3. Capture

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3.8 Gaps in knowledge

Executive Summary

The purpose of CO_2 capture is to produce a concentrated stream that can be readily transported to a CO_2 storage site. CO_2 capture and storage is most applicable to large, centralized sources like power plants and large industries. Capture technologies also open the way for large-scale production of low-carbon or carbon-free electricity and fuels for transportation, as well as for small-scale or distributed applications. The energy required to operate CO_2 capture systems reduces the overall efficiency of power generation or other processes, leading to increased fuel requirements, solid wastes and environmental impacts relative to the same type of base plant without capture. However, as more efficient plants with capture become available and replace many of the older less efficient plants now in service, the net impacts will be compatible with clean air emission goals for fossil fuel use. Minimization of energy requirements for capture, together with improvements in the efficiency of energy conversion processes will continue to be high priorities for future technology development in order to minimize overall environmental impacts and cost.

At present, CO_2 is routinely separated at some large industrial plants such as natural gas processing and ammonia production facilities, although these plants remove CO_2 to meet process demands and not for storage. CO_2 capture also has been applied to several small power plants. However, there have been no applications at large-scale power plants of several hundred megawatts, the major source of current and projected CO_2 emissions. There are three main approaches to CO_2 capture, for industrial and power plant applications. *Post-combustion* systems separate CO_2 from the flue gases produced by combustion of a primary fuel (coal, natural gas, oil or biomass) in air. *Oxy-fuel combustion* uses oxygen instead of air for combustion, producing a flue gas that is mainly H_2O and CO_2 and which is readily captured. This is an option still under development. *Pre-combustion* systems process the primary fuel in a reactor to produce separate streams of CO_2 for storage and H_2 which is used as a fuel. Other industrial processes, including processes for the production of low-carbon or carbon-free fuels, employ one or more of these same basic capture methods. The monitoring, risk and legal aspects associated with CO_2 capture systems appear to present no new challenges, as they are all elements of long-standing health, safety and environmental control practice in industry.

For all of the aforementioned applications, we reviewed recent studies of the performance and cost of commercial or near-commercial technologies, as well as that of newer CO_2 capture concepts that are the subject of intense R&D efforts worldwide. For power plants, current commercial CO_2 capture systems can reduce CO_2 emissions by 80−90% kWh^{-1} (85−95% capture efficiency). Across all plant types the cost of electricity production (COE) increases by 12−36 US$ MWh^{-1} (US$ 0.012−0.036 kWh^{-1}) over a similar type of plant without capture, corresponding to a 40−85% increase for a supercritical pulverized coal (PC) plant, 35−70% for a natural gas combined cycle (NGCC) plant and 20−55% for an integrated gasification combined cycle (IGCC) plant using bituminous coal. Overall the COE for fossil fuel plants with capture, ranges from 43−86 US$ MWh^{-1}, with the cost per tonne of CO_2 ranging from 11−57 US$/t$CO_2$ captured or 13−74 US$/t$CO_2$ avoided (depending on plant type, size, fuel type and a host of other factors). These costs include CO_2 compression but not additional transport and storage costs. NGCC systems typically have a lower COE than new PC and IGCC plants (with or without capture) for gas prices below about 4 US$ GJ^{-1}. Most studies indicate that IGCC plants are slightly more costly without capture and slightly less costly with capture than similarly sized PC plants, but the differences in cost for plants with CO_2 capture can vary with coal type and other local factors. The lowest CO_2 capture costs (averaging about 12 US$/t CO_2 captured or 15 US$/t$CO_2$ avoided) were found for industrial processes such as hydrogen production plants that produce concentrated CO_2 streams as part of the current production process; such industrial processes may represent some of the earliest opportunities for CO_2 Capture and Storage (CCS). In all cases, CO_2 capture costs are highly dependent upon technical, economic and financial factors related to the design and operation of the production process or power system of interest, as well as the design and operation of the CO_2 capture technology employed. Thus, comparisons of alternative technologies, or the use of CCS cost estimates, require a specific context to be meaningful.

New or improved methods of CO_2 capture, combined with advanced power systems and industrial process designs, can significantly reduce CO_2 capture costs and associated energy requirements. While there is considerable uncertainty about the magnitude and timing of future cost reductions, this assessment suggests that improvements to commercial technologies can reduce CO_2 capture costs by at least 20−30% over approximately the next decade, while new technologies under development promise more substantial cost reductions. Realization of future cost reductions, however, will require deployment and adoption of commercial technologies in the marketplace as well as sustained R&D.

3.1 Introduction

3.1.1 The basis for CO_2 capture

The main application of CO_2 capture is likely to be at large point sources: fossil fuel power plants, fuel processing plants and other industrial plants, particularly for the manufacture of iron, steel, cement and bulk chemicals, as discussed in Chapter 2.

Capturing CO_2 directly from small and mobile sources in the transportation and residential & commercial building sectors is expected to be more difficult and expensive than from large point sources. Small-scale capture is therefore not further discussed in this chapter. An alternative way of avoiding emissions of CO_2 from these sources would be by use of energy carriers such as hydrogen or electricity produced in large fossil fuel-based plants with CO_2 capture or by using renewable energy sources. Production of hydrogen with CO_2 capture is included in this chapter.

The possibility of CO_2 capture from ambient air (Lackner, 2003) is not discussed in this chapter because the CO_2 concentration in ambient air is around 380 ppm, a factor of 100 or more lower than in flue gas. Capturing CO_2 from air by the growth of biomass and its use in industrial plants with CO_2 capture is more cost-effective based on foreseeable technologies, and is included in this chapter.

In an analysis of possible future scenarios for anthropogenic greenhouse-gas emissions it is implicit that technological innovations will be one of the key factors which determines our future path (Section 2.5.3). Therefore this chapter deals not only with application of existing technology for CO_2 capture, but describes many new processes under development which may result in lower CO_2 capture costs in future.

3.1.2 CO_2 capture systems

There are four basic systems for capturing CO_2 from use of fossil fuels and/or biomass:
- Capture from industrial process streams (described in Section 3.2);
- Post-combustion capture (described in Section 3.3);
- Oxy-fuel combustion capture (described in Section 3.4);
- Pre-combustion capture (described in Section 3.5).

These systems are shown in simplified form in Figure 3.1.

Figure 3.1. CO_2 capture systems (adapted from BP).

3.1.2.1 Capture from industrial process streams

CO_2 has been captured from industrial process streams for 80 years (Kohl and Nielsen, 1997), although most of the CO_2 that is captured is vented to the atmosphere because there is no incentive or requirement to store it. Current examples of CO_2 capture from process streams are purification of natural gas and production of hydrogen-containing synthesis gas for the manufacture of ammonia, alcohols and synthetic liquid fuels. Most of the techniques employed for CO_2 capture in the examples mentioned are also similar to those used in pre-combustion capture. Other industrial process streams which are a source of CO_2 that is not captured include cement and steel production, and fermentation processes for food and drink production. CO_2 could be captured from these streams using techniques that are common to post-combustion capture, oxy-fuel combustion capture and pre-combustion capture (see below and Section 3.2).

3.1.2.2 Post-combustion capture

Capture of CO_2 from flue gases produced by combustion of fossil fuels and biomass in air is referred to as post-combustion capture. Instead of being discharged directly to the atmosphere, flue gas is passed through equipment which separates most of the CO_2. The CO_2 is fed to a storage reservoir and the remaining flue gas is discharged to the atmosphere. A chemical sorbent process as described in Section 3.1.3.1 would normally be used for CO_2 separation. Other techniques are also being considered but these are not at such an advanced stage of development.

Besides industrial applications, the main systems of reference for post-combustion capture are the current installed capacity of 2261 GWe of oil, coal and natural gas power plants (IEA WEO, 2004) and in particular, 155 GWe of supercritical pulverized coal fired plants (IEA CCC, 2005) and 339 GWe of natural gas combined cycle (NGCC) plants, both representing the types of high efficiency power plant technology where CO_2 capture can be best applied (see Sections 3.3 and 3.7).

3.1.2.3 Oxy-fuel combustion capture

In oxy-fuel combustion, nearly pure oxygen is used for combustion instead of air, resulting in a flue gas that is mainly CO_2 and H_2O. If fuel is burnt in pure oxygen, the flame temperature is excessively high, but CO_2 and/or H_2O-rich flue gas can be recycled to the combustor to moderate this. Oxygen is usually produced by low temperature (cryogenic) air separation and novel techniques to supply oxygen to the fuel, such as membranes and chemical looping cycles are being developed. The power plant systems of reference for oxy-fuel combustion capture systems are the same as those noted above for post-combustion capture systems.

3.1.2.4 Pre-combustion capture

Pre-combustion capture involves reacting a fuel with oxygen or air and/or steam to give mainly a ‘synthesis gas (syngas)’ or ‘fuel gas’ composed of carbon monoxide and hydrogen. The carbon monoxide is reacted with steam in a catalytic reactor, called a shift converter, to give CO_2 and more hydrogen. CO_2 is then separated, usually by a physical or chemical absorption process, resulting in a hydrogen-rich fuel which can be used in many applications, such as boilers, furnaces, gas turbines, engines and fuel cells. These systems are considered to be strategically important (see Section 3.5) but the power plant systems of reference today are 4 GWe of both oil and coal-based, integrated gasification combined cycles (IGCC) which are around 0.1% of total installed capacity worldwide (3719 GWe; IEA WEO, 2004). Other reference systems for the application of pre-combustion capture include substantially more capacity than that identified above for IGCC in existing natural gas, oil and coal-based syngas/hydrogen production facilities and other types of industrial systems described in more detail in Sections 3.2 and 3.5.

3.1.3 Types of CO_2 capture technologies

CO_2 capture systems use many of the known technologies for gas separation which are integrated into the basic systems for CO_2 capture identified in the last section. A summary of these separation methods is given below while further details are available in standard textbooks.

3.1.3.1 Separation with sorbents

The separation is achieved by passing the CO_2-containing gas in intimate contact with a liquid absorbent or solid sorbent that is capable of capturing the CO_2. In the general scheme of Figure 3.2a, the sorbent loaded with the captured CO_2 is transported to a different vessel, where it releases the CO_2 (regeneration) after being heated, after a pressure decrease or after any other change in the conditions around the sorbent. The sorbent resulting after the regeneration step is sent back to capture more CO_2 in a cyclic process. In some variants of this scheme the sorbent is a solid and does not circulate between vessels because the sorption and regeneration are achieved by cyclic changes (in pressure or temperature) in the vessel where the sorbent is contained. A make-up flow of fresh sorbent is always required to compensate for the natural decay of activity and/or sorbent losses. In some situations, the sorbent may be a solid oxide which reacts in a vessel with fossil fuel or biomass producing heat and mainly CO_2 (see Section 3.4.6). The spent sorbent is then circulated to a second vessel where it is re-oxidized in air for reuse with some loss and make up of fresh sorbent.

Figure 3.2. General schemes of the main separation processes relevant for CO_2 capture. The gas removed in the separation may be CO_2, H_2 or O_2. In Figures 3.2b and 3.2c one of the separated gas streams (A and B) is a concentrated stream of CO_2, H_2 or O_2 and the other is a gas stream with all the remaining gases in the original gas (A+B).

The general scheme of Figure 3.2 governs many important CO_2 capture systems, including leading commercial options like chemical absorption and physical absorption and adsorption. Other emerging processes based on new liquid sorbents, or new solid regenerable sorbents are being developed with the aim of overcoming the limitations of the existing systems. One common problem of these CO_2 capture systems is that the flow of sorbent between the vessels of Figure 3.2a is large because it has to match the huge flow of CO_2 being processed in the power plant. Therefore, equipment sizes and the energy required for sorbent regeneration are large and tend to translate into an important efficiency penalty and added cost. Also, in systems using expensive sorbent materials there is always a danger of escalating cost related to the purchase of the sorbent and the disposal of sorbent residues. Good sorbent performance under high CO_2 loading in many repetitive cycles is obviously a necessary condition in these CO_2 capture systems.

3.1.3.2 Separation with membranes

Membranes are specially manufactured materials that allow the selective permeation of a gas through them. The selectivity of the membrane to different gases is intimately related to the nature of the material, but the flow of gas through the membrane is usually driven by the pressure difference across the membrane. Therefore, high-pressure streams are usually preferred for membrane separation. There are many different types of membrane materials (polymeric, metallic, ceramic) that may find application in CO_2 capture systems to preferentially separate H_2 from a fuel gas stream, CO_2 from a range of process streams or O_2 from air with the separated O_2 subsequently aiding the production of a highly concentrated CO_2 stream. Although membrane separation finds many current commercial applications in industry (some of a large scale, like CO_2 separation from natural gas) they have not yet been applied for the large scale and demanding conditions in terms of reliability and low-cost required for CO_2 capture systems. A large worldwide R&D effort is in progress aimed at the manufacture of more suitable membrane materials for CO_2 capture in large-scale applications.

3.1.3.3 Distillation of a liquefied gas stream and refrigerated separation

A gas can be made liquid by a series of compression, cooling and expansion steps. Once in liquid form, the components of the gas can be separated in a distillation column. In the case of air, this operation is currently carried out commercially on a large scale. Oxygen can be separated from air following the scheme of Figure 3.2c and be used in a range of CO_2 capture systems (oxy-fuel combustion and pre-combustion capture). As in the previous paragraphs, the key issue for these systems is the large flow of oxygen required. Refrigerated separation can also be used to separate CO_2 from other gases. It can be used to separate impurities from relatively high purity CO_2 streams, for example, from oxy-fuel combustion and for CO_2 removal from natural gas or synthesis gas that has undergone a shift conversion of CO to CO_2.

3.1.4 Application of CO_2 capture

The CO_2 capture systems shown in Figure 3.1 can be cross-referenced with the different separation technologies of Figure 3.2, resulting in a capture toolbox. Table 3.1 gives an overview of both current and emerging technologies in this toolbox. In the next sections of this chapter a more detailed description of all these technological options will be given, with more emphasis on the most developed technologies for which the CO_2 capture cost can be estimated most reliably. These leading commercial options are shown in bold in Table 3.1. An overview of the diverse range of emerging options being investigated worldwide for CO_2 capture applications will also be provided. All of these options are aimed at more efficient and lower cost CO_2-capture systems (compared with the leading options). It is important to understand that this wide variety of approaches for CO_2 capture will tend to settle with time as the expected benefits (and potential weaknesses) in the technological portfolio of Table 3.1 becomes obvious with new results from current and future research and demonstration projects. Only a few of these options will prove truly cost-effective in the medium to long term.

Table 3.1. Capture toolbox.

CO_2 capture may be installed in new energy utilization plants or it may be retrofitted to existing plants. In principle, if CO_2 capture is to be introduced rapidly, it may have to be retrofitted to some existing plants or these plants would have to be retired prematurely and replaced by new plants with capture. Disadvantages of retrofits are:

- There may be site constraints such as availability of land for the capture equipment;
- A long remaining plant life may be needed to justify the large expense of installing capture equipment;
- Old plants tend to have low energy efficiencies. Including CO_2 capture will have a proportionally greater impact on the net output than in high efficiency plants.

To minimize the site constraints, new energy utilization plants could be built 'capture-ready', that is with the process design initially factoring in the changes necessary to add capture and with sufficient space and facilities made available for simple installation of CO_2 capture at a later date. For some types of capture retrofit, for example pre-combustion capture and oxy-fuel combustion, much of the retrofit equipment could be built on a separate site if necessary.

The other barriers could be largely overcome by upgrading or substantially rebuilding the existing plant when capture is retrofitted. For example, old inefficient boilers and steam turbines could be replaced by modern, high-efficiency supercritical boilers and turbines or IGCC plants. As the efficiencies of power generation technologies are increasing, the efficiency of the retrofitted plant with CO_2 capture could be as high as that of the original plant without capture.

3.2 Industrial process capture systems

3.2.1 Introduction

There are several industrial applications involving process streams where the opportunity exists to capture CO_2 in large quantities and at costs lower than from the systems described in the rest of this chapter. Capture from these sources will not be the complete answer to the needs of climate change, since the volumes of combustion-generated CO_2 are much higher, but it may well be the place where the first capture and storage occurs.

3.2.2 Natural gas sweetening

Natural gas contains different concentration levels of CO_2, depending on its source, which must be removed. Often pipeline specifications require that the CO_2 concentration be lowered to around 2% by volume (although this amount varies in different places) to prevent pipeline corrosion, to avoid excess energy for transport and to increase the heating value of the gas. Whilst accurate figures are published for annual worldwide natural gas production (BP, 2004), none seem to be published on how much of that gas may contain CO_2. Nevertheless, a reasonable assumption is that about half of raw natural gas production contains CO_2 at concentrations averaging at least 4% by volume. These figures can be used to illustrate the scale of this CO_2 capture and storage opportunity. If half of the worldwide production of 2618.5 billion m^3 of natural gas in 2003 is reduced in CO_2 content from 4 to 2% mol, the resultant amount of CO_2 removed would be at least 50 Mt $CO_2\,yr^{-1}$. It is interesting to note that there are two operating natural gas plants capturing and storing CO_2, BP's In Salah plant in Algeria and a Statoil plant at Sleipner in the North Sea. Both capture about 1 $MtCO_2\,yr^{-1}$ (see Chapter 5). About 6.5 million $tCO_2\,yr^{-1}$ from natural gas sweetening is also currently being used in enhanced oil recovery (EOR) in the United States (Beecy and Kuuskraa, 2005) where in these commercial EOR projects, a large fraction of the injected CO_2 is also retained underground (see Chapter 5).

Depending on the level of CO_2 in natural gas, different processes for natural gas sweetening (i.e., H_2S and CO_2 removal) are available (Kohl and Nielsen, 1997 and Maddox and Morgan, 1998):
- Chemical solvents
- Physical solvents
- Membranes

Natural gas sweetening using various alkanolamines (MEA, DEA, MDEA, etc.; See Table 3.2), or a mixture of them, is the most commonly used method. The process flow diagram for CO_2 recovery from natural gas is similar to what is presented for flue gas treatment (see Figure 3.4, Section 3.3.2.1), except that in natural gas processing, absorption occurs at high pressure, with subsequent expansion before the stripper column, where CO_2 will be flashed and separated. When the CO_2 concentration in natural gas is high, membrane systems may be more economical. Industrial application of membranes for recovery of CO_2 from natural gas started in the early 1980s for small units, with many design parameters unknown (Noble and Stern, 1995). It is now a well-established and competitive technology with advantages compared to other technologies, including amine treatment in certain cases (Tabe-Mohammadi, 1999). These advantages include lower capital cost, ease of skid-mounted installation, lower energy consumption, ability to be applied in remote areas, especially offshore and flexibility.

Table 3.2. Common solvents used for the removal of CO_2 from shifted syngas in pre-combustion capture processes.

Figure 3.4. Process flow diagram for CO_2 recovery from flue gas by chemical absorption.

3.2.3 Steel production

The iron and steel industry is the largest energy-consuming manufacturing sector in the world, accounting for 10−15% of total industrial energy consumption (IEA GHG, 2000a). Associated CO_2 emissions were estimated at 1442 $MtCO_2$ in 1995. Two types of iron- and steel-making technologies are in operation today. The integrated steel plant has a typical capacity of 3−5 Mtonnes yr^{-1} of steel and uses coal as its basic fuel with, in many cases, additional natural gas and oil. The mini-mill uses electric arc furnaces to melt scrap with a typical output of 1 Mtonnes yr^{-1} of steel and an electrical consumption of 300−350 kWh $tonne^{-1}$ steel. Increasingly mini-mills blend direct-reduced iron (DRI) with scrap to increase steel quality. The production of direct-reduced iron involves reaction of high oxygen content iron ore with H_2 and CO to form reduced iron plus H_2O and CO_2. As a result, many of the direct reduction iron processes could capture a pure CO_2 stream.

An important and growing trend is the use of new iron-making processes, which can use lower grade coal than the coking coals required for blast furnace operation. A good example is the COREX process (von Bogdandy, 1989), which produces a large additional quantity of N_2-free fuel gas which can be used in a secondary operation to convert iron ore to iron. Complete CO_2 capture from this process should be possible with this arrangement since the CO_2 and H_2O present in the COREX top gas must be removed to allow the CO plus H_2 to be heated and used to reduce iron oxide to iron in the secondary shaft kiln. This process will produce a combination of molten iron and iron with high recovery of CO_2 derived from the coal feed to the COREX process.

Early opportunities exist for the capture of CO_2 emissions from the iron and steel industry, such as:

- CO_2 recovery from blast furnace gas and recycle of CO-rich top gas to the furnace. A minimum quantity of coke is still required and the blast furnace is fed with a mixture of pure O_2 and recycled top gas. The furnace is, in effect, converted from air firing to oxy-fuel firing with CO_2 capture (see Section 3.4). This would recover 70% of the CO_2 currently emitted from an integrated steel plant (Dongke *et al.*, 1988). It would be feasible to retrofit existing blast furnaces with this process.
- Direct reduction of iron ore, using hydrogen derived from a fossil fuel in a pre-combustion capture step (see Section 3.5) (Duarte and Reich, 1998). Instead of the fuel being burnt in the furnace and releasing its CO_2 to atmosphere, the fuel would be converted to hydrogen and the CO_2 would be captured during that process. The hydrogen would then be used as a reduction agent for the iron ore. Capture rates should be 90−95% according to the design of the pre-combustion capture technique (see Section 3.5).

Other novel process routes for steel making to which CO_2 capture can be applied are currently in the research and development phase (Gielen, 2003; IEA, 2004)

3.2.4 Cement production

Emissions of CO_2 from the cement industry account for 6% of the total emissions of CO_2 from stationary sources (see Chapter 2). Cement production requires large quantities of fuel to drive the high temperature, energy-intensive reactions associated with the calcination of the limestone – that is calcium carbonate being converted to calcium oxide with the evolution of CO_2.

At present, CO_2 is not captured from cement plants, but possibilities do exist. The concentration of CO_2 in the flue gases is between 15−30% by volume, which is higher than in flue gases from power and heat production (3−15% by volume). So, in principle, the post-combustion technologies for CO_2 capture described in Section 3.3 could be applied to cement production plants, but would require the additional generation of steam in a cement plant to regenerate the solvent used to capture CO_2. Oxy-fuel combustion capture systems may also become a promising technique to recover CO_2 (IEA GHG, 1999). Another emerging option would be the use of calcium sorbents for CO_2 capture (see Sections 3.3.3.4 and 3.5.3.5) as calcium carbonate (limestone) is a raw material already used in cement plants. All of these capture techniques could be applied to retrofit, or new plant applications.

3.2.5 Ammonia production

CO_2 is a byproduct of ammonia (NH_3) production (Leites, Sama and Lior, 2003). Two main groups of processes are used:

- Steam reforming of light hydrocarbons (natural gas, liquefied petroleum gas, naphtha)
- Partial oxidation or gasification of heavy hydrocarbons (coal, heavy fuel oil, vacuum residue).

Around 85% of ammonia is made by processes in the steam methane reforming group and so a description of the process is useful. Although the processes vary in detail, they all comprise the following steps:

1. Purification of the feed;
2. Primary steam methane reforming (see Section 3.5.2.1);
3. Secondary reforming, with the addition of air, commonly called auto thermal reforming (see Section 3.5.2.3);
4. Shift conversion of CO and H_2O to CO_2 and H_2;
5. Removal of CO_2;
6. Methanation (a process that reacts and removes trace CO and CO_2);
7. Ammonia synthesis.

The removal of CO_2 as a pure stream is of interest to this report. A typical modern plant will use the amine solvent process to treat 200,000 $Nm^3 h^{-1}$ of gas from the reformer, to produce 72 tonnes h^{-1} of concentrated CO_2 (Apple, 1997). The amount of CO_2 produced in modern plants from natural gas is about 1.27 tCO_2/tNH_3. Hence, with a world ammonia production of about 100 Mtonnes yr^{-1}, about 127 $MtCO_2 yr^{-1}$ is produced. However, it should be noted that this is not all available for storage, as ammonia plants are frequently combined with urea plants, which are capable of utilizing 70−90% of the CO_2. About 0.7 $MtCO_2 yr^{-1}$captured from ammonia plants is currently used for enhanced oil recovery in the United States (Beecy and Kuuskraa, 2005) with a large fraction of the injected CO_2 being retained underground (see Chapter 5) in these commercial EOR projects.

3.2.6 Status and outlook

We have reviewed processes – current and potential − that may be used to separate CO_2 in the course of producing another product. One of these processes, natural gas sweetening, is already being used in two industrial plants to capture and store about 2 $MtCO_2 yr^{-1}$ for the purpose of climate change mitigation. In the case of ammonia production, pure CO_2 is already being separated. Over 7 $MtCO_2 yr^{-1}$ captured from both natural gas sweetening and ammonia plants is currently being used in enhanced oil recovery with some storage (see also Chapter 5) of the injected CO_2 in these commercial EOR projects. Several potential processes for CO_2 capture in steel and cement production exist, but none have yet been applied. Although the total amount of CO_2 that may be captured from these industrial processes is insignificant in terms of the scale of the climate change challenge, significance may arise in that their use could serve as early examples of solutions that can be applied on larger scale elsewhere.

3.3 Post-combustion capture systems

3.3.1 Introduction

Current anthropogenic CO_2 emissions from stationary sources come mostly from combustion systems such as power plants, cement kilns, furnaces in industries and iron and steel production plants (see Chapter 2). In these large-scale processes, the direct firing of fuel with air in a combustion chamber has been (for centuries, as it is today) the most economic technology to extract and use the energy contained in the fuel. Therefore, the strategic importance of post-combustion capture systems becomes evident when confronted with the reality of today's sources of CO_2 emissions. Figure 2.1 (Chapter 2) shows that any attempt to mitigate CO_2 emissions from stationary sources on a relevant scale using CO_2 capture and storage, will have to address CO_2 capture from combustion systems. All the CO_2 capture systems described in this section are aimed at the separation of CO_2 from the flue gases generated in a large-scale combustion process fired with fossil fuels. Similar capture systems can also be applied to biomass fired combustion processes that tend to be used on a much smaller scale compared to those for fossil fuels.

Flue gases or stack gases found in combustion systems are usually at atmospheric pressure. Because of the low pressure, the large presence of nitrogen from air and the large scale of the units, huge flows of gases are generated, the largest example of which may be the stack emissions coming from a natural gas combined cycle power plant having a maximum capacity of around 5 million normal $m^3 h^{-1}$. CO_2 contents of flue gases vary depending on the type of fuel used (between 3% for a natural gas combined cycle to less than 15% by volume for a coal-fired combustion plant See Table 2.1). In principle post-combustion capture systems can be applied to flue gases produced from the combustion of any type of fuel. However, the impurities in the fuel are very important for the design and costing of the complete plant (Rao and Rubin, 2002). Flue gases coming from coal combustion will contain not only CO_2, N_2, O_2 and H_2O, but also air pollutants such as SO_x, NO_x, particulates, HCl, HF, mercury, other metals and other trace organic and inorganic contaminants. Figure 3.3 shows a general schematic of a coal-fired power plant in which additional unit operations are deployed to remove the air pollutants prior to CO_2 capture in an absorption-based process. Although capture of CO_2 in these flue gases is in principle more problematic and energy intensive than from other gas streams, commercial experience is available at a sufficiently large scale (see Section 3.3.2) to provide the basis for cost estimates for post-combustion CO_2 capture systems (see Section 3.7). Also, a large R&D effort is being undertaken worldwide to develop more efficient and lower cost post-combustion systems (see Section 3.3.3), following all possible approaches for the CO_2 separation step (using sorbents, membranes or cryogenics; see Section 3.1.3).

Figure 3.3. Schematic of a pulverized coal-fired power plant with an amine-based CO_2 capture system and other emission controls.

3.3.2 Existing technologies

There are several commercially available process technologies which can in principle be used for CO_2 capture from flue gases. However, comparative assessment studies (Hendriks, 1994; Riemer and Ormerod, 1995; IEA GHG, 2000b) have shown that absorption processes based on chemical solvents are currently the preferred option for post-combustion CO_2 capture. At this point in time, they offer high capture efficiency and selectivity, and the lowest energy use and costs when compared with other existing post-combustion capture processes. Absorption processes have reached the commercial stage of operation for post-combustion CO_2 capture systems, albeit not on the scale required for power plant flue gases. Therefore, the following paragraphs are devoted to a review of existing knowledge of the technology and the key technical and environmental issues relevant to the application of this currently leading commercial option for CO_2 capture. The fundamentals of the CO_2 separation step using commercial chemical absorption processes are discussed first. The requirements of flue gas pretreatment (removal of pollutants other than CO_2) and the energy requirements for regeneration of the chemical solvent follow.

3.3.2.1 Absorption processes

Absorption processes in post-combustion capture make use of the reversible nature of the chemical reaction of an aqueous alkaline solvent, usually an amine, with an acid or sour gas. The process flow diagram of a commercial absorption system is presented in Figure 3.4. After cooling the flue gas, it is brought into contact with the solvent in the absorber. A blower is required to overcome the pressure drop through the absorber. At absorber temperatures typically between 40°C and 60°C, CO_2 is bound by the chemical solvent in the absorber. The flue gas then undergoes a water wash section to balance water in the system and to remove any solvent droplets or solvent vapour carried over, and then it leaves the absorber. It is possible to reduce CO_2 concentration in the exit gas down to very low values, as a result of the chemical reaction in the solvent, but lower exit concentrations tend to increase the height of the absorption vessel. The 'rich' solvent, which contains the chemically bound CO_2 is then pumped to the top of a stripper (or regeneration vessel), via a heat exchanger. The regeneration of the chemical solvent is carried out in the stripper at elevated temperatures (100°C–140°C) and pressures not very much higher than atmospheric pressure. Heat is supplied to the reboiler to maintain the regeneration conditions. This leads to a thermal energy penalty as a result of heating up the solvent, providing the required desorption heat for removing the chemically bound CO_2 and for steam production which acts as a stripping gas. Steam is recovered in the condenser and fed back to the stripper, whereas the CO_2 product gas leaves the stripper. The 'lean' solvent, containing far less CO_2 is then pumped back to the absorber via the lean-rich heat exchanger and a cooler to bring it down to the absorber temperature level.

Figure 3.4 also shows some additional equipment needed to maintain the solution quality as a result of the formation of degradation products, corrosion products and the presence of particles. This is generally done using filters, carbon beds and a thermally operated reclaimer. Control of degradation and corrosion has in fact been an important aspect in the development of absorption processes over the past few decades.

The key parameters determining the technical and economic operation of a CO_2 absorption system are:

- **Flue gas flow rate** − The flue gas flow rate will determine the size of the absorber and the absorber represents a sizeable contribution to the overall cost.
- **CO_2 content in flue gas** − Since flue gas is usually at atmospheric pressure, the partial pressure of CO_2 will be as low as 3−15 kPa. Under these low CO_2 partial pressure conditions, aqueous amines (chemical solvents) are the most suitable absorption solvents (Kohl and Nielsen, 1997).

- **CO_2 removal** − In practice, typical CO_2 recoveries are between 80% and 95%. The exact recovery choice is an economic trade-off, a higher recovery will lead to a taller absorption column, higher energy penalties and hence increased costs.
- **Solvent flow rate** − The solvent flow rate will determine the size of most equipment apart from the absorber. For a given solvent, the flow rate will be fixed by the previous parameters and also the chosen CO_2 concentrations within the lean and the rich solutions.
- **Energy requirement** − The energy consumption of the process is the sum of the thermal energy needed to regenerate the solvents and the electrical energy required to operate liquid pumps and the flue gas blower or fan. Energy is also required to compress the CO_2 recovered to the final pressure required for transport and storage.
- **Cooling requirement** − Cooling is needed to bring the flue gas and solvent temperatures down to temperature levels required for efficient absorption of CO_2. Also, the product from the stripper will require cooling to recover steam from the stripping process.

The purity and pressure of CO_2 typically recovered from an amine-based chemical absorption process are as follows (Sander and Mariz, 1992):

CO_2 purity	99.9% by volume or more (water saturated conditions)
CO_2 pressure	50 kPa (gauge)

A further CO_2 purification step makes it possible to bring the CO_2-quality up to food-grade standard. This is required for use in beverages and packaging.

Since combustion flue gases are generally at atmospheric pressure and the CO_2 is diluted, the CO_2 partial pressure is very low. Also, flue gas contains oxygen and other impurities; therefore an important characteristic of an absorption process is in the proper choice of solvent for the given process duty. High CO_2 loading and low heat of desorption energy are essential for atmospheric flue gas CO_2 recovery. The solvents must also have low byproduct formation and low decomposition rates, to maintain solvent performance and to limit the amount of waste materials produced. The important effect of other contaminants on the solvent is discussed in Section 3.3.2.2.

The following three absorption processes are commercially available for CO_2 capture in post-combustion systems:

- The Kerr-McGee/ABB Lummus Crest Process (Barchas and Davis, 1992) − This process recovers CO_2 from coke and coal-fired boilers, delivering CO_2 for soda ash and liquid CO_2 preparations. It uses a 15−20% by weight aqueous MEA (Mono-Ethanolamine) solution. The largest capacity experienced for this process is 800 $tCO_2\ d^{-1}$utilizing two parallel trains (Arnold *et al.*, 1982).
- The Fluor Daniel ® ECONAMINE ™ Process (Sander and Mariz, 1992, Chapel *et al.*, 1999) − This process was acquired by Fluor Daniel Inc. from Dow Chemical Company in 1989. It is a MEA-based process (30% by weight aqueous solution) with an inhibitor to resist carbon steel corrosion and is specifically tailored for oxygen-containing gas streams. It has been used in many plants worldwide recovering up to 320 $tCO_2\ d^{-1}$ in a single train for use in beverage and urea production.
- The Kansai Electric Power Co., Mitsubishi Heavy Industries, Ltd., KEPCO/MHI Process (Mimura *et al.*, 1999 and 2003) − The process is based upon sterically-hindered amines and already three solvents (KS-1, KS-2 and KS-3) have been developed. KS-1 was commercialized in a urea production application. In this process, low amine losses and low solvent degradation have been noted without the use of inhibitors or additives. As shown in Figure 3.5, the first commercial plant at 200 $tCO_2\ d^{-1}$recovery from a flue gas stream has been operating in Malaysia

since 1999 for urea production (equivalent to the emissions from a 10 MWt coal-fired power plant).

Figure 3.5. CO_2 capture plant in Malaysia using a 200 tonne d^{-1} KEPCO/MHI chemical solvent process (Courtesy of Mitsubishi).

The performance of the chemical solvent in the operation is maintained by replacement, filtering and reclaiming, which leads to a consumables requirement. Typical values for the solvent consumption are between 0.2 and 1.6 kg/tCO_2. In addition, chemicals are needed to reclaim the amine from the heat stable salt (typically 0.03–0.13 kg NaOH/tCO_2) and to remove decomposition products (typically 0.03−0.06 kg activated carbon/tCO_2). The ranges are primarily dependent on the absorption process, with KS-1 being at the low end of the range and ECONAMINE ™ at the high end.

3.3.2.2 Flue gas pretreatment

Flue gases from a combustion power plant are usually above 100°C, which means that they need to be cooled down to the temperature levels required for the absorption process. This can be done in a cooler with direct water contact, which also acts as a flue gas wash with additional removal of fine particulates.

In addition to the above, flue gas from coal combustion will contain other acid gas components such as NO_x and SO_x. Flue gases from natural gas combustion will normally only contain NO_x. These acidic gas components will, similar to CO_2, have a chemical interaction with the alkaline solvent. This is not desirable as the irreversible nature of this interaction leads to the formation of heat stable salts and hence a loss in absorption capacity of the solvent and the risk of formation of solids in the solution. It also results in an extra consumption of chemicals to regenerate the solvent and the production of a waste stream such as sodium sulphate or sodium nitrate. Therefore, the pre-removal of NO_x and SO_x to very low values before CO_2 recovery becomes essential. For NO_x it is the NO_2 which leads to the formation of heat stable salts. Fortunately, the level of NO_2 is mostly less than 10% of the overall NO_x content in a flue gas (Chapel *et al.*, 1999).

The allowable SO_x content in the flue gas is primarily determined by the cost of the solvent − as this is consumed by reaction with SO_x. SO_2 concentrations in the flue gas are typically around 300−5000 ppm. Commercially available SO_2-removal plants will remove up to 98−99%. Amines are relatively cheap chemicals, but even cheap solvents like MEA (with a price around 1.25 US\$ kg^{-1} (Rao and Rubin, 2002) may require SO_x concentrations of around 10 ppm, to keep solvent consumption (around 1.6 kg of MEA/tCO_2 separated) and make up costs at reasonable values, which often means that additional flue gas desulphurization is needed. The optimal SO_2 content, before the CO_2 absorption process is a cost trade-off between CO_2-solvent consumption and SO_2-removal costs. For the Kerr-Mcgee/ABB Lummus Crest Technology, SO_2-removal is typically not justified for SO_2 levels below 50 ppm (Barchas and Davis, 1992). For the Fluor Daniel Econamine FG process a maximum of 10 ppm SO_2 content is generally set as the feed gas specification (Sander and Mariz, 1992). This can be met by using alkaline salt solutions in a spray scrubber (Chapel *et al.*, 1999). A SO_2 scrubber might also double as a direct contact cooler to cool down the flue gas.

Careful attention must also be paid to fly ash and soot present in the flue gas, as they might plug the absorber if contaminants levels are too high. Often the requirements of other flue gas treatment are such that precautions have already been taken. In the case of CO_2 recovery from a coal-fired boiler flue gas, the plant typically has to be equipped with a $DeNO_x$ unit, an electrostatic precipitator or a bag house filter and a $DeSO_x$ or flue gas desulphurization unit as part of the environmental protection of the power plant facilities. In some cases, these environmental protection facilities are not enough to carry out deep SO_x removal up to the 1−2 ppm level sometimes needed to minimize solvent consumption and its reclamation from sticking of solvent wastes on reclaimer tube surfaces.

3.3.2.3 Power generation efficiency penalty in CO_2 capture

A key feature of post-combustion CO_2 capture processes based on absorption is the high energy requirement and the resulting efficiency penalty on power cycles. This is primarily due to the heat necessary to regenerate the solvent, steam use for stripping and to a lesser extent the electricity required for liquid pumping, the flue gas fan and finally compression of the CO_2 product. Later in this chapter, Sections 3.6 and 3.7 present summaries of CO_2 capture energy requirements for a variety of power systems and discuss the environmental and economic implications of these energy demands.

In principle, the thermal energy for the regeneration process can be supplied by an auxiliary boiler in a retrofit situation. Most studies, however, focus on an overall process in which the absorption process is integrated into the power plant. The heat requirement is at such levels that low-pressure steam, for example condensing at 0.3 MPa(g), can be used in the reboiler. The steam required for the regeneration process is then extracted from the steam cycle in the power plant. For a coal-fired power station, low-pressure steam will be extracted prior to the last expansion stage of the steam turbine. For a natural gas fired combined cycle, low-pressure steam will be extracted from the last stage in the heat recovery steam generator. Some of this heat can be recovered by preheating the boiler feed water (Hendriks, 1994). Values for the heat requirement for the leading absorption technologies are between 2.7 and 3.3 GJ/tCO_2, depending on the solvent process. Typical values for the electricity requirement are between 0.06 and 0.11 GJ/tCO_2 for post-combustion capture in coal-fired power plants and 0.21 and 0.33 GJ/tCO_2 for post-combustion capture in natural gas fired combined cycles. Compression of the CO_2 to 110 bar will require around 0.4 GJ/tCO_2 (IEA GHG, 2004).

