions with the Cansolv® representative and pro-rated according to gas and sulphur flows for each fuel;
- ❑ Amine system costs were obtained from discussions with process Licensors;
- ❑ 10% of capital costs was used for ASU units;
- ❑ 5% of capital costs for gas separation membranes;
- ❑ 5-8% of capital costs for compressors and other large rotating equipment;

The above costs include operations, repairs, maintenance, lubricants, solvent make-up, and other operating costs. Note that the above costs do not include actual power cost. In all cases the power required for each system was deducted from the gross power output of the plant, which was set at 750 MW.

The cost of CO_2 avoided was calculated as:

A = (B - C) / (D - E)

Where:

A = Cost of CO_2 Avoided
B = Cost of power with capture, $/kWh
C = Cost of power with no capture, $/kWh
D = CO_2 emitted with no capture, t CO_2/kWh
E = CO_2 emitted with capture, t CO_2/kWh

Note that the reference power costs and CO_2 emissions are based on the No Capture case for *that fuel*.

Table 2.3

Parameter	C1 Base	C2 Hi S	C3 Lo S	C4 Hi Ash	C7 UltraLoS	ORI Orimul.	HF2 HFO	PC1 Pet Coke	C8 Lignite
Proximate and Ultimate Analysis of Each Fuel									
LHV, kJ/kg	25,870	29,668	17,942	14,341	20,923	27,830	39,776	28,344	11,374
Proximate Analysis									
Volatile Matter	31.7%	39.2%	30.3%	21.2%	36.0%	incl below	incl below	0.0%	19.1%
Fixed Carbon	46.6%	48.2%	35.3%	29.1%	40.1%	70.7%	98.9%	96.8%	22.4%
Ash	12.2%	7.9%	5.2%	39.2%	1.5%	0.2%	0.5%	3.2%	6.5%
Moisture	9.5%	4.7%	29.2%	10.5%	22.4%	29.1%	0.5%	0.0%	52.0%
Ultimate Analysis									
Carbon	64.6%	73.8%	48.0%	38.1%	57.2%	60.0%	86.2%	84.6%	29.1%
Hydrogen	4.4%	4.7%	3.4%	2.7%	3.9%	7.3%	10.1%	2.2%	2.5%
Oxygen	7.0%	5.0%	13.2%	7.9%	14.2%	0.2%	0.0%	2.6%	8.1%
Nitrogen	1.4%	1.4%	0.5%	0.9%	0.8%	0.5%	0.4%	1.5%	0.3%
Sulfur	0.9%	2.5%	0.5%	0.6%	0.1%	2.7%	2.3%	6.1%	1.4%
Chlorine	0.02%	0.10%	0.00%	0.11%	0.00%	0.01%	0.00%	0.00%	0.10%

Table 2.4

Trace Elements (mg/kg)	C1 Base	C2 Hi S	C3 Lo S	ORI Orimul.	HF2 HFO	PC1 Pet Coke
Trace Elements in Fuels						
Hg	0.042	0.140	0.092	-	0.010	0.020
Se	0.050	3.500	3.500	-	1.000	0.400
As	1.260	9.800	5.800	-	0.020	1.100
Cd	0.067	0.090	0.120	-	0.200	0.200
Pb	6.80	6.50	7.70	-	2.00	2.10
Sb	0.473	0.740	0.690	-	0.020	0.600
Cr	15.9	16.0	22.0	1.1	3.0	18.0
Ni	11.5	15.0	13.0	65	120	279
V	20.6	17.0	39.0	300	180	1,560

3.   CONVENTIONAL BOILERS

3.1   No Capture Case (Case D) - Basis

A supercritical conventional boiler with nominal 750 MW power output and no CO_2 capture was selected as the base case. Design considerations were based on a principle of: *"the type of boiler that could be expected to be chosen for a state-of-the-art base load plant to be built in the EU today"*.

In addition to those identified in section 2.3, the following parameters were used to develop the conventional boiler cases:

❑ Single base-loaded stand-alone unit;
❑ Steam cycle parameters used were single reheat, supercritical, with steam conditions of 247bar/575ºC/595ºC. The cycle included eight stages of regenerative feedwater heating and two turbine-driven boiler feed pumps;
❑ Steam turbine generator gross capacity ranged between 810-830 MW, giving a net station output of 750 MW;
❑ The steam generator was either a PF fired supercritical unit or an oil-fired unit, with either wall or tangential firing, depending on the fuel in question. Down-shot firing was used for the Pet Coke case.

Based on a literature review of current industry practices and projections of what clean-up equipment may be required to meet more stringent emissions levels, SLI modelled the "no capture" case for conventional boilers based on the flue gas controls below. Figure 3.1 shows a Block Flow Diagram (BFD) for this case.

❑ Lo-NOx burners and over-fire air for initial NOx reduction;
❑ Selective Catalytic Reduction (SCR) for final NOx reduction;
❑ Fabric filter or baghouse for particulate reduction;
❑ Flue Gas Desulphurization (FGD) for SO_2 reduction (omitted for fuel C7).

The flue gas clean-up equipment described above was used for all fuels except C7, Ultra-Low Sulphur coal, where FGD was omitted. No specific measures were included for mercury control. The FGD system was assumed to be a wet limestone forced-oxidation (LSFO) system, regardless of the fuel considered.

The criteria below were used as a guide to determine the appropriate flue gas controls required. As the study was conceptual in nature and cost estimates were used only to screen and compare co-capture alternatives, clean-up requirements were not precisely designed for each fuel.

❑ SO_2 emissions to be about 600 g/MWh;
❑ NOx emissions to be about 1000 g/MWh;
❑ Particulate emissions about 50 g/MWh.

The above levels are about half of what we understand to be the current targets for new power plants in the EU. SLI does not represent that sufficient detailed design was performed in each case to determine if the above targets would be met. The targets were primarily used as a guide to ensure that the "no capture" case for conventional boilers would adequately represent the level of flue gas clean-up that might be mandated for this size of plant to be built within the next ten years in the EU, i.e. within the window in which we might expect CO_2 recovery systems begin to be implemented on a large scale.

In our opinion achieving these standards is technically feasible and the above levels do not represent a significant step beyond what controls might be expected to be included in a plant built in the EU today.

Figure 3.1
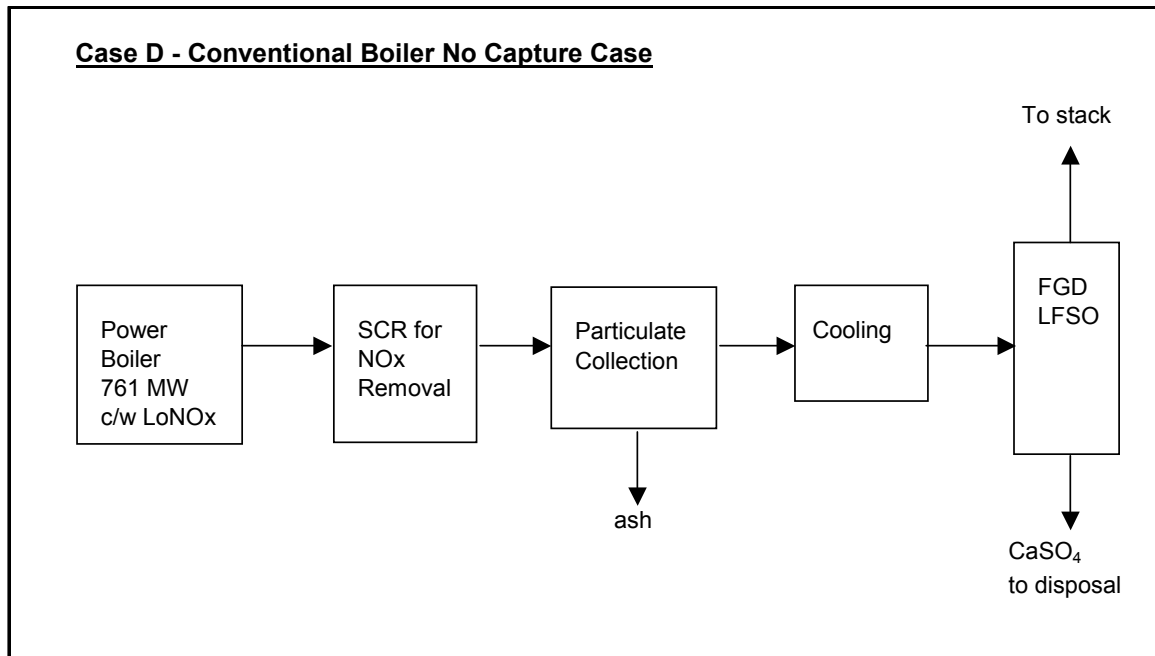
3.2 No Capture Case - Results

Table 3.1 shows summary results for each fuel for this case.

With the exception of SO_2 levels, most other key parameters (flue gas volume, CO_2 content) did not vary significantly among the fuels evaluated (a factor of 1.2-1.3 from low to high). There was a significant difference in the SO_2 levels across the range of fuels studied. SO_2 varied from 55 ppm for the Ultra Low Sulphur coal C7 (prior to sulphur removal) to 4400 ppm for Pet Coke, a factor of 80 from low to high.

Power and capital costs varied over a relatively narrow range (+/- 20%). The lowest capital cost was estimated for the Ultra Low Sulphur coal C7 (no FGD), and the Orimulsion and Heavy Fuel Oil plants (light duty particulate removal and no solid fuel handling system). The most expensive plant was estimated to be the Pet Coke plant, due to the FGD system required to achieve the higher level of SO_2 removal (roughly 98%) to meet the SOx emissions target.

CO_2 emissions ranged from 680 kg/MWh for the Heavy Fuel Oil case, to 930 kg/MWh for the Pet Coke case. CO_2 content in the wet flue gas varied from 12.8% for Orimulsion to 15.3% for the Pet Coke. CO_2 content averaged 14.5% for the five non-lignite coals.

We were told that European Directive emissions targets were around 1200 g/MWh for SOx, 2100 g/MWh for NOx, and 100 g/MWh for particulates. In general, emissions for the no capture case are estimated at around 50% of the current European Directive levels.

Table 3.1

Summary Results by Fuel									
Case D - Conventional Boiler No Capture									
Parameter	**C1** **Base**	**C2** **Hi S**	**C3** **Lo S**	**C4** **Hi Ash**	**C7** **UltraLoS**	**ORI** **Orimul.**	**HF2** **HFO**	**PC1** **Pet Coke**	**C8** **Lignite**
Gross Power Output, MW	761	769	759	760	750	770	768	774	764
Net Power Output, MW	750	750	750	750	750	750	750	750	750
Total Capital Cost, $millions	830	860	820	850	720	770	740	890	870
Specific Capital Cost, $/kW	1,100	1,140	1,100	1,130	960	1,030	980	1,180	1,160
Power Cost, $/kWh	0.040	0.044	0.040	0.042	0.036	0.045	0.045	0.044	0.046
Plant Efficiency, %	41.8%	41.9%	40.9%	41.5%	41.8%	40.4%	42.3%	40.2%	40.9%
CO_2 Emitted, kg/MWh	790	780	860	840	860	700	680	930	830
SO_2 Emitted, g/MWh	520	650	380	640	360	770	650	690	520
NOx Emitted, g/MWh	1,010	930	770	1,140	810	400	500	1,030	650
Particulate Matter, g/MWh	160	90	100	190	50	50	90	80	200
CO_2 Content in Wet Flue Gas,	14.5%	14.6%	14.4%	14.6%	14.7%	12.8%	13.1%	15.3%	12.4%

3.3 Flue Gas Capture (Case C) - Basis

For this base case SLI included an Electrostatic Precipitator (ESP) for particulate removal upstream of the compression equipment. Based on our discussions with equipment manufacturers particulate removal will be required to protect the compressor internals from plugging or excessive abrasion. Inlet cooling and water knock-out was also included upstream of the compressors. Conventional high-efficiency, low excess air burners were included as there was no need to limit NOx emissions. A BFD for this case is shown in Figure 2.

Figure 3.2

Case C - Capture of Entire Flue Gas Stream

Note: Dehydration simplified, typically done after third stage of compression
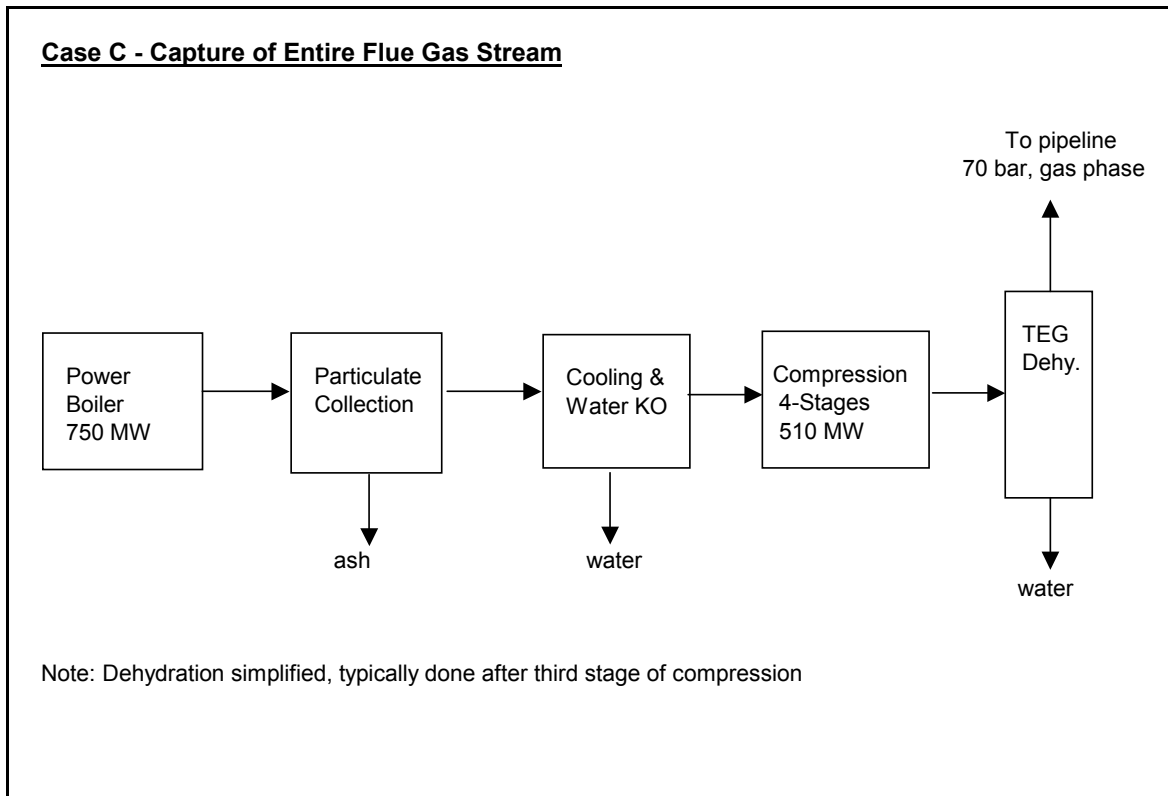
3.4 Flue Gas Capture - Results

Table 3.2 shows summary results for each fuel. The composition of the captured stream for each fuel is shown in Table 3.3.

The power required to compress the entire flue gas stream varied directly with the flue gas volume and ranged from 420 MW (565,000 Hp) for Heavy Fuel Oil to 520 MW (700,000 Hp) for Pet Coke. Compression capital costs ranged from $510 – 640 million, or roughly 80% - 95% the cost of the base power plant. The cost to run and maintain the enormous amount of compression required was estimated to be 25% - 30% of total plant operating cost including fuel.

The above figures are based on an assumed pipeline inlet pressure of 70 bar. Pipeline costs did not change significantly among the fuels, and were based on either 40" or 42" diameter lines (1000 – 1050mm).

Compressor vendors we spoke with told us that three-four trains of the largest compressors and drivers built to date would be required to compress this volume of flue gas from atmospheric pressure to 70 bar. Although compression over this range is common, according to vendors the combination of large compression

ratio and large gas volumes would create significant challenges. They also expressed concerns over the corrosive nature of the flue gas in the temperature regime being considered.

Given the large capital and operating costs for the compressors and inter-stage cooling, the results are very sensitive to the compressor costs. Because of the impracticality and challenging technical aspects of the problem (corrosive nature of the gas, high compression ratio, very high volume), none of the compressor vendors contacted by SLI were willing to give a budget price for the system. Therefore there could be a high degree of uncertainty in the estimated costs for this case.

Based on the above results, we conclude that capture of the entire flue gas stream is not practical or competitive with other CO_2 capture systems. Power and capture costs are on the order of \$0.15/kWh and \$130/t CO_2 avoided on average respectively. Further, the net station output drops to around 40% of the designed gross power output.

Capture of the entire flue gas stream is very sensitive to the pipeline inlet pressure assumed. SLI assumed an inlet pressure of 70 bar because this is a typical operating pressure for a natural gas transmission line. Because the captured stream is in the gas phase, the minimum pressure assumed for dense phase CO_2 transmission of 110 bar is not applicable, and considerations such as optimum wall thickness and allowable pressure drop would take precedence.

One possible application of this scheme could be for ECBM projects where the coal seam is located close to the power plant ("mine-mouth" power plants). Due to the short distance and potentially lower injection pressure in this scenario, if the pipeline inlet pressure could be reduced to 10 bar, compression requirements would be reduced significantly. Power and capture costs at 10 bar are estimated at \$0.08/kWh and \$45/t CO_2 avoided respectively, or comparable to CO_2 capture using an FGD-amines system. Figure 3.3 shows the relationship between capture cost and discharge pressure for this case for the base coal.

With the exception of mercury, most trace elements are believed to be removed during particulate removal prior to compression, or during the free liquid removal stages and dehydration during compression. Therefore trace elements in the captured flue gas are not expected to be a significant concern. It should be noted that the fate of any trace elements still needs to be resolved, as we could be trading an air emissions problem for a solids or liquid waste disposal problem.

In SLI's opinion, the following conclusions can be drawn about Case C, capture of the entire flue gas stream:

- Capture of the entire flue gas stream is generally not practical and leads to significantly higher power and capture costs: $0.15/kWh and $130/t CO_2 avoided on average;
- Three-four trains of the largest compressors and drivers built to date would be required for this service;
- Transport costs would also be more costly due to higher pressure losses associated with gas phase transport;
- Power and capture costs for this case are very sensitive to the pipeline pressure assumed. Using 10 bar reduces power and capture costs to $0.08/kWh and $45/t CO_2 avoided;
- One application for flue gas capture could be ECBM projects close to the capture source;
- Most trace elements are removed during particulate removal prior to compression, or during free liquid removal during compression.