Integration of the absorption process with an existing power plant will require modifications of the low-pressure part of the steam cycle, as a sizeable fraction of the steam will be extracted and hence will not be available to produce power (Nsakala *et al.*, 2001, Mimura *et al.*,1995, Mimura *et al.*, 1997). To limit the required modifications, small back-pressure steam turbines using medium pressure steam to drive the flue gas fan and boiler feed water pumps can be used. The steam is then condensed in the reboiler (Mimura *et al.*, 1999). Furthermore, in power plants based on steam cycles more than 50% thermal energy in the steam cycle is disposed off in the steam condenser. If the steam cycle system and CO_2 recovery can be integrated, part of the waste heat disposed by the steam condenser can be utilized for regeneration of the chemical solvent.

The reduction of the energy penalty is, nevertheless, closely linked to the chosen solvent system. The IEA Greenhouse Programme (IEA GHG) has carried out performance assessments of power plants with post-combustion capture of CO_2, taking into consideration the most recent improvements in post-combustion CO_2 capture processes identified by technology licensors (IEA GHG, 2004). In this study, Mitsui Babcock Energy Ltd. and Alstom provided information on the use of a high efficiency, ultra-supercritical steam cycle (29 MPa, 600°C, 620°C reheat) boiler and steam turbine for a coal-fired power plant, while for the NGCC case, a combined cycle using a GE 9FA gas turbine was adopted. Fluor provided information on the Fluor Econamine + process based on MEA, and MHI provided information on KEPCO/MHI process based on the KS-1 solvent for CO_2 capture. CO_2 leaving these systems were compressed to a pressure of 11 MPa. The overall net power plant efficiencies with and without CO_2 capture are shown in Figure 3.6, while Figure 3.7 shows the efficiency penalty for CO_2 capture. Overall, results from this study show that the efficiency penalty for post-combustion capture in coal and gas fired plant is lower for

KEPCO/MHI's CO_2 absorption process. For the purpose of comparison, the performance of power plants with pre-combustion and oxy-fuel capture, based on the same standard set of plant design criteria are also shown in Figures 3.6 and 3.7.

Figure 3.6. Thermal efficiencies of power plants with and without CO_2 capture, % LHV-basis (Source data: Davison 2005, IEA GHG 2004, IEA GHG 2003; IEA GHG, 2000b; Dillon *et al.*, 2005).

Figure 3.7. Percentage increase in fuel use per kWh of electricity due to CO_2 capture, compared to the same plant without capture (Source data: Davison, 2005; IEA GHG, 2004; IEA GHG, 2003; IEA GHG, 2000b; Dillon *et al.*, 2005).

3.3.2.4 Effluents

As a result of decomposition of amines, effluents will be created, particularly ammonia and heat-stable salts. Rao and Rubin (2002) have estimated these emissions for an MEA-based process based on limited data. In such processes, heat stable salts (solvent decomposition products, corrosion products etc.) are removed from the solution in a reclaimer and a waste stream is created and is disposed of using normal HSE (Health, Safety and Environmental) practices. In some cases, these reclaimer bottoms may be classified as a hazardous waste, requiring special handling (Rao and Rubin, 2002). Also a particle filter and carbon filter is normally installed in the solvent circuit to remove byproducts. Finally, some solvent material will be lost to the environment through evaporation and carry over in the absorber, which is accounted for in the solvent consumption. It is expected that acid gases other than CO_2, which are still present in the flue gas (SO_x and NO_2) will also be absorbed in the solution. This will lower the concentration of these components further and even the net emissions in some cases depending on the amount of additional energy use for CO_2 capture (see Tables 3.4 and 3.5). As SO_2-removal prior to CO_2-removal is very likely in coal-fired plants, this will lead to the production of a waste or byproduct stream containing gypsum and water from the FGD unit.

3.3.3 Emerging technologies

3.3.3.1 Other absorption process

Various novel solvents are being investigated, with the object of achieving a reduced energy consumption for solvent regeneration (Chakma, 1995; Chakma and Tontiwachwuthikul, 1999; Mimura *et al.*, 1999; Zheng *et al.*, 2003; Cullinane and Rochelle, 2003; Leites, 1998; Erga *et al.*, 1995; Aresta and Dibenedetto, 2003; Bai and Yeh, 1997).

Besides novel solvents, novel process designs are also currently becoming available (Leites *et al.* 2003). Research is also being carried out to improve upon the existing practices and packing types (Aroonwilas *et al.*, 2003). Another area of research is to increase the concentration levels of aqueous MEA solution used in absorption systems as this tends to reduce the size of equipment used in capture plants (Aboudheir *et al.*, 2003). Methods to prevent oxidative degradation of MEA by de-oxygenation of the solvent solutions are also being investigated (Chakravarti *et al.*, 2001). In addition to this, the catalytic removal of oxygen in flue gases from coal firing has been suggested (Nsakala *et al.*, 2001) to enable operation with promising solvents sensitive to oxygen.

3.3.3.2 Adsorption process

In the adsorption process for flue gas CO_2 recovery, molecular sieves or activated carbons are used in adsorbing CO_2. Desorbing CO_2 is then done by the pressure swing operation (PSA) or temperature swing operation (TSA). Most applications are associated with pressure swing adsorption (Ishibashi *et al.*, 1999 and Yokoyama, 2003). Much less attention has been focused on CO_2 removal via temperature swing adsorption, as this technique is less attractive compared to PSA due to the longer cycle times needed to heat up the bed of solid particles during sorbent regeneration. For bulk separations at large scales, it is also essential to limit the length of the unused bed and therefore opt for faster cycle times.

Adsorption processes have been employed for CO_2 removal from synthesis gas for hydrogen production (see Section 3.5.2.9). It has not yet reached a commercial stage for CO_2 recovery from flue gases. The following main R&D activities have been conducted:

- Study of CO_2 removal from flue gas of a thermal power plant by physical adsorption (Ishibashi *et al.*, 1999);
- Study of CO_2 removal from flue gas of a thermal power plant by a combined system with pressure swing adsorption and a super cold separator (Takamura *et al.*, 1999);
- Pilot tests on the recovery of CO_2 from a coal and oil fired power plant, using pressure temperature swing adsorption (PTSA) and an X-type zeolite as an adsorbent (Yokoyama, 2003).

Pilot test results of coal-fired flue gas CO_2 recovery by adsorption processes show that the energy consumption for capture (blowers and vacuum pumps) has improved from the original 708 kWh/tCO_2 to 560 kWh/tCO_2. An energy consumption of 560 kWh/tCO_2 is equivalent to a loss corresponding to 21% of the energy output of the power plant. Recovered CO_2 purity is about 99.0% by volume using two stages of a PSA and PTSA system (Ishibashi *et al.*, 1999).

It can be concluded that based on mathematical models and data from pilot-scale experimental installations, the design of a full-scale industrial adsorption process might be feasible. A serious drawback of all adsorptive methods is the necessity to treat the gaseous feed before CO_2 separation in an adsorber. Operation at high temperature with other sorbents (see Section 3.3.3.4) can circumvent this requirement (Sircar and Golden, 2001). In many cases gases have to be also cooled and dried, which limits the attractiveness of PSA, TSA or ESA (electric swing adsorption) vis-à-vis capture by chemical absorption described in previous sections. The development of a new generation of materials that would efficiently adsorb CO_2 will undoubtedly enhance the competitiveness of adsorptive separation in a flue gas application.

3.3.3.3 Membranes

Membrane processes are used commercially for CO_2 removal from natural gas at high pressure and at high CO_2 concentration (see Section 3.2.2). In flue gases, the low CO_2 partial pressure difference provides a low driving force for gas separation. The removal of carbon dioxide using commercially available polymeric gas separation membranes results in higher energy penalties on the power generation efficiency compared to a standard chemical absorption process (Herzog *et al.*, 1991, Van der Sluijs *et al.*, 1992 and Feron, 1994). Also, the maximum percentage of CO_2 removed is lower than for a standard chemical absorption processes. Improvements can be made if more selective membranes become available, such as facilitated membranes, described below.

The membrane option currently receiving the most attention is a hybrid membrane – absorbent (or solvent) system. These systems are being developed for flue gas CO_2 recovery. Membrane/solvent systems employ membranes to provide a very high surface area to volume ratio for mass exchange between a gas stream and a solvent resulting in a very compact system. This results in a membrane contactor system in which the membrane forms a gas permeable barrier between a liquid and a gaseous phase. In general, the membrane is not involved in the separation process. In the case of porous membranes, gaseous components diffuse through the pores and are absorbed by the liquid; in cases of non-porous membranes they dissolve in the membrane and diffuse through the membrane. The contact surface area between gas and liquid phase is maintained by the membrane and is independent of the gas and liquid flow rate. The selectivity of the partition is primarily determined by the absorbent (solvent). Absorption in the liquid phase is determined either by physical partition or by a chemical reaction.

The advantages of membrane/solvent systems are avoidance of operational problems occurring in conventional solvent absorption systems (see Section 3.3.2.1) where gas and liquid flows are in direct contact. Operational problems avoided include foaming, flooding entrainment and channelling, and result in the free choice of the gas and liquid flow rates and a fixed interface for mass transfer in the membrane/solvent system. Furthermore, the use of compact membranes result in smaller equipment sizes with capital cost reductions. The choice of a suitable combination of solvent and membrane material is very important. The material characteristics should be such that the transfer of solvent through the membrane is avoided at operating pressure gradients of typically 50–100 kPa, while the transfer of gas is not hindered. The overall process configuration in terms of unit operations would be very similar to a conventional chemical absorption/desorption process (see Figure 3.4). Membrane/solvent systems can be both used in the absorption as well as in the desorption step. Feron and Jansen (2002) and Falk-Pedersen *et al.* (1999) give examples of suitable membrane/solvent systems.

Research and development efforts have also been reported in the area of facilitated transport membranes. Facilitated transport membranes rely on the formation of complexes or reversible chemical reactions of components present in a gas stream with compounds present in the membrane. These complexes or reaction products are then transported through the membrane. Although solution and diffusion still play a role in the transport mechanism, the essential element is the specific chemical interaction of a gas component with a compound in the membrane, the so-called carrier. Like other pressure driven membrane processes, the driving force for the separation comes from a difference in partial pressure of the component to be transported. An important class of facilitated transport membranes is the so-called supported liquid membrane in which the carrier is dissolved into a liquid contained in a membrane. For CO_2 separations, carbonates, amines and molten salt hydrates have been suggested as carriers (Feron, 1992). Porous membranes and ion-exchange membranes have been employed as the support. Until now, supported liquid membranes have only been studied on a laboratory scale. Practical problems associated with supported liquid membranes are membrane stability and liquid volatility. Furthermore, the selectivity for a gas decreases with increasing partial pressure on the feed side. This is a result of saturation of the carrier in the liquid. Also, as the total feed pressure is increased, the permeation of unwanted components is increased. This also results in a decrease in selectivity. Finally, selectivity is also reduced by a reduction in membrane thickness. Recent development work has focused on the following technological options that are applicable to both CO_2/N_2 and CO_2/H_2 separations:

- Amine-containing membranes (Teramoto *et al.*, 1996);
- Membranes containing potassium carbonate polymer gel membranes (Okabe *et al.*, 2003);
- Membranes containing potassium carbonate-glycerol (Chen *et al.*, 1999);
- Dendrimer-containing membranes (Kovvali and Sirkar, 2001).
- Poly-electrolyte membranes (Quinn and Laciak, 1997);

Facilitated transport membranes and other membranes can also be used in a preconcentration step prior to the liquefaction of CO_2 (Mano *et al.*, 2003).

3.3.3.4 *Solid sorbents*

There are post-combustion systems being proposed that make use of regenerable solid sorbents to remove CO_2 at relatively high temperatures. The use of high temperatures in the CO_2 separation step has the potential to reduce efficiency penalties with respect to wet-absorption methods. In principle, they all follow the scheme shown in Figure 3.2a, where the combustion flue gas is put in contact with the sorbent in a suitable reactor to allow the gas-solid reaction of CO_2 with the sorbent (usually the carbonation of a metal oxide). The solid can be easily separated from the gas stream and sent for regeneration in a different reactor. Instead of moving the solids, the reactor can also be switched between sorption and regeneration modes of operation in a batch wise, cyclic operation. One key component for the development of these systems is obviously the sorbent itself, that has to have good CO_2 absorption capacity and chemical and mechanical stability for long periods of operation in repeated cycles. In general, sorbent performance and cost are critical issues in all post-combustion systems, and more elaborate sorbent materials are usually more expensive and will have to demonstrate outstanding performance compared with existing commercial alternatives such as those described in 3.3.2.

Solid sorbents being investigated for large-scale CO_2 capture purposes are sodium and potassium oxides and carbonates (to produce bicarbonate), usually supported on a solid substrate (Hoffman *et al.*, 2002; Green *et al.*, 2002). Also, high temperature Li-based and CaO-based sorbents are suitable candidates. The use of lithium-containing compounds (lithium, lithium-zirconia and lithium-silica oxides) in a carbonation-calcination cycle, was first investigated in Japan (Nakagawa and Ohashi, 1998). The reported performance of these sorbents is very good, with very high reactivity in a wide range of temperatures below 700ºC, rapid regeneration at higher temperatures and durability in repeated capture-regeneration cycles. This is essential because lithium is an intrinsically expensive material.

The use of CaO as a regenerable CO_2 sorbent has been proposed in several processes dating back to the 19th century. The carbonation reaction of CaO to separate CO_2 from hot gases ($T > 600$ºC) is very fast and the regeneration of the sorbent by calcining the $CaCO_3$ into CaO and pure CO_2 is favoured at $T > 900$ºC (at a partial pressure of CO_2 of 0.1 MPa). The basic separation principle using this carbonation-calcination cycle was successfully tested in a pilot plant (40 tonne d^{-1}) for the development of the Acceptor Coal Gasification Process (Curran *et al.*, 1967) using two interconnected fluidized beds. The use of the above cycle for a post-combustion system was first proposed by Shimizu *et al.* (1999) and involved the regeneration of the sorbent in a fluidized bed, firing part of the fuel with O_2/CO_2 mixtures (see also Section 3.4.2). The effective capture of CO_2 by CaO has been demonstrated in a small pilot fluidized bed (Abanades *et al.*, 2004a). Other combustion cycles incorporating capture of CO_2 with CaO that might not need O_2 are being developed, including one that works at high pressures with simultaneous capture of CO_2 and SO_2 (Wang *et al.*, 2004). One weak point in all these processes is that natural sorbents (limestones and dolomites) deactivate rapidly, and a large make-up flow of sorbent (of the order of the mass flow of fuel entering the plant) is required to maintain the activity in the capture-regeneration loop (Abanades *et al.*, 2004b). Although the deactivated sorbent may find application in the cement industry and the sorbent cost is low, a range of methods to enhance the activity of Ca-based CO_2 sorbents are being pursued by several groups around the world.

3.3.4 Status and outlook

Virtually all the energy we use today from carbon-containing fuels is obtained by directly burning fuels in air. This is despite many decades of exploring promising and more efficient alternative energy conversion cycles that rely on other fuel processing steps prior to fuel combustion or avoiding direct fuel combustion (see pre-combustion capture – Section 3.5). In particular, combustion-based systems are still the competitive choice for operators aiming at large-scale production of electricity and heat from fossil fuels, even under more demanding environmental regulations, because these processes are reliable and well proven in delivering electricity and heat at prices that often set a benchmark for these services. In addition, there is a continued effort to raise the energy conversion efficiencies of these systems through advanced materials and component development. This will allow these systems to operate at higher temperature and higher efficiency.

As was noted in Section 3.1, the main systems of reference for post-combustion capture are the present installed capacity of coal and natural gas power plants, with a total of 970 GWe subcritical steam and 155 GWe of supercritical/ultra-supercritical steam-based pulverized coal fired plants, 339 GWe of natural gas combined cycle, 333 GWe natural gas steam-electric power plants and 17 GWe of coal-fired, circulating, fluidized-bed combustion (CFBC) power plants. An additional capacity of 454 GWe of oil-based power plant, with a significant proportion of these operating in an air-firing mode is also noted (IEA WEO, 2004 and IEA CCC, 2005). Current projections indicate that the generation efficiency of commercial, pulverized coal fired power plants based on ultra-supercritical steam cycles would exceed 50% lower heating value (LHV) over the next decade (IEA, 2004), which will be higher than efficiencies of between 36 and 45% reported for current subcritical and supercritical steam-based plants without capture (see Section 3.7). Similarly, natural gas fired combined cycles are expected to have efficiencies of 65% by 2020 (IEA GHG, 2002b) and up from current efficiencies between 55 and 58% (see Section 3.7). In a future carbon-constrained world, these independent and ongoing developments in power cycle efficiencies will result in lower CO_2-emissions per kWh produced and hence a lower loss in overall cycle efficiency when post-combustion capture is applied.

There are proven post-combustion CO_2 capture technologies based on absorption processes that are commercially available at present . They produce CO_2 from flue gases in coal and gas-fired installations for food/beverage applications and chemicals production in capacity ranges between 6 and 800 $tCO_2\ d^{-1}$. They require scale up to 20–50 times that of current unit capacities for deployment in large-scale power plants in the 500 MW_e capacity range (see Section 3.3.2). The inherent limitations of currently available absorption technologies when applied to post-combustion capture systems are well known and their impact on system cost can be estimated relatively accurately for a given application (see Section 3.7). Hence, with the dominant role played by air-blown energy conversion processes in the global energy infrastructure, the availability of post-combustion capture systems is important if CO_2 capture and storage becomes a viable climate change mitigation strategy.

The intense development efforts on novel solvents for improved performance and reduced energy consumption during regeneration, as well as process designs incorporating new contacting devices such as hybrid membrane-absorbent systems, solid adsorbents and high temperature regenerable sorbents, may lead to the use of more energy efficient post-combustion capture systems. However, all these novel concepts still need to prove their lower costs and reliability of operation on a commercial scale. The same considerations also apply to other advanced CO_2 capture concepts with oxy-fuel combustion or pre-combustion capture reviewed in the following sections of this chapter. It is generally not yet clear which of these emerging technologies, if any, will succeed as the dominant commercial technology for energy systems incorporating CO_2 capture.

3.4 Oxy-fuel combustion capture systems

3.4.1 Introduction

The oxy-fuel combustion process eliminates nitrogen from the flue gas by combusting a hydrocarbon or carbonaceous fuel in either pure oxygen or a mixture of pure oxygen and a CO_2-rich recycled flue gas (carbonaceous fuels include biomass). Combustion of a fuel with pure oxygen has a combustion temperature of about 3500°C which is far too high for typical power plant materials. The combustion temperature is limited to about 1300−1400°C in a typical gas turbine cycle and to about 1900°C in an oxy-fuel coal-fired boiler using current technology. The combustion temperature is controlled by the proportion of flue gas and gaseous or liquid-water recycled back to the combustion chamber.

The combustion products (or flue gas) consist mainly of carbon dioxide and water vapour together with excess oxygen required to ensure complete combustion of the fuel. It will also contain any other components in the fuel, any diluents in the oxygen stream supplied, any inerts in the fuel and from air leakage into the system from the atmosphere. The net flue gas, after cooling to condense water vapour, contains from about 80−98% CO_2 depending on the fuel used and the particular oxy-fuel combustion process. This concentrated CO_2 stream can be compressed, dried and further purified before delivery into a pipeline for storage (see Chapter 4). The CO_2 capture efficiency is very close to 100% in oxy-fuel combustion capture systems. Impurities in the CO_2 are gas components such as SO_x, NO_x, HCl and Hg derived from the fuel used, and the inert gas components, such as nitrogen, argon and oxygen, derived from the oxygen feed or air leakage into the system. The CO_2 is transported by pipeline as a dense supercritical phase. Inert gases must be reduced to a low concentration to avoid two-phase flow conditions developing in the pipeline systems. The acid gas components may need to be removed to comply with legislation covering co-disposal of toxic or hazardous waste or to avoid operations or environmental problems with disposal in deep saline reservoirs, hydrocarbon formations or in the ocean. The carbon dioxide must also be dried to prevent water condensation and corrosion in pipelines and allow use of conventional carbon-steel materials.

Although elements of oxy-fuel combustion technologies are in use in the aluminium, iron and steel and glass melting industries today, oxy-fuel technologies for CO_2 capture have yet to be deployed on a commercial scale. Therefore, the first classification between *existing technologies* and *emerging technologies* adopted in post-combustion (Section 3.3) and pre-combustion (Section 3.5) is not followed in this section. However, it is important to emphasize that the key separation step in most oxy-fuel capture systems (O_2 from air) is an 'existing technology' (see Section 3.4.5). Current methods of oxygen production by air separation comprise cryogenic distillation, adsorption using multi-bed pressure swing units and polymeric membranes. For oxy-fuel conversions requiring less than 200 $tO_2\ d^{-1}$, the adsorption system will be economic. For all the larger applications, which include power station boilers, cryogenic air separation is the economic solution (Wilkinson *et al.*, 2003a).

In the following sections we present the main oxy-fuel combustion systems classified according to how the heat of combustion is supplied and whether the flue gas is used as a working fluid (Sections 3.4.2, 3.4.3, 3.4.4). A brief overview of O_2 production methods relevant for these systems is given (Section 3.4.5). In Section 3.4.6, the emerging technology of chemical looping combustion is presented, in which pure oxygen is supplied by a metal oxide rather than an oxygen production process. The section on oxy-fuel systems closes with an overview of the status of the technology (Section 3.4.7).

3.4.2 Oxy-fuel indirect heating − steam cycle

In these systems, the oxy-fuel combustion chamber provides heat to a separate fluid by heat transfer through a surface. It can be used for either process heating, or in a boiler with a steam cycle for power generation. The indirect system can be used with any hydrocarbon or carbon-containing fuel.

The application of oxy-fuel indirect heating for CO_2 capture in process heating and power generation has been examined in both pilot-scale trials evaluating the combustion of carbonaceous fuels in oxygen and CO_2-rich recycled flue gas mixtures and engineering assessments of plant conversions as described below.

3.4.2.1 Oxy-fuel combustion trials

Work to demonstrate the application of oxy-fuel recycle combustion in process heating and for steam generation for use in steam power cycles have been mostly undertaken in pilot scale tests that have looked at the combustion, heat transfer and pollutant-forming behaviour of natural gas and coal.

One study carried out (Babcock Energy Ltd. *et al.*, 1995) included an oxy-fuel test with flue gas recycle using a 160kW, pulverized coal, low NO_x burner. The system included a heat-transfer test section to simulate fouling conditions. Test conditions included variation in recycle flow and excess O_2 levels. Measurements included all gas compositions, ash analysis and tube fouling after a 5-week test run. The work also included a case study on oxy-fuel operation of a 660 MW power boiler with CO_2 capture, compression and purification. The main test results were that NO_x levels reduced with increase in recycle rate, while SO_2 and carbon in ash levels were insensitive to the recycle rate. Fouling in the convective test section was greater with oxy-fuel firing than with air. High-slagging UK coal had worse slagging when using oxy-fuel firing, the higher excess O_2 level lowered carbon in ash and CO concentration.

For the combustion of pulverized coal, other pilot-scale tests by Croiset and Thambimuthu (2000) have reported that the flame temperature and heat capacity of gases to match fuel burning in air occurs when the feed gas used in oxy-fuel combustion has a composition of approximately 35% by volume O_2 and 65% by volume of dry recycled CO_2 (c.f. 21% by volume O_2 and the rest nitrogen in air). In practice, the presence of inerts such as ash and inorganic components in the coal, the specific fuel composition and moisture in the recycled gas stream and the coal feed will result in minor adjustments to this feed mixture composition to keep the flame temperature at a value similar to fuel combustion in air.

At conditions that match O_2/CO_2 recycle combustion to fuel burning in air, coal burning is reported to be complete (Croiset and Thambimuthu, 2000), with operation of the process at excess O_2 levels in the flue gas as low as 1−3% by volume O_2, producing a flue gas stream of 95−98% by volume dry CO_2 (the rest being excess O_2, NO_x, SO_x and argon) when a very high purity O_2 stream is used in the combustion process with zero leakage of ambient air into the system. No differences were detected in the fly ash formation behaviour in the combustor or SO_2 emissions compared to conventional air firing conditions. For NO_x on the other hand, emissions were lower due to zero thermal NO_x formation from the absence of nitrogen in the feed gas − with the partial recycling of NO_x also reducing the formation and net emissions originating from the fuel bound nitrogen. Other studies have demonstrated that the level of NO_x reduction is as high as 75% compared to coal burning in air (Chatel-Pelage *et al.*, 2003). Similar data for natural gas burning in O_2/CO_2 recycle mixtures report zero thermal NO_x emissions in the absence of air leakage into the boiler, with trace amounts produced as thermal NO_x when residual nitrogen is present in the natural gas feed (Tan *et al.*, 2002).

The above and other findings show that with the application of oxy-fuel combustion in modified utility boilers, the nitrogen-free combustion process would benefit from higher heat transfer rates (McDonald and Palkes, 1999), and if also constructed with higher temperature tolerant materials, are able to operate at higher oxygen concentration and lower flue gas recycle flows – both of which will considerably reduce overall volume flows and size of the boiler.

It should be noted that even when deploying a 2/3 flue gas recycle gas ratio to maintain a 35% by volume O_2 feed to a pulverized coal fired boiler, hot recycling of the flue gas prior to CO_2 purification and compression also reduces the size of all unit operations in the stream leaving the boiler to 1/5 that of similar equipment deployed in conventional air blown combustion systems (Chatel-Pelage *et al.*, 2003). Use of a low temperature gas purification step prior to CO_2 compression (see Section 3.4.2.2) will also eliminate the need to deploy conventional selective catalytic reduction for NO_x removal and flue gas desulphurization to purify the gas, a practice typically adopted in conventional air-blown combustion processes (see Figure 3.3). The overall reduction in flow volumes, equipment scale and simplification of gas purification steps will thus have the benefit of reducing both capital and operating costs of equipment deployed for combustion, heat transfer and final gas purification in process and power plant applications (Marin *et al.*, 2003).

As noted above for pulverized coal, oil, natural gas and biomass combustion, fluidized beds could also be fired with O_2 instead of air to supply heat for the steam cycle. The intense solid mixing in a fluidized bed combustion system can provide very good temperature control even in highly exothermic conditions, thereby minimizing the need for flue gas recycling. In principle, a variety of commercial designs for fluidized combustion boilers exist that could be retrofitted for oxygen firing. A circulating fluidized bed combustor with O_2 firing was proposed by Shimizu *et al.* (1999) to generate the heat required for the calcination of $CaCO_3$ (see also Section 3.3.3.4). More recently, plans for pilot testing of an oxy-fired circulating fluidized bed boiler have been published by Nsakala *et al.* (2003).

3.4.2.2 Assessments of Plants converted to oxy-fuel combustion

We now discuss performance data from a recent comprehensive design study for an application of oxy-fuel combustion in a new build pulverized coal fired power boiler using a supercritical steam cycle (see Figure 3.8; Dillon *et al.*, 2005. The overall thermal efficiency on a lower heating value basis is reduced from 44.2% to 35.4%. The net power output is reduced from 677 MW_e to 532 MW_e.

Figure 3.8. Schematic of an oxy-fuel, pulverized coal fired power plant.

Important features of the system include:

- Burner design and gas recycle flow rate have been selected to achieve the same temperatures as in air combustion (compatible temperatures with existing materials in the boiler).
- The CO_2-rich flue gas from the boiler is divided into three gas streams: one to be recycled back to the combustor, one to be used as transport and drying gas of the coal feed, and the third as product gas. The first recycle and the product stream are cooled by direct water scrubbing to remove residual particulates, water vapour and soluble acid gases such as SO_3 and HCl. Oxygen and entrained coal dust together with the second recycle stream flow to the burners.
- The air leakage into the boiler is sufficient to give a high enough inerts level to require a low temperature inert gas removal unit to be installed, even if pure O_2 were used as the oxidant in the

boiler. The cryogenic oxygen plant will, in this case, produce 95% O_2 purity to minimize power consumption and capital cost.

- The low temperature (-55°C) CO_2 purification plant (Wilkinson *et al.*, 2003b) integrated with the CO_2 compressor will not only remove excess O_2, N_2, argon but can also remove all NO_x and SO_2 from the CO_2 stream, if high purity CO_2 is required for storage. Significantly, removal of these components before final CO_2 compression eliminates the need to otherwise incorporate upstream NO_x and SO_x removal equipment in the net flue gas stream leaving the boiler. Elimination of N_2 from the flue gas results in higher SO_x concentrations in the boiler and reduced NO_x levels. Suitable corrosion resistant materials of construction must be chosen.
- The overall heat transfer is improved in oxy-fuel firing because of the higher emissivity of the CO_2/H_2O gas mixture in the boiler compared to nitrogen and the improved heat transfer in the convection section. These improvements, together with the recycle of hot flue gas, increase the boiler efficiency and steam generation by about 5%.
- The overall thermal efficiency is improved by running the O_2 plant air compressor and the first and final stages of the CO_2 compressor without cooling, and recovering the compression heat for boiler feed water heating prior to de-aeration.

Engineering studies have also been reported by Simbeck and McDonald (2001b) and by McDonald and Palkes (1999). This work has confirmed that the concept of retrofitting oxy-fuel combustion with CO_2 capture to existing coal-fired power stations does not have any technical barriers and can make use of existing technology systems.

It has been reported (Wilkinson *et al.*, 2003b) that the application of oxy-fuel technology for the retrofit of power plant boilers and a range of refinery heaters in a refinery complex (Grangemouth refinery in Scotland) is technically feasible at a competitive cost compared to other types of CO_2 capture technologies. In this case, the existing boiler is adapted to allow combustion of refinery gas and fuel oil with highly enriched oxygen and with partial flue gas recycling for temperature control. Oxy-fuel boiler conversions only needed minor burner modifications, a new O_2 injection system and controls, and a new flue gas recycle line with a separate blower. These are cheap and relatively simple modifications and result in an increase in boiler/heater thermal efficiency due to the recycle of hot gas. Modifications to a coal-fired boiler are more complex. In this study, it was found to be more economic to design the air separation units for only 95% O_2 purity instead of 99.5% to comply with practical levels of air leakage into boilers and to separate the associated argon and nitrogen in the CO_2 inert gas removal system to produce a purity of CO_2 suitable for geological storage. After conversion of the boiler, the CO_2 concentration in the flue gas increases from 17 to 60% while the water content increases from 10 to 30%. Impurities (SO_x, NO_x) and gases (excess O_2, N_2, argon) representing about 10% of the stream are separated from CO_2 at low temperature (−55°C). After cooling, compression and drying of the separated or non-recycled flue gas, the product for sequestration comprises 96% CO_2 contaminated with 2% N_2, 1% argon and less than 1% O_2 and SO_2. Production of ultra-pure CO_2 for storage would also be possible if distillation steps are added to the separation process.

3.4.2.3 Advanced zero emission power plant

The advanced zero emission power plant (or AZEP as outlined in Figure 3.9; Griffin *et al.*, 2003) is an indirect heating gas turbine cycle that incorporates a high-temperature oxygen transport membrane, operating at about 800°C −1000°C (see Section 3.4.5.2). This process uses a standard air-based gas turbine in a combined cycle arrangement. Three process steps take place in a reactor system that replaces the combustion chamber of a standard gas turbine: 1) separation of oxygen from hot air using the membrane and transport to the combustion section; 2) combustion and 3) heat exchange from the combustion products to the compressed air.

Figure 3.9. Principle flow scheme of the advanced zero emission power plant cycle.

A net efficiency for advanced zero emission power cycle of around 49–50% LHV is claimed including CO_2 compression for transport. In order to get full advantage of the potential of the most advanced gas turbines, which have inlet temperatures of 1300°C−1400°C, an afterburner fired with natural gas in air may be added behind the reactor system. The efficiency then climbs up to 52% but now 15% of the CO_2 generated by combustion is released at the stack and is not captured.

3.4.3 Oxy-fuel direct heating − gas turbine cycle

Oxy-fuel combustion takes place in a pressurized CO_2-rich recirculating stream in a modified gas turbine. The hot gas is expanded in the turbine producing power. The turbine exhaust is cooled to provide heat for a steam cycle and water vapour is condensed by further cooling. The CO_2-rich gas is compressed in the compressor section. The net CO_2-rich combustion product is removed from the system. Only natural gas, light hydrocarbons and syngas ($CO + H_2$) can be used as fuel.

3.4.3.1 Cycle description and performance

Figure 3.10 shows how a gas turbine can be adapted to run with oxy-fuel firing using CO_2 as a working fluid. Exhaust gas leaving the heat recovery steam generator is cooled to condense water. The net CO_2 product is removed and the remaining gas is recycled to the compressor. Suitable fuels are natural gas, light to medium hydrocarbons or (H_2 + CO) syngas, which could be derived from coal. The use of CO_2 as the working fluid in the turbine will necessitate a complete redesign of the gas turbine (see Section 3.4.3.2). A recent study (Dillon *et al.*, 2005) gives an overall efficiency including CO_2 compression of 45%.

Figure 3.10. Principle of the oxy-fuel gas turbine combined cycle. Exhaust gas is recycled, compressed and used in the combustion chamber to control the temperature entering the turbine.

Two typical variants of this configuration are the so-called Matiant and Graz cycles (Mathieu, 2003; Jericha *et al.*, 2003). The Matiant cycle uses CO_2 as the working fluid, and consists of features like intercooled compressor and turbine reheat. The exhaust gas is preheating the recycled CO_2 in a heat exchanger. The CO_2 generated in combustion is extracted from the cycle behind the compressor. The net overall LHV efficiency is expected to be 45−47% and can increase above 50% in a combined cycle configuration similar to that shown in Figure 3.10. The Graz cycle consists of an integrated gas turbine and steam turbine cycle. A net LHV efficiency of above 50% has been calculated for this cycle (Jericha *et al.*, 2003).

A recent comprehensive review of gas turbine cycles with CO_2 capture provides efficiencies of different cycles on a common basis (Kvamsdal *et al.*, 2004).

3.4.3.2 The CO_2/oxy-fuel gas turbine

In existing gas turbines the molecular weight of the gases in the compressor and turbine are close to that of air (28.8). In the case of oxy-fuel combustion with CO_2-recycle the compressor fluid molecular weight is about 43 and about 40 in the turbine. The change in working fluid from air to a CO_2-rich gas results in a number of changes in properties that are of importance for the design of the compressor, combustor and the hot gas path including the turbine:

- The speed of sound is 80% of air;
- The gas density is 50% higher than air;

- The specific heat ratio is lower than air resulting in a lower temperature change on adiabatic compression or expansion. An oxy-fuel gas turbine in a combined cycle has a higher optimal pressure ratio, typically 30 to 35 compared to 15 to 18 used with air in a combined cycle system. With the highest turbine inlet temperature consistent with material limitations, the rather high-pressure ratio results in an exhaust gas temperature of about 600°C, which is optimal for the steam cycle.

These changes in the fundamental properties of the working fluid will have a significant impact on gas turbine components, requiring completely new designs of compressors, combustors (to account for aerodynamic changes and acoustic feedbacks) and hot gas path (O_2 partial pressure must be low in oxy-fuel systems but it is also important to avoid reducing conditions for the materials of the turbine or the change to materials allowing much lower O_2 partial pressures).

3.4.4 Oxy-fuel direct heating − steam turbine cycle

In an oxy-fuel steam turbine cycle, water is pressurized as a liquid and is then evaporated, heated by the direct injection and combustion of a fuel with pure oxygen and expanded in a turbine. Most of the water in the low pressure turbine exhaust gas is cooled and condensed, prior to pumping back to a high pressure while the CO_2 produced from combustion is removed and compressed for pipeline transport. A variant of this cycle in which the heat is provided by burning natural gas fuel in-situ with pure oxygen was proposed by Yantovskii *et al.* (1992).

The direct combustion of fuel and oxygen has been practised for many years in the metallurgical and glass industries where burners operate at near stoichiometric conditions with flame temperatures of up to 3500°C. A water quenched H_2/O_2 burner capable of producing 60 tonne h^{-1}, 6 MPa super heated steam was demonstrated in the mid-1980s (Ramsaier *et al.*, 1985). A recent development by Clean Energy Systems incorporating these concepts where a mixture of 90 % by volume superheated steam and 10% CO_2 is produced at high temperature and pressure to power conventional or advanced steam turbines is shown in Figure 3.11. The steam is condensed in a low-pressure condenser and recycled, while CO_2 is extracted from the condenser, purified and compressed. (Anderson *et al.*, 2003 and Marin *et al.*, 2003).

Figure 3.11. Principle of the Clean Energy Systems cycle. The combustion of the fuel and oxygen is cooled by injection of liquid-water, which is recycled in the process.

Plants of this type require a clean gaseous or liquid fuel and will operate at 20 to 50 MPa pressure. The steam plus CO_2 generator is very compact. Control systems must be very precise as start-up and increase to full flow in a preheated plant can take place in less than 2 seconds. Precise control of this very rapid start was demonstrated (Ramsaier *et al.*, 1985) in a 60 tonne steam h^{-1} unit. The Clean Energy Systems studies claim efficiencies as high as 55% with CO_2 capture depending on the process conditions used.

The Clean Energy Systems technology can be initially applied with current steam turbines (565°C inlet temperature). The main technical issue is clearly the design of the steam turbines which could be used at inlet temperatures up to 1300°C by applying technology similar to that used in the hot path of gas turbines. The combustor itself (the 'gas generator') is adapted from existing rocket engine technology. In 2000, Clean Energy Systems proved the concept with a 110 kW pilot project conducted at the University of California Davis. A 20 MW thermal gas generator was successfully operated in a test run of the order of a few minutes in early 2003. A zero emissions demonstration plant (up to 6 MW electrical) is now on-line. US Department of Energy's National Energy Technology Laboratory designed the reheater (Richards, 2003) and NASA tested it in 2002. Much more technology development and demonstration is needed on this proposed power cycle, but it shows significant potential for low capital cost and high efficiency.

3.4.5 Techniques and improvements in oxygen production

Oxygen is the key requirement for any oxy-fuel combustion system. It is also a key technology for pre-combustion CO_2 capture (see Section 3.5). In the next paragraphs, existing large-scale O_2 production methods are described first, followed by emerging concepts aimed at reducing the energy consumption and cost.

3.4.5.1 Cryogenic oxygen production

The very large quantities of oxygen required for CO_2 capture using the techniques of oxy-fuel combustion and pre-combustion de-carbonization can only be economically produced, at present, by using the established process of oxygen separation from air by distillation at cryogenic temperatures (Latimer, 1967). This is a technology that has been practiced for over 100 years.

In a typical cryogenic air separation plant (Castle, 1991; Figure 3.12), air is compressed to a pressure of 0.5 to 0.6 MPa and purified to remove water, CO_2, N_2O and trace hydrocarbons which could accumulate to dangerous levels in oxygen-rich parts of the plant, such as the reboiler condenser. Two or more switching fixed bed adsorbers are used, which can be regenerated by either temperature or pressure swing, using in each case, a low pressure waste nitrogen stream. The air is cooled against returning products (oxygen and nitrogen) in a battery of aluminium plate-fin heat exchangers and separated into pure oxygen and nitrogen fractions in a double distillation column, which uses aluminium packing.

Figure 3.12. Oxygen production by distillation of liquid air. The illustration shows a 3000 tonnes d^{-1} oxygen plant (courtesy of Air Products).

Oxygen can be pumped as liquid and delivered as a high-pressure gas at up to 10 MPa. Pumped oxygen plants have largely replaced the oxygen gas compression systems. They have virtually identical power consumptions but in a pumped cycle, a high-pressure air booster compressor provides a means of efficiently vaporizing and heating the liquid oxygen stream to ambient temperature. Current plant sizes range up to 3500 $tO_2\ d^{-1}$ and larger single train plants are being designed. Typical power consumption for the delivery of 95% O_2 at low pressure (0.17 MPa, a typical pressure for an oxy-fuel application) is 200 to 240 kWh/tO_2. There are numerous process cycle variations particularly for the production of oxygen at less than 97.5% purity which have been developed to reduce power and capital cost. Note that adsorption and polymeric membrane methods of air separation are only economic for small oxygen production rates.

3.4.5.2 High temperature oxygen ion transport membranes

Ceramic mixed metal oxides have been developed which exhibit simultaneous oxygen ion and electron conduction at temperatures above 500°C and preferably above 700°C (Skinner and Kilner 2003; Bouwmeester and Van Laar, 2002; Dyer *et al.*, 2000; Bredesen *et al.*, 2004). Typical crystal structures which exhibit these properties include the perovskites and the brownmillerites. The selectivity of these materials for oxygen is infinite. The oxygen permeability is primarily controlled by the oxygen ion vacancies in the metal oxide lattice. A difference in oxygen partial pressure across the membrane will cause oxygen molecules to ionize on the ceramic surface and pass into the crystal structure while simultaneously on the permeate side of the membrane, the oxygen ions give up their electrons and leave the ceramic in the region of lower activity. The electron conduction path is through the metal ions in the lattice. Unlike conventional membranes, the flux through the ceramic is a function of the partial pressure ratio. In the technical literature, the engineered structures of these ceramic mixed metal oxides are referred to as *ion transport membranes, ITM* or *oxygen transport membranes, OTM*.

The oxygen transport membrane can be fabricated in the form of plain tubes or as hollow fins on a central collector tube (Armstrong *et al.*, 2002). The finned elements are then mounted in tube sheets within a pressure vessel with high-pressure air flowing over the fins. There are several new concepts that have been proposed for using oxygen transport membranes in power cycles with CO_2 capture. A prime example of an oxy-fuel gas turbine cycle that incorporates an oxygen transport membrane for oxygen production is the advanced zero emission power plant described in Section 3.4.2.3. Another example is found in Sundnes (1998).

3.4.5.2.1 Development status

Oxygen transport membrane systems for oxygen production are currently in the early stages of development by at least two consortia receiving research funding from the US Department of Energy and the European Commission. The concept has now reached the pilot plant stage and projected cost, manufacturing procedures and performance targets for full size systems have been evaluated. Systems capable of large-scale production are projected to be available after industrial demonstration in about 7 years time (Armstrong *et al.*, 2002).

3.4.6 Chemical looping combustion

Originally proposed by Richter and Knoche (1983) and with subsequent significant contributions by Ishida and Jin (1994), the main idea of chemical looping combustion is to split combustion of a hydrocarbon or carbonaceous fuel into separate oxidation and reduction reactions by introducing a suitable metal oxide as an oxygen carrier to circulate between two reactors (Figure 3.13). Separation of oxygen from air is accomplished by fixing the oxygen as a metal oxide. No air separation plant is required. The reaction between fuel and oxygen is accomplished in a second reactor by the release of oxygen from the metal oxide in a reducing atmosphere caused by the presence of a hydrocarbon or carbonaceous fuel. The recycle rate of the solid material between the two reactors and the average solids residence time in each reactor, control the heat balance and the temperature levels in each reactor. The effect of having combustion in two reactors compared to conventional combustion in a single stage is that the CO_2 is not diluted with nitrogen gas, but is almost pure after separation from water, without requiring any extra energy demand and costly external equipment for CO_2 separation.

Figure 3.13. The chemical looping combustion principle − in a gas turbine cycle.

Possible metal oxides are some oxides of common transition-state metals, such as iron, nickel, copper and manganese (Zafar *et al.*, 2005). The metal/metal oxide may be present in various forms, but most studies so far have assumed the use of particles with diameter 100−500 μm. In order to move particles between the two reactors, the particles are fluidized. This method also ensures efficient heat and mass transfer between the gases and the particles. A critical issue is the long-term mechanical and chemical stability of the particles that have to undergo repeated cycles of oxidation and reduction, to minimize the make-up requirement. When a chemical looping cycle is used in a gas turbine cycle, the mechanical strength for crushing and the filtration system is important to avoid damaging carry-over to the turbine.

The temperature in the reactors, according to available information in the literature, may be in the range 800°C−1200°C. NO_x formation at these typical operating temperatures will always be low.

The fuel conversion in the reduction reactor may not be complete, but it is likely (Cho *et al.*, 2002) that the concentrations of methane and CO when burning natural gas are very small. In order to avoid deposit of carbon in the reduction reactor, it is necessary to use some steam together with the fuel.

The chemical looping principle may be applied either in a gas turbine cycle with pressurized oxidation and reduction reactors, or in a steam turbine cycle with atmospheric pressure in the reactors. In the case of a gas turbine cycle, the oxidation reactor replaces the combustion chamber of a conventional gas turbine. The exothermic oxidation reaction provides heat for increasing the air temperature entering the downstream expansion turbine. In addition, the reduction reactor exit stream may also be expanded in a turbine together with steam production for power generation. The cooled low pressure CO_2 stream will then be compressed to pipeline pressure. Another option is to generate steam using heat transfer surfaces in the oxidation reactor. Current circulating fluidized bed combustion technology operating at atmospheric pressure in both the oxidation and reduction stages necessitates the use of a steam turbine cycle for power generation. Using natural gas as fuel in a chemical looping combustion cycle which supplies a gas turbine combined cycle power plant and delivering CO_2 at atmospheric pressure, the potential for natural gas fuel-to-electricity conversion efficiency is estimated to be in the range 45−50% (Brandvoll and Bolland, 2004). Work on chemical looping combustion is currently in the pilot plant and materials research stage.

3.4.7 Status and outlook

Oxy-fuel combustion applied to furnaces, process heaters, boilers and power generation systems is feasible since no technical barriers for its implementation have been identified. Early use of this capture technology is likely to address applications involving indirect heating in power generation and process heating (Section 3.4.2), since these options involve the minimal modification of technologies and infrastructure that have hitherto been already developed for the combustion of hydrocarbon or carbonaceous fuels in air. However, several novel applications proposed for direct heating in steam turbine cycles or gas turbine cycles for power generation (Sections 3.4.3 and 3.4.4) still require the development of new components such as oxy-fuel combustors, higher temperature tolerant components such as CO_2- and H_2O-based turbines with blade cooling, CO_2 compressors and high temperature ion transport membranes for oxygen separation. As for Chemical Looping Combustion, it is currently still at an early stage of development.

The potential for thermal efficiencies for oxy-fuel cycles with CO_2 capture, assuming the current state of development in power plant technology, is depicted in Figures 3.6 and 3.7. Power generation from pulverized coal fired systems, using supercritical steam conditions presently operate at efficiencies around 45% (LHV), while projections to the 2010−2020 time frame are predicting efficiencies above 50% (IEA, 2004) for plants using ultra-supercritical steam conditions. An increase in efficiency of more than 5% can therefore be expected for future oxy-fuel capture systems based on coal firing that could potentially match the best efficiencies realisable today for pulverized coal-fired plants without CO_2 capture. Similarly, natural gas fired combined cycles will have efficiencies of 65% in 2020 (IEA GHG, 2000b and up from current efficiencies between 55 and 58%), which will enable plant efficiencies for natural gas fired oxy-fuel cycles with CO_2 capture above 50%. The energy penalty for producing oxygen is by far the most important cause for reduced efficiency in an oxy-fuel cycle compared to a conventional power plant.

Current technology development envisages very high efficiency separation of NO_x, SO_x, and Hg, as part of the CO_2 compression and purification system. Improved separation efficiencies of these contaminants are possible based on further process and heat integration in the power cycle.

Current cryogenic oxygen technology is showing continuing cost reduction based on improved compressor efficiencies, more efficient process equipment and larger scale plants. The new high temperature oxygen membrane could significantly improve power generation efficiency and reduce capital cost.

Future oxy-fuel demonstration plants could be based on retrofits to existing equipment such as process heaters and boilers, in order to minimize development costs and achieve early market entry. In this respect, power systems of reference for oxy-fuel combustion capture are mainly the steam-based pulverized coal and natural gas fired plants that currently represent up to 1468 GWe, or 40% (IEA WEO, 2004) of the existing global infrastructure (see also Section 3.1.2.3). Several demonstration units may be expected within the next few years particularly in Europe, USA, Canada and Australia where active research initiatives are currently underway. As these developments proceed and the technologies achieve market penetration they may become competitive relative to alternate options based on pre- and post-combustion CO_2 capture. A significant incentive to the development of oxy-fuel combustion technology, as well as for pre- and post-combustion capture technologies, is the introduction of environmental requirements and/or fiscal incentives to promote CO_2 capture and storage.

3.5 Pre-combustion capture systems

3.5.1 Introduction

A pre-combustion capture process typically comprises a first stage of reaction producing a mixture of hydrogen and carbon monoxide (syngas) from a primary fuel. The two main routes are to add steam (reaction 1), in which case the process is called ‘steam reforming’, or oxygen (reaction 2) to the primary fuel. In the latter case, the process is often called ‘partial oxidation’ when applied to gaseous and liquid fuels and ‘gasification’ when applied to a solid fuel, but the principles are the same.

Steam Reforming

$C_xH_y + xH_2O \Leftrightarrow xCO + (x+y/2)H_2 \quad \Delta H \text{ +ve}$ [1]

Partial Oxidation

$C_xH_y + x/2O_2 \Rightarrow xCO + y/2H_2 \quad \Delta H \text{ –ve}$ [2]

This is followed by the ‘shift’ reaction to convert CO to CO_2 by the addition of steam (reaction 3):

Water Gas Shift Reaction

$CO + H_2O \Leftrightarrow CO_2 + H_2 \quad \Delta H -41 \text{ kJ mol}^{-1}$ [3]

Finally, the CO_2 is removed from the CO_2/H_2 mixture. The concentration of CO_2 in the input to the CO_2/H_2 separation stage can be in the range 15−60% (dry basis) and the total pressure is typically 2−7 MPa. The separated CO_2 is then available for storage.

It is possible to envisage two applications of pre-combustion capture. The first is in producing a fuel (hydrogen) that is essentially carbon-free. Although the product H_2 does not need to be absolutely pure and may contain low levels of methane, CO or CO_2, the lower the level of carbon-containing compounds, the greater the reduction in CO_2 emissions. The H_2 fuel may also contain inert diluents, such as nitrogen (when air is typically used for partial oxidation), depending on the production process and can be fired in a range of heaters, boilers, gas turbines or fuel cells.

Secondly, pre-combustion capture can be used to reduce the carbon content of fuels, with the excess carbon (usually removed as CO_2) being made available for storage. For example, when using a low H:C ratio fuel such as coal it is possible to gasify the coal and to convert the syngas to liquid Fischer-Tropsch fuels and chemicals which have a higher H:C ratio than coal. In this section, we consider both of these applications.

This section reports on technologies for the production of H_2 with CO_2 capture that already exist and those that are currently emerging. It also describes enabling technologies that need to be developed to enhance the pre-combustion capture systems for power, hydrogen or synfuels and chemicals production or combination of all three.

3.5.2 Existing technologies

3.5.2.1 Steam reforming of gas and light hydrocarbons

Steam reforming is the dominant technology for hydrogen production today and the largest single train plants produce up to 480 $tH_2\ d^{-1}$. The primary energy source is often natural gas, Then the process is referred to as steam methane reforming (SMR), but can also be other light hydrocarbons, such as naphtha. The process begins with the removal of sulphur compounds from the feed, since these are poisons to the current nickel-based catalyst and then steam is added. The reforming reaction (1), which is endothermic, takes place over a catalyst at high temperature (800°C−900°C). Heat is supplied to the reactor tubes by burning part of the fuel (secondary fuel). The reformed gas is cooled in a waste heat boiler which generates the steam needed for the reactions and passed into the CO shift system. Shift reactors in one or two stages are used to convert most of the CO in the syngas to CO_2 (Reaction 3, which is exothermic). The conventional two-stage CO conversion reduces the CO concentration in syngas (or in hydrogen) down to 0.2−0.3%. High temperature shift reactors operating between 400°C and 550°C and using an iron-chromium catalyst leave between 2% and 3% CO in the exit gas (dry basis). Copper-based catalyst can be used at temperatures from 180°C−350°C and leave from 0.2−1% CO in the exhaust. Lower CO content favours higher CO_2 recovery. The gas is then cooled and hydrogen is produced by a CO_2/H_2 separation step. Until about 30 years ago, the CO_2 was removed using a chemical (solvent) absorption process such as an amine or hot potassium carbonate and was rejected to atmosphere as a pure stream from the top of the regenerator. There are many of these plants still in use and the CO_2 could be captured readily.

Modern plants, however, use a pressure swing adsorber (PSA), where gases other than H_2 are adsorbed in a set of switching beds containing layers of solid adsorbent such as activated carbon, alumina and zeolites (see the fuller description of PSA in Section 3.5.2.9). The H_2 exiting the PSA (typically about 2.2 MPa) can have a purity of up to 99.999%, depending on the market need. The CO_2 is contained in a stream, from the regeneration cycle, which contains some methane and H_2. The stream is used as fuel in the reformer where it is combusted in air and the CO_2 ends up being vented to atmosphere in the reformer flue gas. Hence, to capture CO_2 from modern SMR plants would require one of the post-combustion processes described above in Section 3.3. Alternatively, the PSA system could be designed not only for high recovery of pure H_2 but also to recover pure CO_2 and have a fuel gas as the third product stream.

In a design study for a large modern plant (total capacity 720 $tH_2\ d^{-1}$), the overall efficiency of making 6.0 MPa H_2 from natural gas with CO_2 vented that is without CO_2 capture, is estimated to be 76%, LHV basis, with emissions of 9.1 kg CO_2/kg H_2 (IEA GHG, 1996). The process can be modified (at a cost) to provide a nearly pure CO_2 co-product. One possibility is to remove most of the CO_2 from the shifted, cooled syngas in a 'wet' CO_2 removal plant with an appropriate amine solvent. In this case the CO_2-deficient syngas exiting the amine scrubber is passed to a PSA unit from which relatively pure H_2 is recovered and the PSA purge gases are burned along with additional natural gas to provide the needed reformer heat. The CO_2 is recovered from the amine solvent by heating and pressurized for transport. Taking into account the power to compress the CO_2 (to 11.2 MPa) reduces the efficiency to about 73% and the emission rate to 1.4 kg CO_2/kg H_2, while the CO_2 removal rate is 8.0 kg CO_2/kg H_2.

3.5.2.2 Partial oxidation of gas and light hydrocarbons

In the partial oxidation (POX) process (reaction 2), a fuel reacts with pure oxygen at high pressure. The process is exothermic and occurs at high temperatures (typically 1250°C−1400°C). All the heat required for the syngas reaction is supplied by the partial combustion of the fuel and no external heat is required. As with SMR, the syngas will be cooled, shifted and the CO_2 removed from the mixture. The comments made on the separation of CO_2 from SMR syngas above apply equally to the POX process. POX is a technology in common use today, the efficiency is lower than SMR, but the range of fuels that can be processed is much wider.

For large-scale hydrogen production, the oxygen is supplied from a cryogenic air separation unit (ASU). The high investment and energy consumption of the ASU is compensated by the higher efficiency and lower cost of the gasification process and the absence of N_2 (from the air) in the syngas, which reduces the separation costs considerably. However for pre-combustion de-carbonization applications, in which the hydrogen would be used as fuel in a gas turbine, it will be necessary to dilute the H_2 with either N_2 or steam to reduce flame temperature in the gas turbine combustor and to limit NO_x emission levels. In this case the most efficient system will use air as the oxidant and produce a H_2/N_2 fuel mixture (Hufton *et al.* 2005)

3.5.2.3 Auto-thermal reforming of gas and light hydrocarbons

The autothermal reforming (ATR) process can be considered as a combination of the two processes described above. The heat required in the SMR reactor is generated by the partial oxidation reaction (2) using air or oxygen, but because steam is supplied to the reactor as well as excess natural gas, the endothermic reforming reaction (1) occurs in a catalytic section of the reactor downstream of the POX burner. The addition of steam enables a high conversion of fuel to hydrogen at a lower temperature. Operating temperatures of the autothermal process are typically 950°C−1050°C, although this depends on the design of the process. An advantage of the process, compared to SMR, is the lower investment cost for the reactor and the absence of any emissions of CO_2 since all heat release is internal, although this is largely offset by investment and operating cost for the oxygen plant. The range of fuels that can be processed is similar to the SMR process, but the feed gas must be sulphur free. CO_2 capture is accomplished as described above for the steam methane reforming.

3.5.2.4 Gas heated reformer

Each of the three syngas generation technologies, SMR, ATR and POX produce high temperature gas which must be cooled, producing in each case a steam flow in excess of that required by the reforming and shift reactions. It is possible to reduce this excess production by, for example, using preheated air and a pre-reformer in an SMR plant. Another technique is to use the hot syngas, leaving the primary reactor, as the shell-side heating fluid in a tubular steam/hydrocarbon reforming reactor which can operate in series, or in parallel, with the primary reactor (Abbott *et al.*, 2002). The addition of a secondary gas heated reformer will increase the hydrogen production by up to 33% and eliminate the excess steam production. The overall efficiency is improved and specific capital cost is typically reduced by 15%. Again, CO_2 capture is accomplished as described previously for steam methane reforming.

3.5.2.5 Gasification of coal, petroleum residues, or biomass

Gasification (see Figure 3.14) is a chemical process aimed at making high-value products (chemicals, electricity, clean synthetic fuels) out of low-value solid feedstocks such as coal, oil refining residues, or biomass. Gasification is basically partial oxidation (reaction 2), although steam is also supplied to the reactor in most processes. Fixed bed, fluidized bed or entrained flow gasifiers can be used. These can have very different characteristics with respect to oxidant (air or O_2), operating temperature (up to 1350°C), operating pressure (0.1−7 MPa), feed system (dry or water slurry), syngas cooling method (water quench or via radiative and convective heat exchangers) and gas clean-up system deployed. These alternative design options determine the fraction of feedstock converted to syngas, syngas composition and cost. As economics depend strongly on scale, gasification is generally considered to be suitable only for large plants. The gasifier output contains CO, H_2, CO_2, H_2O and impurities (e.g., N_2, COS, H_2S, HCN, NH_3, volatile trace minerals and Hg) that must be managed appropriately.

Figure 3.14. Simplified schematic of a gasification process showing options with CO_2 capture and electricity, hydrogen or chemical production.

A worldwide survey of commercial gasification projects identified 128 operating gasification plants with 366 gasifiers producing 42,700 MW_t of syngas (NETL-DOE, 2002 and Simbeck, 2001a). There are also about 24,500 MW_t of syngas for projects under development or construction, with 4000−5000 MW_t of syngas added annually. The feedstocks are mainly higher rank coals and oil residues. Most commercial gasification growth for the last 20 years has involved entrained-flow gasifiers, for which there are three competing systems on the market. Recent commercial gasification development has been mainly with industrial ammonia production, industrial polygeneration (in which clean syngas is used to make electricity and steam along with premium syngas chemicals) and IGCC power plants. Commercial experience with biomass gasification and fluidized bed gasification has been limited.

CO_2 capture technology is well established for gasification systems that make chemicals and synthetic fuels (NETL-DOE, 2002). Gasification-based NH_3 plants (many in China) include making pure H_2 and CO_2 separation at rates up to 3500 $tCO_2\ d^{-1}$ per plant. South African plants making Fischer-Tropsch fuels and chemicals and a North Dakota plant making synthetic natural gas (SNG) from coal also produce large streams of nearly pure CO_2. Figure 3.15 shows a picture of the North Dakota gasification plant in which 3.3 $MtCO_2\ yr^{-1}$ is captured using a refrigerated methanol-based, physical solvent scrubbing process (Rectisol process, see Section 3.5.2.11 and Table 3.2). Most of this captured CO_2 is vented and about 1.5 Mtonnes yr^{-1} of this stream is currently pipelined to the Weyburn, Canada enhanced oil recovery and CO_2 storage project (see Chapter 5).

Figure 3.15. North Dakota coal gasification plant with 3.3 $MtCO_2\ yr^{-1}$ capture using a cold methanol, physical solvent process (cluster of 4 tall columns in the middle of the picture represent the H_2S and CO_2 capture processes; part of the captured stream is used for EOR with CO_2 storage in Weyburn, Saskatchewan, Canada).

When CO_2 capture is an objective, O_2-blown and high-pressure systems are preferred because of the higher CO_2 partial pressures. De-carbonization via gasification entails lower energy penalties for CO_2 capture than does post-combustion capture when considering only the separation stage, because the CO_2 can be recovered at partial pressures up to 3 orders of magnitude higher. This greatly reduces CO_2 absorber size, solvent circulation rates and CO_2 stripping energy requirements. However, additional energy penalties are incurred in shifting the CO in the syngas to CO_2 and in other parts of the system (see examples for IGCC plant with CO_2 capture in Figures 3.6 and 3.7). Recent analyses for bituminous coals (see, for example, IEA GHG, 2003) suggest using simple high-pressure entrained-flow gasifiers with water slurry feed and direct water quench followed by 'sour' (sulphur-tolerant) shift reactors and finally co-removal of CO_2 and H_2S by physical absorption. With sour shifting, hot raw syngas leaving the gasifier requires only one cooling cycle and less processing. Oxygen requirements increase for slurry fed gasifiers and conversion efficiencies decline with higher cycle efficiency losses with quench cooling. Similar trends are also noted with a shift from bituminous to lower rank sub-bituminous coal and lignite (Breton and Amick, 2002). Some analyses (e.g., Stobbs and Clark, 2005) suggest that the advantages of pre-combustion over post-combustion de-carbonization may be small or disappear for such coals converted with entrained-flow gasifiers. High-pressure, fluidized-bed gasifiers may be better suited for use with low-rank coals, biomass and various carbonaceous wastes. Although there are examples of successful demonstration of such gasifiers (e.g., the high temperature Winkler, Renzenbrink *et al*., 1998), there has been little commercial-scale operating experience.

The H_2S in syngas must be removed to levels of tens of ppm for IGCC plants for compliance with SO_2 emissions regulations and to levels much less than 1 ppm for plants that make chemicals or synthetic fuels, so as to protect synthesis catalysts. If the CO_2 must be provided for storage in relatively pure form, the common practice would be to recover first H_2S (which is absorbed more readily than CO_2) from syngas (along with a small amount of CO_2) in one recovery unit, followed by reduction of H_2S to elemental sulphur in a Claus plant and tail gas clean-up, and subsequent recovery of most of the remaining CO_2 in a separate downstream unit. An alternative option is to recover sulphur in the form of sulphuric acid (McDaniel and Hormick, 2002). If H_2S/CO_2 co-storage is allowed, however, it would often be desirable to recover H_2S and CO_2 in the same physical absorption unit, which would lead to moderate system cost savings (IEA GHG, 2003; Larson and Ren, 2003; Kreutz *et al.*, 2005) especially in light of the typically poor prospects for selling byproduct sulphur or sulphuric acid. Although co-storage of H_2S and CO_2 is routinely pursued in Western Canada as an acid gas management strategy for sour natural gas projects (Bachu and Gunter, 2005), it is not yet clear that co-storage would be routinely viable at large scales − a typical gasification-based energy project would involve an annual CO_2 storage rate of 1−4 Mtonnes yr^{-1}, whereas the total CO_2 storage rate for all 48 Canadian projects is presently only 0.48 Mtonnes yr^{-1} (Bachu and Gunter, 2005).

3.5.2.6 Integrated gasification combined cycle (IGCC) for power generation

In a coal IGCC, syngas exiting the gasifier is cleaned of particles, H_2S and other contaminants and then burned to make electricity via a gas turbine/steam turbine combined cycle. The syngas is generated and converted to electricity at the same site, both to avoid the high cost of pipeline transport of syngas (with a heating value only about 1/3 of that for natural gas) and to cost-effectively exploit opportunities for making extra power in the combined cycle's steam turbine using steam from syngas cooling. The main drivers for IGCC development were originally the prospects of exploiting continuing advances in gas turbine technology, the ease of realizing low levels of air-pollutant emissions when contaminants are removed from syngas, and greatly reduced process stream volumes compared to flue gas streams from combustion which are at low pressure and diluted with nitrogen from air.

Since the technology was initially demonstrated in the 1980s, about 4 GW_e of IGCC power plants have been built. Most of this capacity is fuelled with oil or petcoke; less than 1 GWe of the total is designed for coal (IEA CCC, 2005) and 3 out of 4 plants currently operating on coal and/or petcoke. This experience has demonstrated IGCC load-following capability, although the technology will probably be used mainly in base load applications. All coal-based IGCC projects have been subsidized, whereas only the Italian oil-based IGCC projects have been subsidized. Other polygeneration projects in Canada, the Netherlands and the United States, as well as an oil-based IGCC in Japan, have not been subsidized (Simbeck, 2001a).

IGCC has not yet been deployed more widely because of strong competition from the natural gas combined cycle (NGCC) wherever natural gas is readily available at low prices, because coal-based IGCC plants are not less costly than pulverized coal fired steam-electric plants and because of availability (reliability) concerns. IGCC availability has improved in recent years in commercial-scale demonstration units (Wabash River Energy, 2000; McDaniel and Hornick, 2002). Also, availability has been better for industrial polygeneration and IGCC projects at oil refineries and chemical plants where personnel are experienced with the chemical processes involved. The recent rise in natural gas prices in the USA has also increased interest in IGCC.

Because of the advantages for gasification of CO_2 capture at high partial pressures discussed above, IGCC may be attractive for coal power plants in a carbon-constrained world (Karg and Hannemann, 2004). CO_2 capture for pre-combustion systems is commercially ready, however, no IGCC plant incorporating CO_2 capture has yet been built. With current technology, average estimates of the energy penalties and the impact of increased fuel use for CO_2 removal are compared with other capture systems in Figures 3.6 and 3.7 and show the prospective potential of IGCC options. The data in Figures 3.6 and 3.7 also show that some IGCC options may be different from others (i.e., slurry fed and quench cooled versus dry feed and syngas cooling) and their relative merits in terms of the capital cost of plant and the delivered cost of power are discussed in Section 3.7.

3.5.2.7 Hydrogen from coal with CO_2 capture

Relative to intensively studied coal IGCC technology with CO_2 capture, there are few studies in the public domain on making H_2 from coal via gasification with CO_2 capture (NRC, 2004; Parsons 2002a, b; Gray and Tomlinson, 2003; Chiesa *et al.*, 2005; Kreutz *et al.*, 2005), even though this H_2 technology is well established commercially, as noted above. With commercial technology, H_2 with CO_2 capture can be produced via coal gasification in a system similar to a coal IGCC plant with CO_2 capture. In line with the design recommendations for coal IGCC plants described above (IEA GHG, 2003), what follows is the description from a design study of a coal H_2 system that produces, using best available technology, 1070 MW_t of H_2 from high-sulphur (3.4%) bituminous coal (Chiesa *et al.*, 2005; Kreutz *et al.*, 2005). In the base case design, syngas is produced in an entrained flow quench gasifier operated at 7 MPa. The syngas is cooled, cleaned of particulate matter, and shifted (to primarily H_2 and CO_2) in sour water gas shift reactors. After further cooling, H_2S is removed from the syngas using a physical solvent (Selexol). CO_2 is then removed from the syngas, again using Selexol. After being stripped from the solvents, the H_2S is converted to elemental S in a Claus unit and a plant provides tail gas clean-up to remove residual sulphur emissions; and the CO_2 is either vented or dried and compressed to 150 atm for pipeline transport and underground storage. High purity H_2 is extracted at 6 MPa from the H_2-rich syngas via a pressure swing adsorption (PSA) unit. The PSA purge gas is compressed and burned in a conventional gas turbine combined cycle, generating 78 MW_e and 39 MW_e of electricity in excess of onsite electricity needs in the without and with CO_2 capture cases, respectively. For this base case analysis, the effective efficiency of H_2 manufacture was estimated to be 64% with CO_2 vented and 61% with CO_2 captured, while the corresponding emission rates are 16.9 kg CO_2 and 1.4 kg CO_2/kg H_2, respectively. For the capture case, the CO_2 removal rate was 14.8 kg CO_2/kg H_2. Various alternative system configurations were explored. It was found that there are no thermodynamic or cost advantages from increasing the electricity/H_2 output ratio, so this ratio would tend to be determined by relative market demands for electricity and H_2. One potentially significant option for reducing the cost of H_2 with CO_2 capture to about the same level as with CO_2 vented involves H_2S/CO_2 co-capture in a single Selexol unit, as discussed above.

3.5.2.8 Carbon-based fluid fuels and multi-products

As discussed in Chapter 2, clean synthetic high H/C ratio fuels can be made from syngas via gasification of coal or other low H/C ratio feedstocks. Potential products include synthetic natural gas, Fischer-Tropsch diesel/gasoline, dimethyl ether, methanol and gasoline from methanol via the Mobil process. A byproduct is typically a stream of relatively pure CO_2 that can be captured and stored.

Coal derived Fischer-Tropsch synfuels and chemicals have been produced on a commercial scale in South Africa; coal methanol is produced in China and at one US plant; and coal SNG is produced at a North Dakota (US) plant (NETL-DOE, 2002). Since 2000, 1.5 $MtCO_2\ yr^{-1}$ from the North Dakota synthetic natural gas plant (see Figure 3.15) have been transported by pipeline, 300 km to the Weyburn oil field in Saskatchewan, Canada for enhanced oil recovery with CO_2 storage.

Synfuel manufacture involves O_2-blown gasification to make syngas, gas cooling, gas clean-up, water gas shift and acid gas (H_2S/CO_2) removal. Subsequently cleaned syngas is converted catalytically to fuel in a synthesis reactor and unconverted syngas is separated from the liquid fuel product. At this point either most unconverted gas is recycled to the synthesis reactor to generate additional liquid fuel and the remaining unconverted gas is used to make electricity for onsite needs, or syngas is passed only once through the synthesis reactor, and all unconverted syngas is used for other purposes, for example, to make electricity for sale to the electric grid as well as for onsite use. The latter *once through* option is often more competitive as a technology option (Williams, 2000; Gray and Tomlinson, 2001; Larson and Ren, 2003; Celik *et al.*, 2005).

New slurry-phase synthesis reactors make the once through configuration especially attractive for CO-rich (e.g., coal-derived) syngas by making high once through conversion possible. For once through systems, a water gas shift reactor is often placed upstream of the synthesis reactor to generate the H_2/CO ratio that maximizes synfuel conversion in the synthesis reactor. It is desirable to remove most CO_2 from shifted syngas to maximize synthetic fuel conversion. Also, because synthesis catalysts are extremely sensitive to H_2S and various trace contaminants, these must be removed to very low levels ahead of the synthesis reactor. Most trace metals can be removed at low-cost using an activated carbon filter. CO_2 removal from syngas upstream of the synthesis reactor is a low-cost, partial de-carbonization option, especially when H_2S and CO_2 are co-captured and co-stored as an acid gas management strategy (Larson and Ren, 2003). Further de-carbonization can be realized in once through systems, at higher incremental cost, by adding additional shift reactors downstream of the synthesis reactor, recovering the CO_2, and using the CO_2-depleted, H_2-rich syngas to make electricity or some mix of electricity plus H_2 in a 'polygeneration' configuration (see Figure 3.16). The relative amounts of H_2 and electricity produced would depend mainly on relative demands, as there do not seem to be thermodynamic or cost advantages for particular H_2/electricity production ratios (Chiesa *et al.*, 2005; Kreutz *et al.*, 2005). When syngas is de-carbonized both upstream and downstream of the synthesis reactor (see Figure 3.16) it is feasible to capture and store as CO_2 up to 90% of the carbon in the original feedstock except that contained in the synthetic fuel produced.

Figure 3.16. Making liquid fuel, electricity and hydrogen from coal via gasification, with CO_2 capture and storage.

An example of such a system (Celik *et al.*, 2005) is one making 600 MW of dimethyl ether (containing 27% of coal input energy and 20% of coal input carbon) plus 365 MW of electricity (no H_2) from coal. For this system the CO_2 storage rate (equivalent to 74% of C in coal) is 3.8 Mtonnes yr^{-1} (39% from upstream of the synthesis reactor). The estimated fuel cycle-wide GHG emissions for dimethyl ether are 0.9 times those for crude oil-derived diesel and those for electricity are 0.09 times those for a 43% efficient coal-fired power plant with CO_2 vented.

3.5.2.9 Pressure swing adsorption

Pressure Swing Adsorption (PSA) is the system of choice for the purification of syngas, where high purity H_2 is required. However, it does not selectively separate CO_2 from the other waste gases and so for an SMR application the CO_2 concentration in the waste gas would be 40−50% and require further upgrading to produce pure CO_2 for storage. Simultaneous H_2 and CO_2 separation is possible by using an additional PSA section to remove the CO_2 prior to the H_2 separation step, such as the Air Products Gemini Process (Sircar, 1979).

The PSA process is built around adsorptive separations of cyclic character. The cycles consist of two basic steps: adsorption, in which the more adsorbable species are selectively removed from the feed gas and regeneration (desorption), when these species are removed from the adsorbent so that it can be ready for the next cycle. It is possible to obtain useful products during both adsorption and regeneration. The principal characteristic of PSA processes is the use of a decrease in pressure and/or the purge by a less adsorbable gas to clean the adsorbent bed. Apart from adsorption and regeneration, a single commercial PSA cycle consists of a number of additional steps, including co- and counter-current pressurization, pressure equalization and co- and counter-current depressurization. A detailed description of the PSA technique, along with its practical applications can be found elsewhere (Ruthven *et al.*, 1994).

3.5.2.10 Chemical solvent processes

Chemical solvents are used to remove CO_2 from syngas at partial pressures below about 1.5 MPa (Astarita *et al.*, 1983) and are similar to those used in post-combustion capture (see Section 3.3.2.1). The solvent removes CO_2 from the shifted syngas by means of a chemical reaction, which can be reversed by pressure reduction and heating. The tertiary amine methyldiethanolamine (MDEA, see Table 3.2) is widely used in modern industrial processes, due to the high CO_2 loading possible and the low regenerator heating load, relative to other solvents. Hot potassium carbonate (the most common commercial version of which is known as Benfield) was used for CO_2 removal in most hydrogen plants until about 15 years ago.

3.5.2.11 Physical solvent processes

Physical solvent (or absorption) processes are mostly applicable to gas streams which have a high CO_2 partial pressure and/or a high total pressure. They are often used to remove the CO_2 from the mixed stream of CO_2 and H_2 that comes from the shift reaction in pre-combustion CO_2 capture processes, such as product from partial oxidation of coal and heavy hydrocarbons.

The leading physical solvent processes are shown in Table 3.2. The regeneration of solvent is carried out by release of pressure at which CO_2 evolves from the solvent, in one or more stages. If a deeper regeneration is required the solvent would be stripped by heating. The process has low energy consumption, as only the energy for pressurizing the solvent (liquid pumping) is required.

The use of high sulphur fossil fuels in a pre-combustion capture process results in syngas with H_2S. Acid gas components must be removed. If transport and storage of mixed CO_2 and H_2S is possible then both components can be removed together. Sulphinol was developed to achieve significantly higher solubilities of acidic components compared to amine solvents, without added problems of excessive corrosion, foaming, or solution degradation. It consists of a mixture of sulpholane (tetrahydrothiophene 1,1-dioxide), an alkanolamine and water in various proportions depending on the duty. If pure CO_2 is required, then a selective process is required using physical solvents − often Rectisol or Selexol. The H_2S must be separated at sufficiently high concentration (generally >50%) to be treated in a sulphur recovery plant.

3.5.2.12 Effect on other pollutants

Pre-combustion capture includes reforming, partial oxidation or gasification. In order to maintain the operability of the catalyst of reformers, sulphur (H_2S) has to be removed prior to reforming. In gasification, sulphur can be captured from the syngas, and in the case when liquid or solid fuels are gasified, particulates, NH_3, COS and HCN are also present in the system that need to be removed. In general, all of these pollutants can be removed from a high-pressure fuel gas prior to combustion, where combustion products are diluted with nitrogen and excess oxygen. In the combustion of hydrogen or a hydrogen-containing fuel gas, NO_x may be formed. Depending upon combustion technology and hydrogen fraction, the rate at which NO_x is formed may vary. If the volumetric fraction of hydrogen is below approximately 50−60%, NO_x formation is at the same level as for natural gas dry low-NO_x systems (Todd and Battista, 2001).

In general, with the exception of H_2S that could be co-removed with CO_2, other pollutants identified above are separated in additional pretreatment operations, particularly in systems that gasify liquid or solid fuels. High temperature pretreatment operations for these multi-pollutants that avoid cooling of the syngas have the advantage of improving the cycle efficiency of the overall gasification process, but these separation processes have not been commercially demonstrated. Although it is not yet regulated as a ‘criteria pollutant’, mercury (Hg), is currently the focus of considerable concern as a pollutant from coal power systems. For gasification systems Hg can be recovered from syngas at ambient temperatures at very low-cost, compared to Hg recovery from flue gases (Klett *et al.*, 2002).

3.5.3 Emerging technologies

Emerging options in both natural gas reforming and coal gasification incorporate novel combined reaction/separation systems such as sorption-enhanced reforming and sorption-enhanced water gas shift, membrane reforming and membrane water gas shift. Finally there is a range of technologies that make use of the carbonation of CaO for CO_2 capture.

3.5.3.1 Sorption enhanced reaction

A concept called Sorption Enhanced Reaction (SER) uses a packed bed containing a mixture of a catalyst and a selective adsorbent to remove CO_2 from a high temperature reaction zone, thus driving the reaction to completion. (Hufton *et al.*, 1999). The adsorbent is periodically regenerated by using a pressure swing, or temperature swing adsorption system with steam regeneration (Hufton *et al.*, 2005).

High temperature CO_2 adsorbents such as hydrotalcites (Hufton *et al.*, 1999) or lithium silicate (Nakagawa and Ohashi, 1998) can be mixed with a catalyst to promote either the steam methane reforming reaction (Reaction 1) or water gas shift reaction (Reaction 3) producing pure hydrogen and pure CO_2 in a single process unit. The continuous removal of the CO_2 from the reaction products by adsorption shifts each reaction towards completion.