Table 3.2

Summary Results by Fuel
Case C - Capture of Entire Flue Gas Stream

Parameter	C1 Base	C2 Hi S	C3 Lo S	C4 Hi Ash	C7 UltraLoS	ORI Orimul.	HF2 HFO	PC1 Pet Coke	C8 Lignite
Net Power Output, MW	292	295	267	269	262	323	328	227	269
Total Capital Cost, $millions	1,350	1,350	1,380	1,400	1,370	1,230	1,210	1,420	1,380
Specific Capital Cost, $/kW	4,630	4,560	5,190	5,190	5,250	3,800	3,700	6,270	5,130
Power Cost, $/kWh	0.142	0.140	0.159	0.158	0.161	0.125	0.130	0.174	0.157
Plant Efficiency, %	16.5%	16.9%	14.7%	15.1%	14.6%	17.8%	18.9%	12.5%	14.9%
$/t CO_2 Avoided	128	122	138	138	145	113	126	141	135
$/t CO_2 Captured	48	47	47	47	47	48	54	42	47
CO_2 Emitted, kg/MWh	-	-	-	-	-	-	-	-	-

Table 3.3

Captured Stream Composition
Case C - Capture of Entire Flue Gas Stream

% by Vol.	C1 Base	C2 Hi S	C3 Lo S	C4 Hi Ash	C7 UltraLoS	ORI Orimul.	HF2 HFO	PC1 Pet Coke	C8 Lignite
CO_2	16.0%	15.8%	16.5%	16.3%	16.6%	15.1%	14.7%	16.0%	15.9%
Nitrogen	80.9%	81.0%	80.4%	80.6%	80.3%	83.2%	82.9%	79.9%	80.4%
Oxygen	2.8%	2.8%	2.8%	2.8%	2.8%	1.1%	2.0%	3.5%	2.8%
SO_2, ppm	800	1,970	580	940	50	2,690	1,460	4,320	2,920
NOx, ppm	380	350	290	420	300	170	210	320	280
Water	0.2%	0.2%	0.3%	0.2%	0.3%	0.4%	0.2%	0.1%	0.6%
CO	trace	trace	trace	trace	trace	trace	trace	trace	trace

Figure 3.3

Case C - Capture of Entire Flue Gas Stream
CO$_2$ Capture Cost vs. Required Pressure
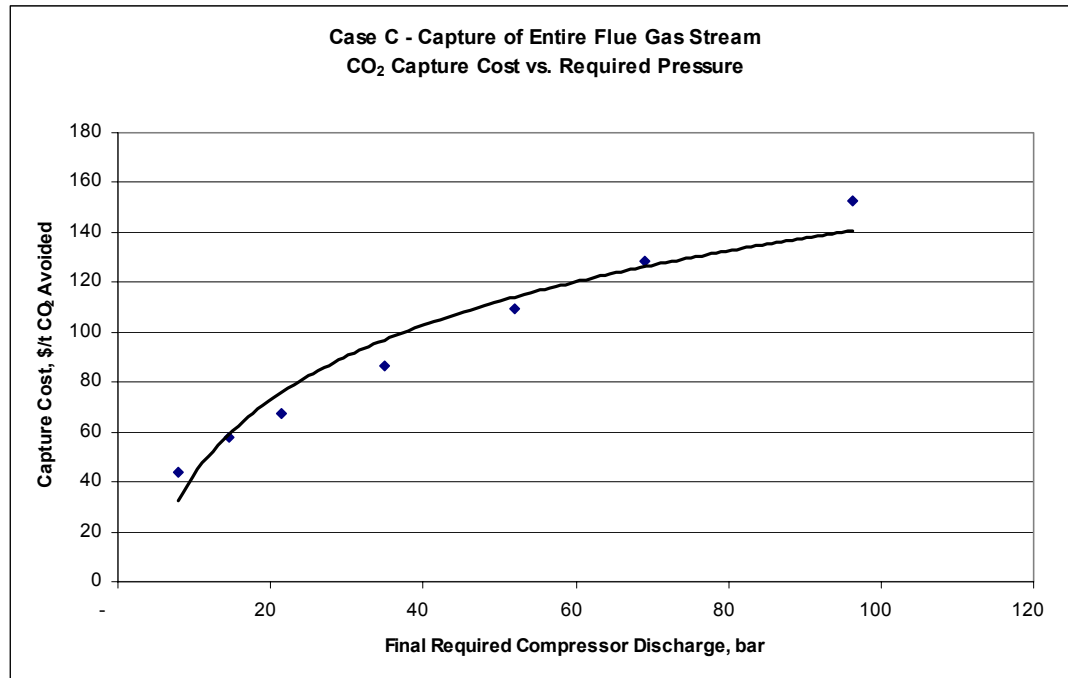
3.5 Baseline Capture Using FGD-Amines (Case E) - Basis

Even though this process produces a relatively pure CO$_2$ stream (+99.9%), this case was evaluated to establish a baseline for "pure" CO$_2$ capture for conventional boilers, and because an amine system was a part of Case F.

Figure 3.4 shows the BFD for this process. These systems are currently used for CO$_2$ recovery from flue gases, albeit at an order-of-magnitude scale less than required for a 750 MW power plant. In our opinion, they represent the "baseline" for current CO$_2$ recovery technology.

A recovery rate of 90% of the CO$_2$ in the amine system was assumed, based on discussions with the Licensors. In our opinion 90% recovery is accepted as a "standard" target. The optimum recovery will depend on site specifics. Higher recovery can be achieved with higher capital and operating costs.

From previous studies and from our discussions with licensors of this technology, depending on the fuel selected, SO$_2$ removal upstream of the amine unit would almost certainly be required, and NO$_2$ reduction would *probably* be required. Therefore for this case we included both SOx and NOx reduction for all fuels in order to meet typical maximum SOx and NOx levels at the absorber inlet. The addition of NOx reduction would tend to increase the cost penalties associated

with this option and could make some co-capture alternatives more attractive relative to amine scrubbing.

SOx reduction was not included for the ultra-low sulphur coal C7. Particulate control systems were included for all fuels.

Figure3.4

Case E - Capture of CO$_2$ Only Using FGD-Amines (for reference)

Note: Dehydration simplified, typically done after third stage of compression
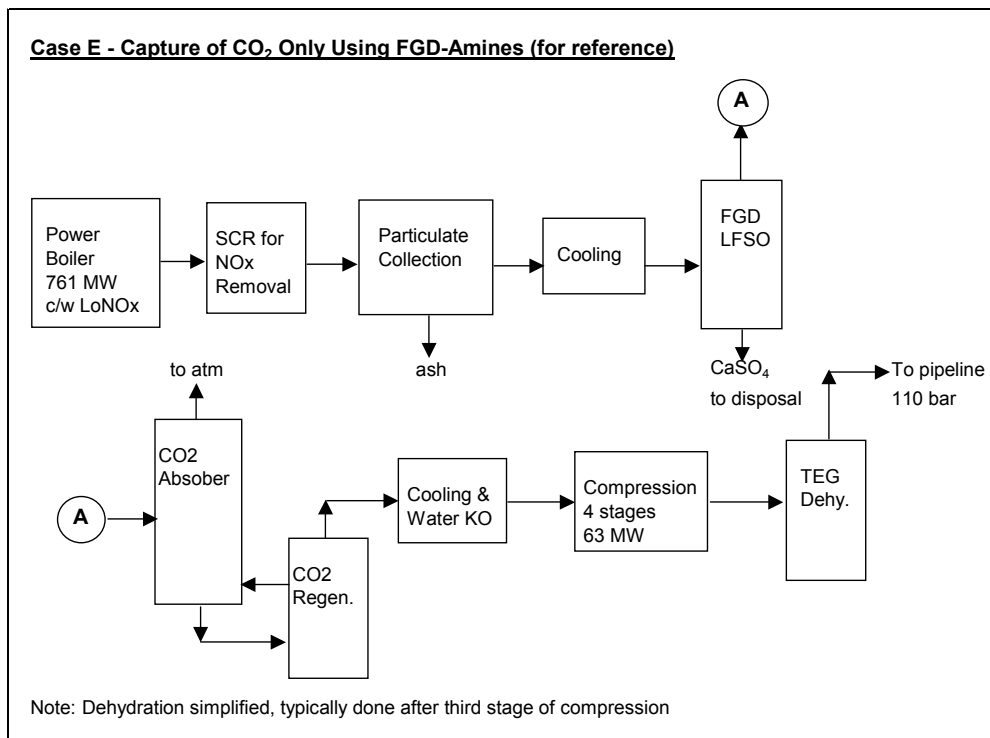
3.6   Baseline Capture Using FGD-Amines - Results

Table 3.4 shows summary results for each fuel for this case.

After discussions with Licensors of amine systems it was assumed that the maximum allowable level of SO$_2$ at the inlet to the amine absorber would be 20 ppm. To meet this low target, SO$_2$ removal in the FGD system needs to be at least 95% for the Low Sulphur coal C3, and between 97% - 99% for all the other coals. SO$_2$ removal would need to be 99.5% for the 6% S Pet Coke.

For this case the basis of $100/kW for FGD for medium sulphur coals was increased by 20% to reflect the additional quantity of sulphur handled, and the additional costs required to achieve the higher sulphur removal level. For the cases where sulphur removal was over 97%, an additional 5% was added to the capital cost of the FGD system, and for the cases where sulphur removal was

over 99%, an additional 10% was added. A caustic wash was included for the Ultra Low Sulphur coal C7 to reduce the SO_2 from 48 to 20 ppm.

In our opinion the area of SO_2 removal for the purpose of CO_2 recovery from flue gas for high sulphur fuels needs further study. Previous studies may be avoiding the issue of how the target level of 20 ppm SO_2 is achieved using conventional FGD, particularly with high sulphur fuels. In our opinion this level of SO_2 removal may not be attainable using a conventional single-stage FGD system. Co-capture of SO_2 using Cansolv® or similar system could resolve this issue.

Power and capture costs generally varied over a narrow range, with the Pet Coke having the highest costs due to the additional sulphur removal requirements and higher gas flow rate.

Table 3.4

Summary Results by Fuel **Case E (Reference Case) - Capture of CO2 Using FGD-Amines**									
Parameter	**C1** **Base**	**C2** **Hi S**	**C3** **Lo S**	**C4** **Hi Ash**	**C7** **UltraLoS**	**ORI** **Orimul.**	**HF2** **HFO**	**PC1** **Pet Coke**	**C8** **Lignite**
Gross Power Output, MW	761	769	759	760	750	770	768	774	764
Net Power Output, MW	551	552	532	537	532	572	580	517	542
Total Capital Cost, $millions	1,120	1,150	1,130	1,150	1,040	1,050	1,010	1,240	1,180
Specific Capital Cost, $/kW	2,030	2,090	2,120	2,140	1,950	1,840	1,740	2,390	2,190
Power Cost, $/kWh	0.074	0.079	0.078	0.079	0.073	0.076	0.075	0.088	0.084
Plant Efficiency, %	30.7%	30.8%	29.0%	29.8%	29.6%	30.8%	32.7%	27.7%	29.5%
$/t CO_2 Avoided	50	52	51	51	50	51	51	56	54
$/t CO_2 Captured	32	34	31	32	31	35	35	33	34
CO_2 Emitted, kg/MWh	110	110	120	120	120	90	90	130	110

3.7 Co-capture of SO_2 Using Cansolv® or Similar (Case F) - Basis

This alternative is based on the substitution of a conventional FGD system by a sulphur removal system that produces a relatively pure SO_2 stream. This stream can then be combined with the pure CO_2 stream produced from the conventional amine-based CO_2 recovery system described in Case E.

Wet limestone based FGD systems are used extensively for sulphur removal at power plants worldwide. They are commercially proven and have well defined capital and operating costs. They add significant costs to conventional CO_2 recovery systems however, and may have difficulty achieving the high level of sulphur removal required for amine-based CO_2 recovery.

The sulphur recovered from a conventional wet lime FGD system is in solid form (gypsum) and is not amenable to combining with a CO_2 stream for storage. Other sulphur removal systems reviewed by SLI were the Wellman-Lord system that produces a sodium sulphite product, and the SNOX system that produces sulphuric acid. These systems were not considered further as they produce streams that are not amenable to combining with CO_2 for geological storage.

The Cansolv® system uses a water soluble amine that is highly selective to SO_2. It is similar to the amine-based CO_2 systems in that the gas is contacted with the amine in an absorber, then the SO_2 is stripped from the rich amine in a second smaller tower with low grade heat. An electro-dialysis unit treats a slipstream of the amine to reduce the level of heat stable salts. A BFD is shown in Figure 3.5.

We would describe the Cansolv® system as near-commercial. According to Cansolv® they have had three units in operation for the last one-two years. All these units are at least an order of magnitude smaller than what would be required for a 750 MW power plant, based on inlet gas flows.

SLI also briefly reviewed a competing system, Labsorb®, which operates on a similar basis. We based our calculations on Cansolv® because this system appears to be more established commercially. In our opinion using Labsorb® as a basis would likely have produced similar results.

Figure 3.5

Case F - Co-Capture of SO₂ Using Cansolv or Similar System

Note: Dehydration simplified, typically done after third stage of compression
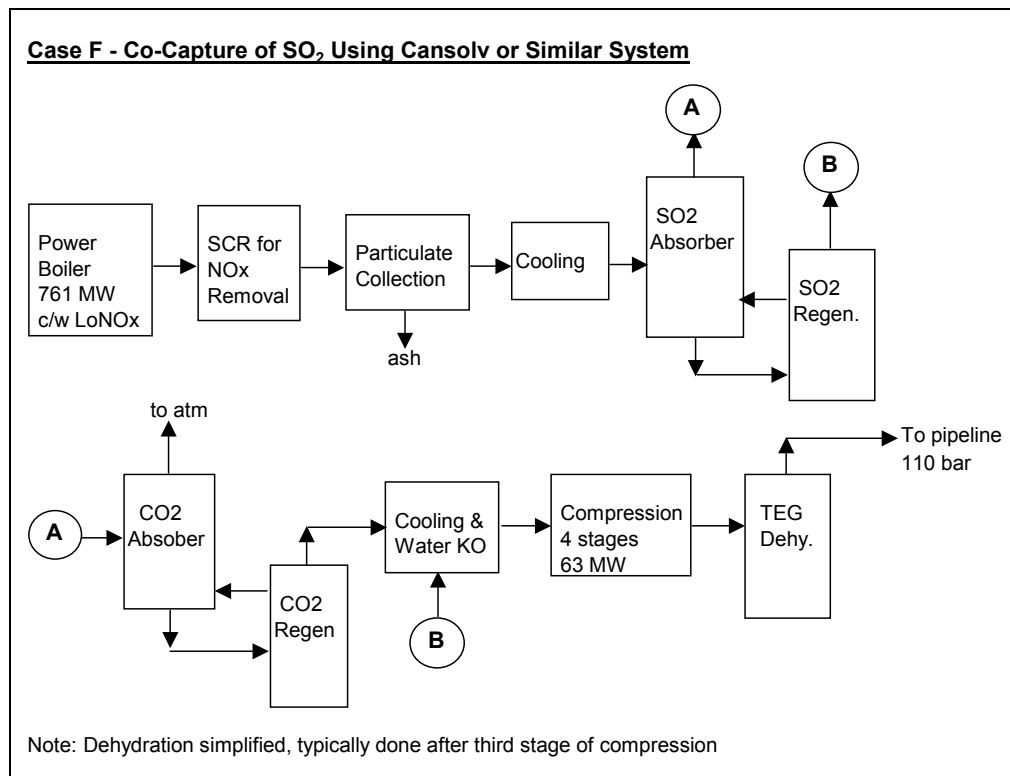
3.8 Co-capture of SO$_2$ Using Cansolv® or Similar - Results

Table 3.5 shows summary results for each fuel. The composition of the captured stream for each fuel is shown in Table 3.6.

There appears to be an opportunity to reduce CO$_2$ capture costs using this type of system. Estimated savings based on the medium sulphur base coal C1 relative to a conventional FDG-amines system are on the order of $6/t CO$_2$ avoided and $0.005/kWh.

Savings for the above system relative to a conventional FGD-amines system appear to increase as the sulphur content of the fuel increases. With the 6% S pet coke savings are estimated to be on the order of $0.016/kWh and $20/t CO$_2$ avoided. Figures 3.6 and 3.7 plot the estimated savings using this system relative to conventional CO$_2$ capture, against sulphur content of the fuel for CO$_2$ recovery cost and power costs respectively.

The cost savings for the high sulphur fuels may be understated in our estimates due to the difficulty of achieving the required levels of sulphur removal using conventional FGD systems (+99% for high-S fuels). To our knowledge there are few commercial scale systems with this level of SO$_2$ removal. From our discussions with Cansolv® and Labsorb®, it appears that the cost of their systems may be less sensitive to increasing sulphur content than conventional FGD systems.

To date this type of SO$_2$ recovery system has not been extensively used. Most plants use conventional FGD. It is our understanding that SO$_2$ gas is generally not in high demand, and this may have hindered introduction of this type of system. In our case, the value of the SO$_2$ is its ability to reduce the cost of capturing CO$_2$ compared to other systems.

FGD costs are generally well defined. Cansolv® systems have not yet been built that are within an order of magnitude of the scale considered for this study (based on gas flow), and hence both the capital and operating costs estimated for a Cansolv or similar system may be less accurate than FGD costs.

No adjustment was made to the fuel cost based on sulphur content. High sulphur coals for power generation generally trade at a discount and this discount increases as sulphur content increases.

This alternative produces a CO$_2$/SO$_2$ stream of nominally 97% - 99.5% CO$_2$ and 0.5% - 3% SO$_2$ with very little other components. Most trace elements are expected to be removed during either the particulate removal stage, through both the Cansolv and CO$_2$ amines scrubbing systems, and through the free liquid removal, cooling and dehydration required for compression and transport. In our

opinion very little metal compounds including mercury end up in the final CO_2 stream.

In SLI's opinion, the following conclusions can be drawn about Case F, co-capture SO_2 using a Cansolv® or similar system:

❑ There could be an opportunity to lower CO_2 capture costs through co-capture of SO_2 using a Cansolv® or similar system;
❑ The composition of this stream would be primarily 97% - 99.5% CO_2 and 0.5% - 3% SO_2, with the SO_2 level being dependent on sulphur content of the fuel;
❑ Savings relative to a conventional FGD-amines system could be on the order of $6/t CO_2 avoided and $0.005/kWh, for a 2% sulphur fuel;
❑ Savings for the above system relative to an FGD-amines system appear to increase as the sulphur content of the fuel increases. For example, with the 6% S pet coke, the savings were estimate to be $0.016/kWh and $16/t CO_2;
❑ The cost savings for the high sulphur fuels may be understated due to the difficulty of achieving the level of sulphur removal using conventional FGD systems (+99% for high-S fuels) required for amine scrubbing systems;
❑ Most trace elements are expected to be removed during either particulate removal, the Cansolv® or amine scrubbing systems, or during the free liquid removal, cooling and dehydration during compression.

Table 3.5

Parameter	C1 Base	C2 Hi S	C3 Lo S	C4 Hi Ash	C7 UltraLoS	ORI Orimul.	HF2 HFO	PC1 Pet Coke	C8 Lignite
Net Power Output, MW	547	549	528	533	532	569	576	514	536
Total Capital Cost, $millions	1,070	1,070	1,080	1,100	1,040	960	940	1,120	1,100
Specific Capital Cost, $/kW	1,950	1,960	2,050	2,060	1,950	1,700	1,630	2,180	2,060
Power Cost, $/kWh	0.071	0.070	0.075	0.074	0.073	0.066	0.069	0.072	0.074
Plant Efficiency, %	30.9%	31.4%	29.1%	29.9%	29.6%	31.4%	33.3%	28.4%	29.7%
$/t CO_2 Avoided	44	39	47	45	50	35	40	35	41
$/t CO_2 Captured	29	26	29	29	31	25	29	23	27
CO_2 Emitted, kg/MWh	110	100	120	120	120	90	90	130	110

Summary Results by Fuel
Case F - Co-Capture of SO_2 Using Cansolv or Similar System

Table 3.6

Captured Stream Composition for Each Fuel
Case F - Co-Capture of SO$_2$ Using Cansolv or Similar System

% by Vol.	C1 Base	C2 Hi S	C3 Lo S	C4 Hi Ash	C7 UltraLoS	ORI Orimul.	HF2 HFO	PC1 Pet Coke	C8 Lignite
CO$_2$	99.4%	98.6%	99.6%	99.4%	100.0%	98.1%	98.9%	97.1%	98.0%
Nitrogen	trace	trace	trace	trace	trace	trace	trace	trace	trace
Oxygen	trace	trace	trace	trace	trace	trace	trace	trace	trace
SO$_2$, ppm	0.6%	1.4%	0.4%	0.6%	0.0%	1.9%	1.1%	2.9%	2.0%
NOx, ppm	trace	trace	trace	trace	trace	trace	trace	trace	trace
Water	trace	trace	trace	trace	trace	trace	trace	trace	trace
CO	trace	trace	trace	trace	trace	trace	trace	trace	trace

Figure 3.6

Capture Costs for SO$_2$ Co-Capture Using Cansolv
(relative to FGD-Amines)

Figure 3.7

Power Cost Savings for SO$_2$ Co-Capture Using Cansolv (relative to FGD-Amines)

Y-axis: Savings in Power Cost vs. FGD-Amines, $/kWh

X-axis: % Sulphur in Fuel
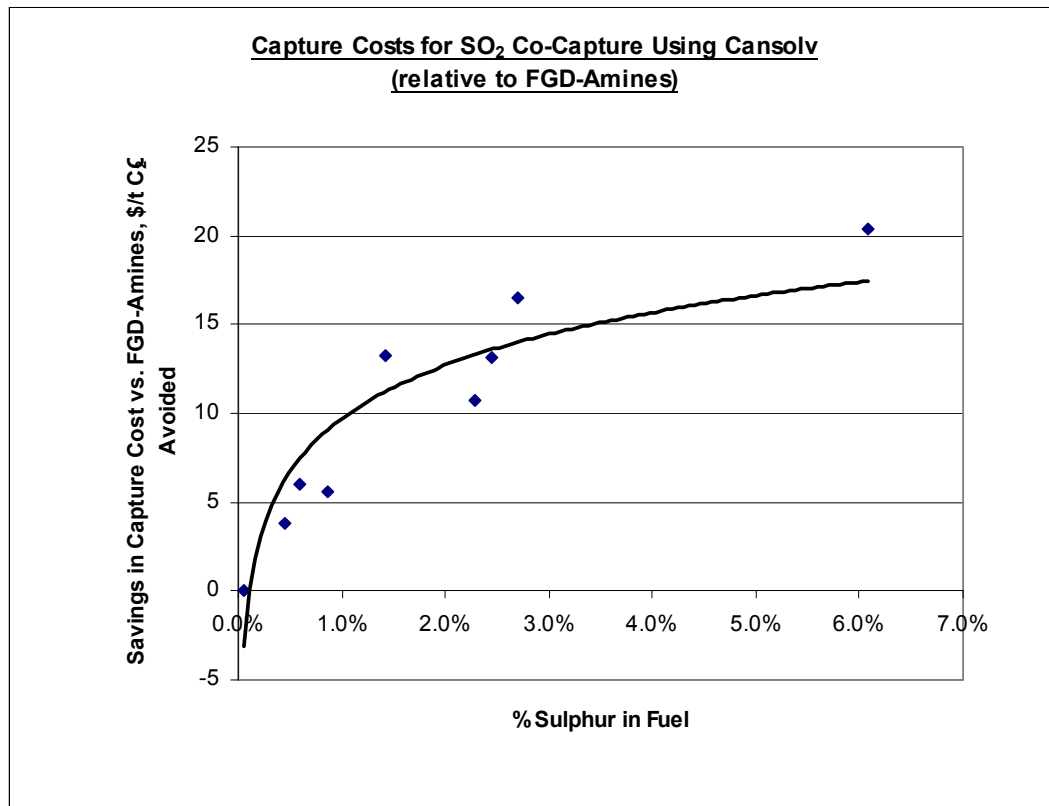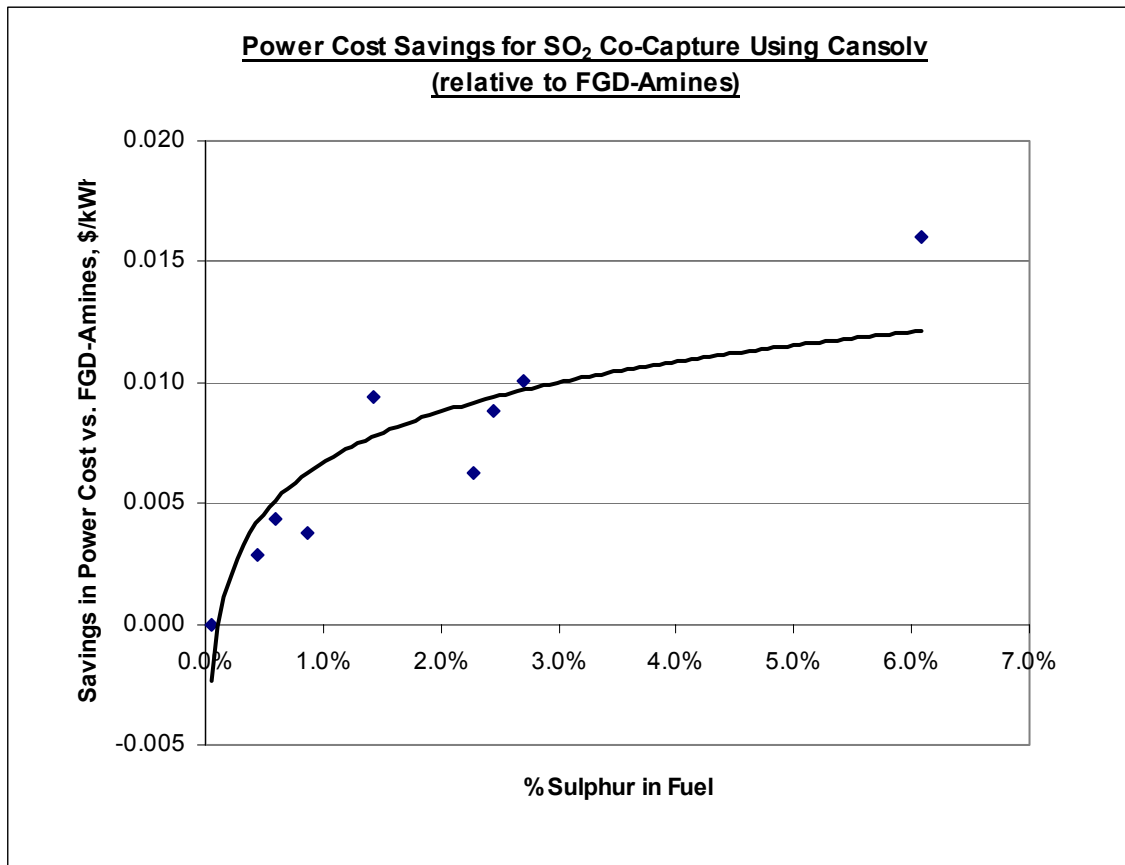
3.9 Co-capture Using Combustion in Oxygen Enriched Air (Case G) - Basis

This alternative uses combustion in an oxygen enriched stream instead of air ("oxy-fuel"), to obtain a flue gas with a much higher concentration of CO_2 and much smaller gas volume. Impurities such as SO_2 and some NOx components remain in the captured flue gas and are stored along with the CO_2. Figure 3.8 shows a BFD of the process.

Combustion in air results in a flue gas with roughly 80% inert nitrogen, which must be heated, handled, and then separated from the CO_2 for effective storage. Separation of nitrogen and CO_2 is difficult and the presence of 80% nitrogen results in large volumes of gas being handled.

Because combustion in oxygen results in a much higher boiler temperature than combustion in air, a portion of the flue gas is recycled back to the boiler. The most significant drawback to oxy-fuel systems is the cost and parasitic power demand of oxygen production. Cryogenic separation is typically used for

production of very large volumes of oxygen, particularly where oxygen purity is not critical. These systems are commercially available.

In our opinion CO_2 capture using oxy-fuel is promising if significant reductions in oxygen production costs could be achieved. We see these as happening through development of other air separation technologies rather than through improvements in cryogenic separation. Less expensive oxygen separation would also benefit IGCC/CO_2 capture schemes.

Figure 3.8

Case G - Co-Capture of SO$_2$ Using Oxygen-Enriched Combustion

Recycle Fan

Power Boiler 750 MW → Particulate Collection → Cooling & Water KO → Compression, Cooling & Pumping, 5 stages → TEG Dehy. → To pipeline 110 bar

Cryogenic ASU 143 MW

ash   water   water

Note: Dehydration simplified, typically follows third stage of compression
Separation of O_2 & N_2 not shown - prior to final pumping of dense phase to 110 bar

3.10 Co-capture Using Oxy-fuel - Results

Table 3.7 shows summary results for each fuel. The composition of the captured stream for each fuel is shown in Table 3.8.

Power and capture costs appear to be lower than conventional FGD-amine scrubbing systems. For the IEA Base coal for this case, power costs and capture costs were $0.066/kWh and $33/t CO_2 avoided respectively, versus

$0.074/kWh and $44/t CO_2 avoided for conventional CO_2 capture, or savings of $0.01/kWh and $17/t CO_2 avoided.

Similar to the Cansolv case, the savings relative to conventional capture appear to increase for Oxy-fuel as the sulphur content of the fuel increases. For the 6% sulphur Pet Coke, the relative savings were $0.023/kWh and $32/t CO_2 avoided respectively. These relationships are shown in Figures 3.9 and 3.10.

Our analysis assumed the plant runs on oxygen only. No provision was included for also running the plant on air. Some studies have made this provision which in our opinion significantly increases the costs for this option, as the flue gas system needs to be designed for air flow rates also, which are four-five times the oxy-fuel flow rates. If the plant is designed to run on air also, a case could be made for including conventional flue gas controls, such as SCR and FGD, for the periods of discharging to the atmosphere under air operation. This also would significantly increase costs for this option.

In SLI's opinion, the issue of whether to design oxy-fuel systems for operation under both air and oxygen needs further study as it has a large impact on expected power and capture costs.

We used a recycle rate 40% to produce an oxygen level of around 30% in the boiler. Some studies have used higher oxygen levels but 30% seems to be a consensus figure. Increasing the level of oxygen in the boiler should potentially reduce the cost by reducing the recycled gas volumes. We included a small reduction to the power plant base cost, due to the flue gas volumes being much smaller than for combustion with air.

We assumed an oxygen feed of 95% oxygen and 5% argon. It is our understanding that oxygen production costs increase significantly above this purity level. Two per cent air leakage into the boiler and recycle system was also included.

Production of oxygen was assumed to be via a commercially available cryogenic system. The amount of oxygen required for a 750 MW boiler was calculated to be on the order of 12,000 tpd, which to our knowledge, equates to about 3-4 times the largest single train oxygen unit built to date.

Power and capture costs for this case are sensitive to the cost of oxygen production. The power required to run the cryogenic units leads to a significant reduction in net power plant output. It is our opinion that oxygen production through cryogenic means is a mature technology that would not be expected to yield significantly lower production costs in the future.

SLI believes that a key to making this alternative more attractive is to significantly lower the oxygen cost through a radically different technology, such as oxygen

separation membranes. In our opinion, none of these are close to being commercially proven at this point.

The CO_2 stream produced primarily consists of about 97.5% - 99.5% CO_2 and about 0.5% - 2.5% SO_2, the SO_2 content varying directly with the sulphur content of the fuel. NOx level in the stored stream was estimated at about 10 – 50 ppm. The source of virtually all the NOx produced is from nitrogen compounds in the fuel itself. We have assumed that half of the NOx gases condense with the CO_2 and SO_2 during compression to 110 bar, the remainder being emitted to atmosphere with nitrogen and oxygen. This would result in plant-wide NOx emissions being 10 – 50 ppm, or about 10 - 50 g/MWh. In our opinion the amount of NOx produced in an oxygen combustion scenario needs further study.

We assumed 85% of the SO_2 and SO_3 produced is captured in the stored gas, with the remainder dissolving in water streams from free liquid removal vessels. These streams could require further treatment if they are too acidic.

In SLI's opinion, the following conclusions can be drawn about Case G, co-capture using oxygen enriched combustion:

❑ There could be an opportunity to lower CO_2 capture costs through co-capture of SO_2 through combustion in an oxygen-rich environment ("oxy-fuel");
❑ The composition of this stream would be primarily 97.5% - 99.5% CO_2 and 0.5% - 2.5% SO_2, with the SO_2 level being dependent on sulphur content of the fuel. There could also be 10 - 50 ppm of NOx present;
❑ Power and capture costs for oxy-fuel appear to be lower than conventional FGD-amine systems ($0.067/kWh and $31/t CO_2 avoided on average);
❑ Power and capture costs appear to be very sensitive to oxygen production costs and efficiency;
❑ Power and capture costs could significantly increase if provision is included for running the power plant on both air and oxygen;
❑ Benefits for an oxy-fuel system vs. conventional FGD-amine scrubbing appear to increase as the sulphur content of the fuel increases. For the 6% S pet coke, the savings for oxy-fuel system vs. FGD-amines was on the order of $32/t CO_2 avoided and $0.023/kWh;
❑ Power and capture costs for this case do not appear to be sensitive to the sulphur content of the fuel;
❑ The oxygen flow required for this size of boiler would result in on the order of 3-4 trains of the largest single-train ASU's built to date;
❑ Most trace elements are expected to be removed during either particulate removal, dehydration, or during the four stages of free liquid removal during compression. However, this case will likely have the highest trace element content in general, because the separation process is much simpler with fewer unit operations and no chemical absorption units.

Table 3.7

	C1	**C2**	**C3**	**C4**	**C7**	**ORI**	**HF2**	**PC1**	**C8**
Summary Results by Fuel **Case G - Co-Capture of SO_2 Using Combustion in Oxygen-Enriched Air**									
Parameter	**Base**	**Hi S**	**Lo S**	**Hi Ash**	**UltraLoS**	**Orimul.**	**HFO**	**Pet Coke**	**Lignite**
Net Power Output, MW	533	536	511	516	509	556	565	500	519
Total Capital Cost, $millions	1,010	1,000	1,040	1,040	1,030	900	880	1,050	1,020
Specific Capital Cost, $/kW	1,890	1,870	2,040	2,020	2,030	1,620	1,560	2,100	1,970
Power Cost, $/kWh	0.066	0.065	0.071	0.070	0.071	0.061	0.064	0.065	0.068
Plant Efficiency, %	30.1%	30.6%	28.1%	28.9%	28.3%	30.7%	32.6%	27.6%	28.8%
$/t CO_2 Avoided	33	27	36	34	41	24	29	24	29
$/t CO_2 Captured	22	19	23	22	26	18	21	16	20
CO_2 Emitted, kg/MWh	22	21	25	24	25	19	18	27	23

Table 3.8

	C1	**C2**	**C3**	**C4**	**C7**	**ORI**	**HF2**	**PC1**	**C8**
Captured Stream Composition for Each Fuel **Case G - Co-Capture of SO_2 Using Combustion in Oxygen-Enriched Air**									
% by Vol.	**Base**	**Hi S**	**Lo S**	**Hi Ash**	**UltraLoS**	**Orimul.**	**HFO**	**Pet Coke**	**Lignite**
CO_2	99.6%	98.9%	99.7%	99.5%	100.0%	98.6%	99.1%	97.7%	98.4%
Nitrogen	trace	trace	trace	trace	trace	trace	trace	trace	trace
Oxygen	trace	trace	trace	trace	trace	trace	trace	trace	trace
SO_2, ppm	0.4%	1.1%	0.3%	0.5%	0.0%	1.4%	0.9%	2.3%	1.6%
NOx, ppm	50	40	20	50	30	20	10	40	20
Hydrogen	trace	trace	trace	trace	trace	trace	trace	trace	trace
H_2S	trace	trace	trace	trace	trace	trace	trace	trace	trace
CO	trace	trace	trace	trace	trace	trace	trace	trace	trace

Figure 3.9

Capture Costs for SO₂ Co-Capture Using "Oxy-fuel"
(relative to FGD-Amines)

Figure 3.10

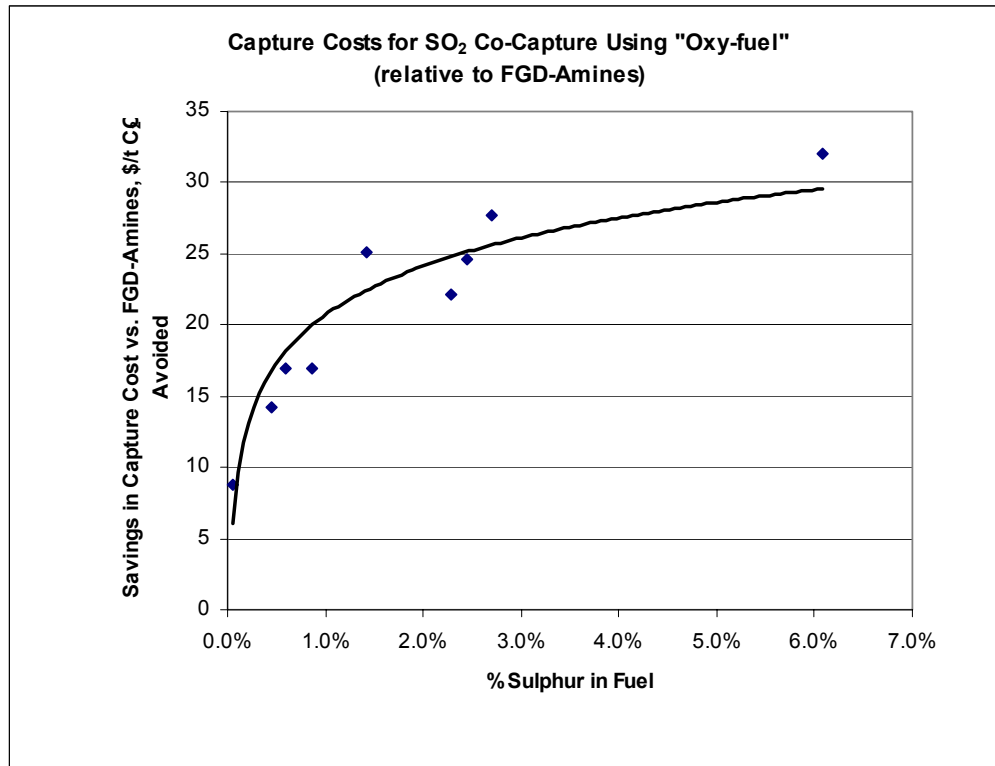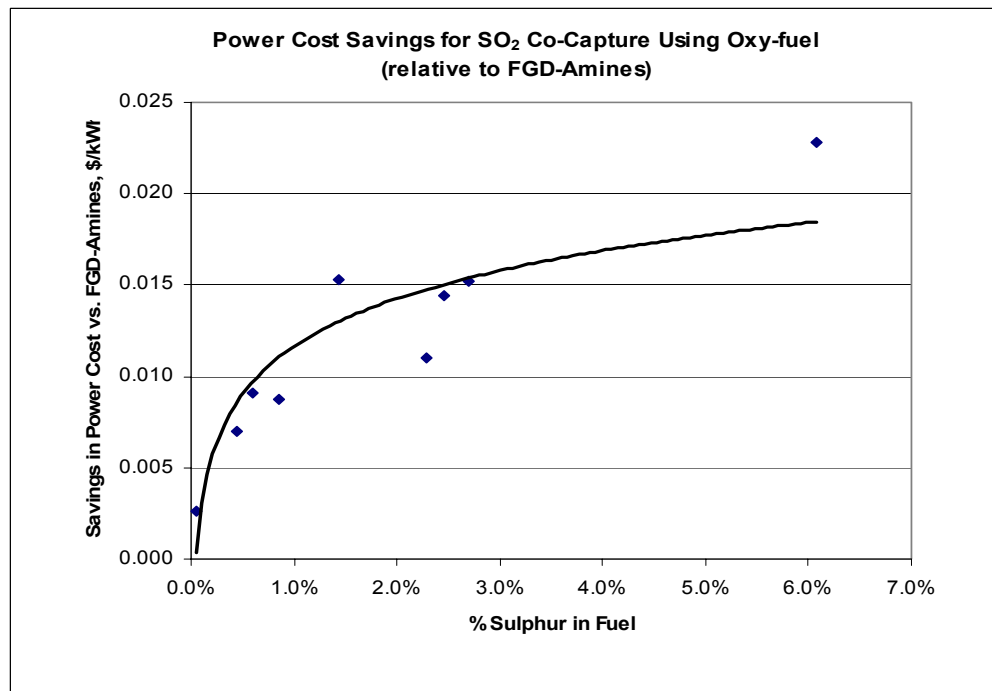Power Cost Savings for SO₂ Co-Capture Using Oxy-fuel
(relative to FGD-Amines)

3.11 Co-capture Using Gas Separation Membranes (Case H) - Basis

There are two types of membranes for this application:

❑ Gas separation membranes;
❑ Gas absorption membranes.