The SER can be used to produce hydrogen at 400°C−600°C to fuel a gas turbine combined cycle power generation system. A design study based on a General Electric 9FA gas turbine with hot hydrogen, produced from an air blown ATR with a sorption enhanced water gas shift reactor, gave a theoretical net efficiency of 48.3% with 90% CO_2 capture at 99% purity and 150 bar pressure (Hufton *et al.*, 2005). The process is currently at the pilot plant stage.

3.5.3.2 Membrane reactors for hydrogen production with CO_2 capture

Inorganic membranes with operation temperatures up to 1000°C offer the possibility of combining reaction and separation of the hydrogen in a single stage at high temperature and pressure to overcome the equilibrium limitations experienced in conventional reactor configurations for the production of hydrogen. The combination of separation and reaction in membrane steam reforming and/or membrane water gas shift offers higher conversion of the reforming and/or shift reactions due to the removal of hydrogen from these equilibrium reactions as shown in Reactions (1) and (3) respectively. The reforming reaction is endothermic and can, with this technique, be forced to completion at lower temperature than normal (typically 500°C−600°C). The shift reaction being exothermic can be forced to completion at higher temperature (500°C−600°C).

Another reason to incorporate H_2 separation membranes in the hydrogen production system is that CO_2 is also produced without the need for additional separation equipment. Membrane reactors allow one-step reforming, or a single intermediate water gas shift reaction, with hydrogen separation (the permeate) leaving behind a retentate gas which is predominantly CO_2 and a small amount of non-recovered hydrogen and steam. This CO_2 remains at the relatively high pressure of the reacting system (see Figure 3.17). Condensation of the steam leaves a concentrated CO_2 stream at high pressure, reducing the compression energy for transport and storage. Membrane reforming will benefit from high-pressure operation due to the increased H_2 partial pressure differential across the membrane which is the driving force for hydrogen permeation. Therefore membrane reactors are also seen as a good option for pre-combustion de-carbonization where a low-pressure hydrogen stream for fuel gas and a high-pressure CO_2-rich stream for transport and storage are required. The use of the membrane reformer reactor in a gas turbine combined cycle means that the hydrogen needs to be produced at such pressure that the significant power consumption for the hydrogen compression is avoided. This could be done by increasing the operating pressure of the membrane reactor or by using a sweep gas, for instance steam, at the permeate side of the membrane (Jordal *et al.*, 2003).

Figure 3.17. Operating principle of a membrane reactor.

For these membrane reactor concepts, a hydrogen selective membrane capable of operating in a high-temperature, high-pressure environment is needed. In the literature a number of membrane types have been reported that have these capabilities and these are listed in Table 3.3. Microporous inorganic membranes based upon surface diffusion separation exhibit rather low separation factors (e.g., H_2/CO_2 separation factor of 15). However, the separation ability of the current commercially available gamma-alumina and silica microporous membranes (which have better separation factors, up to 40) depends upon the stability of the membrane pore size, which is adversely affected by the presence of steam in the feed streams. The dense ceramic membranes based on inorganic perovskite oxides (also called proton conducting) need high temperatures, higher than 800°C, to achieve practical hydrogen flux rates. Palladium-based dense membranes are also known for their high hydrogen selectivity and permeability over other gases in the temperature range 300°C−600°C that is appropriate for these two reactions. Palladium alloy tubes have been available for several decades, but for CCS applications they are too expensive due to the membrane thickness needed for structural stability and consequently low hydrogen flux rates. In order to be suitable for the target application, a hydrogen separation membrane must have adequate selectivity and flux rate and must be stable in the reducing coal gas or fuel-reforming environment containing steam and hydrogen sulphide.

Table 3.3. Membrane materials, operating conditions and characteristics for H_2 separation.

A number of membrane reactor developments have been reported for hydrogen production with CO_2 capture. Several groups have evaluated methane steam reforming membrane reactors based on palladium alloy membranes (Middleton *et al.*, 2002, Damle and Dorchak, 2001). These evaluations showed that membrane reactors could achieve 90% CO_2 recovery and that at this moment the projected cost is nearly identical to that for a conventional system. However, a cost-reduction can be achieved by either reducing the material cost of the membrane or by increasing the permeability. Similar evaluations of membrane reactors for the shift conversion and separation of CO_2 from syngas produced from heavy feeds by gasification have been reported (Bracht *et al.*, 1997; Middleton 2002; Lowe *et al.*, 2003). For these gasifier systems the membrane reactors could reduce the costs for capturing CO_2 and the cost reduction would be more significant if they could be made sulphur tolerant.

3.5.3.3 Microchannel reformer

Microreactor technology can be used to produce a SMR, or low temperature air-based POX system using a multichannel plate-fin heat exchanger, fabricated in stainless steel or high nickel alloy by vacuum brazing or diffusion bonding.

An SMR reactor consists of alternate passages having fins, which are coated with catalyst or porous catalyst insets. Heat is produced by catalytic combustion of fuel gas premixed with air and transferred by conduction to the adjacent passage fed with the steam/hydrocarbon mixture, where the reforming reaction takes place (Babovic *et al.*, 2001). Very compact high efficiency systems can be produced. Although these units are being currently developed by a number of groups for small-scale H_2 production for fuel cell applications, they also show promise in larger H_2 plants.

3.5.3.4 Conversion to hydrogen and carbon

Thermal cracking or pyrolysis of methane is the reaction where methane reacts to carbon and hydrogen through:

Methane pyrolysis:

$$CH_4 \rightarrow C + 2\,H_2 \qquad [4]$$

The main advantage of the process is that it can potentially yield a clean gas (free of carbon oxides) that could be used directly for power production, but a disadvantage is that the chemical energy from the oxidation of carbon to CO_2 is not released. The cracking reaction is endothermic and so heat has to be supplied to the reaction. If the natural gas is converted fully, the theoretical yield of hydrogen corresponds to 60% of the heating value of the natural gas. The amount of carbon, which can be obtained, corresponds to 49% of the heating value, with the extra 9% of the energy in this calculation being provided as endothermic heat shown by Reaction (4) above. Therefore full conversion can be achieved only if heat is supplied from an external source. If full conversion of methane is not achieved, the remaining methane will be combusted to produce heat. There are many different methods under development for reactors based on this principle, including thermal catalytic, thermal non-catalytic and plasma cracking.

In the plasma cracking process natural gas or other hydrocarbons are supplied to a plasma reactor where the hydrocarbons are cracked under pyrolysis conditions (i.e., in absence of oxides, e.g., steam, which can supply oxygen to form CO or CO_2). The plasma arc, for which electricity is used, supplies the heat for the cracking reaction. Advantages of the process are its flexibility with respect to the fuel and the high quality carbon black which can be produced. Two small-scale plasma cracking processes for hydrogen/syngas production have been in development. The Glid Arc process has been developed by the Canadian Synergy Technologies Corporation. The second process is the Kvaerner CB&H process. Kvaerner has reported results for a pilot plant producing 1000 Nm^3 hydrogen per hour and 270 kg or 500 kg carbon black using natural gas and aromatic oil respectively (IEA GHG, 2001).

3.5.3.5 *Technologies based on calcium oxide*

There is a range of pre-combustion systems that make use of the carbonation reaction of CaO at high pressures and temperatures, to further integrate the gasification of the fuel (if solid), the shift reaction, and in-situ CO_2 removal with CaO. The overall reaction aimed in the system is:

Carbonation of calcium oxide:

$$CaO + C + 2\,H_2O \leftrightarrow CaCO_3 + 2H_2 \qquad [5]$$

The regeneration of the sorbent produces pure CO_2 when carried out in a separate reactor by calcining $CaCO_3$. A range of systems can be developed under this general reaction scheme depending on the technology adopted for gasification, carbonation-calcination, hydrogen utilization route and storage option for CO_2. The first of these concepts was proposed at the Los Alamos National Laboratory (USA) and is currently under development as the Zero Emission Coal Alliance (ZECA) process. The full system includes (Lackner *et al.*, 2001) a hydro-gasification reactor, solid oxide fuel cell and a technology for mineral carbonation. However, the fuel cell will require more development and mineral carbonation is only at the laboratory investigation stage (see Section 7.2 for a discussion of mineral carbonation).

The HyPrRing process (Lin *et al.*, 2002) is being developed by the Center for Coal Utilization of Japan. It integrates gasification, reforming and *in situ* CO_2 capture in a single reactor at pressures above 12 MPa and temperature above 650ºC. Projects in Norway using natural gas and in Germany using brown coal (Bandi *et al.*, 2002) are also underway developing pre-combustion systems using capture of CO_2 with CaO. Finally, General Electric (Rizeq *et al.*, 2002) is developing an innovative system involving the capture of CO_2 in the gasification reactor by a high temperature sorbent and with calcination in a separate reactor by burning part of the fuel with an oxygen carrier.

All these systems are at an early stage of development. Detailed process simulations show that the efficiencies are potentially high because most of the energy employed for sorbent regeneration is effectively transferred to the H_2 generated in Reaction (5). The systems are aimed at very large-scale generation of electricity and/or H_2 and cement production (from the deactivated sorbent, CaO). However, many uncertainties remain concerning the performance of the individual units and their practical integration. The main challenge may be the regeneration of the sorbent at very high temperatures (>900^0 C), to produce a pure stream of CO_2. Another is the operating conditions to achieve sufficient conversion towards hydrogen, without the use of a catalyst for the shift reaction.

3.5.4 Enabling technologies

The performance and cost of a pre-combustion capture system is linked to the availability of the enabling technologies that complete the system. In this section we consider the availability of industrial systems, to produce heat from the de-carbonized fuel and gas turbines and fuel cells to produce power.

3.5.4.1 Use of de-carbonized fuel in industrial systems

The use of hydrogen as a fuel for conventional fired heaters and boilers is considered to be proven and indeed it is practiced at certain industrial sites. There is a very large stock of capital equipment of this type and so the use of hydrogen as a fuel might be considered a valuable technology option in a carbon-constrained world. A study (IEA GHG, 2000c) has looked at the cost of converting an existing refinery to use hydrogen fuel.

3.5.4.2 Use of de-carbonized fuel in gas turbine systems

There is extensive commercial experience with hydrogen-rich fuel gas firing in gas turbines. For example, General Electric reports over 450,000 hours of operating experience with high hydrogen (52−95% by volume) content fuel gas in gas turbines (Shilling and Jones, 2003). Unfortunately, most of that experience is for ‘refinery gas’ where methane is the other main component of the fuel gas and is utilized in older lower firing temperature gas turbines, not the state-of-the-art over 1300°C gas turbines normally considered for large de-carbonization power plants.

Norsk Hydro and General Electric collaborated to perform full-scale combustion system testing for modern gas turbines firing hydrogen-rich gas with combustion exit temperatures of above 1400°C (Todd and Battista, 2001). The results showed good combustion conditions with low NO_x emission and acceptable hot metal temperatures for mixtures with 54−77% by volume hydrogen with most of the additional gas being nitrogen. Dilution of the hydrogen with nitrogen or steam reduces the NO_x emission.

For pre-combustion capture of CO_2 from natural gas, air-blown gasification or autothermal reforming is usually preferred (IEA GHG, 2000b; Wilkinson and Clarke, 2002). Nitrogen dilution of the hydrogen required for firing in modern gas turbines comes from the gasification air. High-pressure air is usually extracted from the gas turbine to feed the air-blown gasifier, or autothermal reformer to reduce costs and avoid a separate air compressor. The balance between the amount of air withdrawn from the gas turbine and the amount provided from a separate air compressor is determined by the particular characteristics of the gas turbine used. Some gas turbines can accept a higher ratio of expander to compressor flow, allowing greater volumes of dilution gas or smaller air-side draw flow and giving higher power output.

For pre-combustion capture of CO_2 from coal, oxygen-blown gasification is usually preferred (IEA GHG, 2003). Nitrogen dilution of the hydrogen required for firing in modern gas turbines comes from the cryogenic air separation unit (used to make the oxygen; see Section 3.4.5.1). The nitrogen is added to the hydrogen after the gasification, CO shifting and CO_2 capture to reduce the equipment sizes and cost. High-pressure air is usually extracted from the gas turbine to supply a higher than normal pressure cryogenic air separation unit to reduce costs plus air, oxygen and nitrogen compression power. An alternative IGCC scheme that incorporates newly emerging ion transport membranes for oxygen production is also described below in Section 3.5.4.3.

3.5.4.3 Syngas production using oxygen membranes

Oxygen required for a coal-fired IGCC process (Section 3.5.2.6) can be generated in an oxygen transport membrane system by using a heated, high-pressure air stream produced by heating the discharge air from the compressor section of a gas turbine (Allam *et al.*, 2002), typically at 1.6 MPa or 420°C, to the precise inlet temperature of the oxygen transport membrane module which is above 700°C. The oxygen, which permeates to the low-pressure side passes through a heat recovery section and is compressed to the final pressure of use. The O_2 depleted air leaving the oxygen transport membrane module then enters the gas turbine combustor where it is used to burn fuel before entering the gas turbine expander at the required temperature. Note that due to the necessity to have excess air in a gas turbine to limit turbine inlet temperature, removing one mole of oxygen can be compensated by injection of the equivalent thermal capacity of steam to maintain gas turbine power output. Studies have been carried out (Armstrong *et al.*, 2002) to compare oxygen transport membrane and cryogenic oxygen production in an IGCC power plant using coal as fuel. The oxygen plant projected cost was reduced by 35% and the power consumption by 37%. An LHV efficiency of 41.8% without CO_2 capture and compression is reported for this cycle compared to 40.9% when a conventional cryogenic oxygen plant is used.

For autothermal reforming or the partial oxidation of natural gas, if the permeate side of the oxygen transport membrane is exposed to a natural gas plus water vapour stream in the presence of a reforming catalyst, the oxygen will react as it leaves the membrane in an exothermic reaction (Dyer *et al.*, 2001; Carolan *et al.*, 2001), which will provide heat for the endothermic steam/natural gas reforming reaction. The oxygen partial pressure at these highly-reducing, high temperature conditions is extremely low, allowing heated air at close to atmospheric pressure to be used on the feed side of the membrane while producing a H_2 + CO mixture at high pressure from the permeate side. This system can be used to produce H_2 following CO shift reaction and CO_2 removal.

3.5.4.4 Chemical looping gasification/reforming

The chemical looping concept described in 3.4.6 is being considered for reforming of a fuel to produce H_2 and CO (Zafar *et al.*, 2005). When the amount of oxygen brought by the metal oxide into the reduction reactor is below stoichiometric requirements, the chemical reaction with the fuel produces H_2 and CO. The reaction products may subsequently be shifted with steam to yield CO_2 and more H_2.

3.5.4.5 Use of de-carbonized fuel in fuel cells

Fuel cells offer the possibility for highly efficient power production since the conversion process is not controlled by heat to work Carnot cycle restrictions (Blomen and Mugerwa, 1993). In general fuel cells feature the electrochemical oxidation of gaseous fuels directly into electricity, avoiding the mixture of the air and the fuel flows and thus the dilution with nitrogen and excess oxygen of the oxidized products (Campanari, 2002). As a result, the anode outlet stream of a fuel cell already has a very high CO_2 content that simplifies the CO_2 capture subsystem. The fuel is normally natural gas, though some concepts can also be incorporated into coal gasification systems. The systems concepts can be classified into two main groups (Goettlicher, 1999):

- Systems with pre-fuel cell CO_2 capture;
- Systems with post-fuel cell CO_2 capture.

In pre-fuel cell CO_2 capture systems (see Figure 3.18a) the fuel is first converted into hydrogen using steam reforming or coal gasification, followed by the water gas shift conversion. This system approach has been first proposed both for low temperature and for high temperature fuel cells.

Figure 3.18a. Fuel cell system with pre-fuel cell CO_2 capture. The carbon-containing fuel is first completely converted into a mixture of hydrogen and CO_2. Hydrogen and CO_2 are then separated and the H_2-rich fuel is oxidized in the fuel cell to produce electricity. The CO_2 stream is dried and compressed for transport and storage.

The post-fuel cell capture system (see Figure 3.18b) is proposed for high temperature fuel cell systems (Dijkstra and Jansen, 2003). These systems make use of the internal reforming capabilities of the high temperature fuel cells resulting in an anode off-gas that has a high CO_2-content, but also contains H_2O and unconverted CO and H_2. The water can easily be removed by conventional techniques (cooling, knock-out, additional drying). Oxidizing the H_2 and CO from the (SOFC) anode with air will result in a too high dilution of the stream with nitrogen. Haines (1999) chooses to use an oxygen-transport membrane reactor placed after the SOFC. The anode off-gas is fed to one side of the membrane, the cathode off-gas is fed to the other side of the membrane. The membrane is selective to oxygen, which permeates from the cathode off-gas stream to the anode-off gas. In the membrane unit the H_2 and CO are oxidized. The retenate of the membrane unit consist of CO_2 and water. Finally a concept using a water gas shift membrane reactor has been proposed (Jansen and Dijkstra, 2003).

Figure 3.18b. Fuel cell system with post-fuel cell CO_2 capture. The carbon-containing fuel is first converted into a syngas. The syngas is oxidized in the fuel cell to produce electricity. At the outlet of the fuel cell CO_2 is separated from the flue gas, dried and compressed for transport and storage.

3.5.5 Status and outlook

This section reviewed a wide variety of processes and fuel conversion routes that share a common objective: to produce a cleaner fuel stream from the conversion of a raw carbonaceous fuel into one that contains little, or none, of the carbon contained in the original fuel. This approach necessarily involves the separation of CO_2 at some point in the conversion process. The resulting H_2-rich fuel can be fed to a hydrogen consuming process, oxidized in a fuel cell, or burned in the combustion chamber of a gas turbine to produce electricity. In systems that operate at high pressure, the energy conversion efficiencies tend to be higher when compared to equivalent systems operating at low pressures following the combustion route, but these efficiency improvements are often obtained at the expense of a higher complexity and capital investment in process plants (see Section 3.7).

In principle, all pre-combustion systems are substantially similar in their conversion routes, allowing for differences that arise from the initial method employed for syngas production from gaseous, liquid or solid fuels and from the subsequent need to remove impurities that originate from the fuel feed to the plant. Once produced, the syngas is first cleaned and then reacted with steam to produce more H_2 and CO_2. The separation of these two gases can be achieved with well-known, commercial absorption-desorption methods, producing a CO_2 stream suitable for storage. Also, intense R&D efforts worldwide are being directed towards the development of new systems that combine CO_2 separation with some of the reaction steps, such as the steam reforming of natural gas or water gas shift reaction stages, but it is not yet clear if these emerging concepts (see Section 3.5.3) will deliver a lower CO_2 capture cost.

In power systems, pre-combustion CO_2 capture in natural gas combined cycles has not been demonstrated. However, studies show that based on current state of the art gas turbine combined cycles, pre-combustion CO_2 capture will reduce the efficiency from 56% LHV to 48% LHV (IEA, 2000b). In natural gas combined cycles, the most significant area for efficiency improvement is the gas turbine and it is expected that by 2020, the efficiency of a natural gas combined cycle could be as high as 65% LHV (IEA GHG, 2000d). For such systems the efficiency with CO_2 capture would equal the current state-of-the-art efficiency for plants without CO_2 capture, that is, 56% LHV.

Integrated Gasification Combined Cycles (IGCC) are large scale, near commercial examples of power systems that can be implemented with heavy oil residues and solid fuels like coal and petroleum coke. For the embryonic coal-fired IGCC technology with the largest unit rated at 331 MW_e, future improvements are expected. A recent study describes improvements potentially realisable for bituminous coals by 2020 that could reduce both energy and cost-of-electricity penalties for CO_2 capture to 13% compared to a same base plant without capture. For such systems the generation efficiency with capture would equal the best efficiency realisable today without CO_2 capture (i.e., 43% LHV; IEA GHG, 2003). Notably, all the innovations considered, with the exception of ion transport membrane technology for air separation (which is motivated by many market drivers other than IGCC needs) involve 'non- breakthrough' technologies, with modest continuing improvements in components that are already established commercially − improvements that might emerge as a natural result of growing commercial experience with IGCC technologies.

All fuel cell types are currently in the development phase. The first demonstration systems are now being tested, with the largest units being at the 1 MW scale. However, it will take at least another 5 to 10 years before these units become commercially available. In the longer term, these highly efficient fuel cell systems are expected to become competitive for power generation. Integrating CO_2 capture in these systems is relatively simple and therefore fuel cell power generation systems offer the prospect of reducing the CO_2 capture penalty in terms of efficiency and capture costs. For instance, for high temperature fuel cell systems without CO_2 capture, efficiencies that exceed 67% are calculated with an anticipated 7% efficiency reduction when CO_2 capture is integrated into the system (Jansen and Dijkstra, 2003). However, fuel cell systems are too small to reach a reasonable level of CO_2 transport cost (IEA GHG, 2002a), but in groups of a total of capacity 100MWe, the cost of CO_2 transport is reduced to a more acceptable level.

Most studies agree that pre-combustion systems may be better suited to implement CO_2 capture at a lower incremental cost compared to the same type of base technology without capture (Section 3.7), but with a key driver affecting implementation being the absolute cost of the carbon emission-free product, or service provided. Pre-combustion systems also have a high strategic importance, because their capability to deliver, in a large scale and at high thermal efficiencies, a suitable mix of electricity, hydrogen and lower carbon-containing fuels or chemical feedstocks in an increasingly carbon-constrained world.

3.6 Environmental, monitoring, risk and legal aspects of capture systems

The previous sections of this chapter focused on each of the major technologies and systems for CO_2 capture. Here we summarize the major environmental, regulatory and risk issues associated with the use of CO_2 capture technology and the handling of carbon dioxide common to all of these systems. Issues related to the subsequent transport and storage of carbon dioxide are discussed in Chapters 4 to 7.

3.6.1 Emissions and resource use impacts of CO_2 capture systems

3.6.1.1 Overview of emissions from capture systems

Plants with CO_2 capture would produce a stream of concentrated CO_2 for storage, plus in most cases a flue gas or vent gas emitted to the atmosphere and liquid wastes. In some cases solid wastes will also be produced.

The captured CO_2 stream may contain impurities which would have practical impacts on CO_2 transport and storage systems and also potential health, safety and environmental impacts. The types and concentrations of impurities depend on the type of capture process, as shown in Table 3.4, and detailed plant design. The major impurities in CO_2 are well known but there is little published information on the fate of any trace impurities in the feed gas such as heavy metals. If substances are captured along with the CO_2 then their net emissions to the atmosphere will be reduced, but impurities in the CO_2 may result in environmental impacts at the storage site.

Table 3.4. Concentrations of impurities in dried CO_2, % by volume (Source data: IEA GHG, 2003; IEA GHG, 2004; IEA GHG, 2005).

CO_2 from most capture processes contains moisture, which has to be removed to avoid corrosion and hydrate formation during transportation. This can be done using conventional processes and the costs of doing so are included in published costs of CO_2 capture plants.

CO_2 from post-combustion solvent scrubbing processes normally contains low concentrations of impurities. Many of the existing post-combustion capture plants produce high purity CO_2 for use in the food industry (IEA GHG, 2004).

CO_2 from pre-combustion physical solvent scrubbing processes typically contains about 1−2% H_2 and CO and traces of H_2S and other sulphur compounds (IEA GHG, 2003). IGCC plants with pre-combustion capture can be designed to produce a combined stream of CO_2 and sulphur compounds, to reduce costs and avoid the production of solid sulphur (IEA GHG, 2003). Combined streams of CO_2 and sulphur compounds (primarily hydrogen sulphide, H_2S) are already stored, for example in Canada, as discussed in Chapter 5. However, this option would only be considered in circumstances where the combined stream could be transported and stored in a safe and environmentally acceptable manner.

The CO_2-rich gas from oxy-fuel processes contains oxygen, nitrogen, argon, sulphur and nitrogen oxides and various other trace impurities. This gas will normally be compressed and fed to a cryogenic purification process to reduce the impurities concentrations to the levels required to avoid two-phase flow conditions in the transportation pipelines. A 99.99% purity could be produced by including distillation in the cryogenic separation unit. Alternatively, the sulphur and nitrogen oxides could be left in the CO_2 fed to storage in circumstances where that is environmentally acceptable as described above for pre-combustion capture and when the total amount of all impurities left in the CO_2 is low enough to avoid two-phase flow conditions in transportation pipelines.

Power plants with CO_2 capture would emit a CO_2-depleted flue gas to the atmosphere. The concentrations of most harmful substances in the flue gas would be similar to or lower than in the flue gas from plants without CO_2 capture, because CO_2 capture processes inherently remove some impurities and some other impurities have to be removed upstream to enable the CO_2 capture process to operate effectively. For example, post-combustion solvent absorption processes require low concentrations of sulphur compounds in the feed gas to avoid excessive solvent loss, but the reduction in the concentration of an impurity may still result in a higher rate of emissions per kWh of product, depending upon the actual amount removed upstream and the capture system energy requirements. As discussed below (Section 3.6.1.2), the latter measure is more relevant for environmental assessments. In the case of post-combustion solvent capture, the flue gas may also contain traces of solvent and ammonia produced by decomposition of solvent.

Some CO_2 capture systems produce solid and liquid wastes. Solvent scrubbing processes produce degraded solvent wastes, which would be incinerated or disposed of by other means. Post-combustion capture processes produce substantially more degraded solvent than pre-combustion capture processes. However, use of novel post-combustion capture solvents can significantly reduce the quantity of waste compared to MEA solvent, as discussed in Section 3.3.2.1. The waste from MEA scrubbing would normally be processed to remove metals and then incinerated. The waste can also be disposed of in cement kilns, where the waste metals become agglomerated in the clinker (IEA GHG, 2004). Pre-combustion capture systems periodically produce spent shift and reforming catalysts ant these would be sent to specialist reprocessing and disposal facilities.

3.6.1.2 Framework for evaluating capture system impacts

As discussed in Chapter 1, the framework used throughout this report to assess the impacts of CO_2 capture and storage is based on the material and energy flows needed to produce a unit of product from a particular process. As seen earlier in this chapter, CO_2 capture systems require an increase in energy use for their operation. As defined in this report (see Section 1.5 and Figure 1.5.1), the energy requirement associated with CO_2 capture is expressed as the additional energy required to produce a unit of useful product, such as a kilowatt-hour of electricity (for the case of a power plant). As the energy and resource requirement for CO_2 capture (which includes the energy needed to compress CO_2 for subsequent transport and storage) is typically much larger than for other emission control systems, it has important implications for plant resource requirements and environmental emissions when viewed from the 'systems' perspective of Figure 1.5.

In general, the CCS energy requirement per unit of product can be expressed in terms of the change in net plant efficiency (η) when the reference plant without capture is equipped with a CCS system:[1]

$$\Delta E = (\eta_{\mathrm{ref}} / \eta_{\mathrm{ccs}}) - 1 \qquad [6]$$

where ΔE is the fractional increase in plant energy input per unit of product and η_{ccs} and η_{ref} are the net efficiencies of the capture plant and reference plant, respectively. The CCS energy requirement directly determines the increases in plant-level resource consumption and environmental burdens associated with producing a unit of useful product (like electricity) while capturing CO_2. In the case of a power plant, the larger the CCS energy requirement, the greater the increases per kilowatt-hour of in-plant fuel consumption and other resource requirements (such as water, chemicals and reagents), as well as environmental releases in the form of solid wastes, liquid wastes and air pollutants not captured by the CCS system. The magnitude of ΔE also determines the magnitude of additional upstream environmental impacts associated with the extraction, storage and transport of additional fuel and other resources consumed at the plant. However, the additional energy for these

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[1] A different measure of the 'energy penalty' commonly reported in the literature is the fractional decrease in plant output (plant derating) for a fixed energy input. This value can be expressed as: $\Delta E^* = 1 - (\eta_{\mathrm{ccs}}/\eta_{\mathrm{ref}})$. Numerically, ΔE^* is smaller than the value of ΔE given by Equation (6). For example, a plant derating of $\Delta E^* = 25\%$ corresponds to an increase in energy input per kWh of $\Delta E = 33\%$.

upstream activities is not normally included in the reported energy requirements for CO_2 capture systems.[2]

Recent literature on CO_2 capture systems applied to electric power plants quantifies the magnitude of CCS energy requirements for a range of proposed new plant designs with and without CO_2 capture. As elaborated later in Section 3.7 (Tables 3.7 to 3.15), those data reveal a wide range of ΔE values. For new supercritical pulverized coal (PC) plants using current technology, these ΔE values range from 24−40%, while for natural gas combined cycle (NGCC) systems the range is 11%–22% and for coal-based gasification combined cycle (IGCC) systems it is 14%–25%. These ranges reflect the combined effects of the base plant efficiency and capture system energy requirements for the same plant type with and without capture.

3.6.1.3 Resource and emission impacts for current systems

Only recently have the environmental and resource implications of CCS energy requirements been discussed and quantified for a variety of current CCS systems. Table 3.5 displays the assumptions and results from a recent comparison of three common fossil fuel power plants employing current technology to capture 90% of the CO_2 produced (Rubin *et al.*, 2005). Increases in specific fuel consumption relative to the reference plant without CO_2 capture correspond directly to the ΔE values defined above. For these three cases, the plant energy requirement per kWh increases by 31% for the PC plant, 16% for the coal-based IGCC plant and 17% for the NGCC plant. For the specific examples used in Table 3.5, the increase in energy consumption for the PC and NGCC plants are in the mid-range of the values for these systems reported later in Tables 3.7 to 3.15 (see also Section 3.6.1.2), whereas the IGCC case is nearer the low end of the reported range for such systems. As a result of the increased energy input per kWh of output, additional resource requirements for the PC plant include proportionally greater amounts of coal, as well as limestone (consumed by the FGD system for SO_2 control) and ammonia (consumed by the SCR system for NO_x control). All three plants additionally require more sorbent make-up for the CO_2 capture units. Table 3.5 also shows the resulting increases in solid residues for these three cases. In contrast, atmospheric emissions of CO_2 decrease sharply as a result of the CCS systems, which also remove residual amounts of other acid gases, especially SO_2 in flue gas streams. Thus, the coal combustion system shows a net reduction in SO_2 emission rate as a result of CO_2 capture. However, because of the reduction in plant efficiency, other air emission rates per kWh increase relative to the reference plants without capture. For the PC and NGCC systems, the increased emissions of ammonia are a result of chemical reactions in the amine-based capture process. Not included in this analysis are the incremental impacts of upstream operations such as mining, processing and transport of fuels and other resources. Other studies, however, indicate that these impacts, while not insignificant, tend to be small relative to plant-level impacts (Bock *et al.*, 2003).

Table 3.5. Illustrative impacts of CCS energy requirements on plant-level resource consumption and non-CO_2 emission rates for three current power plant systems. Values shown are mass flow rates per MWh for the capture plant, plus increases over the reference plant rates for the same plant type. See footnotes for additional details. (Source: Rubin *et al.*, 2005)

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[2] Those additional energy requirements, if quantified, could be included by re-defining the system boundary and system efficiency terms in Equation (6) to apply to the full life cycle, rather than only the power plant. Such an analysis would require additional assumptions about the methods of fuel extraction, processing, transport to the power plant, and the associated energy requirements of those activities; as well as the CO_2 losses incurred during storage.

For the most part, the magnitude of impacts noted above − especially impacts on fuel use and solid waste production − is directly proportional to the increased energy per kWh resulting from the reduction in plant efficiency, as indicated by Equation (6). Because CCS energy requirements are one to two orders of magnitude greater than for other power plant emission control technologies (such as particulate collectors and flue gas desulphurization systems), the illustrative results above emphasize the importance of maximizing overall plant efficiency while controlling environmental emissions.

3.6.1.4 Resource and emission impacts of future systems

The analysis above compared the impacts of CO_2 capture for a given plant type based on current technology. The magnitude of actual future impacts, however, will depend on four important factors: (1) the performance of technologies available at the time capture systems are deployed; (2) the type of power plants and capture systems actually put into service; (3) the total capacity of each plant type that is deployed; and, (4) the characteristics and capacity of plants they may be replacing.

Analyses of both current and near-future post-combustion, pre-combustion and oxy-fuel combustion capture technology options reveal that some of the advanced systems currently under development promise to significantly reduce the capture energy requirements − and associated impacts − while still reducing CO_2 emissions by 90% or more, as shown in Figure 3.19. Data in this figure was derived from the studies previously reported in Figures 3.6 and 3.7.

Figure 3.19. Fuel use for a reduction of CO_2 emissions from capture plants (data presented from design studies for power plants with and without capture shown in Figures 3.6 and 3.7).

The timetable for deploying more efficient plants with CO_2 capture will be the key determinant of actual environmental changes. If a new plant with capture replaces an older, less efficient and higher-emitting plant currently in service, the net change in plant-level emission impacts and resource requirements would be much smaller than the values given earlier (which compared identical new plants with and without capture). For example, the efficiency of a modern coal-based plant with capture is close to many older coal-burning plants currently in service. Replacing the latter with the former would thus reduce CO_2 emissions significantly with little or no net change in plant coal consumption or related solid waste impacts. In some cases, there could in fact be net reductions in other plant emissions, in support of clean air goals. If, however, the deployment of new CCS plants is delayed significantly, older existing plants could well be replaced by modern high-efficiency plants without capture. Such plants also would be built to provide additional capacity in regions with high electricity growth rates, such as in China and other parts of Asia today. A decade or two from now, the fleet of ‘existing’ plants in those regions would thus look very different from the present. Accordingly, the environmental and resource impacts of additional new plants with CO_2 capture would have to be assessed in the context of the future situation.

Because comparisons of different plant types require a specific context (or scenario) to be meaningful, this chapter has only focused on characterizing the effects of CO_2 capture systems relative to the same type of power plant and not the type of infrastructure it would replace (either currently, or in a future carbon-constrained world). If other systems such as the use of renewable energy, or electricity and synfuels cogenerated from coal, find significant applications, those systems too would require more comprehensive comparative life-cycle assessments of resource use and impacts that are not currently available. Chapter 8, however, assesses overall energy use impacts for illustrative scenarios of CCS deployment in competition with other carbon mitigation options.

3.6.2 Issues related to the classification of carbon dioxide as a product

As a current commercial product, carbon dioxide is subject to classification and regulations. The classification of carbon dioxide is dependent on its physical state (gas, liquid or solid), its concentration, impurities present and other criteria established by national legislative classification in different regions of the world. During the capture and concentration process, the quality properties can change the classification of the substance. A detailed assessment of carbon dioxide physical and chemical properties is provided in Appendix I.

The environmental, monitoring, risk and legal aspects associated with carbon dioxide handling and storage are well established in the processing industry. However, much larger volumes are targeted for carbon dioxide processing for purposes of CCS than the volumes handled at present. On a local and regional level, additional emergency response and other regulatory measures can be expected in the future, depending on the rate of development of CCS. It is anticipated that human capacity will be developed to assess the monitoring, risk and legal aspects as required by the market.

At present, carbon dioxide typically occurs and is mainly traded as a non-flammable gas (US Department of Transportation classification class 2.2). The classification system of Transport Dangerous Goods, International Maritime Organization / International Maritime Dangerous Goods and International Civil Aviation Organization / International Air Transport Association, all classify carbon dioxide in class 2.2, non-flammable, non-corrosive and non-poisonous gases. In US federal regulations, carbon dioxide is not listed as a product in the Clean Water Act (CWA 307 and 311), Clean Air Act (CAA 112) or the Toxics Release Inventory. In other international regulations carbon dioxide is not classified in the European Inventory of Existing Commercial Chemical Substance or other international lists, but in Canada is classified as a compressed gas (class A) on the Canadian Energy Pipeline Association Dangerous Substances List (Hazardous Substances Data Bank, 2002). The physical-chemical properties of carbon dioxide are described in detail in Appendix I.

3.6.3 Health and safety risks associated with carbon dioxide processing

The effects of exposure to carbon dioxide are described in Appendix 1. However, a risk assessment that includes an understanding of both exposure and effects is required to characterize the risk for various situations associated with carbon dioxide processing (European Chemicals Bureau, 2003); see the following two sections for established risk management practices. The most probable routes of human exposure to carbon dioxide are inhalation or skin contact. The need for a risk-based approach is clear from the following two descriptions. Carbon dioxide and its products of degradation are not legally classified as a toxic substance; is non-hazardous on inhalation, is a non-irritant and does not sensitize or permeate the skin. However, chronic effects on humans follow from long-term exposure to airborne carbon dioxide concentrations of between 0.5 and 1% resulting in metabolic acidosis and increased calcium deposits in soft tissues. The substance is toxic to the cardiovascular system and upper respiratory tract at concentrations above 3%. Sensitive populations to elevated carbon dioxide levels are described in Appendix 1. The product risk assessment process is therefore necessary as with any other chemical use to determine the risk and establish the necessary risk management processes.

As an asphyxiate carbon dioxide presents the greatest danger. If atmospheric oxygen is displaced such that oxygen concentration is 15−16%, signs of asphyxia will be noted. Skin contact with dry ice has caused serious frostbites and blisters (Hazardous Substances Data Bank, 2002). Protective equipment and clothing required in the processing industries include full face-piece respirators to prevent eye contact and appropriate personal protective clothing to protect the skin from becoming frozen by the liquid.

3.6.4 Plant design principles and guidelines used by governments, industries and financiers

New plant facilities like those envisioned for carbon dioxide are subject to design guidelines for the petrochemical industry as determined by relevant authorities. One example is the European Unions' Integrated Pollution Prevention and Control (IPPC) directive requiring the application of the principles of Best Available Technology Not Entailing Excessive Cost (BATNEEC). Carbon dioxide capture and compression processes are listed in several guidelines as gas-processing facilities. Typically the World Bank guidelines and other financial institutions have specific requirements to reduce risk and these require monitoring (World Bank, 1999) which is part of routine plant monitoring to detect accidental releases. Investor guidelines like the World Bank guidelines are particularly important for developing countries where there is less emphasis on monitoring and legislation. National and regional legislation for plant design and specifications from organizations like the US Environmental Protection Agency are available to guide the development of technology.

3.6.5 Commissioning, good practice during operations and sound management of chemicals

The routine engineering design, commissioning and start-up activities associated with petrochemical facilities are applicable to the capture and compression of carbon dioxide; for example Hazard Operability studies are conducted on a routine basis for new facilities (Sikdar and Diwekar, 1999).

The management of carbon dioxide and reagents inside factory battery limits will be in accordance with the relevant practices in use for carbon dioxide. For carbon dioxide, US Occupational Health and Safety Act standards and National Institute for Occupational Safety and Health recommendations exist, which are applied widely in industry to guide safe handling of carbon dioxide and the same applies to reagents and catalysts used. Well established and externally audited management systems such as International Standards Organization's ISO 14001 (environment) and ISO 9001 (quality) and Occupational Health and Safety (OHSAS 18000) exist to provide assurance that environment, safety, health and quality management systems are in place (American Institute of Chemical Engineers, 1995). Tools like life-cycle assessment (ISO 14040 series) with the necessary boundary expansion methodology are useful to determine the overall issues associated with a facility and assist with selection of parameters such as energy carriers, operational conditions and materials used in the process. The life-cycle assessment will also indicate if a trouble-free capture system does generate environmental concerns elsewhere in the product life cycle.