Gas absorption membranes employ an amine solution and use the membrane essentially as the mass transfer medium, where the CO_2 migrates across the membrane and into the amine solution for absorption. These systems appear to be a promising alternative to CO_2 absorption using conventional packed towers, offering reduced equipment sizes, which could be an important issue in offshore oil platforms.

Although gas absorption membranes are promising, we did not review them in detail because they are effectively a sub-set of conventional amine-based absorption systems and do not offer any co-capture opportunities.

Gas separation membranes rely on differences in the rate of physical or chemical interaction between each component in the flue gas and the membrane material. Membranes designed for the recovery of CO_2 ideally should have a high selectivity between CO_2 and other flue gas components, particularly nitrogen. This means the CO_2 permeates faster through the membrane than the nitrogen, leaving the permeate stream more concentrated in CO_2. For co-capture applications, membranes which are also selective between nitrogen and other pollutants, such as SO_2 and NOx, would be ideal, as SO_2 and NOx, would also concentrate in the permeate stream, *and not be released to atmosphere*.

The driving force for gas separation membranes is the partial pressure across the membrane. This poses a challenge for flue gas applications because flue gas must be compressed to a level that can provide an effective differential pressure across the membrane. The permeate is collected at the lower pressure, resulting in additional costs to compress the permeate stream for storage. If more than one stage is required, then secondary compression is also required. Some of the required compression horsepower may be recovered through use of a turbo-expander on the high pressure side of the membrane.

Some authors consider gas separation membranes to be a promising area for CO_2 recovery, and we observe that there appears to be significant research and development activity occurring in this area. A BFD for our system is shown in figure 3.11.

Figure 3.11

Case H - C-Capture Using Gas Separation Membranes

Note: Dehydration simplified, typically done after third stage of compression
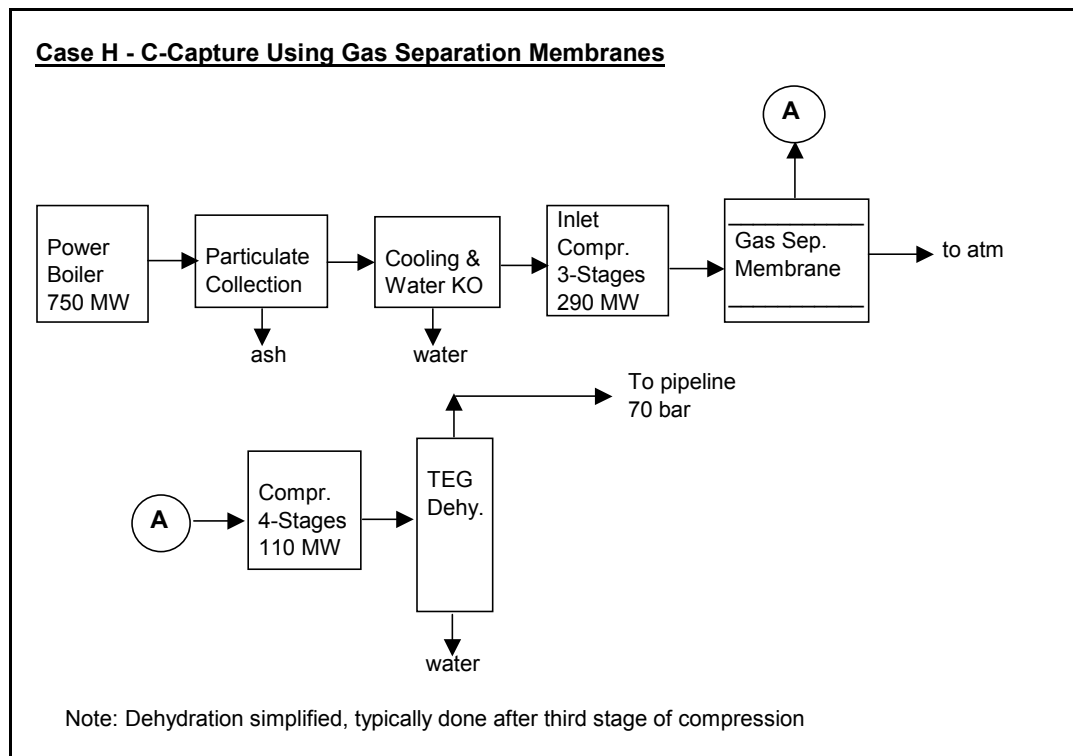
3.12 Co-capture Using Gas Separation Membranes - Results

Table 3.9 shows summary results for each fuel for the one-stage system. The composition of the captured stream for each fuel is shown in Table 3.10.

Co-capture using one-stage gas separation membranes without further CO_2 purification could produce a capture stream of about 50% CO_2 and 45% N_2, with 0.5% - 2.0% oxygen, 0.2% - 1.4% SO_2 and 600-1400 ppm of NOx. Power and capture costs for this case were significantly higher than the conventional FGD-amines case, or about $0.110/kWh and $120/t CO_2 avoided on average respectively.

The above figures are based on the following key parameters:

❑ Inlet pressure to the membranes was 18 bar;
❑ Approximately 85% of the CO_2 captured in the permeate stream;
❑ CO_2 content in the permeate of about 50%, leading to a calculated CO_2-N_2 selectivity of 6.0;
❑ CO_2-O_2 selectivity assumed to be the same as nitrogen;
❑ N_2-SO_2 and N_2-NOx selectivity assumed to be the same as CO_2;
❑ Turbo-expander used to recover a portion of the energy in the retentate stream.

Power and capture costs for this case appeared to be very sensitive to the inlet pressure assumed, and sensitive to the CO_2-N_2 selectivity. Reducing the inlet pressure from 18 to 5 bar while maintaining selectivity at 6.0 resulted in power and capture costs of about $0.08/kWh and $60/t CO_2 avoided for the base coal, or maybe 10% higher than conventional CO_2 capture. None of the vendors we contacted indicated that operating at this pressure was feasible given current or near-term status of membrane technology. The relationships between inlet pressure and capture and power costs are shown in Figures 3.12 and 3.13.

Maintaining the inlet pressure at 18 bar but increasing the CO_2-N_2 selectivity resulted in lower power and capture costs, but because of the large power draw required to compress the inlet stream to 18 bar, power and capture costs levelled off, becoming asymptotic at $0.105/kWh and $102/t CO_2. The relationships between selectivity and capture and power costs are shown in Figures 3.14 and 3.15.

Although inlet pressure and selectivity are related functions (i.e. – selectivity may increase as inlet pressure increases), the above analysis was done to gauge the impact that order-of-magnitude improvements in membrane performance might have on power and capture costs. The results are not intended to be rigorous. For example, membrane capital costs were not re-calculated as the inlet pressure was reduced, although we would expect these costs to increase as actual gas flows increased with lower pressure.

Use of gas separation membranes for CO_2 recovery from flue gas is an emerging application and the on-going research and development in this field could result in large gains in membrane performance. However, the above calculations indicate that order-of-magnitude improvements would be required in membrane technology to reduce costs to a level where they are comparable to an FGD-amines system, and that even with large improvements in membrane performance, the minimum operating pressure could be a limiting factor.

Responses from vendors concerning membrane performance and costs were not satisfactory. From our literature review, it appears that few studies to date have looked at membrane systems for flue gas recovery at anything more than a superficial level. Some studies had cost and performance data but these were used with caution as in our opinion they may have used optimistic assumptions regarding operating pressures. Vendors gave varying opinions on the optimum operating pressure. In the end, the inlet pressure of 18 bar was based on our professional judgement.

SLI believes there could be problems with gas separation membranes under coal-fired boiler service due to the presence of sub-micron particles not removed in the particulate removal operations. We included costs for additional particulate removal, but this was an allowance only and it was not clear how this

potential problem might be solved. It is possible that gas separation membranes could be limited to oil or gas fired boiler service for this application.

We also briefly reviewed a two-stage scenario where the permeate stream is re-compressed and fed to a second membrane stage. This scenario could produce a capture stream of about 90% - 93% CO_2 and 7% - 8% nitrogen, with 0.1% - 0.3% oxygen, 0.3% - 2.4% SO_2 and 1000-2400 ppm of NOx. Power and capture costs for this case were significantly higher, due to the additional equipment and compression power required. There could be an opportunity to use multiple-stage membranes to produce a gas that could be suitable for EOR (+95% CO_2).

In SLI's opinion, the following conclusions can be drawn about Case H, co-capture using gas separation membranes:

❑ CO_2 capture using a single stage of gas separation membranes without further CO_2 processing could produce a capture stream of about 50% CO_2, 45% nitrogen, with 0.6% - 2.0% oxygen, 0.2% - 1.4% SO_2 and 600-1400 ppm of NOx.
❑ Power and capture costs appear to be significantly higher than amine scrubbing, or about $0.110/kWh and $120/t CO_2 avoided (based on an inlet pressure of 18 bar);
❑ Costs are very sensitive to the inlet pressure assumed. Indications are that an inlet pressure of five bar or less would be needed to make this option competitive with amines;
❑ Use of turbo-expanders to recover some shaft power from the high pressure waste stream could reduce operating costs at the expense of increased capital costs, but not enough to change the overall conclusions regarding this process;
❑ Costs are also sensitive to the CO_2-N_2 selectivity of the membrane;
❑ Capture of CO_2 from coal-fired boiler flue gas may have significant technical concerns regarding membrane fouling with sub-micron fly ash particles.

Table 3.9

Summary Results by Fuel									
Case H - Co-Capture Using Gas Separation Membranes									
Parameter	**C1** **Base**	**C2** **Hi S**	**C3** **Lo S**	**C4** **Hi Ash**	**C7** **UltraLoS**	**ORI** **Orimul.**	**HF2** **HFO**	**PC1** **Pet Coke**	**C8** **Lignite**
Net Power Output, MW	421	423	402	404	399	444	448	373	404
Total Capital Cost, $millions	1,510	1,510	1,560	1,580	1,540	1,370	1,360	1,600	1,550
Specific Capital Cost, $/kW	3,600	3,570	3,880	3,910	3,870	3,090	3,030	4,280	3,830
Power Cost, $/kWh	0.109	0.108	0.118	0.118	0.117	0.100	0.104	0.119	0.116
Plant Efficiency, %	23.8%	24.2%	22.1%	22.6%	22.2%	24.5%	25.8%	20.6%	22.4%
$/t CO_2 Avoided	118	110	124	124	131	103	116	114	118
$/t CO_2 Captured	54	52	52	53	53	52	58	46	51
CO_2 Emitted, kg/MWh	210	200	240	230	240	170	170	270	230

Table 3.10

% by Vol.	C1 Base	C2 Hi S	C3 Lo S	C4 Hi Ash	C7 UltraLoS	ORI Orimul.	HF2 HFO	PC1 Pet Coke	C8 Lignite
Captured Stream Composition for Each Fuel
Case H - Co-Capture Using Gas Separation Membranes
CO_2	53.1%	52.7%	54.1%	53.6%	54.4%	51.2%	50.6%	52.7%	52.7%
Nitrogen	44.9%	44.9%	44.0%	44.3%	43.8%	47.0%	47.6%	43.8%	44.5%
Oxygen	1.5%	1.5%	1.5%	1.5%	1.5%	0.6%	1.1%	1.9%	1.5%
SO_2, ppm	0.3%	0.7%	0.2%	0.3%	0.0%	0.9%	0.5%	1.4%	1.0%
NOx, ppm	1,280	1,168	963	1,389	995	591	726	1,064	917
Water	0.1%	0.1%	0.1%	0.1%	0.1%	0.1%	0.1%	0.0%	0.2%
CO	trace	trace	trace	trace	trace	trace	trace	trace	trace

Figure 3.12

Capture Cost of CO_2 Using Gas Separation Membranes

Figure 3.13

Cost of Power for Capture Using Gas Separation Membranes

[Chart: X-axis "Membrane Inlet Pressure, bar" from 0 to 40; Y-axis "Power cost, $/kWh" from 0.06 to 0.13. Data points showing increasing power cost with pressure, with a fitted curve rising from about 0.09 at 8 bar to 0.12 at 35 bar.]

Figure 3.14

Cost of CO$_2$ Capture Using Gas Separation Membranes

[Chart: X-axis "Membrane Selectivity, CO$_2$-N$_2$ (at fixed inlet pressure of 18 bar)" from 0 to 60; Y-axis "Capture Cost, $/t CO$_2$ Avoided" from 60.0 to 140.0. Data points decreasing from about 128 at selectivity 3 to about 112 at selectivity 50.]

Figure 3.15

Cost of Power for Capture Using Gas Separation Membranes

3.13 Other Co-capture Alternatives for Conventional Boilers

Other CO_2 capture systems that were briefly reviewed for co-capture alternatives were physical absorption, cryogenic separation, and adsorption. These are briefly described in the following sections. In general it was felt that these systems either did not offer any co-capture alternatives, or else the technology itself was not well suited to CO_2 recovery.

3.13.1 Physical Absorption (Selexol® or Similar System)

Physical absorption of CO_2 using a Selexol® (dimethylether of polyethylene glycol) or Rectisol® (cold methanol) is a well established method of removing CO_2 from gaseous streams. The CO_2 is physically absorbed by the solvent, CO_2 being more soluble as the pressure increases. The solvent is regenerated typically by pressure reduction. These systems are commonly used in gas processing and have been proposed as the "baseline" for CO_2 recovery in gasification systems.

The primary difficulty in using physical absorption for flue gas applications lies in the atmospheric pressure of the flue gas and the need to compress it to a level where the $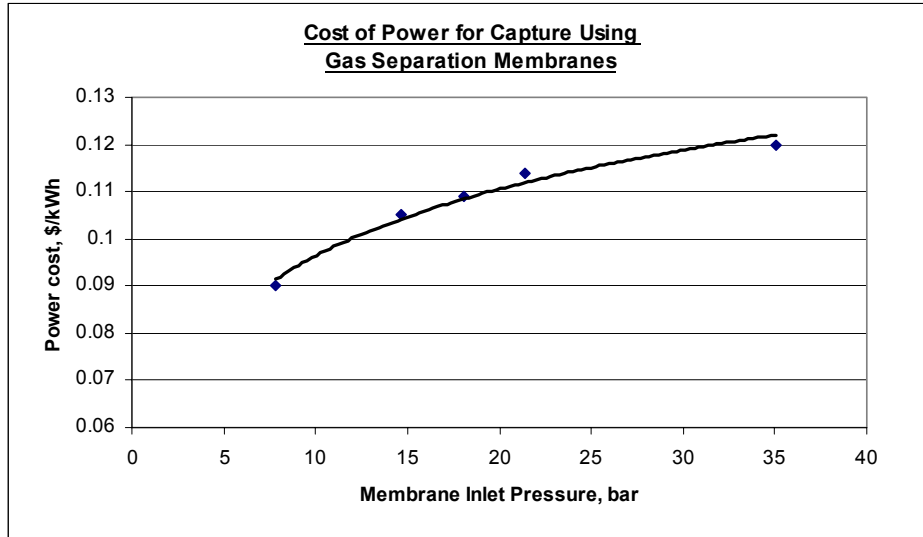CO_2$ solubility is such that the absorption tower can be sized effectively. As the pressure increases the solubility increases and the actual gas flow decreases. Both of these effects result in smaller vessels.

SLI did not investigate co-capture alternatives using physical absorption in detail due to the high inlet pressure required for effective operation. We were told by Selexol's supplier that a pressure of 20 – 25 bar is required, which would result in significant parasitic power demands. Although this pressure is not much higher than the 18 bar used for gas separation membranes, membranes were developed further as they are an emerging technology and would seem to offer a greater chance for breakthroughs in performance.

3.13.2   Cryogenic Separation

Cryogenic separation of CO_2 from flue gas was only briefly reviewed as it is generally accepted that the power required to cool and compress the large volume of flue gas to the region required for CO_2 liquefaction is prohibitive. Further, this is generally a mature technology that is not expected to yield further significant reductions in cost or efficiency.

3.13.3   Adsorption

Pressure Swing Adsorption (PSA) and Temperature Swing Adsorption (TSA) are well established technologies for gas purification. The gas is physically attracted to the solid, and then recovered by either reducing the pressure (PSA) or elevating the temperature (TSA). Alumina and zeolite molecular sieves are used commercially to separate CO_2 from natural gas streams. These systems are used to separate CO_2 in cases where pressures are higher and the CO_2 is at a much lower concentration than typical for flue gas (400 ppm to 1.5% CO_2). It is generally recognized that the low pressure and high CO_2 concentration in flue gas would make adsorption unattractive for flue gas recovery.

3.14   Maximum Level of Impurities

The maximum level of impurities in the co-capture alternatives considered (including gasification alternatives) is shown in Table 3.11. Only the "likely" cases have been included for this analysis (i.e. – Case C, capture of the entire flue gas stream, was omitted). In our opinion the estimated levels of trace elements have a high degree of uncertainty. The fate of trace elements in a complex CO_2 capture process, including dehydration and multiple stages of compression, cooling, and liquid knock-out, is not very well understood, and detailed investigation of this phenomenon is beyond the scope of this conceptual level study.

In our opinion, the focus for co-capture of other impurities should be on SO_2 for conventional boiler cases and on H_2S for gasification systems. In each case the level of these impurities is directly related to the sulphur content of the fuel. The maximum levels of SO_2 and H_2S both resulted from the 6% sulphur Pet Coke used. In our opinion, while this is a high sulphur level, fuels with even higher sulphur content could potentially have been selected.

Given that high sulphur fuels generally trade at a discount, in a carbon-constrained economy, the co-capture of sulphur components particularly for high sulphur fuels could present an opportunity to lower the impact of the CO_2 capture penalty.

Table 3.11

Maximum Level of Impurites in Co-Capture Streams - Likely Cases Only				
Component	**Level**	**Case Description**	**Case ID**	**Fuel**
CO_2	99.9%	FGD-Amines	E	All
Nitrogen	47.6%	Gas Sep. Membranes	H	Heavy Fuel Oil
Oxygen	1.9%	Gas Sep. Membranes	H	Pet Coke
Water	0.0%	Dehydration all cases		
SO_2	2.9%	SO_2 Co-capture using Cansolv	F	Pet Coke
NOx	1400 ppm	Gas Sep. Membranes	H	Coal C4 (India)
Hydrogen	1.8%	Gasification Base Case	B	All fuels
H_2S	3.4%	H_2S Co-capture - Gasification	B1	Pet Coke
CO	0.2%	H_2S Co-capture - Gasification	B1	All fuels
Methane	trace	H_2S Co-capture - Gasification	B1	All fuels
Trace Elements				
Mercury	5.6 E-05 ug/Nm3	SO_2 Co-capture - Oxyfuel	G	Coal C2 (US)
Arsenic	8.6 E-04 ug/Nm3	SO_2 Co-capture - Oxyfuel	G	Coal C2 (US)
Selenium	3.2 E-04 ug/Nm3	SO_2 Co-capture - Oxyfuel	G	Coal C2 (US)
Cadmium	not detectable	SO_2 Co-capture - Oxyfuel	G	Pet Coke
Lead	not detectable	SO_2 Co-capture - Oxyfuel	G	Coal C3 (US)
Antimony	not detectable	SO_2 Co-capture - Oxyfuel	G	Coal C2 (US)
Cromium	not detectable	SO_2 Co-capture - Oxyfuel	G	Pet Coke
Nickel	not detectable	SO_2 Co-capture - Oxyfuel	G	Pet Coke
Vanadium	not detectable	SO_2 Co-capture - Oxyfuel	G	Pet Coke

Notes:
1. Likely cases only shown (e.g. - Case C - flue gas capture is excluded).
2. In some cases higher levels are released to atm.
3. In SLI's opinion, there is a high degree of uncertainty in the trace element levels.