3.6.6 Site closure and remediation

It is not anticipated that carbon dioxide capture will result in a legacy of polluted sites requiring remediation after plant closure, assuming that standard operating procedures and management practices in the previous section are followed. However, depending on the technology used and the materials procured for operations, waste disposal at the facilities and operation according to a formal management system from construction, operation to the development of site closure plans will largely assist to reduce the risk of a polluted site after closure of operations.

3.7 Cost of CO_2 capture

This section of the report deals with the critical issue of CO_2 capture costs. We begin with an overview of the many factors that affect costs and the ability to compare published estimates on a consistent basis. Different measures of CO_2 capture cost also are presented and discussed. The literature on CO_2 capture costs for currently available technologies is then reviewed, along with the outlook for future costs over the next several decades.

3.7.1 Factors affecting CO_2 capture cost

Published estimates for CO_2 capture costs vary widely, mainly as a result of different assumptions regarding technical factors related to plant design and operation (e.g., plant size, net efficiency, fuel properties and load factor), as well as key economic and financial factors such as fuel cost, interest rates and plant lifetime. A number of recent papers have addressed this issue and identified the principal sources of cost differences and variability (Herzog, 1999; Simbeck, 1999; Rubin and Rao, 2003). This section draws heavily on Rubin and Rao (2003) to highlight the major factors affecting the cost of CO_2 capture.

3.7.1.1 Defining the technology of interest

Costs will vary with the choice of CO_2 capture technology and the choice of power system or industrial process that generates the CO_2 emissions. In engineering-economic studies of a single plant or CO_2 capture technology, such definitions are usually clear. However, where larger systems are being analyzed, such as in regional, national or global studies of CO_2 mitigation options, the specific technologies assumed for CO_2 production and capture may be unclear or unspecified. In such cases, the context for reported cost results also may be unclear.

3.7.1.2 Defining the system boundary

Any economic assessment should clearly define the ‘system’ whose CO_2 emissions and cost is being characterized. The most common assumption in studies of CO_2 capture is a single facility (most often a power plant) that captures CO_2 and transports it to an off-site storage area such as a geologic formation. The CO_2 emissions considered are those released at the facility before and after capture. Reported costs may or may not include CO_2 transport and storage costs. *The system boundary of interest in this section of the report includes only the power plant or other process of interest and does not include CO_2 transport and storage systems, whose costs are presented in later chapters. CO_2 compression, however, is assumed to occur within the facility boundary and therefore the cost of compression is included in the cost of capture.*[3]

In some studies the system boundary includes emissions of CO_2 and other greenhouse gases such as methane (expressed as equivalent CO_2) over the complete fuel cycle encompassing not only the power plant or facility in question, but also the ‘upstream’ processes of extraction, refining and transport of fuel used at the facility, plus any ‘downstream’ emissions from the use or storage of captured CO_2. Still larger system boundaries might include all power plants in a utility company’s system; all plants in a regional or national grid; or a national economy where power plant and industrial emissions are but one element of the overall energy system being modelled. In each of these cases it is possible to derive a mitigation cost for CO_2, but the results are not directly comparable because they reflect different system boundaries and considerations. Chapter 8 discusses such differences in more detail and presents results for alternative systems of interest.

3.7.1.3 Defining the technology time frame and maturity

Another factor that is often unclear in economic evaluations of CO_2 capture is the assumed time frame and/or level of maturity for the technology under study. Does the cost estimate apply to a facility that would be built today, or at some future time? This is especially problematic in studies of ‘advanced’ technologies that are still under development and not currently commercial. In most cases, studies of advanced technologies assume that costs apply to an ‘nth plant’ to be built

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[3] Alternatively, compression costs could be attributed wholly or in part to CO_2 transport and storage. Most studies, however, include compression with capture cost. This also facilitates comparisons of capture technologies that operate at different pressures, and thus incur different costs to achieve a specified final pressure.

sometime in the future when the technology is mature. Such estimates reflect the expected benefits of technological learning, but may or may not adequately account for the increased costs that typically occur in the early stages of commercialization. The choice of technology time frame and assumed rate of cost improvements and can therefore make a big difference in CO_2 capture cost estimates.

3.7.1.4 Different cost measures and assumptions

The literature reveals a number of different measures used to characterize CO_2 capture and storage costs, including capital cost, cost of electricity, cost of CO_2 avoided and others. Because some of these measures are reported in the same units (e.g., US dollars per tonne of CO_2) there is great potential for misunderstanding. Furthermore, for any given cost measure, different assumptions about the technical, economic and financial parameters used in cost calculations can also give rise to large differences in reported capture costs. Section 3.7.2 elaborates on some of the common metrics of cost and the parameters they employ.

3.7.2 Measures of CO_2 capture cost

We define four common measures of CO_2 capture cost here: capital cost, incremental product cost (such as the cost of electricity), cost of CO_2 avoided and cost of CO_2 captured or removed. Each of these measures provides a different perspective on CO_2 capture cost for a particular technology or system of interest. All of them, however, represent an 'engineering economic' perspective showing the added cost of capturing CO_2 in a particular application. Such measures are required to address larger questions such as which options or strategies to pursue − a topic addressed later in Chapter 8.

3.7.2.1 Capital cost

Capital cost (also known as investment cost or first cost) is a widely used, albeit incomplete, metric of the cost of a technology. It is often reported on a normalized basis (e.g., cost per kW). For CO_2 capture systems, the capital cost is generally assumed to represent the total expenditure required to design, purchase and install the system of interest. It may also include the additional costs of other plant components not needed in the absence of a CO_2 capture device, such as the costs of an upstream gas purification system to protect the capture device. Such costs often arise in complex facilities like a power plant. Thus, the total incremental cost of CO_2 capture for a given plant design is best determined as the difference in total cost between plants with and without CO_2 capture, producing the same amounts of useful (primary) product, such as electricity.

Different organizations employ different systems of accounts to specify the elements of a capital cost estimate. For electric power plants, one widely used procedure is that defined by the Electric Power Research Institute (EPRI) (EPRI, 1993). However, because there is no universally employed nomenclature or system of accounts, capital costs reported by different organizations or authors may not always include the same items. The terms used to report capital costs may further disguise such differences and lead to misunderstandings about what is and is not included. For example, power plant cost studies often report a value of capital cost that does not include the cost of interest during construction or other so-called 'owners costs' that typically add at least 10−20% (sometimes substantially more) to the 'total capital requirement' of a system. Only if a capital cost breakdown is reported can such omissions be discovered. Studies that fail to report the year of a cost estimate introduce further uncertainty that may affect cost comparisons.

3.7.2.2 Incremental product cost

The effect of CO_2 capture on the cost of electricity (or other product) is one of the most important measures of economic impact. Electric power plants, a major source of CO_2 emissions, are of particular interest in this regard. The cost electricity (COE) for a power plant can be calculated as:[4]

$$COE = [(TCR)(FCF) + (FOM)]/[(CF)(8760)(kW)] + VOM + (HR)(FC) \qquad [7]$$

where, COE = levelized cost of electricity (US\$ kWh^{-1}), TCR = total capital requirement (US\$), FCF = fixed charge factor (fraction yr^{-1}), FOM = fixed operating costs (US\$ yr^{-1}), VOM = variable operating costs (US\$ kWh^{-1}), HR = net plant heat rate (kJ kWh^{-1}), FC = unit fuel cost (US\$ kJ^{-1}), CF = capacity factor (fraction), 8760 = total hours in a typical year and kW = net plant power (kW). In this chapter, the costs in Equation (7) include only the power plant and capture technologies and not the additional costs of CO_2 transport and storage that are required for a complete system with CCS. The incremental COE is the difference in electricity cost with and without CO_2 capture.[5] Again, the values reported here exclude transport and storage costs. Full CCS costs are reported in Chapter 8.

Equation (7) shows that many factors affect this incremental cost. For example, just as the total capital cost includes many different items, so too do the fixed and variable costs associated with plant operation and maintenance (O&M). Similarly, the fixed charge factor (FCF, also known as the capital recovery factor) reflects assumptions about the plant lifetime and the effective interest rate (or discount rate) used to amortize capital costs.[6] Assumptions about any of the factors in Equation (7) can have a pronounced effect on overall cost results. Nor are these factors all independent of one another. For example, the design heat rate of a new power plant may affect the total capital requirement since high-efficiency plants usually are more costly than lower-efficiency designs.

Finally, because several of the parameter values in Equation (7) may change over the operating life of a facility (such as the capacity factor, unit fuel cost, or variable operating costs), the value of COE also may vary from year to year. To include such effects, an economic evaluation would calculate the net present value (NPV) of discounted costs based on a schedule of year-to-year cost variations, in lieu of the simpler formulation of Equation (7). However, most engineering-economic studies use Equation (7) to calculate a single value of 'levelized' COE over the assumed life of the plant. The levelized COE is the cost of electricity, which, if sustained over the operating life of the plant, would produce the same NPV as an assumed stream of variable year-to-year costs. In most economic studies of CO_2 capture, however, all parameter values in Equation (7) are held constant, reflecting (either implicitly or explicitly) a levelized COE over the life of the plant.[7]

3.7.2.3 Cost of CO_2 avoided

One of the most widely used measures for the cost of CO_2 capture and storage is the 'cost of CO_2 avoided.' This value reflects the average cost of reducing atmospheric CO_2 mass emissions by one unit while providing the same amount of useful product as a 'reference plant' without CCS. For an electric power plant the avoidance cost can be defined as:

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[4] For simplicity, the value of FCF in Equation (7) is applied to the total capital requirement. More detailed calculations of COE based on a year-by-year analysis apply the FCF to the total capital cost excluding owner's costs (such as interest during construction), which are separately accounted for in the years prior to plant start-up.

[5] For CO_2 capture systems with large auxiliary energy requirements, the magnitude of incremental cost also depends on whether the plant with capture is assumed to be a larger facility producing the same net output as the reference plant without capture, or whether the reference plant is simply derated to supply the auxiliary energy. While the latter assumption is most common, the former yields a smaller incremental cost due to economy-of-scale effects.

[6] In its simplest form, FCF can be calculated from the project lifetime, n (years), and annual interest rate, i (fraction), by the equation: $FCF = i / [1 - (1 + i)^{-n}]$.

[7] Readers not familiar with these economic concepts and calculations may wish to consult a basic economics text, or references such as (EPRI, 1993) or (Rubin, 2001) for more details.

Cost of CO_2 Avoided (US\$/tCO_2) = $[(COE)_{capture} - (COE)_{ref}] / [(CO_2\ kWh^{-1})_{ref} - (CO_2\ kWh^{-1})_{capture}]$ [8]

where, COE = levelized cost of electricity (US\$ kWh^{-1}) as given by Equation (7) and $CO_2\ kWh^{-1}$ = CO_2 mass emission rate (in tonnes) per kWh generated, based on the net plant capacity for each case. The subscripts 'capture' and 'ref' refer to the plant with and without CO_2 capture, respectively. Note that while this equation is commonly used to report a cost of CO_2 avoided for the capture portion of a full CCS system, strictly speaking it should be applied only to a complete CCS system including transport and storage costs (since all elements are required to avoid emissions to the atmosphere).

The choice of the reference plant without CO_2 capture plays a key role in determining the CO_2 avoidance cost. *Here the reference plant is assumed to be a plant of the same type and design as the plant with CO_2 capture.* This provides a consistent basis for reporting the incremental cost of CO_2 capture for a particular type of facility.

Using Equation (8), a cost of CO_2 avoided can be calculated for any two plant types, or any two aggregates of plants. Thus, special care should be taken to ensure that the basis for a reported cost of CO_2 avoided is clearly understood or conveyed. For example, the avoidance cost is sometimes taken as a measure of the cost to society of reducing GHG emissions.[8] In that case, the cost per tonne of CO_2 avoided reflects the average cost of moving from one situation (e.g., the current mix of power generation fuels and technologies) to a different mix of technologies having lower overall emissions. Alternatively, some studies compare individual plants with and without capture (as we do), but assume different types of plants for the two cases. Such studies, for example, might compare a coal-fired plant with capture to an NGCC reference plant without capture. Such cases reflect a different choice of system boundaries and address very different questions, than those addressed here. However, the data presented in this section (comparing the same type of plant with and without capture) can be used to estimate a cost of CO_2 avoided for any two of the systems of interest in a particular situation (see Chapter 8).

3.7.2.4 *Cost of CO_2 captured or removed*

Another cost measure frequently reported in the literature is based on the mass of CO_2 captured (or removed) rather than emissions avoided. For an electric power plant it can be defined as:

Cost of CO_2 Captured (US\$/tCO_2) = $[(COE)_{capture} - (COE)_{ref}] / (CO_{2,\ captured}\ kWh^{-1})$ [9]

where, $CO_{2,\ captured}\ kWh^{-1}$ = total mass of CO_2 captured (in tonnes) per net kWh for the plant with capture. This measure reflects the economic viability of a CO_2 capture system given a market price for CO_2 (as an industrial commodity). If the CO_2 captured at a power plant can be sold at this price (e.g., to the food industry, or for enhanced oil recovery), the COE for the plant with capture would be the same as for the reference plant having higher CO_2 emissions. Numerically, the cost of CO_2 captured is lower than the cost of CO_2 avoided because the energy required to operate the CO_2 capture systems increases the amount of CO_2 emitted per unit of product.

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[8] As used here, 'cost' refers only to money spent for technology, fuels and related materials, and not to broader societal measures such as macroeconomic costs or societal damage costs associated with atmospheric emissions. Further discussions and use of the term 'cost of CO_2 avoided' appear in Chapter 8 and in the references cited earlier.

3.7.2.5 Importance of CCS energy requirements

As the energy requirement for CCS is substantially larger than for other emission control systems, it has important implications for plant economics as well as for resource requirements and environmental impacts. The energy 'penalty' (as it is often called) enters cost calculations in one of two ways. Most commonly, all energy needed to operate CCS absorbers, compressors, pumps and other equipment is assumed to be provided within the plant boundary, thus lowering the net plant capacity (kW) and output (kWh, in the case of a power plant). The result, as shown by Equation (7), is a higher unit capital cost (US\$ kW^{-1}) and a higher cost of electricity production (US\$ kWh^{-1}). Effectively, these higher unit costs reflect the expense of building and operating the incremental capacity needed to operate the CCS system.

Alternatively, some studies − particularly for industrial processes such as hydrogen production − assume that some or all of the energy needed to operate the CCS system is purchased from outside the plant boundary at some assumed price. Still other studies assume that new equipment is installed to generate auxiliary energy on-site. In these cases, the net plant capacity and output may or may not change and may even increase. However, the COE in Equation (7) again will rise due to the increases in VOM costs (for purchased energy) and (if applicable) capital costs for additional equipment. The assumption of purchased power, however, does not guarantee a full accounting of the replacement costs or CO_2 emissions associated with CCS. In all cases, however, the larger the CCS energy requirement, the greater the difference between the costs of CO_2 captured and avoided.

3.7.2.6 Other measures of cost

The cost measures above characterize the expense of adding CO_2 capture to a single plant of a given type and operating profile. A broader modelling framework is needed to address questions involving multiple plants (e.g., a utility system, regional grid, or national network), or decisions about what type of plant to build (and when). Macroeconomic models that include emission control costs as elements of a more complex framework typically yield cost measures such as the change in gross domestic product (GDP) from the imposition of a carbon constraint, along with changes in the average cost of electricity and cost per tonne of CO_2 abated. Such measures are often useful for policy analysis, but reflect many additional assumptions about the structure of an economy as well as the cost of technology. Chapter 8 provides a discussion of macroeconomic modelling as it relates to CO_2 capture costs.

3.7.3 The context for current cost estimates

Recall that CO_2 capture, while practiced today in some industrial applications, is not currently a commercial technology used at large electric power plants, which are the focus of most CCS studies. Thus, cost estimates for CO_2 capture systems rely mainly on studies of hypothetical plants. Published studies also differ significantly in the assumptions used for cost estimation. Equation (7), for example, shows that the plant capacity factor has a major impact on the cost of electric power generation, as do the plant lifetime and discount rate used to compute the fixed charge factor. The COE, in turn, is a key element of CO_2 avoidance cost, Equation (8). Thus, a high plant capacity factor or a low fixed charge rate will lower the cost of CO_2 capture per kWh. The choice of other important parameters, such as the plant size, efficiency, fuel type and CO_2 removal rate will similarly affect the CO_2 capture cost. Less apparent, but often equally important, are assumptions about parameters such as the 'contingency cost factors' embedded in capital cost estimates to account for unspecified costs anticipated for technologies at an early stage of development, or for commercial systems that have not yet been demonstrated for the application, location, or plant scale under study.

Because of the variability of assumptions employed in different studies of CO_2 capture, a systematic comparison of cost results is not straightforward (or even possible in most cases). Moreover, there is no universally 'correct' set of assumptions that apply to all the parameters affecting CO_2 capture cost. For example, the quality and cost of natural gas or coal delivered to power plants in Europe and the United States may differ markedly. Similarly, the cost of capital for a municipal or government-owned utility may be significantly lower than for a privately-owned utility operating in a competitive market. These and other factors lead to real differences in CO_2 capture costs for a given technology or power generation system. Thus, we seek in this report to elucidate the key assumptions employed in different studies of similar systems and technologies and their resulting impact on the cost of CO_2 capture. Analyses comparing the costs of alternative systems on an internally consistent basis (within a particular study) also are highlighted. Nor are all studies equally credible, considering their vintage, data sources, level of detail and extent of peer review. Thus, the approach adopted here is to rely as much as possible on recent peer-reviewed literature, together with other publicly-available studies by governmental and private organizations heavily involved in the field of CO_2 capture. Later, in Chapter 8, the range of capture costs reported here are combined with cost estimates for CO_2 transport and storage to arrive at estimates of the overall cost of CCS for selected power systems and industrial processes.

3.7.4 Overview of technologies and systems evaluated

Economic studies of CO_2 capture have focused mainly on electric power generation, a major source of CO_2 emissions. To a lesser extent, CO_2 capture from industrial processes also has been subject to economic evaluations, especially processes producing hydrogen, often in combination with other products.

The sections below review and summarize recent estimates of CO_2 capture costs for major systems of interest. Sections 3.7.4 to 3.7.7 focus first on the cost of current CO_2 capture technologies, while Sections 3.7.8 to 3.7.11 go on to discuss improved or 'advanced' technologies promising lower costs in the future. In all cases the system boundary is defined as a single facility at which CO_2 is captured and compressed for delivery to a transport and storage system. To reflect different levels of confidence (or uncertainty) in cost estimates for technologies at different stages of development, the qualitative descriptors shown in Table 3.6 are applied in summarizing published cost estimates.[9] The studies reviewed typically report costs in US dollars for reference years ranging from 2000 to early 2004. Because inflation effects generally have been small during this period no adjustments have been made in summarizing ranges of reported costs.

Table 3.6. Confidence levels for technology and system cost estimates.

3.7.5 Post-combustion CO_2 capture cost for electric power plants (current technology)

Most of the world's electricity is currently generated from the combustion of fossil fuels, especially coal and (to an increasing extent) natural gas. Hence, the ability to capture and sequester the CO_2 emitted by such plants has been a major focus of investigation. This section of the report focuses on the cost of currently available technology for CO_2 capture. Because of the relatively low CO_2 concentration in power plant flue gases, chemical absorption systems have been the dominant technology of interest for post-combustion capture (see Section 3.3.2). However, the cost of CO_2 capture depends not only on the choice of capture technology, but also − and often more

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[9] These descriptions are used in subsequent tables to characterize systems with CO_2 capture. In most cases the cost estimates for reference plants (without capture) would rank as high (e.g., IGCC power plants) or very high (e.g., PC and NGCC power plants).

importantly − on the characteristics and design of the overall power plant. For purposes of cost reporting, we distinguish between coal-fired and gas-fired plant designs and between new and existing facilities.

3.7.5.1 New coal-fired power plants

Table 3.7 summarizes the key assumptions and results of recent studies of post-combustion CO_2 capture at new coal-fired power plants. Assumed plant sizes with CO_2 capture range from approximately 300−700 MW net power output. In all cases, CO_2 capture is accomplished using an amine-based absorption system, typically MEA. Capture efficiencies range from 85−95% with the most common value being 90%. The studies employ different assumptions about other key parameters such as the base power plant efficiency, coal properties, coal cost, plant capacity factor, CO_2 product pressure and financial parameters such as the fixed charge factor. All of these factors have a direct influence on total plant cost and the cost of CO_2 capture.

Table 3.7. CO_2 capture costs: New pulverized-coal power plants using current technology.

Table 3.7 summarizes several measures of CO_2 capture cost, both in absolute and relative terms. Across the full set of studies, CO_2 capture adds 44−87% to the capital cost of the reference plant (US\$ kW^{-1}) and 42−81% to the cost of electricity (US\$ MWh^{-1}), while achieving CO_2 reductions of approximately 80−90% per net kWh produced. The cost of CO_2 avoided for these cases varies from 29−51 US\$/tCO_2. The absolute values of capital cost, COE and incremental cost of electricity in Table 3.7 reflect the different assumptions employed in each study. The result is an incremental COE of 18−38 US\$ MWh^{-1} (or US\$ 0.018−0.038 kWh^{-1}) for CO_2 capture. The total COE for plants with capture ranges from 62−87 US\$ MWh^{-1}. In all cases, a significant portion of the total CO_2 capture cost is due to the energy requirement for CO_2 capture and compression. For the studies in Table 3.7, the plants with CO_2 capture require 24−42% more fuel input per MWh of plant output relative to a similar reference plant without capture. Roughly half the energy is required for solvent regeneration and a third for CO_2 compression.

While many factors contribute to the cost differences observed in Table 3.7, systematic studies of the influence of different factors indicate that the most important sources of variability in reported cost results are assumptions about the CO_2 capture system energy requirement, power plant efficiency, fuel type, plant capacity factor and fixed charge rate (Rao and Rubin, 2002). In this regard, it is useful to note that the lowest-cost capture systems in Table 3.7 (in terms of COE and cost of CO_2 avoided) come from a recent study (IEA GHG, 2004) that combines an efficient supercritical power plant design using bituminous coal, with high plant utilization, lowest fixed charge rate and more energy-efficient amine system designs, as recently announced by two major vendors (but not yet demonstrated on coal-fired power plants). In contrast, the highest reported COE values are for less efficient subcritical plant designs using low rank coal, combined with lower capacity factors, higher fixed charge rates and employing amine system designs typical of units currently in operation at small power plants.

Recent increases in world coal prices, if sustained, also would affect the levelized COE values reported here. Based on one recent study (IEA GHG, 2004), each 1.00 US\$ GJ^{-1} increase in coal price would increase the COE by 8.2 US\$ MWh^{-1} for a new PC plant without capture and by 10.1 US\$ MWh^{-1} for a plant with capture.

These results indicate that new power plants equipped with CO_2 capture are likely to be high-efficiency supercritical units, which yield lowest overall costs. The worldwide use of supercritical units (without capture) with current usage at 155 GWe (Section 3.1.2.2), is rapidly increasing in several regions of the world and, as seen in Table 3.7, the preponderance of recent studies of CO_2 capture are based on supercritical units using bituminous coals. For these plants, Table 3.7 shows that capture systems increase the capital cost by 44−74% and the COE by 42−66% (18−34 US$ MWh^{-1}). The major factors contributing to these ranges were differences in plant size, capacity factor and fixed charge factor. New or improved capture systems and power plant designs that promise to further reduce the costs of CO_2 capture are discussed later in Section 3.7.7. First, however, we examine CO_2 capture costs at existing plants.

3.7.5.2 Existing coal-fired plants

Compared to the study of new plants, CO_2 capture options for existing power plants have received relatively little study to date. Table 3.8 summarizes the assumptions and results of several studies estimating the cost of retrofitting an amine-based CO_2 capture system to an existing coal-fired power plant. Several factors significantly affect the economics of retrofits, especially the age, smaller sizes and lower efficiencies typical of existing plants relative to new builds. The energy requirement for CO_2 capture also is usually higher because of less efficient heat integration for sorbent regeneration. All of these factors lead to higher overall costs. Existing plants not yet equipped with a flue gas desulphurization (FGD) system for SO_2 control also must be retrofitted or upgraded for high-efficiency sulphur capture in addition to the CO_2 capture device. For plants with high NO_x levels, a NO_2 removal system also may be required to minimize solvent loss from reactions with acid gases. Finally, site-specific difficulties, such as land availability, access to plant areas and the need for special ductwork, tend to further increase the capital cost of any retrofit project relative to an equivalent new plant installation. Nonetheless, in cases where the capital cost of the existing plant has been fully or substantially amortized, Table 3.8 shows that the COE of a retrofitted plant with capture (including all new capital requirements) can be comparable to or lower than that of a new plant, although the incremental COE is typically higher because of the factors noted above.

Table 3.8. CO_2 capture costs: Existing pulverized-coal power plants using current technology.

Table 3.8 further shows that for comparable levels of about 85% CO_2 reduction per kWh, the average cost of CO_2 avoided for retrofits is about 35% higher than for the new plants analyzed in Table 3.7. The incremental capital cost and COE depend strongly on site-specific assumptions, including the degree of amortization and options for providing process energy needs. As with new plants, heat and power for CO_2 capture are usually assumed to be provided by the base (reference) plant, resulting in a sizeable (30 to 40%) plant output reduction. Other studies assume that an auxiliary gas-fired boiler is constructed to provide the CO_2 capture steam requirements and (in some cases) additional power. Low natural gas prices can make this option more attractive than plant output reduction (based on COE), but such systems yield lower CO_2 reductions (around 60%) since the emissions from natural gas combustion are typically not captured. For this reason, the avoided cost values for this option are not directly comparable to those with higher CO_2 reductions.

Also reflected in Table 3.8 is the option of rebuilding an existing boiler and steam turbine as a supercritical unit to gain efficiency improvements in conjunction with CO_2 capture. One recent study (Gibbins *et al.*, 2005) suggests this option could be economically attractive in conjunction with CO_2 capture since the more efficient unit minimizes the cost of capture and yields a greater net power output and a lower COE compared to a simple retrofit. The use of a new and less energy-intensive capture unit yields further cost reductions in this study. Another recent study similarly concluded that the most economical approach to CO_2 capture for an existing coal-fired plant was to combine CO_2 capture with repowering the unit with an ultra-supercritical steam system (Simbeck, 2004). One additional option, repowering an existing unit with a coal gasifier, is discussed later in Section 3.7.5.2.

3.7.5.3 Natural gas-fired power plants

Power plants fuelled by natural gas may include gas-fired boilers, simple-cycle gas turbines, or natural gas combined cycle (NGCC) units. The current operating capacity in use globally is 333 GWe for gas-fired boilers, 214 GWe for simple cycle gas turbines and 339 GWe for NGCC (IEA WEO, 2004). The absence of sulphur and other impurities in natural gas reduces the capital costs associated with auxiliary flue gas clean-up systems required for amine-based CO_2 capture technology. On the other hand, the lower concentration of CO_2 in gas-fired units tends to increase the cost per tonne of CO_2 captured or avoided relative to coal-fired units.

Table 3.9 summarizes the assumptions and cost results of several recent studies of CO_2 capture at gas-fired combined cycle power plants ranging in size from approximately 300−700 MW. Relative to reference plants without capture, to achieve net CO_2 reductions (per kWh) of the order of 83−88%, the capital cost per kW increases by 64−100%, while the COE increases by 37−69%, or by 12−24 US\$ MWh^{-1} on an absolute basis. The corresponding cost of CO_2 avoided ranges from 37−74 US\$/tCO_2, while the CCS energy requirement increases plant fuel consumption per kWh by 11−22%.

Table 3.9. CO_2 capture costs: Natural gas-fired power plants using current technology.

As seen earlier in Equations (7) to (9), assumptions about the plant fuel cost have an especially important influence on the COE for gas-fired plants because the contribution of capital costs is relatively low compared to coal plants. The studies in Table 3.9 assume stable gas prices of 2.82−4.44 US\$ GJ^{-1} (LHV basis) over the life of the plant, together with high capacity factors (65−95%) representing base load operation. These assumptions result in relatively low values of COE for both the reference plant and capture plant. Since about 2002, however, natural gas prices have increased significantly in many parts of the world, which has also affected the outlook for future prices. Based on the assumptions of one recent study (IEA GHG, 2004), the COE for an NGCC plant without capture would increase by 6.8 US\$ MWh^{-1} for each 1.00 US\$ GJ^{-1} increase in natural gas price (assuming no change in plant utilization or other factors of production). An NGCC plant with CCS would see a slightly higher increase of 7.3 US\$ MWh^{-1}. The price of natural gas, and its relation to the price of competing fuels like coal, is an important determinant of which type of power plant will provide the lowest cost electricity in the context of a particular situation. However, across a twofold increase in gas price (from 3−6 US\$ GJ^{-1}), the incremental cost of CO_2 capture changed by only 2 US\$ MWh^{-1} (US\$ 0.002 kWh^{-1}) with all other factors held constant.

In countries like the US, higher gas prices have also resulted in lower utilization rates (averaging 30−50%) for plants originally designed for base-load operation, but where lower-cost coal plants are available for dispatch. This further raises the average cost of electricity and CO_2 capture for those NGCC plants, as reflected in one case in Table 3.9 with a capacity factor of 50%. In other parts of the world, however, lower-cost coal plants may not be available, or gas supply contracts might limit the ability to curtail gas use. Such situations again illustrate that options for power generation with or without CO_2 capture should be evaluated in the context of a particular situation or scenario.

Studies of commercial post-combustion CO_2 capture applied to simple-cycle gas turbines have been conducted for the special case of retrofitting an auxiliary power generator in a remote location (CCP, 2005). This study reported a relatively high cost of 88 US\$/tCO_2 avoided. Studies of post-combustion capture for gas-fired boilers have been limited to industrial applications, as discussed later in Section 3.7.8.

3.7.5.4 Biomass-firing and co-firing systems

Power plants can be designed to be fuelled solely by biomass, or biomass can be co-fired in conventional coal-burning plants. The requirement to reduce net CO_2 emissions could lead to an increased use of biomass fuel, because plants that utilize biomass as a primary or supplemental fuel may be able to take credit for the carbon removed from the atmosphere during the biomass growth cycle. If the biomass carbon released during combustion (as CO_2) is then captured and sequestered, the net quantity of CO_2 emitted to the atmosphere could in principle be negative.

The most important factor affecting the economics of biomass use is the cost of the biomass. This can range from a negative value, as in the case of some biomass wastes, to costs substantially higher than coal, as in the case of some purposely-grown biomass fuels, or wastes that have to be collected from diffuse sources. Power plants that use only biomass are typically smaller than coal-fired plants because local availability of biomass is often limited and biomass is more bulky and hence more expensive to transport than coal. The smaller sizes of biomass-fired plants would normally result in lower energy efficiencies and higher costs of CO_2 capture. Biomass can be co-fired with coal in larger plants (Robinson *et al.*, 2003). In such circumstances the incremental costs of capturing biomass-derived CO_2 should be similar to costs of capturing coal-derived CO_2. Another option is to convert biomass into pellets or refined liquid fuels to reduce the cost of transporting it over long distances. However, there are costs and emissions associated with production of these refined fuels. Information on costs of CO_2 capture at biomass-fired plants is sparse but some information is given in Section 3.7.8.4. The overall economics of CCS with biomass combustion will depend very much on local circumstances, especially biomass availability and cost and (as with fossil fuels) proximity to potential CO_2 storage sites.

3.7.6 Pre-combustion CO_2 capture cost for electric power plants (current technology)

Studies of pre-combustion capture for electric power plants have focused mainly on IGCC systems using coal or other solid fuels such as petroleum coke. This section of the report focuses on currently available technology for CO_2 capture at such plants. As before, the cost of CO_2 capture depends not only on the choice of capture technology, but more importantly on the characteristics and design of the overall power plant, including the fuel type and choice of gasifier. Because IGCC is not widely used for electric power generation at the present time, economic studies of IGCC power plants typically employ design assumptions based on the limited utility experience with IGCC systems and the more extensive experience with gasification in industrial sectors such as petroleum refining and petrochemicals. For oxygen-blown gasifiers, the high operating pressure and relatively high CO_2 concentrations achievable in IGCC systems makes physical solvent absorption systems the predominant technology of interest for pre-combustion CO_2 capture (see Section 3.5.2.11). For purposes of cost reporting, we again distinguish between new plant designs and the retrofitting of existing facilities.

3.7.6.1 New coal gasification combined cycle power plants

Table 3.10 summarizes the key assumptions and results of several recent studies of CO_2 capture costs for new IGCC power plants ranging in size from approximately 400−800 MW net power output. While several gasifiers and coal types are represented, most studies focus on the oxygen-blown Texaco quench system,[10] and all but one assume bituminous coals. CO_2 capture efficiencies

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[10] In 2004, the Texaco gasifier was re-named as the GE gasifier following acquisition by GE Energy (General Electric). However, this report uses the name Texaco, as it is referred to in the original references cited.

across these studies range from 85−92% using commercially available physical absorption systems. The energy requirements for capture increase the overall plant heat rate (energy input per kWh) by 16−25%, yielding net CO_2 reductions per kWh of 81−88%. Other study variables that influence total plant cost and the cost of CO_2 capture include the fuel cost, CO_2 product pressure, plant capacity factor and fixed charge factor. Many of the recent studies also include the cost of a spare gasifier to ensure high system reliability.

Table 3.10. CO_2 capture costs: New IGCC power plants using current technology.

Table 3.10 indicates that for studies based on the Texaco or E-Gas gasifiers, CO_2 capture adds approximately 20−40% to both the capital cost (US\$ kW^{-1}) and the cost of electricity (US\$ MWh^{-1}) of the reference IGCC plants, while studies using the Shell gasifier report increases of roughly 30−65%. The total COE reported for IGCC systems ranges from 41−61 US\$ MWh^{-1} without capture and 54−79 US\$ MWh^{-1} with capture. With capture, the lowest COE is found for gasifier systems with quench cooling designs that have lower thermal efficiencies than the more capital-intensive designs with heat recovery systems. Without capture, however, the latter system type has the lowest COE in Table 3.10. Across all studies, the cost of CO_2 avoided ranges from 13−37 US\$/tCO_2 relative to an IGCC without capture, excluding transport and storage costs. Part of the reason for this lower incremental cost of CO_2 capture relative to coal combustion plants is the lower average energy requirement for IGCC systems. Another key factor is the smaller gas volume treated in oxygen-blown gasifier systems, which substantially reduces equipment size and cost.

As with PC plants, Table 3.10 again emphasizes the importance of plant financing and utilization assumptions on the calculated cost of electricity, which in turn affects CO_2-capture costs. The lowest COE values in this table are for plants with a low fixed charge rate and high capacity factor, while substantially higher COE values result from high financing costs and lower plant utilization. Similarly, the type and properties of coal assumed has a major impact on the COE, as seen in a recent Canadian Clean Power Coalition study, which found substantially higher costs for low-rank coals using a Texaco-based IGCC system (Stobbs and Clark, 2005, Table 3.10). EPRI also reports higher IGCC costs for low-rank coals (Holt *et al.*, 2003). On the other hand, where plant-level assumptions and designs are similar across studies, there is relatively little difference in the estimated costs of CO_2 capture based on current commercial technology. Similarly, the several studies in Tables 3.7 and 3.10 that estimate costs for both IGCC and PC plants on an internally consistent basis, all find that IGCC plants with capture have a lower COE than PC plants with capture. There is not yet a high degree of confidence in these cost estimates, however (see Table 3.6).

The costs in Table 3.10 also reflect efforts in some studies to identify least-cost CO_2 capture options. For example, one recent study (IEA GHG, 2003) found that capture and disposal of hydrogen sulphide (H_2S) along with CO_2 can reduce overall capture costs by about 20% (although this may increase transport and storage costs, as discussed in Chapters 4 and 5). The feasibility of this approach depends in a large part on applicable regulatory and permitting requirements. Advanced IGCC designs that may further reduce future CO_2 capture costs are discussed in Section 3.7.7.

3.7.6.2 Repowering of existing coal-fired plants with IGCC

For some existing coal-fired power plants, an alternative to the post-combustion capture systems discussed earlier is repowering with an IGCC system. In this case − depending on site-specific circumstances − some existing plant components, such as the steam turbine, might be refurbished and utilized as part of an IGCC plant. Alternatively, the entire combustion plant might be replaced with a new IGCC system while preserving other site facilities and infrastructure.

Although repowering has been widely studied as an option to improve plant performance and increase plant output, there are relatively few studies of repowering motivated by CO_2 capture. Table 3.8 shows results from one recent study (Chen et *al.*, 2003) which reports CO_2 capture costs for IGCC repowering of a 250 MW coal-fired unit that is assumed to be a fully amortized (hence, a low COE of 21 US\$ MWh^{-1}). IGCC repowering yielded a net plant capacity of 600 MW with CO_2 capture and a COE of 62−67 US\$ MWh^{-1} depending on whether or not the existing steam turbine can be reused. The cost of CO_2 avoided was 46−51 US\$/tCO_2. Compared to the option of retrofitting the existing PC unit with an amine-based capture system and retaining the existing boiler (Table 3.8), the COE for IGCC repowering was estimated to be 10−30% lower. These findings are in general agreement with earlier studies by Simbeck (1999). Because the addition of gas turbines roughly triples the gross plant capacity of a steam-electric plant, candidates for IGCC repowering are generally limited to smaller existing units (e.g., 100−300 MW). Taken together with the post-combustion retrofit studies in Table 3.8, the most cost-effective options for existing plants involve combining CO_2 capture with plant upgrades that increase overall efficiency and net output. Additional studies would be needed to systematically compare the feasibility and cost of IGCC repowering to supercritical boiler upgrades at existing coal-fired plants.

3.7.7 CO_2 capture cost for hydrogen production and multi-product plants (current technology)

While electric power systems have been the dominant technologies of interest for CO_2 capture studies, other industrial processes, including hydrogen production and multi-product plants producing a mix of fuels, chemicals and electricity also are of interest. Because CO_2 capture cost depends strongly on the production process in question, several categories of industrial processes are discussed below.

3.7.7.1 Hydrogen Production Plants

Section 3.5 discussed the potential role of hydrogen as an energy carrier and the technological options for its production. Here we examine the cost of capturing CO_2 normally released during the production of hydrogen from fossil fuels. Table 3.11 shows the key assumptions and cost results of recent studies of CO_2 capture costs for plants with hydrogen production rates of 155,000−510,000 $Nm^3\ h^{-1}$ (466−1531 MW_t), employing either natural gas or coal as a feedstock. The CO_2 capture efficiency for the hydrogen plant ranges from 87−95% using commercially available chemical and physical absorption systems. The CO_2 reduction per unit of product is lower, however, because of the process energy requirements and because of additional CO_2 emitted by an offsite power plant assumed in some of these studies. As hydrogen production requires the separation of H_2 from CO_2, the incremental cost of capture is mainly the cost of CO_2 compression.

Table 3.11. CO_2 capture costs: Hydrogen and multi-product plants using current or near-commercial technology.