4. GASIFICATION

4.1 No Capture Case (Case A) – Basis

Much of the work described in this section was performed under subcontract to SLI by Alberta Research Council Inc.

As per agreement with IEA GHG, the gasification cases used for SLI's study were modelled after the systems in IEA GHG Report PH4/19, 2003, "Potential for Improvement in Gasification Combined Cycle Power Generation with CO_2 Capture". A BFD of the system from the IEA GHG report is shown in Figure 4.1. The main elements assumed for the "no-capture" IGCC system are listed below:

- ❑ Texaco high pressure quench gasifier operating at 65 bar;
- ❑ Four 33% capacity gasifier units;
- ❑ Quench cooler;
- ❑ Cryogenic air separation unit (ASU) with 50% integration into IGCC cycle;
- ❑ Two gas turbines, class 9FA, producing nominal 830MW net power;
- ❑ Steam turbines;
- ❑ COS hydrolysis unit;
- ❑ Separate H_2S recovery using physical absorption (Selexol)[®];
- ❑ Claus plant to convert H_2S to sulphur;
- ❑ Sulphur storage and handling facilities.

Texaco gasifiers were chosen instead of Shell based on the lowest expected capital cost and power cost. The IEA report concluded that the Shell units had higher efficiency than the Texaco units but also higher capital costs. SLI did not attempt to optimize the gasifier selected for each fuel.

Discussions with Texaco indicated that due to the low calorific value of the high-ash Indian coal (C4) and the lignite (C8), neither are considered optimum feeds for a Texaco gasifier. Because of this, cases were not developed in detail for these fuels. These fuels could be more appropriate for a Shell or other type of gasifier.

Figure 4.1
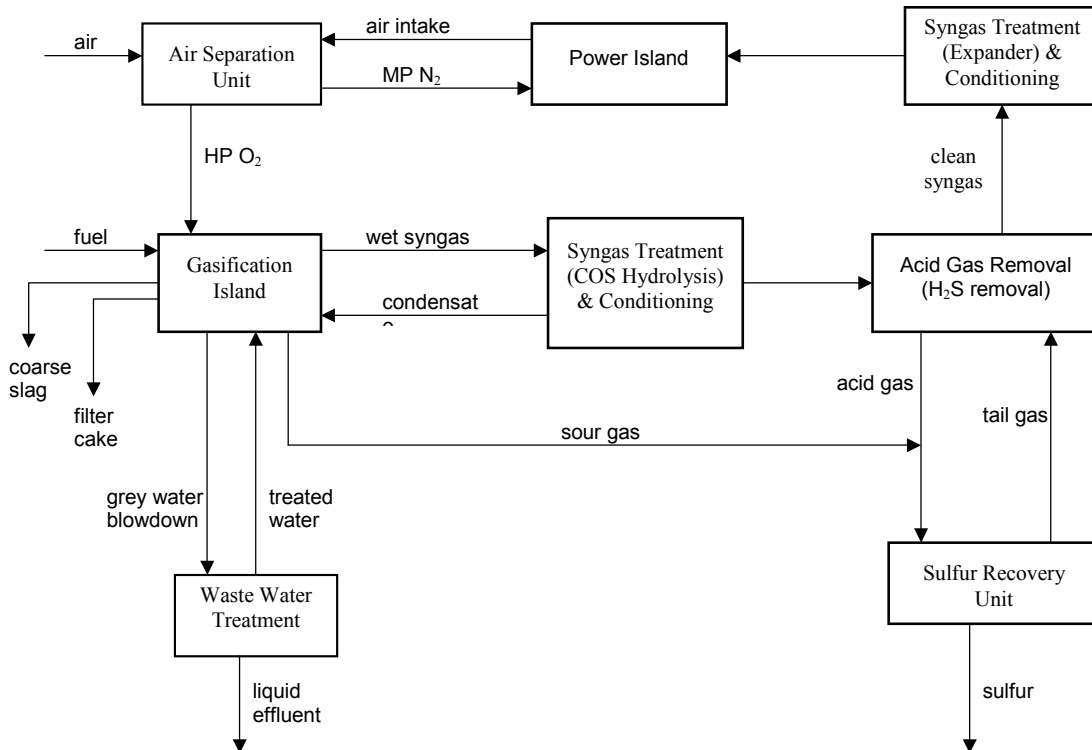
BFD of Gasification No Capture – Case A (ref. IEA report PH4/19)

4.2   No Capture Case A – Results

Table 4.1 shows summary results for the fuels developed in detail.

Power costs were on average about 15% higher than the "no capture" case for conventional boilers, or on average $0.048/kWh versus $0.042/kWh for the seven fuels developed for both cases. Specific capital costs were also about 15% higher, or $1,220/kW versus $1,070/kW. Due to the different estimating bases and procedures used for each, direct comparisons between conventional and gasification cases should be treated with caution.

Power costs for this case varied over a relatively narrow range and appeared to be very sensitive to the fuel cost assigned (i.e. – Pet Coke was $0.9/GJ, while Heavy Fuel Oil was $2.25/GJ). Orimulsion and Pet Coke were estimated to have the highest specific CO_2 emissions, at around 990 kg/MWh, while Heavy Fuel Oil was estimated to have the lowest emissions around 700 kg/MWh.

SOx emissions for this case were estimated around 40 g/MWh, NOx emissions around 200 g/MWh, and particulates at around 16 g/MWh. These levels are significantly below the levels typically associated with conventional boilers.

The fate of trace elements in gasification systems for both capture and no-capture cases, were reviewed by ARC and are discussed in section 4.9.

Table 4.1

| **Summary Results by Fuel** **Case A - IGCC with No Capture** | | | | | | | |
Parameter	C1 Base	C2 Hi S	C3 Lo S	C7 UltraLoS	ORI Orimul.	HF2 HFO	PC1 Pet Coke
Gross Power Output, MW	989	981	1,000	971	928	921	985
Net Power Output, MW	827	818	838	809	765	758	823
Total Capital Cost, $ millions	980	1,010	980	960	950	910	1,060
Specific Capital Cost, $/kW	1,190	1,230	1,170	1,190	1,240	1,200	1,290
Power Cost, $/kWh	0.046	0.047	0.045	0.046	0.051	0.055	0.044
Plant Efficiency, %	38.0%	38.2%	37.7%	38.0%	38.7%	39.0%	38.2%
CO_2 Emitted, kg/MWh	830	830	900	910	1,000	700	990
SO_2 Emitted, g/MWh	38	38	37	39	41	41	38
NOx Emitted, g/MWh	190	200	190	200	210	210	190
Particulate Matter, g/MWh	15	15	15	16	16	17	15

4.3 Baseline CO_2 Capture (Case B) – Basis

Baseline CO_2 capture cases for gasification systems were also modelled after IEA GHG Report PH 4/19, 2003. A BFD of this system from the IEA GHG report is shown in Figure 4.2. The main process elements are listed below:

❑ Gasifiers, power island, and ASU as per no-capture case;
❑ "Sour" gas shift reactor;
❑ Selexol® unit designed for *separate* recovery of H_2S and CO_2;
❑ Claus plant to convert H_2S to sulphur;
❑ Sulphur storage and handling facilities;
❑ CO_2 dehydration and compression to 110 bar;
❑ Overall 85% recovery of CO_2.

Figure 4.2
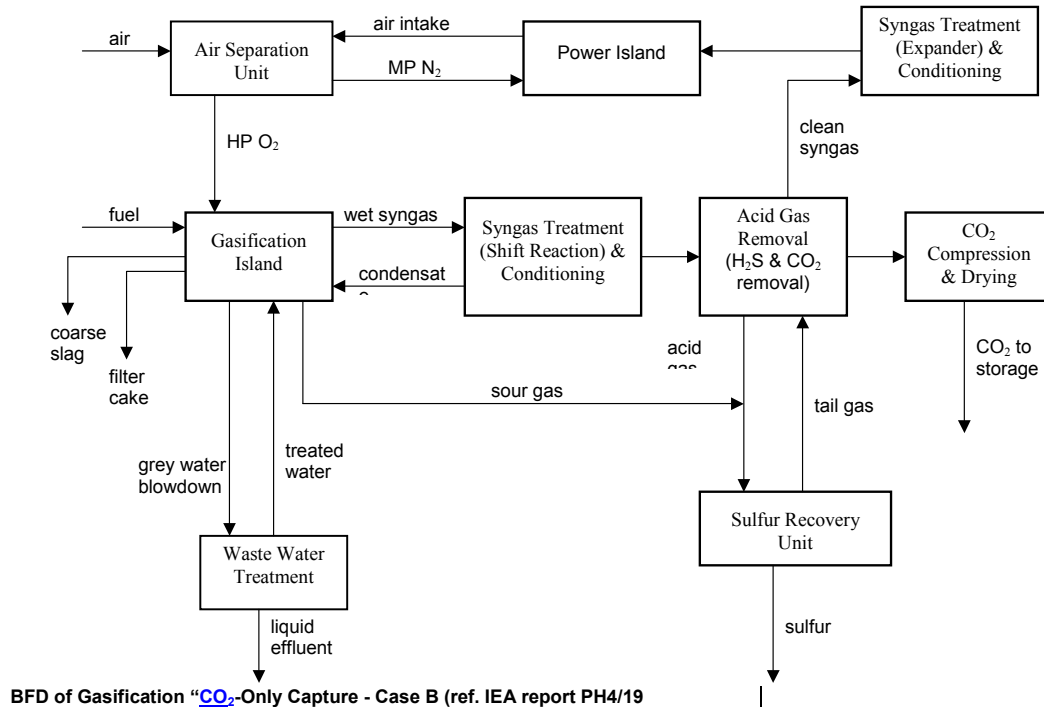
BFD of Gasification "CO₂-Only Capture - Case B (ref. IEA report PH4/19

4.4   Baseline CO$_2$ Capture Case – Results

The composition of the study fuels differ primarily in carbon to hydrogen ratio and in content of moisture, ash, oxygen, nitrogen, sulfur and trace elements. Changes in the fuel composition would have negligible impact on the recovered CO$_2$ composition in the IGCC case with recovery of a pure CO$_2$ stream. In this case, the recovered CO$_2$ composition is primarily determined by the design and operation of the Selexol unit. The composition of the "CO$_2$-only" case is essentially the same for all fuels and is shown in Table 4.4.

The capture of CO$_2$ in these types of systems reduces the plant efficiency from 38.3% to about 31.5% on average, and drops the net plant output from 760 – 830 MW (depending on the fuel) to 670 – 730 MW, or a reduction of about 12%. Table 4.2 shows summary results for the fuels developed in detail.

Capture costs varied from \$16.20–23.20 per tonne CO$_2$, with the average being around \$17.50/t. Power costs varied from \$0.057/kWh to \$0.069/kWh, and on average 25% higher than the "no capture" gasification case, or about \$0.060/kWh. Specific CO$_2$ emissions ranged from 130-180 kg/MWh.

Composition of the capture stream was about 97.5% CO_2 by volume, and also contained 1.8% hydrogen, 0.6% nitrogen, and 0.2% CO. This stream would likely be acceptable for EOR, but hydrogen, nitrogen, and CO all tend to increase the MMP and therefore the suitability of this stream for EOR would likely need to be confirmed for each reservoir. See section 6.2 for further discussion of this topic.

Table 4.2

Summary Results by Fuel							
Case B - IGCC with Conventional CO_2 Capture							
Parameter	**C1** **Base**	**C2** **Hi S**	**C3** **Lo S**	**C7** **UltraLoS**	**ORI** **Orimul.**	**HF2** **HFO**	**PC1** **Pet Coke**
Gross Power Output, MW	973	966	984	956	917	910	971
Net Power Output, MW	730	723	731	705	676	671	713
Total Capital Cost, $ millions	1,090	1,123	1,089	1,060	1,071	1,014	1,206
Specific Capital Cost, $/kW	1,490	1,550	1,490	1,500	1,580	1,510	1,690
Power Cost, $/kWh	0.057	0.059	0.057	0.058	0.064	0.069	0.058
Plant Efficiency, %	31.5%	31.6%	30.8%	31.0%	32.0%	32.3%	31.1%
$/t CO_2 Avoided	16.60	17.70	16.20	16.20	16.30	23.30	16.50
CO_2 Emitted, kg/MWh	150	150	160	170	180	130	180
SO2 Emitted, g/MWh	2.8	2.8	2.8	2.9	3.0	3.0	2.9
NOx Emitted, g/MWh	200	200	200	210	220	220	200
Particulate Matter, g/MWh	17	17	17	18	19	19	18

4.5 Co-capture of H_2S Using Sour Shift and Selexol® (Case B1) – Basis

The sulphur present in the feed fuel can be captured in a gasification system by re-designing the acid gas recovery in the CO_2 capture case (Case B) for combined (versus separate) recovery of CO_2 and H_2S. A Selexol® system is still used, with the combined stream sent to compression and dehydration. A BFD for this system from the IEA GHG report is shown in Figure 4.3, and the main elements are listed below:

❑ Gasifiers, power island, and ASU as per no-capture case;
❑ "Sour" gas shift reactor;
❑ Selexol® unit designed for *combined* recovery of H_2S and CO_2;
❑ Dehydration and compression and of combined CO_2 and H_2S stream to 110 bar.

For the co-capture case, the composition of the fuel will impact the composition of the recovered CO_2/H_2S stream, the primary difference being in the H_2S

content of this stream. To estimate the range of co-captured gas composition, ChevronTexaco supplied a predicted gas composition at the exit of a Texaco gasifier for seven of the nine feeds of this study. ChevronTexaco supplied predicted gas compositions at the exit of the gasifier prior to the quench chamber assuming that 95% purity oxygen was supplied to the gasifier. Fuel C4 (India high ash bituminous coal) and fuel C8 (German lignite) were not appropriate feeds for a ChevronTexaco gasifier unless they are co-fed with pet coke or other high heating value fuel, and were not developed further.

To estimate the composition of the co-capture stream, the following assumptions were made based on the IEA GHG PH4/19 report:

❑ 85% recovery of CO_2 to the co-captured gas stream;
❑ Essentially complete conversion of fuel sulfur to H_2S after gasification and sour gas shift reaction, with 100% recovery of H_2S to the co-capture stream;
❑ Similar concentration of H_2, CO, N_2 and other gases in the co-capture stream due to carryover in the Selexol unit (these components make up <2% of the co-capture gas);
❑ Moisture content of 0.3% in the co-capture gas.

Figure 4.3
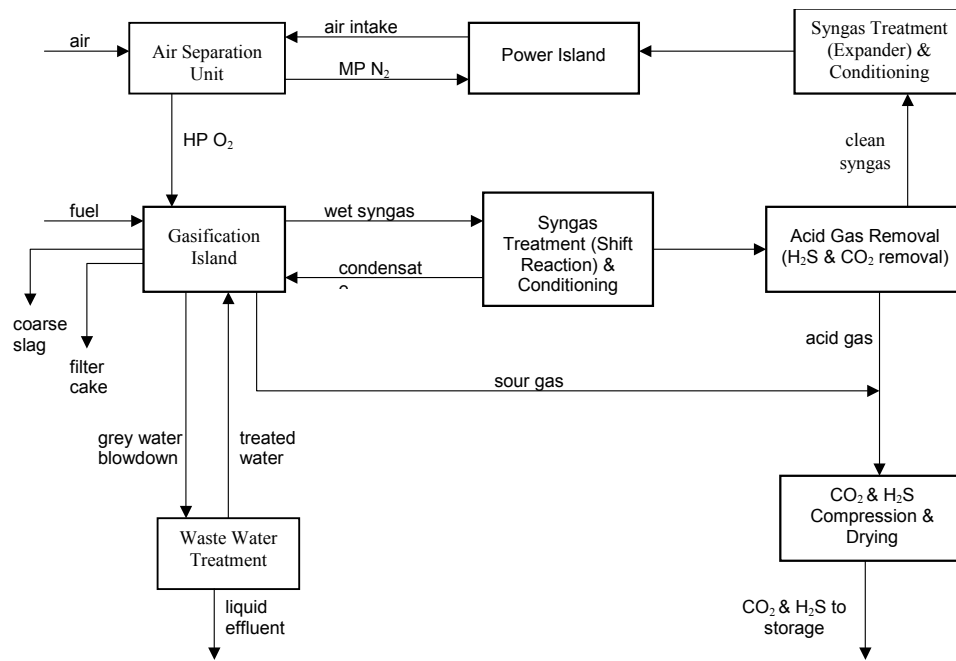
BFD of Gasification Co-Capture – Case B1 (ref. IEA Report PH4/19)

4.6   Co-capture of H_2S Using Sour Shift and Selexol$^®$ – Results

In this case the Selexol system is designed for combined capture of H_2S and CO_2, and the combined stream is sent for compression and dehydration. A primary advantage of this process is that the sulphur processing and handling facilities are eliminated. Disadvantages are that the energy value of the H_2S is lost, and revenue from sulphur sales is also lost.

Table 4.3 shows summary results for each fuel developed in detail for this system. The composition of the captured stream for each fuel is shown in Table 4.4. Table 4.5 summarizes key parameters for the three gasification cases for the base IEA coal C1. The results in Table 4.5 are primarily based on IEA GHG report PH 4/19.

Based on these results there appears to be an opportunity to reduce CO_2 capture costs through co-capture of H_2S using sour shift and Selexol$^®$ or similar system. Estimated savings for power costs and capture costs relative to the "pure CO_2" capture gasification case, were about $0.005/kWh and $6-7/t CO_2 avoided.

Relative savings for co-capture versus the "pure capture" case appear to increase as the sulphur content of the fuel increases. This relationship is shown in Figure 4.4. For the fuels with higher than 3% sulphur, savings were $8-10/t CO_2 avoided and $0.007-0.008/kWh for power costs.

Composition of the capture stream ranged from 94.5-98% CO_2 by volume, and from 0.6-3.4% H_2S. The capture stream also contained 1.7% hydrogen and about 0.2% CO. This stream would likely be acceptable for EOR, as although both hydrogen and CO tend to increase the MMP, the H_2S tends to decrease MMP.

There could be serious concerns regarding injection of a capture stream containing a significant level of H_2S into a "sweet" oil reservoir, due to the additional health and safety and cost concerns. This topic is further discussed in section 5.2.

In SLI's opinion, the following conclusions can be drawn about co-capture of H_2S with CO_2 using a sour shift and a Selexol$^®$ process:

❑ There are opportunities to reduce power and capture costs by using a "sour shift" and Selexol process designed for combined capture of H_2S and CO_2;
❑ Cost savings for the above compared to a CO_2–only capture gasification system are on the order of $0.005/kWh and about 6-7$/t CO_2 avoided;
❑ Savings appear to increase as the sulphur content of the fuel increases;
❑ The above system results in a capture stream consisting of 94.5-98% CO_2 and about 0.5%-3% H_2S for the fuels studied. For the 6% Pet Coke case,

CO_2 content was slightly less than 95% by volume. There is also 1.7% hydrogen and 0.2% CO in these capture streams;

❑ There are some serious corrosion and safety concerns regarding transport and compression of the above stream. Corrosion can be minimized by dehydration to well below the dew point of water, but hydrogen in the presence of H_2S may present other concerns, such as steel embrittlement;

❑ H_2S is an extremely toxic gas which will require additional safety precautions and design allowances. Gas containing H_2S is routinely handled in Western Canada, but it does lead to additional cost and may face resistance in other jurisdictions if shipped through populated areas;

❑ The above stream would likely be acceptable for EOR, but there are some significant concerns with injecting streams containing H2S into a "sweet" reservoir.

Table 4.3

Summary Results by Fuel							
Case B1 - IGCC Co-Capture of H_2S Using Sour Shift & Selexol							
Parameter	**C1** **Base**	**C2** **Hi S**	**C3** **Lo S**	**C7** **UltraLoS**	**ORI** **Orimul.**	**HF2** **HFO**	**PC1** **Pet Coke**
Gross Power Output, MW	980	973	990	963	923	917	977
Net Power Output, MW	742	742	728	710	713	717	713
Total Capital Cost, $ millions	1,050	1,050	1,050	1,050	990	960	1,060
Specific Capital Cost, $/kW	1,410	1,410	1,440	1,480	1,390	1,340	1,490
Power Cost, $/kWh	0.054	0.054	0.055	0.057	0.057	0.062	0.050
Plant Efficiency, %	32.0%	32.5%	30.7%	31.3%	33.8%	34.5%	31.1%
$/t CO_2 Avoided	12.70	10.20	14.50	15.10	7.10	10.70	7.80
CO_2 Emitted, kg/MWh	150	150	170	170	170	120	180
SO2 Emitted, g/MWh	2.6	2.6	2.7	2.7	2.7	2.7	2.7
NOx Emitted, g/MWh	190	190	190	190	190	190	190
Particulate Matter, g/MWh	17	17	17	17	17	17	17

Table 4.4

% by Vol.	C1 (CO₂ only)	C1 Base	C2 Hi S	C3 Lo S	C7 UltraLoS	ORI Orimul.	HF2 HFO	PC1 Pet Coke
	Captured Stream Composition for Each Fuel							
	Case B1 - IGCC Co-Capture of H₂S Using Sour Shift & Selexol							
CO_2	97.1%	97.2%	96.2%	97.3%	97.7%	95.7%	96.5%	94.4%
Nitrogen	0.6%	0.02%	0.02%	0.02%	0.02%	0.02%	0.02%	0.02%
Oxygen	trace	trace	trace	trace	trace	trace	trace	trace
Hydrogen	1.8%	1.7%	1.7%	1.7%	1.7%	1.7%	1.7%	1.7%
Water	0.3%	0.3%	0.3%	0.3%	0.3%	0.3%	0.3%	0.3%
H_2S	> 0.01%	0.6%	1.6%	0.4%	0.0%	2.1%	1.3%	3.4%
CO	0.2%	0.2%	0.2%	0.2%	0.2%	0.2%	0.2%	0.2%
Methane	trace	trace	trace	trace	trace	trace	trace	trace
Argon	0.05%	0.04%	0.04%	0.04%	0.04%	0.04%	0.04%	0.04%

Table 4.5

Parameter	No Capture	CO_2 Capture	H_2S Co-Capture
	Summary Results for Gasification Cases		
	(Using IEA Base Coal - results from IEA GHG Ph 4/19)		
Gross Power Output, MW	989	973	980
Net Power Output, MW	827	730	742
Total Capital Cost, $ millions	980	1,090	1,050
Specific Capital Cost, $/kW	1,190	1,490	1,420
Power Cost, $/kWh	0.046	0.057	0.054
Plant Efficiency, %	38.0%	31.5%	32.0%
$/t CO_2 Avoided	n/a	16.70	12.70
CO_2 Emitted, kg/MWh	830	150	150
SO2 Emitted, g/MWh	38.0	2.8	2.6
NOx Emitted, g/MWh	190	200	190
Particulate Matter, g/MWh	15	17	17

Figure 4.4

Capture Cost Savings for Co-capture of H_2S
(relative to "Pure" CO_2 Capture)

y-axis: Savings in Capture Cost, $/t CQ Avoided
x-axis: % Sulphur in Fuel
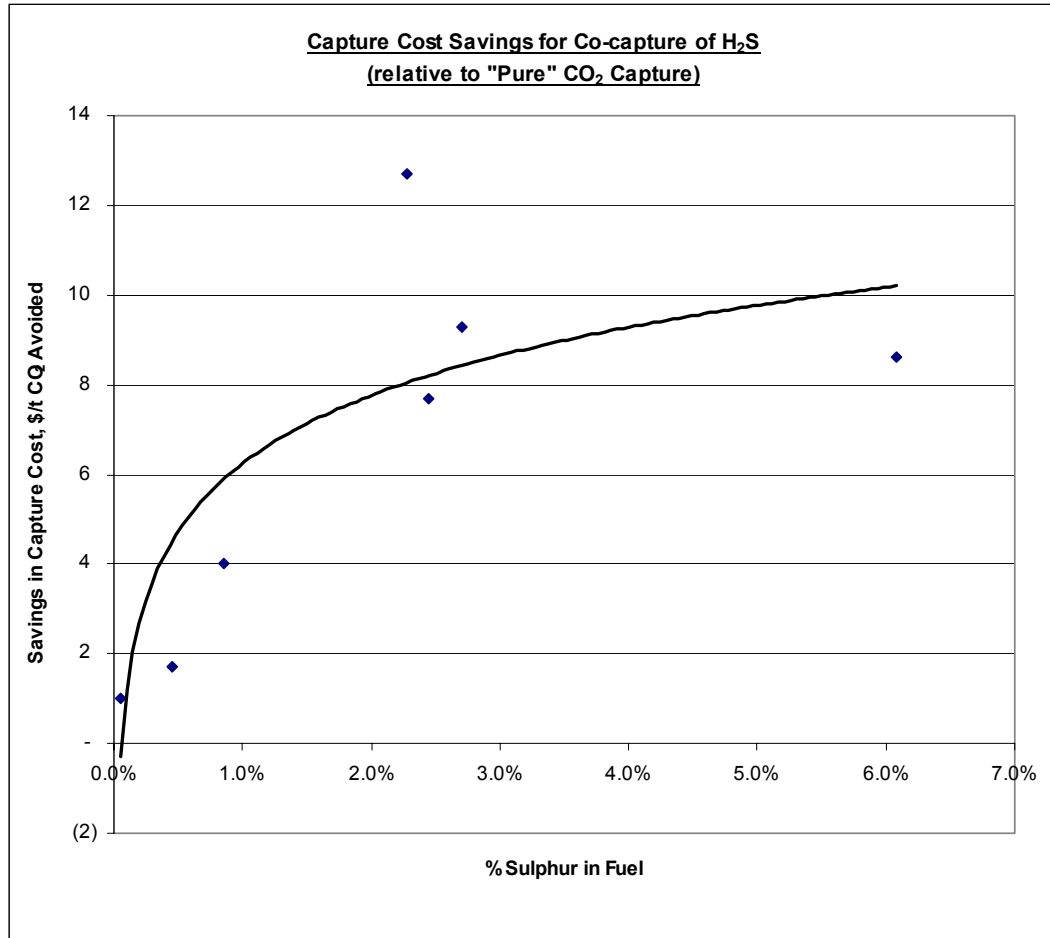
4.7 Co-capture of H_2S Using Sour Shift and Rectisol® (Case B2)

A disadvantage of a Selexol®-based system is that there is about 2% hydrogen in the captured stream. This represents a further potential loss of energy from the base case, in addition to the energy lost through diversion of the H_2S from the Claus plant to the capture stream. ARC proposed that using Rectisol® as the acid gas recovery system the amount of hydrogen could be reduced.

The BFD for this system is similar to Case B1 except that the Rectisol® process is employed. Both Selexol® and Rectisol® are well established commercial systems.

Following discussions with Lurgi, ARC developed compositions for the base capture case (Case B) and the co-capture case (Case B1) based on the IEA

base coal (C1), but using Rectisol® as a solvent. It appears that ARC's conclusions were that the level of hydrogen may be reduced from about 1.7% to roughly 1.2%, but there is an offsetting increase in CO from 0.2% to 0.4%. Both hydrogen and CO in the capture stream represent an energy loss to the overall system, in addition to having a potentially negative impact on EOR.

ARC also concluded that co-capture of H_2S using Rectisol® could offer savings when compared to CO_2–only capture (also using Rectisol® as the solvent). Lurgi commented that further optimization of the system, particularly when designed with co-capture as a primary goal, could result in both lower capture cost and possibly lower hydrogen and CO levels.

Engineering and costs were not developed either by ARC or SLI to the extent where a comparison could be made between Selexol® and Rectisol®. It is possible that the hydrogen content of a co-capture stream obtained using Selexol® could also be reduced. Selexol® licensors were not contacted by us regarding that possibility.

In our opinion, the issue of co-capture for gasification systems has not been extensively studied and we believe that further optimization and research could reduce co-capture costs, and the amount of hydrogen lost in the capture stream.

In SLI's opinion, the following conclusions can be drawn about co-capture of H_2S using a sour shift and combined H_2S and CO_2 recovery with Rectisol®:

❑ There may be opportunities to reduce the amount of hydrogen in the captured stream, either through using different solvents or through optimization of Rectisol® or Selexol®. Reductions in hydrogen or CO levels could come at the expense of other parameters;
❑ The decrease in hydrogen would have little impact on the compression, transport or final storage option, but would reduce the overall energy penalty associated with capture of CO_2 versus the no-capture case.

4.8   Co-capture of H_2S Using Gas Separation Membranes

ARC also reviewed the potential for co-capture in a gasification system using gas separation membranes. The membrane unit would be used to capture CO_2 from the sour water shift reactor. Gas separation membranes for this service would be potentially more attractive than for conventional boilers, because the gas stream from the shift reactor is typically at 25 bar, which would provide the differential pressure needed for the membrane to be effective.

This investigation indicated that polymeric membranes as they exist today or are likely to exist commercially for large scale applications in the near future are not selective enough between CO_2 and hydrogen for this application. While both CO_2 and H_2S would concentrate in the capture stream, it would also have a

much higher hydrogen content than Selexol® and thus higher energy losses, with no other apparent advantages.

4.9 Fate of Trace Elements in Gasification Capture Systems

Coal contains a significant amount of trace contaminants that have the potential to exit the IGCC process through the gas turbines, in the acid gas going to sulfur recovery, and in the recovered CO_2 stream. The fate of volatile trace metals, including lead, mercury and selenium, has been studied extensively in conventional coal-fired power plants. Due to the low concentration of these compounds and the resulting difficulty in accurate measurement and closure of mass balance, controversy remains over the fate of trace elements following combustion and gas clean-up, even for conventional boilers.

IGCC technology is still under development, and even less is known about the fate of trace elements in IGCC systems due to the relative lack of experience and sampling from operating plants. This section reviews information available in the literature on trace elements in IGCC systems, and whether trace elements may have any impact the CO_2 storage option.

A recent US DOE report (Brown et al., 2002) consolidates most of the data collected on the fate of trace elements in coal gasification. This report is based on thermodynamic equilibrium calculations and on sampling and analysis of three commercial scale demonstration plants:

❑ Tampa Electric's 250MW Polk Power Station;
❑ Cinergy's 262MW Wabash River Generation Station;
❑ Dow Chemical's 160 MW Louisiana Gasification Technology Inc. Project.

The Polk Power Station used a Texaco entrained bed gasifier. The other two plants used an E-Gas entrained bed gasifier. All three plants were oxygen blown with cold gas cleanup (water scrubber for particulate removal and amine scrubber for acid gas removal). Data collected from these plants should represent process conditions similar to the base case plant of our study.

General conclusions that were well supported by measurements at these operating IGCC plants included:

❑ No trace organics were found (dioxins and furans were below detection limits) in the cleaned synthesis gas;
❑ Although HCl, HFl, HCN and NH_3 compounds were produced at low concentrations in the gasifier, essentially all of these materials were removed from the gas stream during the water quench, water scrubbing and acid gas control equipment;
❑ Essentially all of the non-volatile metals were retained in the ash from the gasifier, or in particulates from the water quench and water scrubbers;

❑ Less than 5% of the lead was present in the gas turbine exhaust, with 30% removed in the gasifier slag and the remainder discharged in waste streams from the particulate and acid gas cleanup systems;

❑ Volatile metals such as mercury, were partially removed during gas clean-up, but a large portion remained in the cleaned gas and in the separated acid gas.

Mercury has received special attention in both equilibrium predictions and measurements at operating plants. Equilibrium calculations and plant testing indicates that the mercury is mainly present as elemental mercury in entrained flow gasification systems. Due to the difficulty of measuring the low concentrations of mercury present, and the potential for accumulation in gas cleaning equipment, mass balance closures for mercury in the plant sampling were poor (<50%). In sampling at the POLK and Wabash River plants, about 60% of the mercury in the feed exited in the gas turbine flue gas. In contrast, sampling at the LGTI plant found 23% of the mercury in the gas turbine exhaust and 8.2% in the exhaust of the tail gas incinerator of the sulfur recovery plant (Brown et al., 2002).

In summary, up to 60% of the mercury in the feed may be retained in the clean gas feed to the gas turbine and a significant amount of the mercury may be present in a CO_2 or combined CO_2-H_2S stream. Considering the increasing restrictions on mercury release, future IGCC plants may require mercury removal prior to combustion of the synthesis gas. Systems for mercury removal from synthesis gas are commercially available and may not add a significant cost to the plant. Rutkowski et al. (2002) estimated that mercury removal from an IGCC plant would add less than 1% to the cost of electricity. Some cost benefits may be possible for the CO_2-H_2S co-capture scheme as the mercury removal would only be required on the fuel gas going to the gas turbine.

Although mercury may be present in the recovered CO_2 stream, the concentrations would be very low (in the order of 10's of µg/Nm3). At these low concentrations, no negative effects for pipeline transportation, EOR or ECBM are anticipated.

Iron and nickel carbonyls ($Fe(CO)_5$ and $Ni(CO)_4$) can form during gasification, particularly with heavy oil and coke feeds high in metal content (Lagas, 1999). These compounds may lead to plugging problems in gas treating and sulfur recovery and also must be removed from the fuel gas prior to the gas turbine.

5. TRANSPORTATION & COMPRESSION ISSUES

5.1 Compression and Transportation Considerations

The potential impacts on compression and transportation for the co-captured streams were evaluated against the following criteria:

- ❑ Safety;
- ❑ Corrosion and operational issues;
- ❑ Environmental considerations;
- ❑ Additional costs due to gas phase versus dense phase transport.

Each case was evaluated as if that stream was being transported from the capture source directly to a dedicated storage site. The impact of co-mingling any of the identified streams was also evaluated (e.g. – mixing streams containing both SO_2 and H_2S).

It was assumed that all streams were dehydrated prior to pipeline transmission to well below the dew-point of water at the operating conditions using a commercially available tri-ethylene-glycol (TEG) based system.

A summary of the key compression and transport issues for the above streams is shown in Table 5.1. The capture stream resulting from a conventional FGD-Amines system is included for reference.

Table 5.1

Summary of Compression and Transport Issues

Parameter	Case E FGD-Amines (Reference)	Case C Flue Gas Injection	Case F SO₂ Capture (Cansolv)	Case G Oxy-fuel	Case H1 Membranes (One-stage)	Case B Gasification (capture)	Case B1 Gasification (co-capture)
CO_2 Purity, %	+ 99.9%	15 - 17%	97 - 99.5%	97.5 - 99.5%	50 - 55%	98 - 99%	97 - 99%
Other Components	none	Typical flue gas	Up to 3% SO_2	Up to 2.5% SO_2, some NOx	Mainly N_2, some O_2, SO_2 & NOx	Some H_2, N_2 & CO	Up to 3% H_2S, 2% H_2
Transport Phase	Dense	Gas	Dense	Dense	Gas	Dense	Dense
Safety Concerns	Minor	Minor	SO_2 toxic	SO_2 toxic	Minor	Minor	H_2S very toxic
Corrosion & Operating Issues	Dehydrate below dew point	Minor	More stringent dewpoint	More stringent dewpoint	Must be dry	Must be dry	Dewpoint, concerns re H_2S + H_2
Cost Implications	Stnd Wt., CS pipe	Much larger pipeline, stnd wt. CS	Extra safety & corrosion issues	Extra safety & corrosion issues	Larger pipeline, stnd wt. CS	Stnd wt. CS pipe	Extra safety & corrosion issues
Environmental Concerns	Minor	Minor	Release of SO_2	Release of SO_2	Minor	Minor	Release of H_2S

5.2 Safety Concerns

CO_2 is a colourless, odourless, non-combustible gas, generally considered to non-toxic except at high concentrations when it can cause asphyxiation and death. Physical discomfort has been reported starting at a concentration of 5%. Concentrations greater than 10% can lead to difficulty in breathing, impaired hearing, nausea, stupor within ten minutes and loss of consciousness within fifteen minutes. Deaths have been attributed to exposure in excess of 20%. At higher concentrations some of these effects can also be attributed to accompanying oxygen deficiency.

During an accidental release of pure CO_2, our calculations indicate that a dilution factor of about 20:1 would be required to lower the CO_2 concentration in air to a level where we would not expect any short-term health impact. We would generally expect high pressure CO_2 leaks, especially jet leaks from above ground equipment, to be dispersed in the atmosphere to where they would not be a health threat. Because CO_2 is denser than air under certain atmospheric conditions it could collect in low-lying areas in dangerous concentrations. Prolonged leaks, especially if from buried pipelines or in enclosed situations where jet entrainment of air to aid dispersion is inhibited, would be the most dangerous in this respect.

In our opinion, the risks associated with the handling and transport of pure CO_2 are technically manageable and low relative to the risks associated with many other industrial gases and chemicals. Properly designed and operated, a high-pressure CO_2 pipeline should present minimal health and safety risk.

The presence of small amounts of H_2S or SO_2 in the captured stream significantly increases the safety issues associated with the transport of "pure" CO_2. In our opinion the risks associated with transport of the streams containing up to 3% H_2S or 3% SO_2 are technically manageable with increased capital and operating costs, but such activities will face increased regulatory scrutiny and may encounter adverse public opinion.

H_2S is a flammable and extremely toxic gas with a strong odour at low concentrations. It has been long recognized in oil and gas production in North America as a very serious health and safety concern. Operators working in facilities handling any streams containing H_2S would need additional safety training and awareness, and additional emergency response training and equipment.

The National Institute of Occupational Safety and Health (NIOSH) recommends a maximum exposure level to H_2S of 10 ppm. Levels of 50 ppm can cause headaches and nausea, while levels over 100 ppm can cause loss of

consciousness within a few minutes. Exposure to levels around 500 – 1000 ppm can lead to immediate unconsciousness and death.

SO_2 is a non-combustible gas with a strong odour and has long been identified as a key component in air pollution. While it is not as deadly as H_2S, even at low concentrations SO_2 can have negative health effects. The Occupational Safety and Health Administration (OSHA) has set a continuous exposure limit of two ppm over an eight-hour workday. According to Material Safety Data Sheets, burning of the nose and throat and breathing difficulties will occur at higher concentrations, and levels of 100 ppm can be considered life-threatening.

Therefore while an accidental release of a captured stream containing SO_2 is not as dangerous as one with H_2S, both present additional health and safety concerns over transport of pure CO_2. The table below helps illustrate the issue.

Component	Parameter	Negative Impacts	Life Threatening
97% CO_2	Maximum Limit	5%	20%
	Dilution Factor Required	20 : 1	5 : 1
3% H_2S	Maximum Limit, ppm	10	50
	Dilution Factor	3000 : 1	600 : 1
3% SO_2	Concentration Required	20	100
	Dilution Factor	1500 : 1	300 : 1

The above table is meant only to give an indication of the health and safety concerns that are significantly increased due to the presence of H_2S or SO_2.

Both SO_2 and H_2S require a significantly higher level of dilution than CO_2 in the atmosphere to where their concentration would fall below the required levels. Both are heavier than air, and tend to concentrate in low lying areas. Under the certain atmospheric conditions a release of these gases in a low-lying area could be dangerous and potentially life threatening.