At present, hydrogen is produced mainly from natural gas. Two recent studies (see Table 3.11) indicate that CO_2 capture would add approximately 18-33% to the unit cost of hydrogen while reducing net CO_2 emissions per unit of H_2 product by 72-83% (after accounting for the CO_2 emissions from imported electricity). The total cost of hydrogen is sensitive to the cost of feedstock, so different gas prices would alter both the absolute and relative costs of CO_2 capture.
For coal-based hydrogen production, a recent study (NRC,2004) projects an 8% increase in the unit cost of hydrogen for an 83% reduction in CO_2 emissions per unit of product. Again, this figure includes the CO_2 emissions from imported electricity.

3.7.7.2 Multi-product plants

Multi-product plants (also known as polygeneration plants) employ fossil fuel feedstocks to produce a variety of products such as electricity, hydrogen, chemicals and liquid fuels. To calculate the cost of any particular product (for a given rate of return), economic analyses of multi-product plants require that the selling price of all other products be specified over the operating life of the plant. Such assumptions, in addition to those discussed earlier, can significantly affect the outcome of cost calculations when there is not one dominant product at the facility.

Several of the coal-based hydrogen production plants in Table 3.11 also produce electricity, albeit in small amounts (in fact, smaller than the electricity quantities purchased by the stand-alone plants). Most of these studies assume that the value of the electricity product is higher under a carbon capture regime than without CO_2 capture. The result is a 5−33% increase in hydrogen production cost for CO_2 reductions of 72−96% per unit of product. The case with the lowest incremental product cost and highest CO_2 reduction assumes co-disposal of H_2S with CO_2, thus eliminating the costs of sulphur capture and recovery. As noted earlier (Section 3.7.6.1), the feasibility of this option depends strongly on local regulatory requirements; nor are higher costs for transport and storage reflected in the Table 3.11 cost estimate for this case.

Table 3.11 also presents examples of multi-product plants producing liquid fuels plus electricity. In these cases the amounts of electricity produced are sizeable compared to the liquid products, so the assumed selling price of electricity has a major influence on the product cost results. So too does the assumption in two of the cases of co-disposal of H_2S with CO_2 (as described above). For these reasons, the incremental cost of CO_2 capture ranges from a 13% decrease to a 13% increase in fuel product cost relative to the no-capture case. Note too that the overall level of CO_2 reductions per unit of product is only 27−56%. This is because a significant portion of carbon in the coal feedstock is exported with the liquid fuel products. Nonetheless, an important benefit of these fuel-processing schemes is a reduction (of 30−38%) in the carbon content per unit of fuel energy relative to the feedstock fuel. To the extent these liquid fuels displace other fuels with higher carbon per unit of energy, there is a net benefit in end-use CO_2 emissions when the fuels are burned. However, no credit for such reductions is taken in Table 3.11 because the system boundary considered is confined to the fuel production plant.

3.7.8 Capture costs for other industrial processes (current technology)

CO_2 can be captured in other industrial processes using the techniques described earlier for power generation. While the costs of capture may vary considerably with the size, type and location of industrial processes, such costs will be lowest for processes or plants having: streams with relatively high CO_2 concentrations; process plants that normally operate at high load factors; plants with large CO_2 emission rates; and, processes that can utilize waste heat to satisfy the energy requirements of CO_2 capture systems. Despite these potential advantages, little detailed work has been carried out to estimate costs of CO_2 capture at industrial plants, with most work focused on oil refineries and petrochemical plants. A summary of currently available cost studies appears in Table 3.12.

Table 3.12. Capture costs: Other industrial processes using current or advanced technology.

3.7.8.1 Oil refining and petrochemical plants

Gas-fired process heaters and steam boilers are responsible for the bulk of the CO_2 emitted from typical oil refineries and petrochemical plants. Although refineries and petrochemical plants emit large quantities of CO_2, they include multiple emission sources often dispersed over a large area.

Economies of scale can be achieved by using centralized CO_2 absorbers or amine regenerators but some of the benefits are offset by the cost of pipes and ducts. Based on Table 3.14, the cost of capturing and compressing CO_2 from refinery and petrochemical plant heaters using current technology is estimated to be 50−60 US$/t$CO_2$ captured. Because of the complexity of these industrial facilities, along with proprietary concerns, the incremental cost of plant products is not normally reported.

Table 3.14. CO_2 capture costs: Multi-product plants using advanced technology.

High purity CO_2 is currently vented to the atmosphere by some gas processing and petrochemical plants, as described in Chapter 2. The cost of CO_2 capture in such cases would be simply the cost of drying and compressing the CO_2 to the pressure required for transport. The cost would depend on various factors, particularly the scale of operation and the electricity price. Based on 2 MtCO_2 yr^{-1} and an electricity price of US$ 0.05 kWh^{-1}, the cost is estimated to be around 10 US$/t$CO_2$ emissions avoided. Electricity accounts for over half of the total cost.

3.7.8.2 Cement plants

As noted in Chapter 2, cement plants are the largest industrial source of CO_2 apart from power plants. Cement plants normally burn lower cost high-carbon fuels such as coal, petroleum coke and various wastes. The flue gas typically has a CO_2 concentration of 14−33% by volume, significantly higher than at power plants, because CO_2 is produced in cement kilns by decomposition of carbonate minerals as well as by fuel combustion. The high CO_2 concentration would tend to reduce the specific cost of CO_2 capture from flue gas. Pre-combustion capture, if used, would only capture the fuel-related CO_2, so would be only a partial solution to CO_2 emissions. Oxy-fuel combustion and capture using calcium sorbents are other options, which are described in Sections 3.2.4 and 3.7.11.

3.7.8.3 Integrated steel mills

Integrated steel mills are some of the world's largest emitters of CO_2, as described in Chapter 2. About 70% of the carbon introduced into an integrated steel mill is contained in the blast furnace gas in the form of CO_2 and CO, each of which comprise about 20% by volume of the gas. The cost of capturing CO_2 from blast furnace gas was estimated to be 35 US$/t$CO_2$ avoided (Farla *et al.*, 1995) or 18 US$/t$CO_2$ captured (Gielen, 2003).

Iron ore can be reacted with synthesis gas or hydrogen to produce iron by direct reduction (Cheeley, 2000). Direct reduction processes are already used commercially but further development work would be needed to reduce their costs so as to make them more widely competitive with conventional iron production processes. The cost of capturing CO_2 from a direct reduction iron (DRI) production processes was estimated to be 10 US$/t$CO_2$ (Gielen, 2003). CO_2 also could be captured from other gases in iron and steel mills but costs would probably be higher as they are more dilute or smaller in scale.

3.7.8.4 Biomass plants

The main large point sources of biomass-derived CO_2 are currently wood pulp mills, which emit CO_2 from black liquor recovery boilers and bark-fired boilers, and sugar/ethanol mills, which emit CO_2 from bagasse-fired boilers. Black liquor is a byproduct of pulping that contains lignin and chemicals used in the pulping process. The cost of post-combustion capture was estimated to be 34 US$/t$CO_2$ avoided in a plant that captures about 1 MtCO_2 yr^{-1} (Möllersten *et al.*, 2003). Biomass gasification is under development as an alternative to boilers.

CO_2 could be captured from sucrose fermentation and from combustion of sugar cane bagasse at a cost of about 53 US\$/tCO_2 avoided for a plant capturing 0.6 MtCO_2 yr^{-1} avoided (Möllersten *et al.*, 2003). CO_2 from sugar cane fermentation has a high purity, so only drying and compression is required. The overall cost is relatively high due to an annual load factor that is lower than that of most power stations and large industrial plants.

CO_2 could be captured at steam-generating plants or power plants that use other biomass byproducts and/or purpose-grown biomass. At present most biomass plants are relatively small. The cost of capturing 0.19 MtCO_2 yr^{-1} in a 24 MW biomass-powered IGCC plant, compared to a biomass IGCC plant without capture, is estimated to be about 70 US\$/tCO_2 (Audus and Freund, 2005). Larger plants using purpose-grown biomass may be built in the future and biomass can be co-fired with fossil fuels to give economies of scale, as discussed in Chapter 2. Biomass fuels produce similar or slightly greater quantities of CO_2 per unit of fuel energy as bituminous coals; thus, the CO_2 concentration of flue gases from these fuels will be broadly similar. This implies that the cost of capturing CO_2 at large power plants using biomass may be broadly similar to the cost of capturing CO_2 in large fossil fuel power plants in cases where plant size, efficiency, load factor and other key parameters are similar. The costs of avoiding CO_2 emissions in power plants that use biomass are discussed in more detail in Chapter 8.

3.7.9 Outlook for future CO_2 capture costs

The following sections focus on 'advanced' technologies that are not yet commercial available, but which promise to lower CO_2 capture costs based on preliminary data and design studies. Earlier sections of Chapter 3 discussed some of the efforts underway worldwide to develop lower-cost options for CO_2 capture. Some of these developments are based on new process concepts, while others represent improvements to current commercial processes. Indeed, the history of technology innovation indicates that incremental technological change, sustained over many years (often decades), is often the most successful path to substantial long-term improvements in performance and reductions in cost of a technology (Alic *et al.*, 2003). Such trends are commonly represented and quantified in the form of a 'learning curve' or 'experience curve' showing cost reductions as a function of the cumulative adoption of a particular technology (McDonald and Schrattenholzer, 2001). One recent study relevant to CO_2 capture systems found that over the past 25 years, capital costs for sulphur dioxide (SO_2) and nitrogen oxides (NO_x) capture systems at US coal-fired power plants have decreased by an average of 12% for each doubling of installed worldwide capacity (a surrogate for cumulative experience, including investments in R&D) (Rubin *et al.*, 2004a). These capture technologies bear a number of similarities to current systems for CO_2 capture. Another recent study (IEA, 2004) suggests a 20% cost reduction for a doubling of the unit capacity of engineered processes due to technological learning. For CCS systems the importance of costs related to energy requirements is emphasized, since reductions in such costs are required to significantly reduce the overall cost of CO_2 capture.

At the same time, a large body of literature on technology innovation also teaches us that learning rates are highly uncertain,[11] and that cost estimates for technologies at the early stages of development are often unreliable and overly optimistic (Merrow *et al.*, 1981). Qualitative

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[11] In their study of 42 energy-related technologies, McDonald and Schrattenholzer (2001) found learning rates varying from -14% to 34%, with a median value of 16%. These rates represent the average reduction in cost for each doubling of installed capacity. A negative learning rate indicates that costs increased rather than decreased over the period studied.

descriptions of cost trends for advanced technologies and energy systems typically show costs increasing from the research stage through full-scale demonstration; only after one or more full-scale commercial plants are deployed do costs begin to decline for subsequent units (EPRI, 1993; NRC, 2003). Case studies of the SO_2 and NO_x capture systems noted above showed similar behaviour, with large (factor of two or more) increases in the cost of early full-scale FGD and SCR installations before costs subsequently declined (Rubin *et al.*, 2004b). Thus, cost estimates for CO_2 capture systems should be viewed in the context of their current stage of development. Here we try to provide a perspective on potential future costs that combines qualitative judgments with the quantitative cost estimates offered by technology developers and analysts. The sections below revisit the areas of power generation and other industrial processes to highlight some of the major prospects for CO_2 capture cost reductions.

3.7.10 CO_2 capture costs for electric power plants (advanced technology)

This section first examines oxy-fuel combustion, which avoids the need for CO_2 capture by producing a concentrated CO_2 stream for delivery to a transport and storage system. Following this we examine potential advances in post-combustion and pre-combustion capture.

3.7.10.1 Oxy-fuel combustion systems

It is first important to distinguish between two types of oxy-fuel systems: an oxy-fuel boiler (either a retrofit or new design) and oxy-fuel combustion-based gas turbine cycles. The former are close to demonstration at a commercial scale, while the latter (such as chemical looping combustion systems and novel power cycles using CO_2/water as working fluid) are still at the design stage. Table 3.13 summarizes the key assumptions and cost results of several recent studies of CO_2 capture costs for oxy-fuel combustion systems applied to new or existing coal-fired units. As discussed earlier in Section 3.4, oxygen combustion produces a flue gas stream consisting primarily of CO_2 and water vapour, along with smaller amounts of SO_2, nitrogen and other trace impurities. These designs eliminate the capital and operating costs of a post-combustion CO_2 capture system, but new costs are incurred for the oxygen plant and other system design modifications. Because oxy-fuel combustion is still under development and has not yet been utilized or demonstrated for large-scale power generation, the design basis and cost estimates for such systems remain highly variable and uncertain. This is reflected in the wide range of oxy-fuel cost estimates in Table 3.13. Note, however, that cost estimates for advanced design concepts based on oxy-fuel combustion gas turbine cycles are more uncertain at this time than cost estimates for new or retrofitted boilers employing oxy-fuel combustion.

Table 3.13. Capture costs: Advanced technologies for electric power plants.

For new plant applications, the data in Table 3.13 indicate that oxy-fuel combustion adds about 30−90% to the capital cost and 30−150% to the COE of a conventional plant, while reducing CO_2 emissions per kWh by 75−100%. Retrofit applications exhibit higher relative costs in cases where the existing plant is wholly or partially amortized. The lowest-cost oxy-fuel system in Table 3.13 is one that employs chemical looping to achieve nearly a 100% reduction in CO_2 emissions. While this concept thus appears promising (see Section 3.4.6), it has yet to be tested and verified at a meaningful scale. Thus cost estimates based on conceptual designs remain highly uncertain at this time.

To judge the potential cost savings of oxy-fuels relative to current CO_2 capture systems, it is useful to compare the costs of alternative technologies evaluated within a particular study based on a particular set of premises. In this regard, the COE for the oxy-fuel retrofit system reported by Alstom *et al.* (2001) in Table 3.13 is 20% lower than the cost of an amine system retrofit (Table

3.13) for the same 255 MW plant, while the cost of CO_2 avoided is 26% lower. In contrast, a recent study by the Canadian Clean Power Coalition (Stobbs and Clark, 2005) reports that the COE for an oxy-fuel system at a large lignite-fired plant (Table 3.13) is 36% higher than for an amine CO_2 capture system, while the cost of CO_2 avoided is more than twice as great. The major source of that cost difference was a specification in the CCPC study that the oxy-fuelled unit also be capable of full air firing. This resulted in a much higher capital cost than for a new unit designed solely for oxy-fuel operation. A more recent study sponsored by IEA GHG (Dillon *et al.*, 2005) found that a large new supercritical coal-fired boiler with oxy-fuel combustion had a COE slightly (2−3%) lower than a state-of-the-art coal plant with post-combustion analyzed in a separate study employing similar assumptions (IEA GHG, 2004). Further cost reductions could be achieved with the successful development of new lower-cost oxygen production technology (see Section 3.4.5). At the current time, the optimum designs of oxy-fuel combustion systems are not yet well established and costs of proposed commercial designs remain uncertain. This is especially true for advanced design concepts that employ components which are not yet available or still in the development stage, such as CO_2 gas turbines or high temperature ceramic membranes for oxygen production.

3.7.10.2 Advanced systems with post-combustion capture

Improvements to current amine-based systems for post-combustion CO_2 capture are being pursued by a number of process developers (Mimura *et al.*, 2003; Muramatsu and Iijima, 2003; Reddy *et al.*, 2003) and may offer the nearest-term potential for cost reductions over the systems currently in use. The newest systems summarized earlier in Table 3.7 reportedly reduce the cost of CO_2 avoided by approximately 20−30% (IEA GHG, 2004). Table 3.13 indicates that additional advances in plant heat integration could further reduce the COE of capture plants by about 5%. These results are consistent with a recent study by Rao *et al.* (2003), who used expert elicitations and a plant simulation model to quantify the improvements likely achievable by 2015 for four key process parameters: sorbent concentration, regeneration energy requirements, sorbent loss and sorbent cost. The ‘most likely’ improvement was an 18% reduction in COE, while the ‘optimistic’ estimates yielded a 36% cost reduction from improvements in just these four parameters. The cost of CO_2 avoided was reduced by similar amounts. Advances in more efficient heat integration (for sorbent regeneration) and higher power plant efficiency could lead to even greater reductions in CO_2 capture cost.

Advances in gas turbine technology produce similar benefits for NGCC systems. Table 3.13 shows several cases based on the H-turbine design. Relative to the cases in Table 3.9, these systems offer higher efficiency and greater CO_2 reductions per kWh. The higher COEs for the advanced NGCC systems reflects the higher natural gas prices assumed in more recent studies.

Table 3.13 indicates that other advanced technologies for post-combustion applications, such as membrane separation systems, may also lower the future cost of CO_2 capture (see Section 3.3.3). Reliable cost estimates for such technologies should await their further development and demonstration.

3.7.10.3 Advanced systems with pre-combustion capture

The cost of gasification-based systems with CO_2 capture also can be expected to fall as a result of continued improvements in gas turbine technology, gasifier designs, oxygen production systems, carbon capture technology, energy management and optimization of the overall facility. One recent study (IEA GHG, 2003) estimates a 20% reduction in the cost of electricity generation from a coal-based IGCC plant with CO_2 capture by 2020. This takes into account improvements in gasification, oxygen production, physical solvent scrubbing and combined cycle processes, but does not take into account any possible radical innovations in CO_2 separation technology. The additional IGCC cases shown in Table 3.13, including recent results of the CO_2 Capture Project (CCP, 2005), foresee similar reductions in the COE of advanced IGCC systems compared to the systems in Table 3.10.

3.7.11 CO_2 capture costs for hydrogen production and multi-product plants (advanced technology)

Table 3.14 shows results of several recent studies that have projected the performance and cost of new or improved ways of producing hydrogen and electricity from fossil fuels.

Compared to the current commercial plants in Table 3.11, the advanced single-product systems with CO_2 capture have hydrogen cost reductions of 16% (for natural gas feedstock) to 26% (for coal feedstock). Additional cases in Table 3.14 show multi-product systems producing hydrogen and electricity. These cases indicate the potential for substantial reductions in the future cost of hydrogen production with CO_2 capture. As before, the results are sensitive to the assumed selling price of co-product electricity. More importantly, these cases assume the successful scale-up and commercialization of technologies that have not yet been demonstrated, or which are still under development at relatively small scales, such as solid oxide fuel cells (SOFC). Published cost estimates for these systems thus have a very high degree of uncertainty.

3.7.12 CO_2 capture costs for other industrial processes (advanced technology)

As noted earlier, CO_2 capture for industrial processes has not been widely studied. The most extensive analyses have focused on petroleum refineries, especially CO_2 capture options for heaters and other combustion-based processes (see Table 3.12). The use of oxy-fuel combustion offers potential cost savings in several industrial applications. The CO_2 Capture Project reports the cost of capturing CO_2 in refinery heaters and boilers, with an ion transport membrane oxygen plant, to be 31 US\$/tCO_2 avoided. The cost of pre-combustion capture based on shift and membrane gas separation was predicted to be 41 US\$/tCO_2 avoided (CCP, 2005).

It also may be possible to apply oxy-fuel combustion to cement plants, but the CO_2 partial pressure in the cement kiln would be higher than normal and the effects of this on the calcination reactions and the quality of the cement product would need to be investigated. The quantity of oxygen required per tonne of CO_2 captured in a cement plant would be only about half as much as in a power plant, because only about half of the CO_2 is produced by fuel combustion. This implies that the cost of CO_2 capture by oxy-fuel combustion at large cement plants would be lower than at power plants, but a detailed engineering cost study is lacking. Emerging technologies that capture CO_2 using calcium-based sorbents, described in Section 3.3.3.4, may be cost competitive in cement plants in the future.

3.7.13 Summary of CO_2 capture cost estimates

Table 3.15 summarizes the range of current CO_2 capture costs for the major electric power systems analyzed in this report. These costs apply to case studies of large new plants employing current commercial technologies. For the PC and IGCC systems, the data in Table 3.15 apply only to plants using bituminous coals and the PC plants are for supercritical units only. The cost ranges for each of the three systems reflect differences in the technical, economic and operating assumptions employed in different studies. While some differences in reported costs can be attributed to differences in the CO_2 capture system design, the major sources of variability are differences in the assumed design, operation and financing of the reference plant to which the capture technology is applied (i.e., factors such as plant size, location, efficiency, fuel type, fuel cost, capacity factor and cost of capital). Because no single set of assumptions applies to all situations or all parts of the world, we display the ranges of cost represented by the studies in Tables 3.8, 3.10, 3.11 and 3.12.

Table 3.15. Summary of new plant performance and CO_2 capture cost based on current technology.

For the power plant studies reflected in Table 3.15, current CO_2 capture systems reduce CO_2 emissions per kilowatt-hour by approximately 85−90% relative to a similar plant without capture. The cost of electricity production attributed to CO_2 capture increases by 35−70% for a natural gas combined cycle plant, 40−85% for a new pulverized coal plant and 20−55% for an integrated gasification combined cycle plant. Overall, the COE for fossil fuel plants with capture ranges from 43−86 US\$ MWh^{-1}, as compared to 31−61 US\$ MWh^{-1} for similar plants without capture. These costs include CO_2 compression but not transport and storage costs. In most studies to date, NGCC systems typically have a lower COE than new PC and IGCC plants (with or without capture) for large base load plants with high capacity factors (75% or more) and gas prices below about 4 US\$ GJ^{-1} over the life of the plant. However, for higher gas prices and/or lower capacity factors, NGCC plants typically have higher COEs than coal-based plants, with or without capture. Recent studies also found that IGCC plants were on average slightly more costly without capture and slightly less costly with capture than similarly sized PC plants. However, the difference in cost between PC and IGCC plants with or without CO_2 capture can vary significantly with coal type and other local factors, such as the cost of capital. Since neither PC nor IGCC systems have yet been demonstrated with CO_2 capture and storage for a large modern power plant (e.g., 500 MW), neither the absolute or relative costs of these systems (nor comparably sized NGCC systems with capture and storage) can be stated with a high degree of confidence at this time, based on the criteria of Table 3.6.

Table 3.15 also shows that the lowest CO_2 capture costs with current technology (as low as 2 US\$/tCO_2 captured or avoided) were found for industrial processes such as coal-based hydrogen production plants that produce concentrated CO_2 streams as part of the production process. Such industrial processes may represent some of the earliest opportunities for CCS.

Figure 3.20 displays the normalized power plant cost and emissions data from Table 3.15 in graphical form. On this graph, the cost of CO_2 avoided corresponds to the slope of a line connecting any two plants (or points) of interest. While Table 3.15 compares a given capture plant to a similar plant without capture, in some cases comparisons may be sought between a given capture plant and a different type of reference plant. Several cases are illustrated in Figure 3.20 based on either a PC or NGCC reference plant. In each case, the COE and CO_2 emission rate are highly dependent upon technical, economic and financial factors related to the design and operation of the power systems of interest at a particular location. The cost of CO_2 avoided is especially sensitive to these site-specific factors and can vary by an order of magnitude or more when different types of plants are compared. Comparisons of different plant types, therefore, require a specific context and geographical location to be meaningful and should be based on the full COE including CO_2 transport and storage costs. Later, Chapter 8 presents examples of full CCS costs for different plant types and storage options.

Figure 3.20. Cost of electricity (excluding transport and storage costs) compared to CO_2 emission rate for different reference and capture plants based on current technology. The shaded areas show the Table 3.15 ranges of CO_2 emission rates and levelized cost of electricity (COE) for new PC, IGCC and NGCC plants with and without CO_2 capture. All coal plant data are for bituminous coals only. PC plants are supercritical units only. (See Tables 3.7, 3.9, 3.10 and 3.15 for additional assumptions.) The cost of CO_2 avoided corresponds to the slope of a line connecting a plant with capture and a reference plant without capture (i.e., the change in electricity cost divided by the

change in emission rate). Avoidance costs for the same type of plant with and without capture plant are given in Table 3.15. When comparing different plant types, the reference plant represents the least-cost plant that would 'normally' be built at a particular location in the absence of a carbon constraint. In many regions today, this would be either a PC plant or an NGCC plant. The cost per tonne of CO_2 avoided can be highly variable and depends strongly on the costs and emissions of new plants being considered in a particular situation. See Chapter 8 for the full COE and full cost of CO_2 avoided for different plant types.

In contrast to new plants, CO_2 capture options and costs for existing power plants have not been extensively studied. Current studies indicate that these costs are extremely site-specific and fall into two categories (see Table 3.8). One is the retrofitting of a post-combustion capture system to the existing unit. The other category combines CO_2 capture with upgrading or repowering the existing plant to significantly improve its efficiency and net power output (see Sections 3.7.4.2 and 3.7.5.2). In general, the latter option appears to be more cost-effective. However, further site-specific studies are required to systematically assess the feasibility and cost of alternative repowering options in conjunction with CO_2 capture for existing power plants.

New or improved methods of CO_2 capture, combined with advanced power systems and industrial process designs, promise to significantly reduce CO_2 capture costs and associated energy requirements. Tables 3.12 to 3.14 summarize the results from recent studies that examine future options. As discussed earlier, there is considerable uncertainty about the magnitude and timing of future cost reductions, as well as the potential for costs to rise above current estimates, especially for technologies still in the early stages of research and development. The current assessment is based on studies of the specific technologies in Tables 3.12 to 3.14 (and the supporting discussions and literature cited in Sections 3.7.9 to 3.7.12), as well as analyses of historical cost trends for related energy and environmental technologies. This assessment suggests that improvements to current commercial technologies can reduce CO_2 capture costs by at least 20−30% over approximately the next decade, while new technologies under development promise more substantial cost reductions. Achieving future cost reductions, however, will require deployment and adoption of commercial technologies in the marketplace as well as sustained R&D.

3.8 Gaps in knowledge

Gaps in knowledge are related to differences in the stages of development of component technologies for the capture systems reviewed in Sections 3.2 to 3.5. For CO_2 capture from industrial processes, a number of technologies that are commonly used in gas sweetening and ammonia production are already used on a commercial scale. For other types of industrial systems capturing CO_2 from steel and cement production, further work is still needed. For CO_2 capture that might be reliant on post-combustion capture or oxy-fuel combustion, options are less well developed, or are available at a smaller scale than those required for applications such as in power generation, where much larger gas flows are handled. For pre-combustion capture many of the required systems have been developed and applied in industry already.

Although many of the component and/or enabling technologies required for CO_2 capture in post-combustion, pre-combustion and oxy-fuel combustion are well known, gaps in knowledge are in the practical and/or commercial demonstration of integrated systems. This demonstration is essential to prove the cost of CO_2 capture and its use on a large scale, particularly in power generation applications, but also for cement, steel and other large industries. Operating experience is also needed to test system reliability, improved methods of system integration, methods to reduce the energy requirements for CO_2 capture, improved process control strategies and the use of optimized functional materials for the implementation of capture processes with advanced, higher efficiency power cycles. As such developments are realized, environmental issues associated with the capture of CO_2 and other deleterious pollutants in these systems should also be re-assessed from a perspective involving the whole capture-transport-storage operation.

In an ongoing search to implement existing, new or improved methods of CO_2 capture, most capture systems also rely on the application of a range of enabling technologies that influence the attractiveness of a given system. These enabling technologies have their own critical gaps of knowledge. For example, improved processes for the effective removal of sulphur, nitrogen, chlorine, mercury and other pollutants are needed for the effective performance of unit operations for CO_2 separation in post- and pre-combustion capture systems, especially when coal is used as the primary fuel. Improved gasification reactors for coals and biomass, the availability of hydrogen-burning gas turbines and fuel cells for stationary power generation also need further development in the pre-combustion route. Combustors and boilers operating at higher temperatures, or a new class of CO_2 turbines and compressors, are important requirements for oxy-fuel systems.

With reference to the development of novel CO_2 capture and/or other enabling technologies, a wide range of options are currently being investigated worldwide. However, many technical details of the specific processes proposed or under development for these emerging technologies are still not well understood. This makes the assessment of their performance and cost highly uncertain. This is where intense R&D is needed to develop and bring to pilot scale testing the most promising concepts for commercial application. Membranes for H_2, CO_2 or O_2 separation, new sorbents, O_2 or CO_2 solid carriers and materials for advanced combustors, boilers and turbines all require extensive performance testing. Multi-pollutant emission controls in these novel systems and the impact of fuel impurities and temperature on the functional materials, should also be an area of future work.

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Tables

Table 3.1. Capture toolbox.

	Process Streams[a]		**Post-Combustion Capture**		**Oxy-Fuel Combustion Capture**		**Pre-Combustion Capture**	
Separation Task	CO_2/CH_4		CO_2/N_2		O_2/N_2		CO_2/H_2	
Capture Technologies	Current	Emerging	Current	Emerging	Current	Emerging	Current	Emerging
Solvents (Absorption)	**Physical solvents** **Chemical solvents**	Improved solvents Novel contacting equipment Improved design of processes	**Chemical solvents**	Improved solvents Novel contacting equipment Improved design of processes	n. a.	Biomimetic solvents, e.g., haemoglobin-derivatives	**Physical solvent** **Chemical solvents**	Improved chemical solvents Novel contacting equipment Improved design of processes
Membranes	**Polymeric**	Ceramic Facilitated transport Carbon Contactors	Polymeric	Ceramic Facilitated transport Carbon Contactors	Polymeric	Ion transport membranes Facilitated transport	Polymeric	Ceramic Palladium Reactors Contactors
Solid Sorbents	Zeolites Activated carbon		Zeolites Activated carbon	Carbonates Carbon-based sorbents	Zeolites Activated carbon	Adsorbents for O_2/N_2 separation, Perovskites Oxygen chemical looping	Zeolites Activated carbon Alumina	Carbonates Hydrotalcites Silicates
Cryogenic	Ryan-Holmes process		Liquefaction	Hybrid processes	**Distillation**	Improved distillation	Liquefaction	Hybrid processes

[a] Notes: Processes shown in bold are commercial processes that are currently preferred in most circumstances. Some process streams involve CO_2/H_2 or CO_2/N_2 separations but this is covered under pre-combustion capture and post-combustion capture. The key separation processes are outlined in Section 3.1.3 and described in Sections 3.2–3.5.

Table 3.2. Common solvents used for the removal of CO_2 from shifted syngas in pre-combustion capture processes.

Solvent name	Type	Chemical Name	Vendors
Rectisol	Physical	Methanol	Lurgi and Linde, Germany Lotepro Corporation, USA
Purisol	Physical	N-methyl-2-pyrolidone (NMP)	Lurgi, Germany
Selexol	Physical	Dimethyl ethers of polyethylene glycol (DMPEG)	Union Carbide, USA
Benfield	Chemical	Potassium carbonate	UOP
MEA	Chemical	Monoethanolamine	Various
MDEA	Chemical	Methyldiethylamine	BASF and others
Sulphinol	Chemical	Tetrahydrothiophene 1,1-dioxide (Sulpholane), an alkaloamine and water	Shell

Table 3.3. Membrane materials, operating conditions and characteristics for H_2 separation.

	Microporous Ceramic	Microporous Ceramic	Microporous Carbon	Zeolites	Metal
Membrane material	Alumina	Silica	Carbon	Silica (Alumina)	Pd/Ag
Temperature range (°C)	<500	<400	<400	<500–700	<600
Pressure range (bar)	>100	>100	10	>100	>100
Pore size distribution (nm)	0.7–2.0	0.7–2.0	0.7–2.0	0.3–0.7	no pores
Separation factors (H_2/CO_2)	15	15	15–25	50	100
Permeability (mol m^{-2} s Pa)	10^{-6}	10^{-6}	10^{-7}	10^{-6}	10^{-7}–10^{-6}
Experim. temp. (°C)	200	200	300–400	300–400	300–400
Pre-clean-up requirements				S	S, HCl, HF (?)
Chemical resistance problem		H_2O	O_2	S	S, HCl, HF
Geometry	Top layer tube	Top layer tube	Top layer tube/fibre	Top layer tube	Top layer tube/plate
Configuration	Cascade /recycle/once through	Cascade /recycle/once through	Cascade /recycle/once through	Once through	Once through
Lifetime	+	-	+	+	0
Costs (US$ m^{-2})	4250	4250	3000?	4000–4250	4000–4250
Scalability	0	0	0	-	0

Table 3.4. Concentrations of impurities in dried CO_2, % by volume (Source data: IEA GHG, 2003; IEA GHG, 2004; IEA GHG, 2005).

	SO_2	NO	H_2S	H_2	CO	CH_4	N_2/Ar/O_2	Total
COAL-FIRED PLANTS								
Post-combustion capture	<0.01	<0.01	0	0	0	0	0.01	0.01
Pre-combustion capture (IGCC)	0	0	0.01−0.6	0.8−2.0	0.03−0.4	0.01	0.03−0.6	2.1−2.7
Oxy-fuel	0.5	0.01	0	0	0	0	3.7	4.2
GAS-FIRED PLANTS								
Post-combustion capture	<0.01	<0.01	0	0	0	0	0.01	0.01
Pre-combustion capture	0	0	<0.01	1.0	0.04	2.0	1.3	4.4
Oxy-fuel	<0.01	<0.01	0	0	0	0	4.1	4.1

a. The SO_2 concentration for oxy-fuel and the maximum H_2S concentration for pre-combustion capture are for cases where these impurities are deliberately left in the CO_2, to reduce the costs of capture (see Section 3.6.1.1). The concentrations shown in the table are based on use of coal with a sulphur content of 0.86%. The concentrations would be directly proportional to the fuel sulphur content.

b. The oxy-fuel case includes cryogenic purification of the CO_2 to separate some of the N_2, Ar, O_2 and NOx. Removal of this unit would increase impurity concentrations but reduce costs.

c. For all technologies, the impurity concentrations shown in the table could be reduced at higher capture costs.

Table 3.5. Illustrative impacts of CCS energy requirements on plant-level resource consumption and non-CO_2 emission rates for three current power plant systems. Values shown are mass flow rates per MWh for the capture plant, plus increases over the reference plant rates for the same plant type. See footnotes for additional details. (Source: Rubin *et al.*, 2005)

Capture Plant Parameter [a]	**PC** [b]		**IGCC** [c]		**NGCC** [d]	
	Rate	**Increase**	**Rate**	**Increase**	**Rate**	**Increase**
Resource Consumption (all values in kg MWh^{-1})						
Fuel	390	93	361	49	156	23
Limestone	27.5	6.8	-	-	-	-
Ammonia	0.80	0.19	-	-	-	-
CCS Reagents	2.76	2.76	0.005	0.005	0.80	0.80
Solid Wastes/Byproduct						
Ash/slag	28.1	6.7	34.2	4.7	-	-
FGD residues	49.6	12.2	-	-	-	-
Sulphur	-	-	7.53	1.04	-	-
Spent CCS sorbent	4.05	4.05	0.005	0.005	0.94	0.94
Atmospheric Emissions						
CO_2	107	*–704*	97	*–720*	43	*–342*
SO_x	0.001	*– 0.29*	0.33	0.05	-	-
NO_x	0.77	0.18	0.10	0.01	0.11	0.02
NH_3	0.23	0.22	-	-	0.002	0.002

[a] Net power output of all plants is approximately 500 MW. Coal plants use Pittsburgh #8 coal with 2.1% S, 7.2% ash, 5.1% moisture and 303.2 MJ kg^{-1} lower heating value basis (LHV). Natural gas LHV = 59.9 MJ kg^{-1}. All plants capture 90% of potential CO_2 emissions and compress to 13.7 MPa.

[b] PC = Pulverized coal fired plant; based on a supercritical unit with SCR, ESP and FGD systems, followed by an amine system for CO_2 capture. SCR system assumes 2 ppmv ammonia slip. SO_2 removal efficiency is 98% for reference plant and 99% for capture plant. Net plant efficiency (LHV basis) is 40.9% without CCS and 31.2% with CCS.

[c] IGCC = Integrated gasification combined cycle system based on Texaco quench gasifiers (2 + 1 spare), two GE 7FA gas turbines, 3-pressure reheat HRSG. Sulphur removal efficiency is 98% via hydrolyzer plus Selexol system; Sulphur recovery via Claus plant and Beavon-Stretford tailgas unit. Net plant efficiency (LHV basis) is 39.1% without CCS and 33.8% with CCS.

[d] NGCC = Natural gas combined cycle plant using two GE 7FA gas turbines and 3-pressure reheat HRSG, with an amine system for CO_2 capture. Net plant efficiency (LHV basis) is 55.8% without CCS and 47.6% with CCS.

Table 3.6. Confidence levels for technology and system cost estimates.

Confidence Level	Description
Very High	Mature technology with multiple commercial replications for this application and scale of operation; considerable operating experience and data under a variety of conditions.
High	Commercially deployed in applications similar to the system under study, but at a smaller scale and/or with limited operating experience; no major problems or issues anticipated in this application; commercial guarantees available.
Moderate	No commercial application for the system and/or scale of interest, but technology is commercially deployed in other applications; issues of scale-up, operability and reliability remain to be demonstrated for this application.
Low	Experience and data based on pilot plant or proof-of-concept scale; no commercial applications or full-scale demonstrations; significant technical issues or cost-related questions still to be resolved for this application.
Very Low	A new concept or process not yet tested, or with operational data limited to the laboratory or bench-scale level; issues of large-scale operability, effectiveness, reliability and manufacturability remain to be demonstrated.

Table 3.7. CO_2 capture costs: New pulverized-coal power plants using current technology.