The handling and transport of gas containing H_2S or SO_2 would require the following precautions and additional equipment:

❑ Plant facilities require leak detection and shut-down equipment;
❑ Operators at the plant or pipeline require personal monitors and need to have access to emergency breathing apparatus;
❑ Operators require additional safety training, including emergency evacuation and revival training;

- Pipelines require a leak detection system, and emergency shutdown valves (ESDs) need to be included at locations determined by line volume, pressure and proximity to population;
- The pipeline operator needs to implement an emergency evacuation program for residents and businesses that could be affected by a leak;
- Public awareness and education programs concerning the dangers of H_2S and SO_2 would likely be required.

Gases containing much higher levels of H_2S are transported in many areas of North America and Europe, and in some cases near populated areas. In SLI's opinion the risks involved in the transport of gases containing H_2S are technically manageable. While we are not experts on public affairs issues, there appears to be a growing concern over handling and releases of H_2S in North America, and it is possible that there could be significant public resistance to the transport of this stream in certain regions.

In our opinion, similar precautions would need to be taken for the streams containing SO_2. We would also expect public concern to be expressed regarding transport of gases containing SO_2.

5.3   Corrosion and Operational Issues

Corrosion is a key issue for all the capture streams evaluated. CO_2, H_2S, and SO_2 all form very corrosive acids in the presence of water. Therefore for all cases we assumed corrosion resistant metallurgy for all equipment where water is present and dehydration of the final stream for transport and storage. These factors are included in all cost estimates. We assumed commercially available systems using TEG would generally be used, as these are effective in reducing the water content of the capture gas to well below the dew point of water at the pipeline conditions. This level is typically specified in natural gas distribution systems to be 0.128 g/m^3 (8 lbs./MMSCFD).

Once dehydration has been performed we do not anticipate corrosion to be a major issue for transport of any of the capture streams. Low-strength carbon steel is generally used for acid gas pipelines in Western Canada.

Dehydration is typically performed at an intermediate stage of compression, generally around 50 bar. Corrosion precautions must be taken in all units upstream of dehydration, including compressor wetted parts and inter-stage cooling an piping, where water tends to condense. Vendors indicated that stainless steel would be used for all compressor wetted parts, inter-stage piping and heat exchanger tubes. These steps are also employed for CO_2 only streams however, due to the corrosive nature of carbonic acid.

The following steps would need to be taken for CO_2 pipeline design, regardless of whether the stream is pure or contains H_2S or SO_2:

❑ Low-strength steel for pipeline materials;
❑ Measures to resist fracture propagation;
❑ Additional design corrosion allowance (i.e. – greater wall thickness);
❑ Additional corrosion monitoring;
❑ Possible injection of corrosion inhibitors.

The presence of H_2S or SO_2 in the quantities we would expect from the indications of this study would mandate the additional design steps noted in section 5.2.

Another transport issue for the gasification capture streams is the tendency of hydrogen under upset conditions to migrate from the dense phase into a separate gas phase, which could cause problems for the multi-stage pumps generally used to transport the stream.

Some concerns have been expressed that the presence of both H_2S and hydrogen can promote stress cracking in the pipeline. This concern can usually be lessened through selection of softer steel for the pipeline.

5.4   Cost Implications

Cost impacts generally fall under the following considerations;

❑ Additional costs to mitigate corrosion effects;
❑ Additional costs for safety precautions due to presence of H_2S or SO_2;
❑ Additional costs for those streams that would probably be transported in the gas phase versus the dense phase.

The actual cost impact will vary depending on location specific issues, population density, pipeline diameter, pressure, etc. In our opinion, as a rule of thumb, the cost of additional corrosion measures to pipeline CO_2 versus "sweet" natural gas is on the order of 10%. With the exception of the gas containing H_2S, we would expect that the presence of impurities identified in this study in the co-captured streams would increase the transport costs by about 5% over "pure" CO_2 pipelines, or on the order of $1-2 million for the case considered here.

Additional costs to transport the H_2S-containing stream might be on the order of 10% compared to "pure" CO_2 pipelines, or about $3-4 million in our case.

5.5   Environmental Issues

Environmental issues associated with the transport and storage of CO_2 generally fall into the following categories:

❑ Concerns over short-term, emergency situations, such as emergency venting or pipeline rupture;
❑ Concerns over leakage from the storage site over the long-term.

In SLI's opinion, with the exception of streams containing a significant level of H_2S, none of the other streams present significant environmental concerns, outside of the long-term issue of CO_2 leakage en-route or from the storage site, which is addressed in section 7.3.

In the case of a controlled emergency release of the captured streams (i.e. – venting due to upset conditions), we would not consider the release of any of these streams, for brief periods, to be of serious environmental concern. With the exception of H_2S, for the most part all the components are released to some degree already at non-capture facilities.

Most controlled releases due to upsets are generally brief, but persistent releases of H_2S or SO_2 near populated areas, could lead to complaints and potential problems for the operator of the capture facility.

In SLI's opinion, the following conclusions can be drawn about the compression and transport of the co-captured streams evaluated:

❑ None of the co-capture streams present significant corrosion or cost implications beyond those associated with the transport of "pure" CO_2;
❑ The streams containing H_2S and to a lesser degree SO_2 present additional but not insurmountable health and safety concerns. H_2S is extremely toxic and dangerous, and SO_2 presents significant health risks also. Additional safety precautions need to be included in the design of all facilities handling and transporting these gases;
❑ Gas containing much higher levels of H_2S are routinely transported in North America however, and in SLI's opinion the risks involved are manageable;
❑ In North America we observe increased concerns over H_2S–related operations, irrespective of population density. In our opinion the higher population density in Europe will result in a higher level of concern;
❑ As there is no experience we are aware of with pipelines containing CO_2 and SO_2, in our opinion the prudent approach would be to expect the same level of concern as raised by the presence of H_2S.

6. SUB-SURFACE STORAGE

6.1 Sub-surface Storage Options

The following storage options were considered:

- ❑ Enhanced Oil Recovery with CO_2 (EOR);
- ❑ Enhanced Coal Bed Methane recovery (ECBM);
- ❑ Storage in depleted oil and gas reservoirs;
- ❑ Storage in deep saline aquifers.

We consider these to be the most likely storage options, based on the large CO_2 volumes, level of technical development, and anticipated acceptance of each option. A summary of the key issues for these options is shown in Table 5.2.

For this study we included only those options that are already established technically (EOR, depleted oil and gas reservoirs, deep saline aquifers), or else expected to be technically established over the next ten years (ECBM). We excluded any options that in our opinion are not feasible at this scale, either due to the low volumes involved (e.g. – industrial gases or food and beverage use), or else are not expected to be proven within the next ten years (if ever), such as mineral carbonization or ocean sequestration.

Table 6.2

Summary of Sub-surface Issues

Parameter	Case E FGD-Amines (Ref.)	Case C Flue Gas Injection	Case F SO_2 Capture (Cansolv)	Case G Oxy-fuel	Case H1 Membranes (One-stage)	Case B Gasification (capture)	Case B1 Gasification (co-capture)
CO_2 Purity, %	+ 99.9%	15 - 17%	97 - 99.5%	97.5 - 99.5%	50 - 55%	98 - 99%	97 - 99%
Other Components	none	Typical flue gas	Up to 3% SO_2	Up to 2.5% SO_2, some NOx	Mainly N_2, some O_2, SO_2 & NOx	Some H_2, N_2 & CO	Up to 3% H_2S, 2% H_2
Transport Phase	Dense	Gas	Dense	Dense	Gas	Dense	Dense
Suitability for EOR	Good	Not suitable	OK, concerns with SO_2	OK, concerns with SO_2	Not suitable	Lowers MMP but should be OK	OK for sour fields, otherwise caution
Suitability for ECBM	Good	OK, quick N_2 break-through	OK, concerns with SO_2	OK, concerns with SO_2	Could be optimum due to N_2	Good	OK, concerns with H_2S
Suitability for Depleted Oil & Gas Reservoirs	Good	Not practical	OK	OK	Good	Not practical	OK
Suitability for Storage in Deep Saline Aquifers	Good	Not practical	OK	OK	Good	Not practical	OK

6.2 Impact on Enhanced Oil Recovery (EOR)

EOR involves the injection of a relatively pure CO_2 stream at elevated pressure into a producing oil field. In order to maximize oil recovery it is best to operate a CO_2 flood as a "miscible" process. Under miscible EOR, the CO_2 dissolves in the oil, which both reduces the oil's viscosity and displaces the oil, increasing oil flow to the producing wells. Significant amounts of CO_2 are also produced and must be separated from the oil and associated hydrocarbon gas before re-injection into the reservoir. Previous studies by SLI have indicated that EOR could be an important CO_2 storage mechanism as it has the best chance of offsetting capture costs through sales of incremental oil.

The main concerns regarding impurities in EOR are summarized below:

❑ Changes in minimum miscibility pressure (MMP) in the reservoir due to the presence of impurities;
❑ Presence of H_2S turning existing "sweet" fields "sour" and the additional costs and safety concerns arising from this change;
❑ Additional corrosion of down-hole equipment due to SO_2, H_2S and oxygen;
❑ Possible effects of injection of SO_2 into a reservoir already containing H_2S;
❑ Unsuitability of gas phase streams in general for EOR;
❑ Additional testing or piloting programs that may be required to confirm that there will be no negative impacts from the impurities in question.

Figure 5.1 shows a theoretical analysis of the impact of impurities in the CO_2 on the minimum miscibility pressure (MMP). The presence of contaminants changes the lowest pressure at which the CO_2 stream will be miscible at the reservoir temperature, and possibly the effectiveness of the EOR operation. Contaminants that reduce the MMP are acceptable or even preferred, while those that increase the MMP significantly could have a negative effect on oil recovery and project economics. An "immiscible" CO_2 flood is also expected to have less CO_2 storage capacity.

Lighter gas components such as oxygen, nitrogen and carbon monoxide increase the MMP. In SLI's opinion a combined total of any more than 5% of these components could have a negative impact on EOR performance. Cases C (flue gas injection; roughly 80% nitrogen) and H (gas separation membranes; roughly 45% nitrogen) would not be suitable for a miscible EOR due to the high nitrogen content. Case B (co-capture from gasification), with total contaminants of about 2% hydrogen, CO, and nitrogen, is probably acceptable for EOR.

Any of the capture streams of course could be used for an immiscible EOR flood or for simple reservoir pressure maintenance schemes. The impurities take up valuable reservoir storage volume however, and increase the specific storage cost for CO_2.

Figure 5.1
Impact of Contaminants on CO_2 MMP
Alston et al correlation
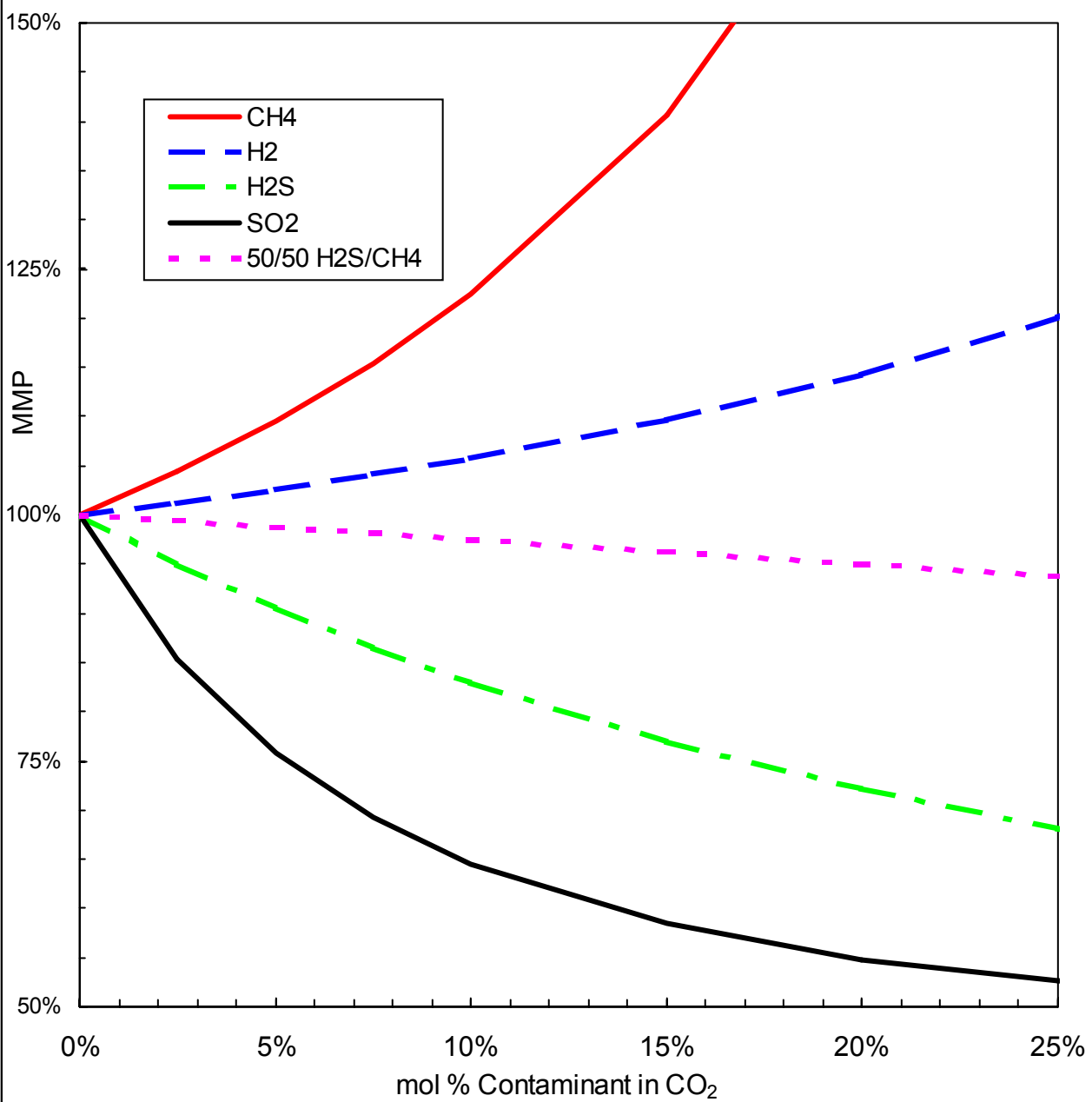
Oxygen is particularly harmful because in addition to raising the MMP, the presence of oxygen and free water could significantly increase corrosion rates in the production and downstream processing equipment. Oxygen may also oxidize the oil, making it more viscous and more difficult to extract and refine.

Heavier components such as H_2S, SO_2, and NOx depress the MMP, and are expected to be beneficial to EOR performance. Some SO_2 may dissolve in the reservoir water, particularly if the field has initially been water-flooded prior to EOR (often the case in North America).

There are significant health and safety concerns associated with H_2S and SO_2. Injection of a capture stream containing up to 3% H_2S or 3% SO_2 into an "sweet" oil field (i.e. – one where the oil contains negligible amounts of H_2S) would present additional safety concerns and could lead to existing pipelines and equipment being "de-rated" or unacceptable for this service, resulting in significant additional capital costs and higher operating costs. Injection of gas containing H_2S into a "sour" field would not present the same problems.

SLI is of the opinion that most of the contaminants would tend to be produced with the oil and gas (and CO_2), and whether they concentrate in the hydrocarbon gas stream or in the recycled CO_2 would depend on the separation process. The membrane manufacturers we contacted indicated that both H_2S and SO_2 would tend to concentrate with the recycled CO_2, and therefore would remain stored in the reservoir after the EOR project has been terminated.

There is also a concern over the potential deposition of elemental sulphur if gases containing SO_2 are injected into oil that already contains H_2S. It is unclear as to what the impact of this might be but most observers expressed caution until the effects are better understood.

It should be noted that the feed CO_2 for the Weyburn EOR project is reported to contain 1% H_2S.

The presence of a combined total of any lighter gases such as nitrogen, oxygen or CO in excess of 5% could result in a much higher pressure than 110 bar being required to keep the CO_2 in the dense phase. This would significantly increase both compression and transmission costs, due to the higher pressures required to maintain the dense phase, or larger pipelines if the stream is transported as a gas instead.

We found no published references to effects of trace components such as mercury in the storage mechanisms we evaluated, but we would expect minor, if any, impact on any geological sequestration scheme.

We note that the presence of impurities could result in additional testing or piloting being required prior to commercial operations, in order to confirm that there will be no negative impact from the impurities. To our knowledge, most EOR schemes have used relatively pure CO_2, with low levels of H_2S and no SO_2.

6.3   Impact on Enhanced Coalbed Methane Recovery

ECBM involves the injection of CO_2 at elevated pressure into a coal seam. Coal seams contain methane, and the injected CO_2 will tend to sorb preferentially onto the coal, releasing the methane and allowing it to be produced. A benchmark value for the ratio of CO_2 sorption to incremental methane production is 2:1. Water is often produced with the gas depending on the coal seam and location.

ECBM is based on the premise that the coal seam is relatively deep, uneconomic to mine, and would not be mined after CO_2 injection. ECBM projects are not well developed technically or commercially. Based on previous work by SLI it is possible that ECBM could be an important CO_2 storage option due to its large potential storage capacity and the potential to offset capture cost through recovered methane gas.

The main issues regarding impurities for use in ECBM are summarized below:

❑ Potential for the presence of H_2S turning the coal seam "sour" and the additional costs and safety concerns arising from this;
❑ Additional down-hole and produced gas handling equipment corrosion due to SO_2, H_2S and oxygen;
❑ Possibility of use of a CO_2-N_2 injection scheme to further enhance methane recovery, and selection of the optimum level of nitrogen;
❑ Potential migration of impurities into fresh water aquifers.

Based on the assumption that most coal seams do not contain significant amounts of H_2S, processing systems for ECBM gas are expected to be simple and relatively low cost, particularly if the production wells are shut in after breakthrough. Introducing either SO_2 or H_2S into the coal seam could increase the cost to process the recovered gas, and increase the cost to separate and compress CO_2 for re-injection. Both H_2S and SO_2 present additional safety concerns.

Based on discussions with the Alberta Research Council (ARC), SLI expects that both SO_2 and H_2S would sorb preferentially onto the coal during ECBM, similar to CO_2. This belief is based on a theoretical understanding of gas behaviour and molecular weights and to our knowledge it has not been confirmed either at a bench scale or in the field. If SO_2 or H_2S tend to sorb at the same or greater rate than CO_2, their presence might not be a significant issue if the production wells are not produced after CO_2 breakthrough.

There is evidence to believe that the presence of nitrogen can significantly increase recovery of methane in ECBM projects, particularly in the short term. Based on this, the stream resulting from Case H, (roughly 50% CO_2, 45% nitrogen) could be attractive for ECBM. It is not clear that overall recovery changes, but more methane is recovered faster, improving project economics. The disadvantage is that nitrogen breakthrough is rapid, leading to additional gas processing costs to separate nitrogen from methane.

Because ECBM candidates are relatively close to the surface there is a higher potential for the injected gases to come into contact with fresh water aquifers, and more chance of gases escaping to the surface. This could be perceived as a potentially harmful situation were injected gases containing either SO_2 or H_2S to migrate to a fresh water aquifer or be released at the surface.

6.4 Impact on Storage in Depleted Oil and Gas Reservoirs

This option involves the injection of the CO_2 stream into a depleted oil and gas reservoir. The primary advantage of this option is that storage costs are expected to be relatively low. This option is already being carried out in Western Canada, albeit for a different purpose ("acid gas injection") and at much lower CO_2 volumes and higher H_2S contents.

The main concerns regarding impurities on storage in depleted oil and gas reservoirs are summarized below:

❑ Gas phase streams are not optimum due to significantly higher pore volume occupied by the gas phase;
❑ Additional costs and safety concerns due to presence of H_2S or SO_2;
❑ Additional corrosion concerns due to SO_2, H_2S and oxygen;
❑ Mixing of streams containing both SO_2 and H_2S could result in possible sulphur conversion.