Study Assumptions and Results	**Parsons**	**Parsons**	**Simbeck**	**IEA GHG**	**IEA GHG**	**Rubin *et al.***	**Range**		**NETL**	**Rao & Rubin**	**Stobbs and Clark**
	2002b	**2002b**	**2002**	**2004**	**2004**	**2005**	**min**	**max**	**2002**	**2002**	**2005**
	SUPERCRITICAL UNITS / BITUMINOUS COALS								**SUBCRIT UNITS / LOW RANK COALS**		
Reference Plant (without capture)			*			*			*	*	
Boiler type (subcritical, super, ultra)	super	ultra	ultra	ultra	ultra	super			subcritcal	subcritcal	super
Coal type (bit, sub-bit, lig) and %S	bit, 2.5% S	bit, 2.5% S	bit, 1% S	bit, 1% S	bit, 1% S	bit, 2.1% S			bit, 2.5%S	sub-bit, 0.5%S	lignite
Emission control technologies (SO_2/NO_x)	FGD, SCR	FGD, SCR	FGD, SCR	FGD, SCR	FGD, SCR	FGD, SCR			FGD	FGD, SCR	FGD, SCR, LoTOx
Reference plant net output (MW)	462	506	520	758	754	524	**462**	**758**	397	462	424
Plant capacity factor (%)	65	65	80	85	85	75	**65**	**85**	85	75	90
Net plant efficiency, LHV (%)	42.2	44.8	44.5	44.0	43.7	40.9	**41**	**45**	38.9	36.1	43.4
Coal cost, LHV ($\text{US\$ GJ}^{-1}$)	1.29		*0.98*	1.50	1.50	1.25	**0.98**	**1.50**	*1.03*	*1.25*	0.88
Reference plant emission rate (tCO_2 MWh^{-1})	0.774	0.736	0.76	0.743	0.747	0.811	**0.74**	**0.81**	0.835	0.941	0.883
Capture Plant Design											
CO_2 capture technology	MEA	MEA	MEA	MEA	KS-1	MEA			MEA	MEA	MEA
Net plant output with capture (MW)	329	367	408	666	676	492	**329**	**676**	283	326	311.0
Net plant efficiency, LHV (%)	30.1	32.5	*34.9*	34.8	35.4	31.1	**30**	**35**	27.7	*25.4*	31.8
CO_2 capture system efficiency (%)	90	90	85	87.5	90	90	**85**	**90**	95	90	95
CO_2 emission rate after capture (t MWh^{-1})	0.108	0.101	0.145	0.117	0.092	0.107	**0.09**	**0.15**	0.059	0.133	0.060
CO_2 captured (Mt yr^{-1})	1.830	2.350	2.360	4.061	4.168	3.102	**1.83**	**4.17**	2.346	2.580	2.795
CO_2 product pressure (MPa)	8.4	8.4	13.7	11.0	11.0	13.9	**8**	**14**	10.3	13.9	13.9
CCS energy requirement (% more input MWh^{-1})	40	38	28	26	24	31	**24**	**40**	40	42	36
CO_2 reduction per kWh (%)	**86**	**86**	**81**	**84**	**88**	**87**	**81**	**88**	**93**	**86**	**93**
Cost Results			***	**	**						***
Cost year basis (constant dollars)	2000	2000	2000	2004	2004	2002			2002	2000	2003
Fixed charge factor (%)	15.5	15.5	*12.7*	11.0	11.0	14.8	**11.0**	**15.5**	14.8	15.0	
Reference plant TCR ($\text{US\$ kW}^{-1}$)	1281	1161	*1486*	*1319*	*1265*	1205	**1161**	**1486**	1268	1236	*1891*
Capture plant TCR ($\text{US\$ kW}^{-1}$)	2219	1943	*2578*	*1894*	*2007*	1936	**1894**	**2578**	2373	2163	*3252*
Incremental TCR for capture ($\text{US\$ kW}^{-1}$)	938	782	*1092*	*575*	*742*	731	**575**	**1092**	1105	927	*1361*
Reference plant COE ($\text{US\$ MWh}^{-1}$)	**51.5**	**51.0**	**42.9**	**43.9**	**42.8**	**46.1**	43	52	**42.3**	**49.2**	**44.5**
Capture plant COE ($\text{US\$ MWh}^{-1}$)	**85.6**	**82.4**	**70.9**	**62.4**	**63.0**	**74.1**	62	86	**76.6**	**87.0**	**74.3**
Incremental COE for capture ($\text{US\$ MWh}^{-1}$)	**34.1**	**31.4**	**28**	**18.5**	**20.2**	**28**	18	34	**37.8**	**37.8**	**29.8**
% increase in capital cost (over ref. plant)	73	67	74	44	59	61	**44**	**74**	87	75	72
% increase in COE (over ref. plant)	**66**	**62**	**65**	**42**	**47**	**61**	**42**	**66**	**81**	**77**	**67**
Cost of CO_2 captured (US$/$tCO_2$)	35	28	34	23	24	29	**23**	**35**	31	31	26
Cost of CO_2 avoided (US$/$tCO_2$)	51	49	43	29	31	40	**29**	**51**	43	47	36
Capture cost confidence level (see Table 3.6)	- moderate -								- moderate -		

Notes: All costs in this table are for capture only and do not include the costs of CO_2 transport and storage; see Chapter 8 for total CCS costs. Values in italics were adjusted from original reported values as explained below. * Reported HHV values converted to LHV assuming LHV/HHV = 0.96 for coal. ** Reported capital costs increased by 8% to include interest during construction. ***Reported capital costs increased by 15% to estimate interest during construction and other owners' costs.

Table 3.8. CO$_2$ capture costs: Existing pulverized-coal power plants using current technology.

Study Assumptions and Results	**Simbeck**	**Alstom *et al.***	**Rao & Rubin**	**Rao & Rubin**	**Chen *et al.***	**Chen *et al.***	**Chen *et al.***	**Singh *et al.***	**Gibbins *et al.***	**Range**		**Gibbins *et al.***	**Gibbins *et al.***	**Chen *et al.***
	2000	**2001**	**2002**	**2002**	**2003**	**2003**	**2003**	**2003**	**2005**	**min**	**max**	**2005**	**2006**	**2003**
	AMINE SYSTEM RETROFITS TO EXISTING BOILERS											**REPOWERING + CO$_2$ CAPTURE**		
Reference Plant (without capture)		*	*	*	*	*	*							
Boiler type (subcritical, super, ultra)	sub	sub	sub	sub	sub	sub	sub		sub			super	super	sub
Coal type (bit, sub-bit, lig) and %S	sub-bit, 0.5%	bit, 2.7%S	sub-bit, 0.5%	sub-bit, 0.5%	sub-bit, 1.1%S	sub-bit, 1.1%S	sub-bit, 1.1%S	sub-bit						
Emission control technologies (SO$_2$/NO$_x$)	none	FGD	none	FGD	FGD	FGD	FGD	not reported	not reported			not reported	not reported	FGD
Reference plant size (MW)	292	434	470	470	248	248	248	400		**248**	**470**			248
Plant capacity factor (%)	80	67	75	75	80	76 (Capture= 80)	76 (Capture=80)	91.3	80	**67**	**91**	80	80	80
Net plant efficiency, LHV (%)	36.2	36.2	36.6		33.1	33.1	33.1		36.0	**33**	**37**	43.5	43.5	
Coal cost, LHV (US\$ GJ^{-1})	0.98	*1.30*	*1.25*	*1.25*	*1.20*	*1.20*	*1.20*		3.07	**0.98**	**3.07**	3.07	3.07	*1.20*
Reference plant emission rate (tCO$_2$ MWh^{-1})	0.901	0.908	0.941	0.95	1.004	1.004	1.004	0.925		**0.90**	**1.00**			1.004
Capture Plant Design														
CO$_2$ capture technology	MEA	MEA	MEA	MEA	MEA	MEA	MEA	MEA	MEA			MEA	KS	Selexol
Other equipment included	new FGD	FGD upgrade	New FGD	FGD upgrade	FGD upgrade	FGD upgrade	FGD upgrade	FGD				Advanced supercrit boiler retrofit	Advanced supercrit boiler retrofit	IGCC (Texaco Q) repower +current steam turbine
Net plant size with capture (MW)	294	255	275	275	140	282	282	400		**140**	**400**			590
Auxiliary boiler/fuel used? (type, LHV cost)	NG. \$4.51 GJ^{-1}	none	none	none	none	*NG. \$2.59 GJ^{-1}*	*NG. \$5.06 GJ^{-1}*	NG. \$3.79 *GJ^{-1}*	none			none	none	none
Net plant efficiency, LHV (%)	25.3	21.3	21.4	21.4	18.7				24.0	**19**	**25**	31.5	34.5	32.6
CO$_2$ capture system efficiency (%)	90	96	90	90	90	90	90	90		**90**	**96**			90
CO$_2$ emission rate after capture (t MWh^{-1})	0.113	0.059	0.155	0.16	0.177	0.369	0.369	0.324		**0.06**	**0.37**			0.099
CO$_2$ captured (Mt yr^{-1})	2.090	2.228			1.480	1.480	1.480	2.664		**1.48**	**2.66**			3.684
CO$_2$ product pressure (MPa)	13.7	13.9	13.9	13.9	13.9	13.9	13.9		10.0	**10**	**14**	10.0	10.0	14.5
CCS energy requirement (% more input MWh^{-1})	43	70	71		77				50	**43**	**77**	38	26	
CO$_2$ reduction per kWh (%)	**87**	**94**	**84**	**83**	**82**	**63**	**63**	**65**		**63**	**94**			
Cost Results	**							**						
Cost year basis (constant dollars)	1999	n/a	2000	2000	2000	2000	2000	2001						
Fixed charge factor (%)	*12.8*	13.0	15.0	15.0	14.8	14.8	14.8	9.4	11.8	**9.4**	**15.0**	11.8	11.8	15
Reference plant TCR (US\$ kW^{-1})	*112*				0	0	0	0	160	**0**	**160**	480	480	0
Capture plant TCR (US\$ kW^{-1})	*1059*	1941			837	647	654	*846*	1028	**647**	**1941**	1282	1170	1493
Incremental TCR for capture (US\$ kW^{-1})	*947*	1602			837	647	654	*846*	868	**647**	**1602**	802	690	1493
Reference plant COE (US\$ MWh^{-1})	**18.8**		**18.0**	**18.0**	**20.6**	**20.6**	**20.6**		**26.0**	**18**	**26**	**27.0**	**27.0**	**21**
Capture plant COE (US\$ MWh^{-1})	**54.3**		**70.4**	**66.7**	**66.8**	**51.1**	**62.2**		**65.0**	**51**	**70**	**58.0**	**53.0**	**62.2**
Incremental COE for capture (US\$ MWh^{-1})	**35.5**	**61.7**	**52.4**	**48.7**	**46.2**	**30.6**	**41.7**	**33.2**	**39.0**	**31**	**62**	**31.0**	**26.0**	**41.2**
% increase in capital cost (over ref. plant)														
% increase in COE (over ref. plant)	**189**		**291**	**271**	**225**	**149**	**203**		**150**	**149**	**291**	**115**	**96**	**196**
Cost of CO$_2$ captured (US\$/tCO$_2$)	35	42			31	41	56	40		**31**	**56**			
Cost of CO$_2$ avoided (US\$/tCO$_2$)	45	73	67	59	56	48	66	55		**45**	**73**			46
Capture cost confidence level (see Table 3.6)	- moderate -											- moderate -		

Notes: All costs in this table are for capture only and do not include the costs of CO$_2$ transport and storage; see Chapter 8 for total CCS costs. Values in italics were adjusted from original reported values as explained below. * Reported HHV values converted to LHV assuming LHV/HHV = 0.96 for coal and 0.90 for natural gas. **Reported capital costs increased by 15% to estimate interest during construction and other owners' costs.

Table 3.9. CO_2 capture costs: Natural gas-fired power plants using current technology.

Study Assumptions and Results	**EPRI**	**NETL**	**IEA GHG**	**IEA GHG**	**CCP**	**Rubin *et al.***	**Rubin *et al.***	**Range**	
	2002	**2002**	**2004**	**2004**	**2005**	**2005**	**2005**	**min**	**max**
Reference Plant (without capture)	*					*	*		
Plant type (boiler, gas turbine, comb. cycle)	comb. cycle	comb. cycle	comb. cycle	comb. cycle	comb. cycle	comb. cycle	comb. cycle		
Reference plant size (MW)	509	379	776	776	392	507	507	**379**	**776**
Plant capacity factor (%)	65	85	85	85	95	75	50	**50**	**95**
Net plant efficiency, LHV (%)	*55.1*	57.9	55.6	55.6	57.6	*55.8*	*55.8*	**55**	**58**
Fuel cost, LHV (US$ GJ^{-1})	*2.82*	3.55	3.00	3.00	*2.96*	*4.44*	*4.44*	**2.82**	**4.44**
Reference plant emission rate (tCO_2 MWh^{-1})	0.364	0.344	0.379	0.379	0.37	0.367	0.367	**0.344**	**0.379**
Capture Plant Design									
CO_2 capture technology	MEA	MEA	MEA	KS-1	MEA	MEA	MEA		
Net plant size with capture (MW)	399	327	662	692	323	432	432	**323**	**692**
Net plant efficiency, LHV (%)	*47.4*	49.9	47.4	49.6	47.4	*47.6*	*47.6*	**47**	**50**
CO_2 capture system efficiency (%)	90	90	85	85	86	90	90	**85**	**90**
CO_2 emission rate after capture (t MWh^{-1})	0.045	0.040	0.066	0.063	0.063	0.043	0.043	**0.040**	**0.066**
CO_2 captured (Mt yr^{-1})	0.949	0.875	1.844	1.844	1.09	1.099	0.733	**0.733**	**1.844**
CO_2 product pressure (MPa)	8.4	10.3	11.0	11.0		13.7	13.7	**8**	**14**
CCS energy requirement (% more input MWh^{-1})	16	16	15	11	22	17	17	**11**	**22**
CO_2 reduction per kWh (%)	**88**	**88**	**83**	**83**	**83**	**88**	**88**	**83**	**88**
Cost Results									
Cost year basis (constant dollars)	2000	2002	2004	2004		2001	2001		
Fixed charge factor (%)			11.0	11.0	11.0	14.8	14.8	**11.0**	**14.8**
Reference plant TCR (US$ kW^{-1})	549	515	539	539	724	554	554	**515**	**724**
Capture plant TCR (US$ kW^{-1})	1099	911	938	958	1261	909	909	**909**	**1261**
Incremental TCR for capture (US$ kW^{-1})	550	396	399	419	537	355	355	**355**	**550**
Reference plant COE (US$ MWh^{-1})	**34.2**	**34.7**	**31.3**	**31.3**	34.2	**43.1**	**50**	**31**	**50**
Capture plant COE (US$ MWh^{-1})	**57.9**	**48.3**	**44**	**43.1**	51.8	**58.9**	**72**	**43**	**72**
Incremental COE for capture (US$ MWh^{-1})	**23.7**	**13.6**	**12.7**	**11.8**	**17.6**	**15.8**	**22**	**12**	**24**
% increase in capital cost (over ref. plant)	100	77	74	78	74	64	64	**64**	**100**
% increase in COE (over ref. plant)	**69**	**39**	**41**	**38**	51	**37**	**44**	**37**	**69**
Cost of CO_2 captured ($/$tCO_2$)	57	38	34	33	46	41	57	**33**	**57**
Cost of CO_2 avoided ($/$tCO_2$)	74	45	41	37	57	49	68	**37**	**74**
Capture cost confidence level (see Table 3.6)	- moderate -								

Notes: All costs in this table are for capture only and do not include the costs of CO_2 transport and storage; see Chapter 8 for total CCS costs. Values in italics were adjusted from original reported values as explained below.

* Reported HHV values converted to LHV assuming LHV/HHV = 0.90 for natural gas.

Table 3.10. CO_2 capture costs: New IGCC power plants using current technology.

Study Assumptions and Results	**NETL**	**NETL**	**NETL**	**Parsons**	**Simbeck**	**Nsakala *et al.***	**IEA GHG**	**IEA GHG**	**IEA GHG**	**Rubin *et al.***	**Rubin *et al.***	**Range**		**Stobbs /Clark**	**Stobbs /Clark**	**Stobbs /Clark**	**IEA GHG**
	2002	**2002**	**2002**	**2002b**	**2002**	**2003**	**2003**	**2003**	**2003**	**2005**	**2005**	**min**	**max**	**2005**	**2005**	**2005**	**2000**
	PLANTS WITH BITUMINOUS COAL FEEDSTOCK													**PLANTS WITH OTHER FEEDSTOCKS**			
Reference Plant (without capture)	*	*	*	*		*											
Gasifier name or type	Shell, O_2 blown, CGCU	E-gas, O_2 blown, CGUC	Texaco quench, O_2 blown	E-gas, O_2 blown	Texaco quench, O_2 blown	Texaco syngas cooler, O_2 blown	Texaco quench, O_2 blown	Texaco quench, O_2 blown	Shell, O_2 blown	Texaco quench, O_2 blown	Texaco quench, O_2 blown				Texaco quench, O_2 blown	Shell, O_2 blown	O_2 blown, partial oxidation
Fuel type (bit, sub-bit, lig; other) and %S	Illinois #6	Illinois #6	Illinois #6	bit, 2.5% S	bit, 1% S	bit	bit, 1%S	bit, 1%S	bit, 1%S	bit, 2.1%S	bit, 2.1%S			bit	sub-bit	lignite	Natural gas
Reference plant size (MW)	413	401	571	425	521		827	827	776	527	527	**401**	**827**	*(No IGCC Reference Plants(*			790
Plant capacity factor (%)	85	85	65	65	80	80	85	85	85	75	65	**65**	**85**	90	90	90	90
Net plant efficiency, LHV (%)	47.4	46.7	*39.1*	44.8	44.6		38.0	38.0	43.1	39.1	39.1	**38**	**47**				56.2
Fuel cost, LHV (US\$ GJ^{-1})	*1.03*	*1.03*	*1.28*	1.29	0.98	*1.23*	1.50	1.50	1.50	1.25	1.25	**0.98**	**1.50**	1.90	0.48	0.88	2.00
Reference plant emission rate (tCO2 MWh^{-1})	0.682	0.692	0.846	0.718	0.725		0.833	0.833	0.763	0.817	0.817	**0.68**	**0.85**				0.370
Capture Plant Design																	
CO_2 capture technology	Selexol	Selexol	Selexol	Selexol	Selexol	Selexol	Selexol	Selexol. NS	Selexol	Selexol	Selexol			Selexol	Selexol	Selexol	Selexol
Net plant size, with capture (MW)	351	359	457	404	455		730	742	676	492	492	**351**	**742**	445	437	361	820
Net plant efficiency, LHV (%)	40.1	40.1	*31.3*	38.5	39.0	31.5	31.5	32.0	34.5	33.8	33.8	**31**	**40**	32.8	27.0	28.3	48.3
CO_2 capture system efficiency (%)	89.2	87.0	89.0	91.0	91.2		85	85	85	90	90	**85**	**91**	87	92	86	85
CO_2 emission rate after capture (t MWh^{-1})	0.087	0.105	0.116	0.073	0.065	0.104	0.152	0.151	0.142	0.097	0.097	**0.07**	**0.15**	0.130	0.102	0.182	0.065
CO_2 captured (Mt yr^{-1})	1.803	1.870	2.368	1.379	2.151		4.682	4.728	4.050	2.749	2.383	**1.38**	**4.73**	3.049	4.040	3.183	2.356
CO_2 product pressure (MPa)	14.5	14.5	8.3	8.3			11.0	11.0	11.0	13.7	13.7	**8**	**14**	13.9	13.9	13.9	11.0
CCS energy requirement (% more input MWh^{-1})	18	16	25	16	14		21	19	25	16	16	**14**	**25**				14
CO_2 reduction per kWh (%)	**87**	**85**	**86**	**90**	**91**		**82**	**82**	**81**	**88**	**88**	**81**	**91**				**82**
Cost Results							**	**	**					***	***	***	**
Cost year basis (constant dollars)	2002	2002	2002	2000	2000		2002	2002	2002	2001	2001			2003	2003	2003	2000
Fixed charge factor (%)	14.8	14.8	15.0	13.8	13.0		11.0	11.0	11.0	14.8	17.3	**11**	**17**				11.0
Reference plant TCR (US\$ kW^{-1})	1370	1374	1169	1251	1486	1565	*1187*	*1187*	*1371*	1311	1311	**1169**	**1565**				447
Capture plant TCR (US\$ kW^{-1})	2270	1897	1549	1844	2067	2179	*1495*	*1414*	*1860*	1748	1748	**1414**	**2270**	*2205*	*2518*	*3247*	*978*
Incremental TCR for capture (US\$ kW^{-1})	900	523	380	593	581	614	*308*	*227*	*489*	437	437	**227**	**900**				*531*
Reference plant COE (US\$ MWh^{-1})	**40.6**	**40.9**	**43.4**	**47.7**	**43.0**	**53.0**	**45.0**	**45.0**	**48.0**	**48.3**	**61**	**41**	**61**				**21.6**
Capture plant COE (US\$ MWh^{-1})	**62.9**	**54.4**	**59.9**	**65.8**	**57.7**	**71.5**	**56.0**	**54.0**	**63.0**	**62.6**	**79**	**54**	**79**	68.4	62.1	83.9	**34.4**
Incremental COE for capture (US\$ MWh^{-1})	**22.3**	**13.5**	**16.5**	**18.1**	**14.7**	**18.5**	**11**	**9**	**15**	**14.3**	**18.2**	**9**	**22**				**12.8**
% increase in capital cost (over ref. plant)	66	38	33	47	39	39	26	19	36	33	33	**19**	**66**				119
% increase in COE (over ref. plant)	55	33	38	38	34	**35**	**24**	**20**	**31**	30	30	**20**	**55**				**59**
Cost of CO_2 captured (US\$/tCO_2)	32	19	18	30	21		13	11	19	17	21	**11**	**32**				35
Cost of CO_2 avoided (US\$/tCO_2)	37	23	23	28	22	23	16	13	24	20	25	**13**	**37**	31	33	56	42
Capture cost confidence level (see Table 3.6)	- moderate -													- moderate -			moderate

Notes: All costs in this table are for capture only and do not include the costs of CO_2 transport and storage; see Chapter 8 for total CCS costs. Values in italics were adjusted from original reported values as explained below. * Reported HHV values converted to LHV assuming LHV/HHV = 0.96 for coal. ** Reported capital costs increased by 8% to include interest during construction. ***Reported capital costs increased by 15% to estimate interest during construction and other owners' costs.

Table 3.11. CO_2 capture costs: Hydrogen and multi-product plants using current or near-commercial technology. *(continued on next page)*

	HYDROGEN AND ELECTRICITY PRODUCTS									
Study Assumptions and Results	**Simbeck**	**NRC**		**NRC**	**Parsons**	**Mitretek**	**Kreutz *et al.***	**Kreutz *et al.***	**Range**	
	2004	**2004**		**2004**	**2002a**	**2003**	**2005**	**2005**	**min**	**max**
Reference Plant (without capture)	*				*	*				
Plant products (primary/secondary)	H_2	H_2		H_2	H_2+electricity	H_2+electricity	H_2+electricity	H_2+electricity		
Production process or type	Steam reforming	Steam reforming		Texaco quench, CGCU	Conv E-Gas, CGCU, H_2SO_4 co-product	Texaco quench, CGCU, Claus/Scot sulphur co-product	Texaco quench	Texaco quench		
Feedstock	Natural gas	Natural gas		Coal	Pgh #8 Coal	Coal	Coal	Coal		
Feedstock cost, LHV ($ GJ^{-1})	*5.26*	4.73		1,20	*0.89*	*1.03*	1.26	1.26	**0.89**	**5.26**
Ref. plant input capacity, LHV ($GJ\ h^{-1}$)	*9848*	7235		8861	*2627*	*2954*	6706	6706	**2627**	**9848**
Ref plant output capacity, LHV: Fuels ($GJ\ h^{-1}$)	*7504*	5513		6004	*1419*	*1579*	3853	3853	**1419**	**7504**
Electricity (MW)	-44	−32		−121	38	20	78	78	**−121**	**78**
Net plant efficiency, LHV (%)	74.6	74.6		62.9	59.2	55.9	61.7	61.7	**55.9**	**74.6**
Plant capacity factor (%)	90	90		90	80	85	80	80	**80**	**90**
CO_2 emitted ($MtCO_2\ yr^{-1}$)	4.693	3.339		7.399	1.795	2.148	4.215	4.215	**1.80**	**7.40**
Carbon exported in fuels ($MtC\ yr^{-1}$)	0	0		0	0	0	0	0	**0**	**0**
Total carbon released (kg CO_2 GJ^{-1} products)	81	78		168	164	174	145	145	**78**	**174**
Capture Plant Design										
CO_2 capture/separation technology	Amine scrubber, SMR flue ga	MEA scrubber		Not reported	Selexol	Not reported	Selexol	CO_2 H_2S co-capture, Selexol		
Capture plant input capacity, LHV ($GJ\ h^{-1}$)	*11495*	8339		8861	*2627*	*2954*	6706	6706	**2627**	**11495**
Capture plant output capacity, LHV: Fuels ($GJ\ h^{-1}$)	*7504*	6004		6004	*1443*	*1434*	3853	3853	**1434**	**7504**
Electricity (MW)	−129	−91		−187	12	27	39	35	**−187**	**39**
Net plant efficiency, LHV (%)	61.2	68.1		60.2	56.6	51.8	59.5	59.3	**51.8**	**68.1**
CO_2 capture efficiency (%)**	90	90		90	92	87	91	95	**87**	**95**
CO_2 emitted ($MtCO_2\ yr^{-1}$)***	1.280	0.604		1.181	0.143	0.279	0.338	0.182	**0.14**	**1.280**
Carbon exported in fuels ($MtC\ yr^{-1}$)	0	0		0	0	0	0	0	**0.00**	**0**
Total carbon released ($kgCO_2$ GJ^{-1} products)	23.0	13.5		28.1	13.7	24.5	12.1	6.5	**6.5**	**28.1**
CO_2 captured ($MtCO_2\ yr^{-1}$)	4.658	3.378		6.385	1.654	1.869	3.882	4.037	**1.7**	**6.4**
CO_2 product pressure (MPa)	13.7	13.7		13.7	13.4	20	15	15	**13.4**	**20.0**
CCS energy requirement (% more input/GJ plant output)	21.8	9.5		4.5	4.7	7.9	3.6	3.9	**3.6**	**21.8**
CO_2 reduction per unit product (%	**72**	**83**		**83**	**92**	**86**	**92**	**96**	**72**	**96**
Cost Results										
Cost year basis (constant dollars)	2003	2000		2000	2000	2000	2002	2002		
Fixed charge rate (%)	20.0	16.0		16.0	14.3	13.0	15.0	15.0	**13.0**	**20.0**
Reference plant TCR (million US$)****	*668*	*469*		*1192*	*357*	*365*	*887*	*887*	**357**	**1192**
Capture plant TCR (million US$)****	*1029*	*646*		*1218*	*415*	*409*	*935*	*872*	**409**	**1218**
% increase in capital cost (%)	54.1	37.7		2.2	16.5	11.9	5.4	−1.7	**−1.7**	**54.1**
Ref. plant electricity price (US$ MWh^{-1})	50.0	45.0		45.0	30.8	35.6	46.2	46.2	**30.8**	**50.0**
Capture plant electricity price (US$ MWh^{-1})	50.0	45.0		45.0	30.8	53.6	62.3	60.5	**30.8**	**62.3**
% increase in assumed electricity price	0.0	0.0		0.0	0.0	50.6	34.8	31.0	**0.0**	**50.6**
Ref. plant fuel product cost, LHV (US$ GJ^{-1})	***10.03***	**8.58**		**7.99**	***6.51***	***7.29***	**7.19**	**7.19**	6.51	10.03
Capture plant fuel product cost, LHV (US$ GJ^{-1})	***13.29***	**10.14**		**8.61**	***7.90***	***8.27***	**7.86**	**7.52**	7.52	13.29
Increase in fuel product cost (US$ GJ^{-1})	**3.26**	**1.56**		**0.62**	**1.38**	**0.98**	**0.67**	**0.32**	0.32	3.26
% increase in fuel product cost	**32.5**	**18.2**		**7.7**	**21.1**	**13.4**	**9.3**	**4.5**	**4.5**	**32.5**
Cost of CO_2 captured (US$/$tCO_2$)	38.9	20.7		4.1	8.7	6.0	4.8	2.2	**2.2**	**38.9**
Cost of CO_2 avoided (US$/$tCO_2$)	56.3	24.1		4.4	9.2	6.5	5.0	2.3	**2.3**	**56.3**
Confidence level (see Table 3.6)	high			high	- moderate -					

Notes: All costs in this table are for capture only and do not include the costs of CO_2 transport and storage; see Chapter 8 for total CCS costs. Values in italics were adjusted from original reported values as explained below. * Reported HHV values converted to LHV assuming LHV/HHV = 0.96 for coal, 0.846 for hydrogen, and 0.93 for F-T liquids. **CO_2 capture efficiency = (C in CO_2 captured/stored)/(C in fossil fuel input to plant − C in carbonaceous fuel products of plant)*100; C associated with imported electricity is not included. ***Includes CO_2 emitted in the production of electricity imported by the plant. ****Reported total plant investment values increased by 3.5% to estimate total capital requirement.

Table 3.11. *Continued.*

	LIQUID FUEL AND ELECTRICITY PRODUCTS										
Study Assumptions and Results	**Mitretek**	**Larson/ Ren**	**Larson/ Ren**	**Larson/ Ren**	**Larson/ Ren**	**Celik *et al.***	**Celik *et al.***	**Celik *et al.***	**Celik *et al.***	**Range**	
	2001	**2003**	**2003**	**2003**	**2003**	**2005**	**2005**	**2005**	**2005**	**min**	**max**
Reference Plant (without capture)	*										
Plant products (primary/secondary)	F-T liquids + electricity	MeOH +electricity	MeOH +electricity	DME +electricity	DME +electricity	DME + electricity	DME + electricity	DME + electricity	DME + electricity		
Production process or type	Unspecified O_2-blown gasifier, unspecified synthesis reactor	Texaco quench, Liquid phase reactor, Once-through config,	Texaco quench, Liquid phase reactor, Once-through config,	Texaco quench, Liquid phase reactor, Once-through config,	Texaco quench, Liquid phase reactor, Once-through config,	Texaco quench, Liquid phase reactor, Once-through config,	Texaco quench, Liquid phase reactor, Once-through config,	Texaco quench, Liquid phase reactor, Once-through config,	Texaco quench, Liquid phase reactor, Once-through config,		
Feedstock	Coal	Coal	Coal	Coal	Coal	Coal	Coal	Coal	Coal		
Feedstock cost, LHV (US$ GJ^{-1})	*1,09*	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	**1.00**	**1.09**
Ref. plant input capacity, LHV (GJ h^{-1})	*16136*	9893	9893	8690	8690	7931	7931	7931	7931	**7931**	**16136**
Ref plant output capacity, LHV: Fuels (GJ h^{-1})	*7161*	2254	2254	2160	2160	2161	2161	2161	2161	**2160**	**7161**
Electricity (MW)	697	625	625	552	552	490	490	490	490	**490**	**697**
Net plant efficiency, LHV (%)	59.9	45.5	45.5	47.7	47.7	49.5	49.5	49.5	49.5	**45.5**	**59.9**
Plant capacity factor (%)	90	85	85	85	85	80	80	80	80	**80**	**90**
CO_2 emitted ($MtCO_2$ yr^{-1})	8.067	5.646	5.646	4.895	4.895	4.077	4.077	4.077	4.077	**4.08**	**8.07**
Carbon exported in fuels (MtC yr^{-1})	1.190	0.317	0.317	0.334	0.334	0.274	0.274	0.274	0.274	**0.27**	**1.19**
Total carbon released ($kgCO_2$ GJ^{-1} products)	163	203	203	198	198	185	185	185	185	**163**	**203**
Capture Plant Design											
CO_2 capture/separation technology	Amine scrubber	Selexol	CO_2 H_2S co-capture. Selexol	Selexol	CO_2 H_2S co-capture. Selexol	CO_2 H_2S co-capture. Rectisol	CO_2 H_2S co-capture. Rectisol	CO_2 H_2S co-capture. Rectisol	CO_2 H_2S co-capture. Rectisol		
Capture plant input capacity, LHV (GJ h^{-1})	*16136*	9893	9893	8690	Coal	7931	7931	7931	7931	**7931**	**16136**
Capture plant output capacity LHV: Fuels (GJ h^{-1})	*7242*	2254	2254	2160	2160	2161	2160	2160	2160	**2160**	**7242**
Electricity (MW)	510	582	577	531	527	469	367	365	353	**353**	**582**
Net plant efficiency, LHV (%)	56.3	44.0	43.8	46.9		48.5	43.9	43.8	43.2	**43**	**56**
CO_2 capture efficiency (%)**	91	58	63	32	37	36	89	92	97	**32**	**97**
CO_2 emitted ($MtCO_2$ yr^{-1})***	0.733	2.377	2.099	3.320	3.076	2.598	0.390	0.288	0.028	**0.03**	**3.32**
Carbon exported in fuels (MtC yr^{-1})	1.2	0.317	0.317	0.294	0.294	0.274	0.274	0.274	0.274	**0.274**	**1.200**
Total carbon released ($kgCO_2$ GJ^{-1} products)	71.7	109.2	101.0	144.9	137.4	134	57	53	43	**43**	**145**
CO_2 captured ($MtCO_2$ yr^{-1})	7.260	3.269	3.547	1.574	1.819	1.479	3.692	3.790	4.021	**1.48**	**7.26**
CO_2 product pressure (MPa)	13.8	15	15	15	15	15	15	15	15	**14**	**15**
CCS energy requirement. (% more input/GJ plant output)	6.5	3.6	4.0	1.9		2.0	12.8	13.0	14.5	**1.9**	**14.5**
CO_2 reduction/unit product (%)	**56**	**46**	**50**	**27**	**31**					**27**	**56**
Cost Results											
Cost year basis (constant dollars)						2003	2003	2003	2003		
Fixed charge rate (%)	12.7	15.0	15.0	15.0	15.0	15.0	15.0	15.0	15.0	**12.7**	**15.0**
Reference plant TCR (million US$)****	*2160*	*1351*	*1351*	*1215*	*1215*	*1161*	*1161*	*1161*	*1161*	**1161**	**2160**
Capture plant TCR (million US$)****	*2243*	*1385*	*1220*	*1237*	*1090*	*1066*	*1128*	*1164*	*1172*	**1066**	**2243**
% increase in capital cost (%)	3.8	2.6	−9.7	1.8	−10.3	−8.1	−2.8	0.2	0.9	**−10.3**	**3.8**
Ref. plant electricity price (US$ MWh^{-1})	35.6	42.9	42.9	42.9	42.9	44.1	44.1	44.1	44.1	**35.6**	**44.1**
Capture plant electricity price (US$ MWh^{-1})	53.6	42.9	42.9	42.9	42.9	58.0	58.0	58.0	58.0	**42.9**	**58.0**
% increase in assumed elec. price	50.5	0.0	0.0	0.0	0.0	31.5	31.5	31.5	31.5	**0.0**	**50.5**
Ref. plant fuel product cost, LHV (US$ GJ^{-1})	***5.58***	**9.12**	**9.12**	**8.68**	**8.68**	**7.41**	**7.41**	**7.41**	**7.41**	**5.6**	**9.1**
Capture plant fuel product cost, LHV (US$ GJ^{-1})	***5.43***	**10.36**	**8.42**	**9.37**	**7.57**	**6.73**	**7.18**	**7.65**	**8.09**	**5.4**	**10.4**
Increase in fuel product cost (US$ GJ^{-}	**−0.15**	**1.24**	**−0.70**	**0.69**	**−1.11**	**−0.68**	**−0.23**	**0.24**	**0.68**	**−1.1**	**1.2**
% increase in fuel product cost	**−5.7**	**13.6**	**−7.7**	**7.9**	**−12.8**	**−9.2**	**−3.1**	**3.2**	**9.2**	**−12.8**	**13.6**
Cost of CO_2 captured (US$/$tCO_2$)		12.3	−6.4	13.3	−18.4	−12.4	−1.5	1.5	4.1	**−18.4**	**13.3**
Cost of CO_2 avoided (US$/$tCO_2$)		13.2	−6.9	13.0	−18.3	−13.3	−1.8	1.8	4.8	**−18.3**	**13.2**
Confidence level (see Table 3.6)	moderate	moderate	moderate	- low to moderate -							

Notes: All costs in this table are for capture only and do not include the costs of CO_2 transport and storage; see Chapter 8 for total CCS costs. Values in italics were adjusted from original reported values as explained below. * Reported HHV values converted to LHV assuming LHV/HHV = 0.96 for coal, 0.846 for hydrogen, and 0.93 for F-T liquids. **CO_2 capture efficiency = (C in CO_2 captured/stored)/(C in fossil fuel input to plant − C in carbonaceous fuel products of plant)*100; C associated with iimported electricity is not included. ***Includes CO_2 emitted in the production of electricity imported by the plant. ****Reported total plant investment values increased by 3.5% to estimate total capital requirement.

Table 3.12. Capture costs: Other industrial processes using current or advanced technology.

	CURRENT TECHNOLOGY									ADVANCED TECHNOLOGY			
Study Assumptions and Cost Results	**Farla *et al.***	**IEA GHG**	**IEA GHG**	**IEA GHG**	**Möllersten *et al.***	**Möllersten *et al.***	**Möllersten *et al.***	**CCP**	**CCP**	**CCP**	**CCP**	**CCP**	**CCP**
	1995	**2000**	**2000**	**2002**	**2003**	**2003**	**2003**	**2005**	**2005**	**2005**	**2005**	**2005**	**2005**
Reference Plant (without capture)													
Industrial process	Iron production	Oil refining petrochemical	Oil refining petrochemical	High purity CO_2 sources	Pulp mill	Pulp mill	Ethanol fermentation	Refinery heaters & boilers	Small gas turbines	Refinery heaters & boilers	Refinery heaters & boilers	Small gas turbines	Small gas turbines
Feedstock type	Coke	Refinery gas / natural gas	Refinery gas / natural gas		Black liquor and bark	Black liquor	Sugar cane	Mixed	NG	Mixed	Mixed	Natural gas	Natural gas
Plant size (specify units)	168 kg s^{-1} iron	315 kg s^{-1} crude oil	315 kg s^{-1} crude oil		17.9 kg s^{-1} pulp	17.9 kg s^{-1} pulp	9.1 kg s^{-1} ethanol	1351	358	1351	1351	358	358
Plant capacity factor (%)	95.3	90	90	90	90.4	90.4	49.3	90.4	98.5	90.4	90.4	98.5	98.5
Feedstock cost (US$/unit specified)					\$3 GJ s^{-1} LHV	\$3 GJ s^{-1} LHV							
Ref. plant emission rate (kgCO_2/unit product)								*2.96*	*2.96*	*2.96*	*2.96*	*2.96*	*2.96*
Capture Plant Design								0.22	0.82	0.22	0.22	0.82	0.82
CO_2 capture/separation technology	MDEA	MEA	Pre-combustion	Compression only	Amine	Physical solvent							
Location of CO_2 capture	Blast furnace gas	Fired heaters and H_2 plant	Fired heaters and H_2 plant		Boiler	IGCC	Fermentation and bagasse boiler	MEA Baseline (post-comb.)	MEA Baseline (post-comb.)	Membrane Water Gas Shift without DOE-membrane. GRACE (pre-comb.)	Flue Gas Recycle & ITM (oxy-fuel)	Very Large-scale ATR (pre-comb.)	Sorption Enhanced Water Gas Shift (pre-comb.)
Capture unit size (specify units)					392 MW fuel	338 MW fuel		1351	358	1351	1351	358	358
CO_2 capture system efficiency (%)	90	95	91		90	90	100 / 90						
Energy source(s) for capture (type +onsite or offsite)													
Are all energy-related CO_2 emissions included?								0.09	0.19	0.09	0.05	0.10	0.14
CO_2 emission rate after capture (kgCO_2/unit product)								2.19	1.90	1.99	2.09	2.88	2.50
CO_2 captured (Mt yr^{-1})	2.795	1.013	1.175	1.970	0.969	0.399	0.560						
CO_2 product pressure (MPa)	11.0	11.0	11.0	8.0	10.0	10.0	10.0						
CO_2 reduction per unit of product (%)								**60.3**	**76.5**	**58.4**	**75.8**	**87.4**	**82.2**
Cost Results													
Cost year basis (constant dollars)													
Fixed charge factor (%)					15	15	15	11.0	11.0	11.0	11.0	11.0	11.0
Ref. plant capital cost (US$/unit capacity)													
Capture plant capital cost (US$/unit capacity)													
Incremental capital cost (million US$ per kg/s CO_2)*	3.8	4.1	4.9	0.3	3.2	1.9	2.6	0	0	0	0	0	0
Ref. plant cost of product (US$/unit)								0.0	0.0	0.0	0.0	0.0	0.0
Capture plant cost of product (US$/unit)								10.2	55.1	6.1	6.8	54.2	48.2
Incremental cost of product (US$/unit)								**10.2**	**55.1**	**6.1**	**6.8**	**54.2**	**48.2**
% increase in capital cost (over ref. plant)													
% increase in unit cost of product (over ref. plant)													
Cost of CO_2 captured (US$/t$CO_2$)		50	60					55.3	90.9	36.4	38.2	59.0	60.5
Cost of CO_2 avoided (US$/t$CO_2$)	35	74	116	10	34	23	53	78.1	88.2	48.1	41.0	76.0	71.8
Capture cost Confidence level (see Table 3.6)	- moderate -									- low -			

Notes: All costs in this table are for capture only and do not include the costs of CO_2 transport and storage; see Chapter 8 for total CCS costs. Values in italics were adjusted from original reported values as explained below.