To maximize use of the reservoir pore space and to minimize infrastructure and field development costs, CO_2 should be injected into a reservoir where the temperature and pressure are such that the injected CO_2 can remain in the dense phase. Because of the large difference in density between the gas and dense phases, and because much of the pore space would be occupied by other gases, primarily nitrogen, neither Case C (80% nitrogen) or Case H (45% nitrogen) would be preferable for storage in depleted oil and gas reservoirs.

Beyond the safety issues associated with the stream containing H_2S or SO_2 and the possible sulphur conversion issue, we do not see any issues with the storage of any of the other streams in depleted oil and gas reservoirs.

6.5　Impact on Storage in Deep Saline Aquifers

Based on previous work by SLI, we expect that the issues associated with storage in deep saline aquifers to be similar to storage in depleted oil and gas reservoirs, except that costs for aquifer storage may be slightly higher. The fate of CO_2 in the aquifer is generally regarded as being less certain, as aquifers do not appear to have well defined natural rock sealing mechanisms.

It needs to be stressed that we are referring to non-potable saline aquifers, typically 1000 m deep or more. The experts we consulted are of the opinion that there is little if any communication between the aquifers targeted for this application and shallower fresh aquifers. A well-known demonstration of this option is the Norwegian Sleipner West project, where CO_2 recovered from offshore gas processing has been injected for three years in a 800-1000 m deep undersea aquifer.

There could be perceived concerns over the long term fate of the impurities in the aquifer, and to our knowledge little work has been done on this issue. We do not believe it is a serious technical concern, but it may need to be demonstrated that the rate of migration, dissolution and mineral carbonization in the host rock are reasonably defined, and the long term fate of the CO_2 and other impurities can assumed to be known.

7. OTHER ISSUES

7.1 Regulatory Implications

Relatively pure CO_2 has been transported considerable distances in the Western U.S. for several decades. To our knowledge this has occurred with no major incidents. A 333 km pipeline transporting 2.8 e6m3/day (100MMCF/D) dense-phase CO_2 containing up to 2% H_2S was recently built and is now operating between Beulah, North Dakota, US and Weyburn, Saskatchewan, Canada. The area through which the pipeline passes is generally flat prairie, with a low population density, and has experienced extensive oil production operations.

In our opinion, the regulatory process and facilities design for the Weyburn EOR project could serve as models for what may happen in other jurisdictions.

The key document we reviewed is The National Energy Board of Canada (NEB), Reasons for Decision, Souris Valley Pipeline Limited, MH-1-98, October, 1998. This document covers the 61km of the pipeline that passes through Canada. We also spoke informally with representatives of the NEB and the Alberta Energy and Utilities Board (AEUB) to discuss to pipelines containing SO_2. Both agencies were noncommittal on the subject of SO_2 and these comments should not be construed to be their official opinions.

The pipeline design standard currently in use in North America, CSA Z662, prescribes design considerations for CO_2 pipelines and those containing H_2S, but not for SO_2. A pipeline can be designed to meet specifications for both H_2S and SO_2. Neither of the above agencies has experience with a CO_2/SO_2 combination. They indicated that they would expect to use the CO_2/H_2S combination as a yardstick and would expect proponents to provide technical support for deviations from that norm, as well as technical confirmation that the design is appropriate to the CO_2/SO_2 environment. The need for public hearings would be determined by the level of public concern expressed.

In all other considerations, both agencies reacted to the CO_2/SO_2 combination as if H_2S was involved, and stated that standard risk assessment/mitigation requirements would be required commensurate with the level of hazard assessed.

Based on these investigations, SLI concludes that there should be no major regulatory impediments to the transportation of co-capture gases as evaluated in this study.

7.2 Public Acceptance Issues

There is growing concern over the health and safety issues regarding H_2S in North America, and an increasing level of resistance to facilities handling gas

containing H_2S near populated areas. We have observed this in Alberta, where there are established oil and gas industry, and a higher level of acceptance of these activities than in other regions.

SLI infers that if there is growing resistance to the transport of gas containing H_2S in North America, we would expect similar reaction in the EU, where awareness of environmental issues and resistance to initiatives perceived as being "unpopular" has historically been much higher.

SLI suggests that the most serious potential impediment to CO_2 pipelines in Europe, with or without H_2S or SO_2, may be the higher population density and the difficulty of avoiding population centres. At the least, this will result in higher capital and operating costs due to more frequent ESD valves and emergency response plans.

One factor in the public acceptance of such schemes is the existence of similar operations in the area. For example, it may be easier to implement a CO_2/H_2S pipeline in western North America where oil and gas production is common, than in eastern areas where it is not.

7.3   Monitoring Issues

With the exception of the streams containing H_2S or SO_2, SLI does not see any further issues associated with monitoring of impure CO_2 streams beyond those associated with the monitoring of "pure" CO_2 streams.

The capture streams containing H_2S or SO_2 present additional monitoring issues when storage in depleted oil and gas reservoirs is considered. The most likely route for CO_2 leaks are through abandoned and improperly sealed production wells. Over time, gases could leak through and around down-hole plugs, and through and around surface valves and casings. This is particularly a concern for sour gas fields, where the presence of H_2S and water can cause excessive corrosion. "Minor" CO_2 leaks which would generally not be a significant concern could be potentially serious if they contained H_2S or SO_2.

7.4 CO_2 Recycle Issues

This issue primarily concerns EOR, where due to the relatively quick CO_2 breakthrough CO_2 needs to be separated from the produced oil and gas and re-injected into the reservoir. From our previous work, it is our opinion that SO_2 and H_2S will tend to concentrate with the CO_2-rich permeate if a gas separation membrane is used for the application. This would imply that the H_2S and SO_2 would tend to be returned to the reservoir with the recycled CO_2 for permanent storage at the end of the project life.

Gas separation membranes typically used in this application still allow a small per cent of CO_2 to escape in the methane-rich retentate stream, and hence there would probably be small portion of H_2S or SO_2 in the methane stream as well. This would result in a minor increase in gas processing costs for a plant already processing H_2S, but could be a significant increase for "sweet" gas operations.

8.0 CONCLUSIONS

Based on our conceptual review of the benefits and issues of co-capture of other flue gas components in addition to CO_2, SLI makes the following conclusions:

Capture of Entire Flue Gas Stream

❑ Capture of the entire flue gas stream is not practical and leads to significantly higher power and CO_2 capture costs: about 60% loss of the plant's nameplate output, $0.14/kWh power cost, and $130/t CO_2 avoided for the base coal. Further, the captured flue gas would not be acceptable for EOR use due to the high nitrogen content, and would not be practical for storage in depleted oil and gas reservoirs or aquifers due to the relatively low volume of CO_2 stored per volume of reservoir;

❑ One potential application might be ECBM projects close to the capture source, where the pipeline inlet pressure could be reduced to 10 bar or less;

Co-Capture of SO_2 from Conventional Boilers

❑ There could be an opportunity to lower CO_2 capture costs through co-capture of SO_2 using either a Cansolv® or similar system or using oxygen-enriched combustion ("Oxy-fuel"). Either alternative results in a CO_2 stream with 0.5-3.0% SO_2 by volume. Based on the BFD we used, the Oxy-fuel stream could also contain 50-100 ppm NOx;

❑ Both alternatives are technically feasible but neither have been demonstrated at the scale required for 750 MW power plant. Oxygen-enriched combustion has been demonstrated at a pilot-scale only;

❑ Most of the NO_2 produced using Oxy-fuel is also captured at the same time, while NO produced with either system is generally released to atmosphere;

❑ Savings using either of the above alternatives with a 2% sulphur coal, compared to a conventional FGD-amines system, could be on the order of $13-25/t CO_2 avoided and $0.010/kWh;

❑ Savings for either alternative relative to an FGD-amines system increase as the sulphur content of the fuel increases. Savings with the 6% sulphur Pet Coke were estimated to be $20-32/t CO_2 avoided and $0.016-0.023/kWh.

❑ The SO_2-CO_2 capture streams would be acceptable for EOR in certain reservoirs. There may be some concerns regarding conversion of SO_2 to sulphur in reservoirs with H_2S. These streams would likely be acceptable for ECBM as we believe SO_2 will sorb preferentially onto the coal.

❑ The SO_2-CO_2 capture streams presents no significant compression or transportation issues, beyond that the stream must be dehydrated to well below the dew point of water at the pipeline conditions to minimize corrosion. This is also a requirement for transport of "pure" CO_2 however;

❑ There are serious health concerns regarding exposure to SO_2. The same concerns that apply to transport of streams containing H_2S would apply to streams containing SO_2 but possibly to a lesser degree;

❑ Power and capture costs for oxy-fuel are very sensitive to oxygen production cost. Our analysis assumed the plant runs on "oxy-fuel" mode only. No provision was included for running the plant on air. Some studies have made this provision which can significantly increase costs;

Co-Capture of H_2S from Gasification Systems

❑ There could be an opportunity to lower CO_2 capture costs by co-capture of H_2S using a sour shift reaction followed by physical absorption of the H_2S and CO_2 using Selexol® or similar process. This option results in a CO_2 stream with about 0.5-3% H_2S and 2% hydrogen by volume;

❑ Savings relative to "CO_2-only" capture for the base coal were estimated to be $4/t CO_2 avoided and $0.003/kWh. Savings appear to increase as the sulphur content of the fuel increases. Savings with the 6% sulphur Pet Coke were estimated to be $12.5/t CO_2 avoided and $0.007/kWh;

❑ A disadvantage of co-capture of H_2S and H_2 is that the energy value of these streams is lost. There may be an opportunity to reduce the amount of hydrogen in the above stream by using Rectisol® instead of Selexol®, which would help overall heat efficiency;

❑ The H_2S-CO_2 streams would be acceptable for EOR in reservoirs that already contain H_2S. There are significant concerns over injection of H_2S into a "sweet" oil field, where existing equipment may not be designed for this service. These streams would likely be acceptable for ECBM use as it is believed that H_2S sorbs preferentially onto coal. However, similar concerns exist over introduction of H_2S into a "sweet" coalbed methane seams;

❑ There are serious but not insurmountable safety concerns regarding transport of the above streams. H_2S is an extremely toxic gas that requires additional safety precautions and design allowances;

❑ Gases containing higher levels of H_2S are transported near populated areas in North America. There appears to be growing resistance to this in North America however, and we expect there to be some level of public resistance to transport of gases containing H_2S in the EU.

Co-Capture Using Membranes

❑ Opportunities for co-capture of other flue gas components using gas separation membranes are not promising. Power and capture costs for this alternative were much higher than conventional FGD-amines, or about $0.110/kWh and $112/t CO_2 avoided on average for the fuels considered.

❑ The above figures are based on an inlet pressure of 18 bar. The results are sensitive to the inlet pressure assumed. Indications are that an inlet pressure of 5 bar or less would be needed to make this option competitive with amines. Vendors we contacted indicated that such low pressures were not feasible given current or near-term developments in membrane technology;

❑ Gas separation membranes without further CO_2 processing could produce a capture stream of about 50% CO_2, 45% nitrogen, 1-2% oxygen, 1-2% SO_2 and 600-1400 ppm of NO_X. This stream would not be acceptable for EOR use but could be attractive for ECBM;

❑ Membrane capture costs are also sensitive to the CO_2/nitrogen selectivity of the membrane. Both power and capture costs decrease as selectivity improves; however, at a constant inlet pressure of 18 bar, costs appear to become asymptotic at $0.105/kWh and $110/t CO_2 avoided. This could lead to the conclusion that anticipated improvements in membranes still may not yield a competitive CO_2 capture process;

❑ Capture of CO_2 from coal-fired boiler flue gas using gas separation membranes may have significant technical concerns regarding membrane fouling with sub-micron fly ash particles;

Flue Gas Parameters

❑ There was a significant difference in the SO_2 levels among the fuels studied (a factor of 80 from high to low). Most other key parameters (flue gas volume, CO_2 content) did not vary significantly among the fuels evaluated;

Conventional Capture Using FGD-Amines Systems

❑ Previous studies on CO_2 recovery using conventional FGD-amine systems may be side-stepping the difficulty of meeting very low SO_2 levels (20 ppm) using conventional FGD, particularly with high sulphur fuels. For example, sulphur removal of 99.5% is needed for the 6% sulphur Pet Coke. This level of SO_2 reduction may be difficult to attain with a single stage unit;

Trace Elements

❑ Trace elements generally do not appear to be an issue in the co-capture systems reviewed as most trace elements are removed by the particulate removal system, or else via the cooling, liquid knock-out, and dehydration required for compression and transport of the CO_2. This may simply be trading an air emissions problem for a solid or liquid waste disposal problem.

❑ With the exception of the oxy-fuel case, mercury is not likely to be co-captured with CO_2.

❑ The highest level of co-capture of trace elements, including mercury, was expected to occur in the Oxy-fuel process, as this system had the fewest processing (and hence removal) steps;

Other Transportation Issues

❑ Given the potential problems of mixing CO_2 containing 3% SO_2 with CO_2 containing 3% H_2S for example, it is likely that for common pipelines, CO_2 specifications will evolve similar to natural gas distribution networks. This could limit co-capture applications to situations where capture sources are feeding dedicated storage locations only;

❑ The regulatory process itself should not be an impediment to co-capture and transport activities. We anticipate regulatory bodies may treat SO_2/CO_2 streams for example similar to transport of H_2S/CO_2 streams, for which there are well established appproval procedures and precedents, at least in North America.

9.0   RECOMMENDATIONS

Based on our conceptual study, SLI makes the following recommendations with respect to the benefits and issues of combined capture of other components of flue gas in addition to CO_2:

❑ A more detailed evaluation of co-capture of SO_2 using Cansolv® or similar systems and oxygen-enriched combustion should be performed to confirm the potential savings relative to conventional FGD-amines systems, particularly for high-sulphur coals. The impact of fuel prices to fuel sulphur content should be included in the analysis;

❑ The impact of CO_2 containing SO_2 or H_2S on EOR and other storage options such as deep saline aquifers projects needs to be studied further;

❑ The issue of whether to design oxygen-enriched combustion systems for operation on both oxygen or air needs to be studied further. In our opinion this has a significant impact on the cost of this alternative, and hence the savings relative to FGD-amine systems. Because very little detailed design work appears to have been done on Oxy-fuel systems, the apparent capture cost savings remain somewhat in question;

❑ The issue of high SO_2 removal capability upstream of conventional amine-based CO_2 recovery systems should be investigated further, particularly the challenge of meeting low SO_2 levels with standard FGD systems for high sulphur fuels;

❑ The applicability of gas separation membranes for capture of CO_2 from flue gas needs further research and development, particularly on regarding potential fouling from sub-micron particles;

❑ Development of emerging oxygen production technologies should be monitored, as lowering the cost of oxygen is the biggest challenge in oxy-fuel systems;

❑ Physical absorption systems such Rectisol® or Selexol® should be further optimized to reduce the amount of hydrogen left in the capture stream.

10.0   REFERENCES

The following sources were used for the coal analyses:

1. "Trace Elements in Coal", ACARP Report, www.acarp.com.au/Newsletters/traceelements.html.

2. Swaine, D,., "Trace Elements in Coal", Butterworths, September 1990.

3. "Orimulsion Fact Sheet" www.orimulsionfuel.com.

4. Miller, A., Dreher, K., Wentsell, R., Nadeau, R. "Environmental Impacts on the Use of Orimulsion", Phase One of the Orimulsion Technology Assessment Program, EPA Report to Congress, EPA-600/R-01-056a, July, 2001.

5. "Coal Properties", www.dti.gov.uk/energy/coal/cfft/cct/pub/pdfs/r195b.

6. "Petroleum Coke Composition", CONCAWE Product Dossier No. 93/105.

7. "Petroleum Coke Reference Materials", CANMET Mining and Mineral Science Laboratories, www.nrcan.gc.ca/mms/canmet-mtb/mmsl-lmsm/ccrmp/certificates.

8. "Fuel Sulphur Effects on Particulate Emissions from Fuel Oil Combustion Systems Under Accelerated Laboratory Conditions", www.ec.gc.ca/energ/fuels/reports/cnslt_rpts/fpq/tables1_e/htm.

9. "Coal Quality Database Version 1.3", US Geological Survey Open File Report 94-205.

10. Davidson, R., "How Coal Properties Influence Emissions", IEA Coal Research – Clean Coal Centre, Report CCC/28, February, 2000.

11. Sloss, L., Smith, I., "Trace Elements Emissions", IEA Coal Research – Clean Coal Centre, Report CCC/28, June, 2000.

The following sources were used to develop the capital and operating cost estimates:

12. "Potential for Improvement in Gasification Combined Cycle Power Generation with CO_2 Capture", IEA Greenhouse Gas R & D Programme, Report PH4/19, May, 2003.

13. "Carbon Dioxide Capture from Power Stations", www.ieagreen.org.uk/sr2p.htm.

14. Chambers, A., "Technical Advice on Air Pollution Control Technologies for Coal-fired Power Plants", Report for Alberta Environment, prepared by Alberta Research Council Inc. and Climate Change Technologies, June, 2001.

15. Bielawski, G., Rogan, J., McDonald, D., "How Low Can We Go" Controlling Emissions in New Coal-Fired Power Plants, US EPA/DOE/EPRI Combined Power Plant Air Pollutant Control Symposium, August 20-23, 2001, Chicago, IL, US.

16. Takahashi, M., "Technologies for Reducing Emissions in Coal-Fired Power Plants", The World Bank Energy Issues, No. 14, August, 1999.

17. Andersson, K., et al, "An 865 MW Lignite Fired CO_2 Free Power Plant – A Technical Feasibility Study", Paper j2-2, GHGT-6, Kyoto, Japan, 2002.

18. Birkestad, H., "Separation and Compression of CO_2 in an CO_2/O_2–fired Power Plant", Report T2002-262, Chalmers University of Technology, Goteborg, Sweden, 2002.

19. Marion, J., Nsakala, N., Griffin, T., Alain, B., "Controlling Power Plant CO_2 Emissions: A Long Range View", www.netl.doe.gov/publications/proceedings/01/carbon_seq/1b2.pdf.

20. Morimoto, S. et al, "Current Review of CO_2 Separation and Recovery Technologies", Research Institute of Innovative Technology for the Earth, GHGT-6, Kyoto, Japan, October, 2002.

21. Dave, N., Duffy, G., Edwards, J., Lowe A., "Economic Evaluation of Capture and Separation of CO_2 from Australian Black Coal-Fired Power Stations", www.energymarketreview.org/submissions/DETduffy.pdf.

22. Marion, J., et al, "Engineering Feasibility and Economics of CO_2 Sequestration on an Existing Coal-Fired Power Plant, Alstom presentation at the 26th International Conference on Coal Utilization and Fuel Systems, Clearwater, FL, US, March 5-8, 2001.

The following sources were used for the gasification work:

23. 'Potential for Improvement in Gasification Combined Cycle Power Generation with CO_2 Capture', IEA Report No. PH4/19, May 2003.

24. Brown, J.R., L. Manfredo, J. Hoffmann and M. Ramezan, 'Major Environmental Aspects of Gasification-based Power Generation Technologies' U.S. DOE/NETL Final Report, December 2002.

25. Koss, U. and M. Meyer, 'Zero Emission IGCC with Rectisol Technology', presented at the Gasification Technology Conference, San Francisco, October 2002.

26. Weiss, H. 'Rectisol wash for purification of partial oxidation gases', Gas Separation and Purification, Vol. 2, December 1988.

27. Lagas, J. 'Desulphurization of gases: a Selection of Sweetening Processes', presented at 1999 Gas Processing Symposium, Dubai, April 1999.

28. Rutkowski, M., M. Klett, R. Maxwell, 'The cost of mercury removal from coal-based IGCC relative to a PC plant', presented at Gasification Technologies Conf., San Francisco, Oct. 2002.

Vikor used the following reference to analyze the impact on miscibility:

29. Alston, R.B., Kokolis, G.P., and James, C.F.; "CO_2 Minimum Miscibility Pressure: A Correlation for Impure CO_2 Streams and Live Oil Systems"; SPE; April 1985.