*Capital costs are incremental costs of capture, excluding cost of make-up steam and power generation and also excluding interest during construction and other owner's costs.

Table 3.13. Capture costs: Advanced technologies for electric power plants. *(continued on next page)*

	OXY-FUEL COMBUSTION								**ADVANCED PC**	
Study Assumptions and Results	**Alstom et *al.***	**Singh et *al.***	**Stobbs/Clark**	**Dillon et *al.***	**Nsakala et *al.***	**Nsakala et *al.***	**Nsakala et *al.***		**Gibbins et *al.***	**Gibbins et *al.***
	2001	**2003**	**2005**	**2005**	**2003**	**2003**	**2003**		**2005**	**2005**
Reference Plant (without capture)	*				*	*	*			
Power plant type	RETROFIT subcrit PC	RETROFIT PC + aux NGCC	RETROFIT PC	New PC	Air-fired CFB	Air-fired CFB	Air-fired CFB			
Fuel type (bit, sub-bit, lig; NG, other) and %S	bit, 2.7%S	sub-bit	lignite	bit	bit, 2.3%S	bit, 2.3%S	bit, 2.3%S			
Reference plant net size (MW)	434	400	300	677	193	193	193			
Plant capacity factor (%)	67	91		85	80	80	80		85	85
Net plant efficiency, LHV (%)				44.2	*37.0*	*37.0*	*37.0*		45,6	45,6
Fuel cost, LHV (US$ GJ^{-1})	*1.30*			1.50	*1.23*	*1.23*	*1.23*		1,50	1,50
Reference plant emission rate (tCO_2 MWh^{-1})	0.908	0.925	0.883	0.722	0.909	0.909	0.909			
Capture Plant Design										
CO_2 capture technology	oxy-fuel	oxy-fuel	oxy-fuel	oxy-fuel	oxy-fuel	oxy-fuel with CMB	chemical looping with CMB			
Net plant size with capture (MW)	273	400		532	135	197	165			
Net plant efficiency, LHV (%)	*23.4*			35.4	*25.8*	*31.3*	*32.2*		34,3	36,5
CO_2 capture system efficiency (%)				*about91*						
CO_2 emission rate after capture (t MWh^{-1})		0.238	0.145	0.085	0.086	0.073	0.005			
CO_2 captured (Mt yr^{-1})		2.664								
CO_2 product pressure (MPa)	13.9	15	13.7	11					11,0	11,0
CCS energy requirement (% more input MWh^{-1})				25	43	18	15		33	25
CO_2 reduction per kWh (%)		**74**		**88.2**	**90.5**	**92.0**	**99.5**			
Cost Results	**	**								
Cost year basis (constant dollars)		2001	2000		2003	2003	2003			
Fixed charge factor (%)	13.0	9.4		*11*					*11,0*	*11,0*
Reference plant TCR (US$ kW^{-1})		0		1260	1500	1500	1500		1022	1022
Capture plant TCR (US$ kW^{-1})	*1527*	*909*	*4570*	1857	2853	2731	1912		1784	1678
Incremental TCR for capture (US$ kW^{-1})	*1198*	*909*		597	1354	1232	413		762	656
Reference plant COE (US$ MWh^{-1})			**44.5**	**44**	**45.3**	**45.3**	**45.3**		**37**	**37**
Capture plant COE (US$ MWh^{-1})			**97.5**	**61.2**	**82.5**	**70.5**	**58.4**		**61**	**57**
Incremental COE for capture (US$ MWh^{-1})	44.5	23.9	53	17.2	37.2	25.2	13.1		24	20
% increase in capital cost (over ref. plant)				47	90	82	28		75	64
% increase in COE (over ref. plant)			**119**	**39**	**82**	**56**	**29**		**65**	**54**
Cost of CO_2 captured (US$/$tCO_2$)		29								
Cost of CO_2 avoided (US$/$tCO_2$)	54	35	*72*	27	45	30	14			
Capture cost confidence level (see Table 3.6)	- low -					very low	very low			

Notes: All costs in this table are for capture only and do not include the costs of CO_2 transport and storage; see Chapter 8 for total CCS costs. Values in italics were adjusted from original reported values as explained below.
**Reported value increased by 15% to estimate interest during construction and other owners' costs.

Table 3.13. *Continued.*

	ADVANCED NGCC								**ADVANCED IGCC**					**ADVANCED HYBRIDS**	
Study Assumptions and Results	**Simbeck**	**Parsons**	**Parsons**	**CCP**	**CCP**	**CCP**	**CCP**	**Dillon *et al.***	**Parsons**	**NETL**	**NETL**	**CCP**	**CCP**	**NETL**	**Parsons**
	2002	**2002b**	**2002b**	**2005**	**2005**	**2005**	**2005**	**2005**	**2002b**	**2002**	**2002**	**2005**	**2005**	**2002**	**2002b**
Reference Plant (without capture)									*	*	*			*	
Power plant type	comb.cycle H-class turbine	comb.cycle H-class turbine	comb.cycle H-class turbine					NGCC	E-gas, O$_2$, water scrubber; H-class turbine	E-gas, O$_2$, CGCU, Hydraulic air compression	E-gas, O$_2$, CGCU, Hydraulic air compression with open loop water system	Canada coke gasification	Canada coke gasification		CHAT SOFC
Fuel type (bit, sub-bit, lig; NG, other) and %S	Nat. gas	Nat. gas	Nat. gas	NG	NG	NG	NG	NG	Illinois #6	Illinois #6	Illinois #6	Coke	Coke	Illinois #6	Nat. gas
Reference plant net size (MW)	480	384	384	392	392	392	507	388	425	326	408	588	588	644	557
Plant capacity factor (%)	80	65	80	95	95	95	95	85	80	85	85	91.3	91.3	85	80
Net plant efficiency, LHV (%)	60.0	59.5	59.5	57.6%	57.6%	57.6%	57.6%	56.0	41.1	43.8	54.9			56.4	66.2
Fuel cost, LHV (US$ GJ^{-1})	4.86	2.82	2.82	*2.96*	*2.96*	*2.96*	*2.96*	3.00	*1.23*	*1.03*	*1.03*	*2.96*	*2.96*	*1.03*	2.82
Reference plant emission rate (tCO$_2$ MWh^{-1})	0.342	0.338	0.338	0.37	0.37	0.37	0.37	0.371	0.720	0.712	0.568	0.95	0.95	0.572	0.302
Capture Plant Design															
CO$_2$ capture technology	MEA	MEA	MEA	MEA low-cost/ CCGT-integrated (post-comb.)	Membrane Contactor; KS-1 (post-comb.)	Hydrogen Membrane Reformer (pre-comb.)	Sorption Enhanced Water Gas Shift- Air ATR (pre-comb)	Oxy-fuel	Selexol	Selexol		IGCC with capture (pre-comb.)	IGCC with advanced capture (pre-comb.)		
Net plant size with capture (MW)	413	311	311	345	335	361	424	440	387	312	404	699	734	755	517
Net plant efficiency, LHV (%)	51.7	48.1	48.1	50.6	49.2	53.0	48.2	44.7	33.8	35.2	45.4			49.7	46.1
CO$_2$ capture system efficiency (%)	85	90	90	86	86	100	90		91.5	92.7	92.7			90	86.8
CO$_2$ emission rate after capture (t/MWh)	0.06	0.042	0.042	0.06	0.06	0.00	0.04	0.011	0.074	0.065	0.050	0.27	0.28	0.046	0.043
CO$_2$ captured (Mt yr^{-1})	0.980	0.669	0.823	1.09	1.09	1.27	1.47		2.074	1.984	1.984	6.80	6.44	3.390	
CO$_2$ product pressure (MPa)	13.7	8.3	8.3					11	8.3	14.5	14.5			14.5	8.3
CCS energy requirement(% more input MWh^{-1})	16	24	24	−99	−99	−99	−99	25	22	24	21			13	44
CO$_2$ reduction per kWh (%)	**82**	**88**	**88**	**84.1**	**83.6**	**100**	**87.9**	**97.0**	**90**	**91**	**91**	**71.2**	**71.1**	**92**	**86**
Cost Results															
Cost year basis (constant dollars)	2001	2000	2000						2000	2002	2002			2002	2000
Fixed charge factor (%)	15.0			11.0	11.0	11.0	11.0	11	15.0	14.8	14.8	11.0	11.0	14.8	
Reference plant TCR (US$ kW^{-1})	*582*	539	496	724	724	724	724	559	1249	1436	881.4	1398	1398	1508	623
Capture plant TCR (US$ kW^{-1})	*1216*	1026	**943**	1002	1225	1058	1089	1034	1698	2189	1450	1919	1823	1822	
Incremental TCR for capture (US$ kW^{-1})	*634*	487	447	278	501	334	365	475	449	753	568	521	425	314	
Reference plant COE (US$ MWh^{-1})	**42.9**	**33.5**	**30.7**	34.2	34.2	34.2	34.2	**33.5**	**41.0**	**47.0**	**28.5**	32.3	32.3	**41.1**	
Capture plant COE (US$ MWh^{-1})	**65.9**	**54.1**	**48.8**	45.1	48.9	43.2	45.4	**50.3**	**53.6**	**65.5**	**41.8**	42.1	40.5	**48.8**	
Incremental COE for capture (US$ MWh^{-1})	23	20.6	18.1	10.9	14.7	9.0	11.2	16.8	12.6	18.5	13.3	9.8	8.2	7.7	
% increase in capital cost (over ref. plant)	109	90	90	38	69	46	50	85	36	52	64	37	30	21	
% increase in COE (over ref. plant)	**54**	**61**	**59**	32	43	26	33	**50**	**31**	**39**	**47**	**30**	**25**	**19**	
Cost of CO$_2$ captured (US$/tCO$_2$)		48		30.2	39.5	22.5	28.2		16	22	20	11	10	13	
Cost of CO$_2$ avoided (US$/tCO$_2$)	82	70	61	35.1	47.5	24.4	34.4	47	19	29	26	14	12	15	
Capture cost confidence level (see Table 3.6)	- low to moderate -				- low to very low -				- low -					- very low -	

Notes: All costs in this table are for capture only and do not include the costs of CO_2 transport and storage; see Chapter 8 for total CCS costs. Values in italics were adjusted from original reported values as explained below.
**Reported value increased by 15% to estimate interest during construction and other owners' costs.

Table 3.14. CO_2 capture costs: Multi-product plants using advanced technology.

Study Assumptions and Results	**Simbeck**	**NRC**	**NRC**	**Parsons**	**Mitretek**	**Mitretek**	**Mitretek**	**Range**	
	2004	**2004**	**2004**	**2002a**	**2003**	**2003**	**2003**	**min**	**max**
Capture Plant Design	*			*	*	*	*		
Plant products (primary/secondary)	H_2	H_2	H_2	H_2 +electricity	H_2 +electricity	H_2 +electricity	H_2 +electricity		
Production process or type	Autothermal reforming with O_2 provided by ITM	78% efficient ATR/SMR, adv CO_2 compressor	Gasifier LHV= 75-->80%, Adv ASU, membrane sep, adv CO_2 compressor	High-pressure E-gas, HGCU, HTMR, H2SO4 co-product	Advanced E-gas, HGCU, HTMR	Advanced E-gas, HGCU, HTMR, large elec. co-product	Advanced E-gas, HGCU, HTMR, SOFC, large elec. co-product		
Feedstock	Natural gas	Natural gas	Coal	Pgh #8 Coal	Coal	Coal	Coal		
Feedstock cost, LHV (US\$ GJ^{-1})	*5.26*	4.73	1.20	*0.89*	*1.03*	*1.03*	*1.03*	**1**	**5**
Plant capacity factor (%)	90	90	90	80	85	85	85	**80**	**90**
CO_2 capture/separation technology	Oxy-fuel	?	?	Oxy-fuel	Oxy-fuel	Oxy-fuel	Oxy-fuel		
Capture plant input capacity, LHV (GJ h^{-1})	*9527*	7697	8121	*2794*	*3020*	*6051*	*6051*	**2794**	**9527**
Capture plant output capacity, LHV: Fuels (GJ h^{-1})	*7504*	6004	6004	*1956*	*1904*	*1844*	*1808*	**1808**	**7504**
Electricity (MW)	−13	−66	−88	7	25	416	519	**−88**	**519**
Net plant efficiency, LHV (%)	78.3	74.9	70.0	70.9	66.0	55.2	60.7	**55**	**78**
CO_2 capture efficiency (%)**	95	90	90	94	100	100	95	**90**	**100**
CO_2 emitted ($MtCO_2\ yr^{-1}$)***	0.086	0.505	0.873	0.117	0.000	0.000	0.191	**0.000**	**0.873**
Carbon exported in fuels (MtC yr^{-1})	0	0	0	0	0	0	0	**0**	**0**
Total carbon released ($kgCO_2\ GJ^{-1}$ products)	1.46	11.10	19.45	8.45	0.00	0.00	6.96	**0.0**	**19.5**
CO_2 captured ($MtCO_2\ yr^{-1}$)	4.074	3.119	5.853	1.855	1.918	3.846	3.652	**1.9**	**5.9**
CO_2 product pressure (MPa)	13.7	13.7	13.7	13.4	20	20	20	**13.4**	**20.0**
Cost Results									
Cost year basis (constant dollars)	2003	2000	2000	2000	2000	2000	2000		
Fixed charge rate (%)	20	16	16	14.3	12.7	12.7	12.7	**12.7**	**20.0**
Capture plant TCR (million US\$)****	*725*	*441*	*921*	*398*	*441*	*950*	*1023*	**398**	**1023**
Capture plant electricity price (US\$ MWh^{-1})	50.0	45.0	45.0	30.8	53.6	53.6	53.6	**31**	**54**
Capture plant fuel product cost, LHV (US\$ GJ^{-1})	*9.84*	8.53	6.39	*5.79*	*6.24*	*3.27*	*1.13*	**1.13**	**9.84**
Capture cost confidence level (see Table 3.6)	low	low	low	- low to very low -			very low		

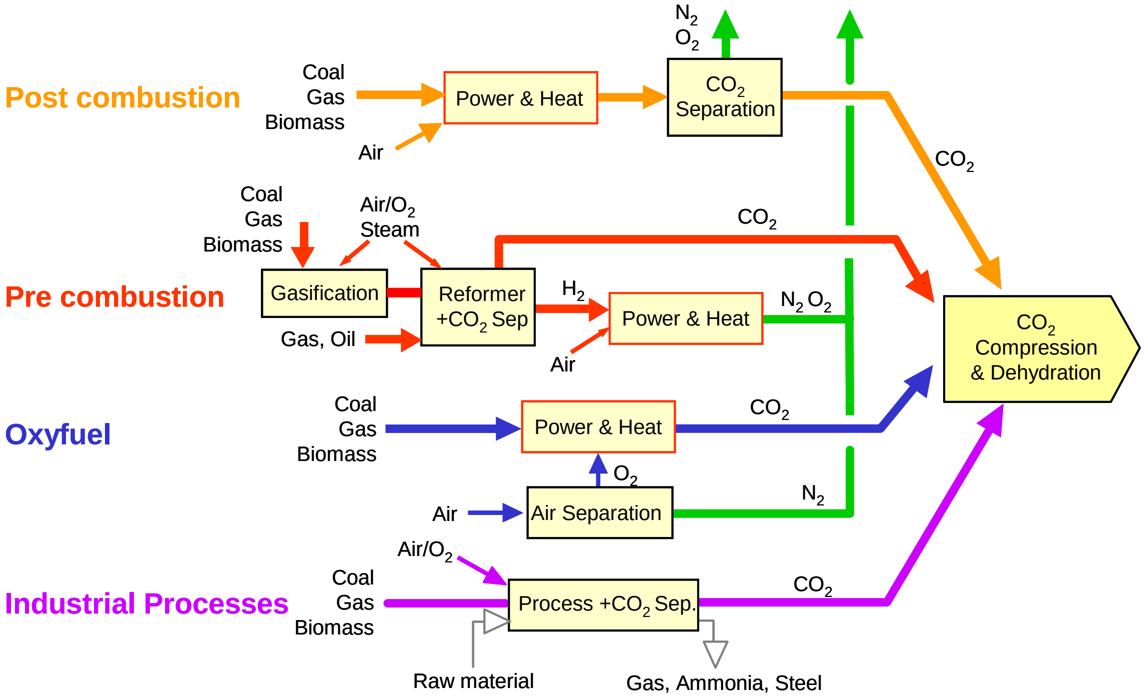
Notes: All costs in this table are for capture only and do not include the costs of CO_2 transport and storage; see Chapter 8 for total CCS costs. Values in italics were adjusted from original reported values as explained below.
* Reported HHV values converted to LHV assuming LHV/HHV = 0.96 for coal and 0.846 for hydrogen. **CO_2 capture efficiency = (C in CO_2 captured/stored)/(C in fossil fuel input to plant − C in carbonaceous fuel products of plant)*100; C associated with imported electricity is not included. ***Includes CO_2 emitted in the production of electricity imported by the plant. ****Reported total plant investment values increased by 3.5% to estimate total capital requirement.

Table 3.15. Summary of new plant performance and CO_2 capture cost based on current technology.

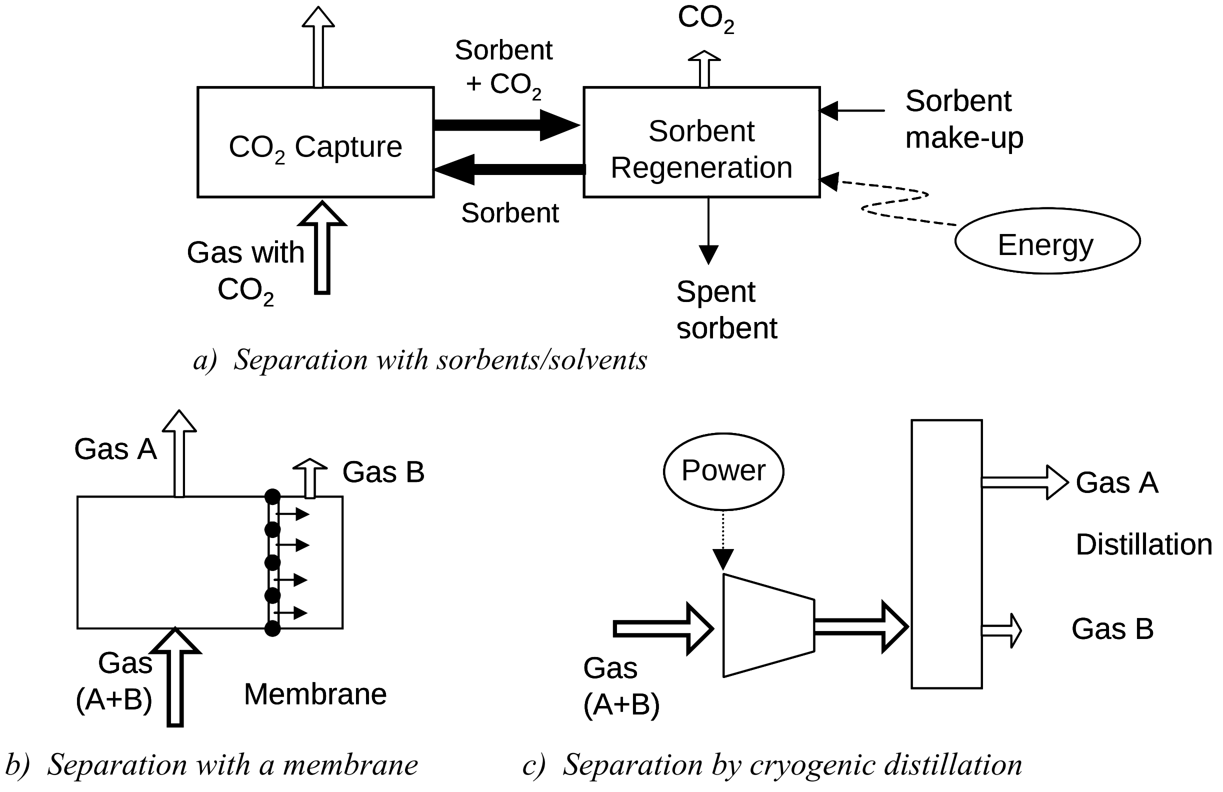
Performance and Cost Measures	**New NGCC Plant**				**New PC Plant**				**New IGCC Plant**				**New Hydrogen Plant**				**(Units for H_2 Plant)**
	Range			*Rep.*	**Range**			*Rep.*	**Range**			*Rep.*	**Range**			*Rep.*	
	low		**high**	*Value*	**low**		**high**	*Value*	**low**		**high**	*Value*	**low**		**high**	*Value*	
Emission rate without capture ($\text{kgCO}_2\ \text{MWh}^{-1}$)	344	-	379	*367*	736	-	811	*762*	682	-	846	*773*	78	-	174	*137*	$\text{kgCO}_2\ \text{GJ}^{-1}$ (without capture)
Emission rate with capture ($\text{kgCO}_2\ \text{MWh}^{-1}$)	40	-	66	*52*	92	-	145	*112*	65	-	152	*108*	7	-	28	*17*	$\text{kgCO}_2\ \text{GJ}^{-1}$ (with capture)
Percent CO_2 reduction per kWh (%)	83	-	88	*86*	81	-	88	*85*	81	-	91	*86*	72	-	96	*86*	% reduction/unit of product
Plant efficiency with capture, LHV basis (%)	47	-	50	*48*	30	-	35	*33*	31	-	40	*35*	52	-	68	*60*	Capture plant efficiency (% LHV)
Capture energy requirement (% more input MWh^{-1})	11	-	22	*16*	24	-	40	*31*	14	-	25	*19*	4	-	22	*8*	% more energy input GJ^{-1} product
Total capital requirement without capture ($\text{US\$ kW}^{-1}$)	515	-	724	*568*	1161	-	1486	*1286*	1169	-	1565	*1326*	*(No unique normalization for multi-product plants)*				Capital requirement without capture
Total capital requirement with capture ($\text{US\$ kW}^{-1}$)	909	-	1261	*998*	1894	-	2578	*2096*	1414	-	2270	*1825*					Capital requirement with capture
Percent increase in capital cost without capture (%)	64	-	100	*76*	44	-	74	*63*	19	-	66	*37*	-2	-	54	*18*	% increase in capital cost
COE without capture ($\text{US\$ MWh}^{-1}$)	31	-	50	*37*	43	-	52	*46*	41	-	61	*47*	6,5	-	10,0	*7,8*	H_2 cost without capture ($\text{US\$ GJ}^{-1}$)
COE with capture only ($\text{US\$ MWh}^{-1}$)	43	-	72	*54*	62	-	86	*73*	54	-	79	*62*	7,5	-	13,3	*9,1*	H_2 cost without capture ($\text{US\$ GJ}^{-1}$)
Increase in COE without capture ($\text{US\$ MWh}^{-1}$)	12	-	24	*17*	18	-	34	*27*	9	-	22	*16*	0,3	-	3,3	*1,3*	Increase in H_2 cost ($\text{US\$ GJ}^{-1}$)
Percent increase in COE without capture (%)	37	-	69	*46*	42	-	66	*57*	20	-	55	*33*	5	-	33	*15*	% increase in H_2 cost
Cost of CO_2 captured ($\text{US\$/tCO}_2$)	33	-	57	*44*	23	-	35	*29*	11	-	32	*20*	2	-	39	*12*	$\text{US\$/tCO}_2$ captured
Cost of CO_2 avoided ($\text{US\$/tCO}_2$)	37	-	74	*53*	29	-	51	*41*	13	-	37	*23*	2	-	56	*15*	$\text{US\$/tCO}_2$ avoided
Capture cost confidence level (see Table 3.6)	moderate				moderate				moderate				moderate to high				Confidence Level (see Table 3.6)

Notes: See Section 3.6.1 for calculation of energy requirement for capture plants. Values in italics were adjusted from original reported values as explained below.
(a) Ranges and representative values are based on data from Tables 3.8, 3.11, 3.11 and 3.12. All costs in this table are for capture only and do not include the costs of CO_2 transport and storage; see Chapter 8 for total CCS costs. (b) All PC and IGCC data are for bituminous coals only at costs of $\text{US\$}1.0{-}1.5\ \text{GJ}^{-1}$ (LHV); all PC plants are supercritical units. (c) NGCC data based on natural gas prices of $\text{US\$}2.8{-}4.4\ \text{GJ}^{-1}$ (LHV basis). (d) Cost are in constant US dollars (approx. year 2002 basis). (e) Power plant sizes range from approximately 400−800 MW without capture and 300−700 MW with capture. (f) Capacity factors vary from 65−85% for coal plants and 50−95% for gas plants (average for each = 80%). (g) Hydrogen plant feedstocks are natural gas ($\text{US\$ }4.7{-}5.3\ \text{GJ}^{-1}$) or coal ($\text{US\$ }0.9{-}1.3\ \text{GJ}^{-1}$); some plants in data set produce electricity in addition to hydrogen. (h) Fixed charge factors vary from 11−16% for power plants and 13−20% for hydrogen plants. (i) All costs include CO_2 compression but not additional CO_2 transport and storage costs (see Chapter 8 for full CCS costs).

Figures



5 **Figure 3.1.** CO_2 capture systems (adapted from BP).



a) Separation with sorbents/solvents

b) Separation with a membrane *c) Separation by cryogenic distillation*

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Figure 3.2. General schemes of the main separation processes relevant for CO_2 capture. The gas removed in the separation may be CO_2, H_2 or O_2. In Figures 3.2b and 3.2c one of the separated gas streams (A and B) is a concentrated stream of CO_2, H_2 or O_2 and the other is a gas stream with all the remaining gases in the original gas (A+B).

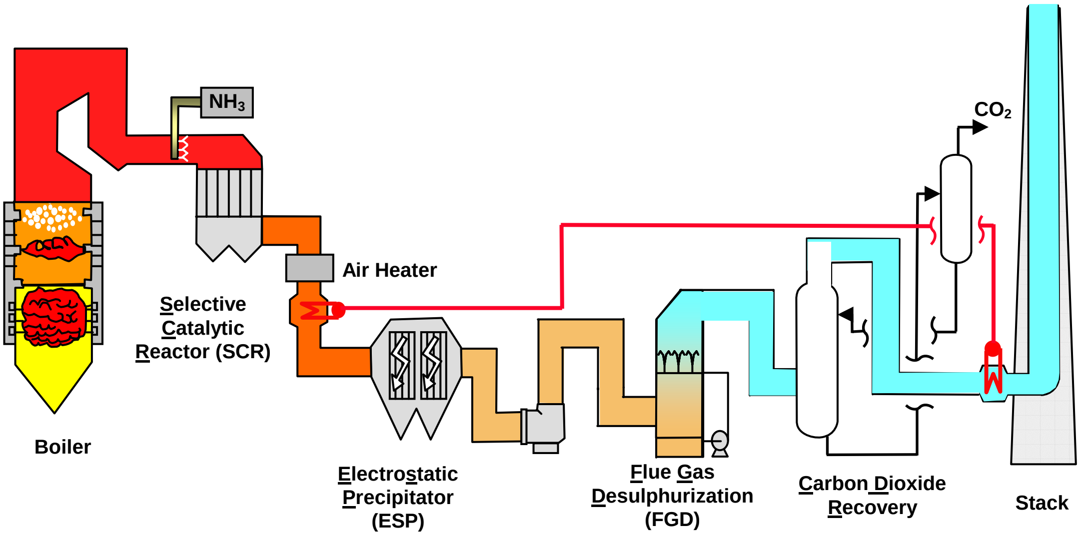

Figure 3.3. Schematic of a pulverized coal-fired power plant with an amine-based CO_2 capture system and other emission controls.

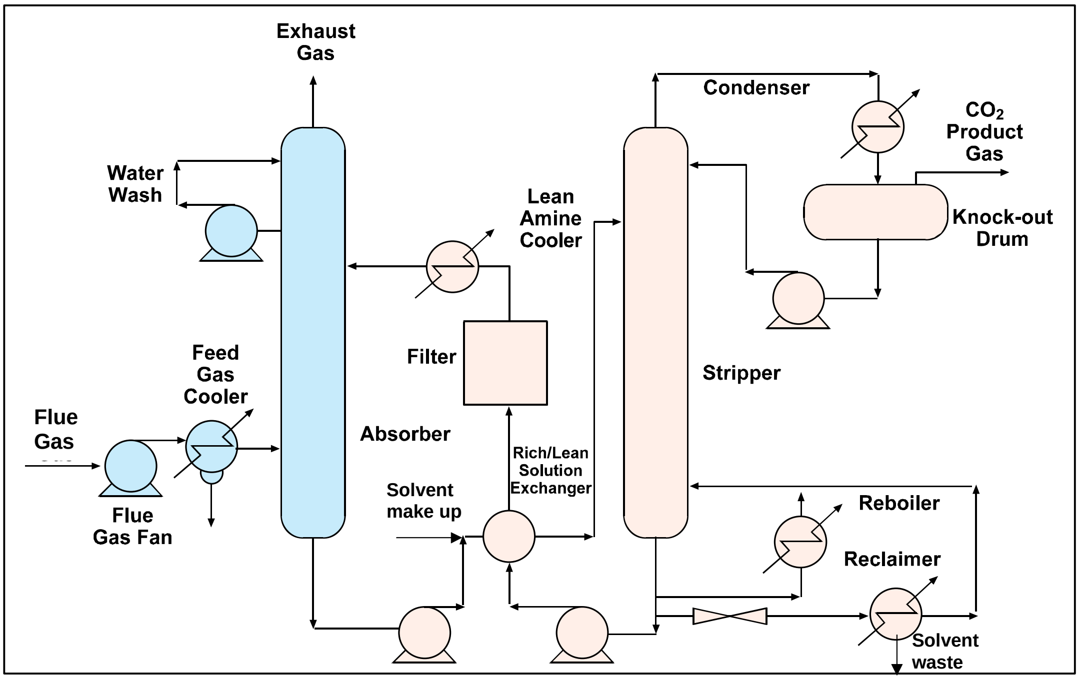

Figure 3.4. Process flow diagram for CO_2 recovery from flue gas by chemical absorption.

Figure 3.5. CO_2 capture plant in Malaysia using a 200 tonne d^{-1} KEPCO/MHI chemical solvent process (Courtesy of Mitsubishi).

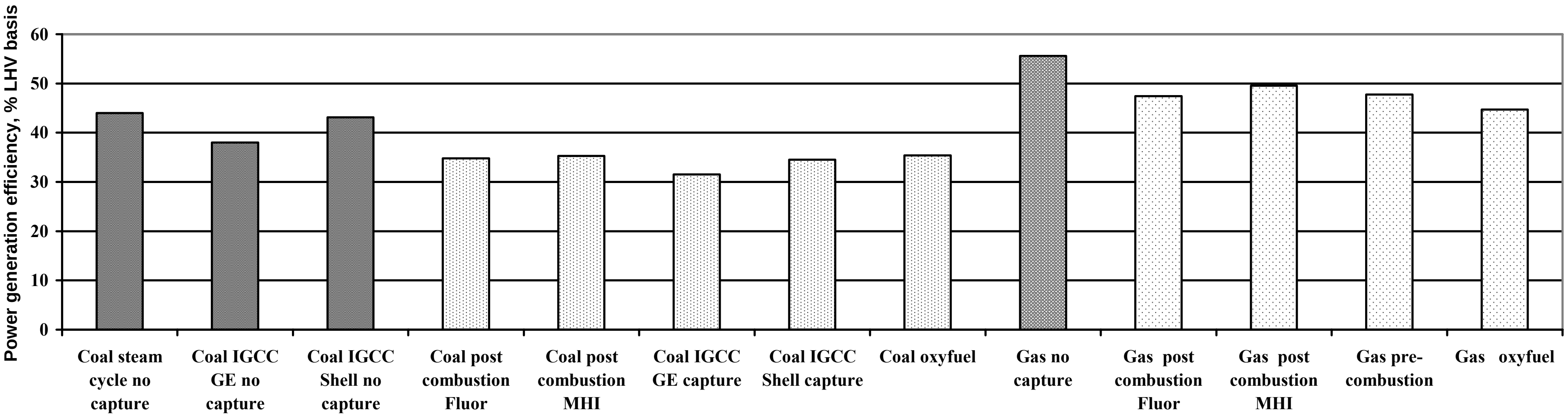

Figure 3.6. Thermal efficiencies of power plants with and without CO_2 capture, % LHV-basis (Source data: Davison 2005, IEA GHG 2004, IEA GHG 2003; IEA GHG, 2000b; Dillon *et al.*, 2005).

a. The efficiencies are based on a standard set of plant design criteria (IEA GHG, 2004).
b. The coal steam cycle plants, including the post-combustion capture and oxy-fuel plants, are based on ultra-supercritical steam (29MPa, 600C superheat, 620C reheat). The IGCC and natural gas pre- and post-combustion capture plants are based on GE 9FA gas turbine combined cycles. The natural gas oxy-fuel plant is based on a CO_2 recycle gas turbine, as shown in Figure 3.10, with different operating pressures and temperatures but similar mechanical design criteria to that of the 9FA.
c. Data are presented for two types of post-combustion capture solvent: MEA (Fluor plant designs) and KS-1 (MHI plant designs). The solvent desorption heat consumptions are 3.2 and 2.7 MJ/kgCO_2 captured respectively for the coal plants and 3.7 and 2.7 MJ kg^{-1} for the natural gas plants.
d. Data are presented for IGCC plants based on two types of gasifier: the Shell dry feed/heat recovery boiler type and the GE (formerly Texaco) slurry feed water quench type.
e. The natural gas pre-combustion capture plant is based on partial oxidation using oxygen.
f. The oxy-fuel plants include cryogenic removal of some of the impurities from the CO_2 during compression. Electricity consumption for oxygen production by cryogenic distillation of air is 200 kWh/ tO_2 at atmospheric pressure for the coal plant and 320 kWh/ tO_2 at 40 bar for the natural gas plant. Oxygen production in the IGCC and natural gas pre-combustion capture plants is partially integrated with the gas turbine compressor, so comparable data cannot be provided for these plants.
g. The percentage CO_2 capture is 85–90% for all plants except the natural gas oxy-fuel plant which has an inherently higher percentage capture of 97%.

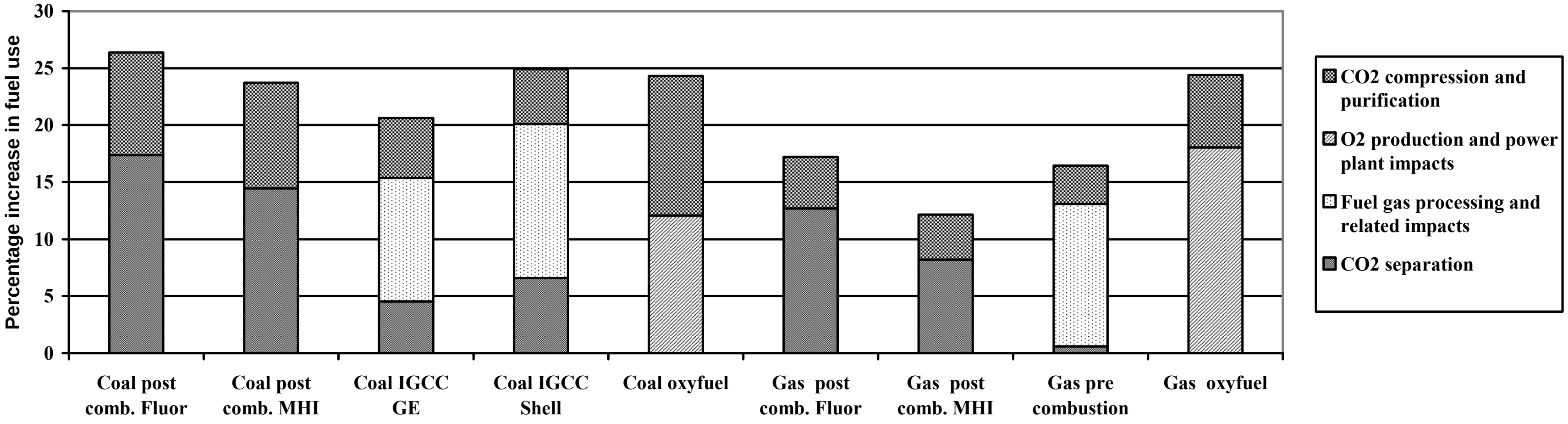

Figure 3.7. Percentage increase in fuel use per kWh of electricity due to CO_2 capture, compared to the same plant without capture (Source data: Davison, 2005; IEA GHG, 2004; IEA GHG, 2003; IEA GHG, 2000b; Dillon *et al.*, 2005).

a. The increase in fuel required to produce a kWh of electricity is calculated by comparing the same type of plant with and without capture. The increase in fuel consumption depends on the type of baseline plant without capture. For example, the increase in energy consumption for a GE IGCC plant with capture compared to a coal steam cycle baseline plant without capture would be 40% as opposed to the lower value shown in the figure that was calculated relative to the same type of baseline plant without capture.
b. The direct energy consumptions for CO_2 separation are lower for pre-combustion capture than for post-combustion capture, because CO_2 is removed from a more concentrated, higher pressure gas, so a physical rather than a chemical solvent can be used.
c. The 'Fuel gas processing and related impacts' category for IGCC includes shift conversion of the fuel gas and the effects on the gas turbine combined cycle of removal of CO_2 from the fuel gas and use of hydrogen as a fuel instead of syngas. For natural gas pre-combustion capture this category also includes partial oxidation/steam reforming of the natural gas.
d. The energy consumption for CO_2 compression is lower in pre-combustion capture than in post-combustion capture because some of the CO_2 leaves the separation unit at elevated pressure.
e. The energy consumption for CO_2 compression in the oxy-fuel processes depends on the composition of the extracted product, namely 75% by volume in the coal-fired plant and 93% by volume in the gas fired plant. Impurities are cryogenically removed from the CO_2 during compression, to give a final CO_2 purity of 96% by volume. The energy consumption of the cryogenic CO_2 separation unit is included in the CO_2 compression power consumption.
f. The 'Oxygen production and power plant impacts' category for oxy-fuel processes includes the power consumption for oxygen production and the impacts of CO_2 capture on the rest of the power plant, that is excluding CO_2 compression and purification. In the coal-fired oxy-fuel plant, the efficiency of the rest of the power plant increases slightly, for example due to the absence of a flue gas desulphurization (FGD) unit. The efficiency of the rest of the gas fired oxy-fuel plant decreases because of the change of working fluid in the power cycle from air to recycled flue gas.

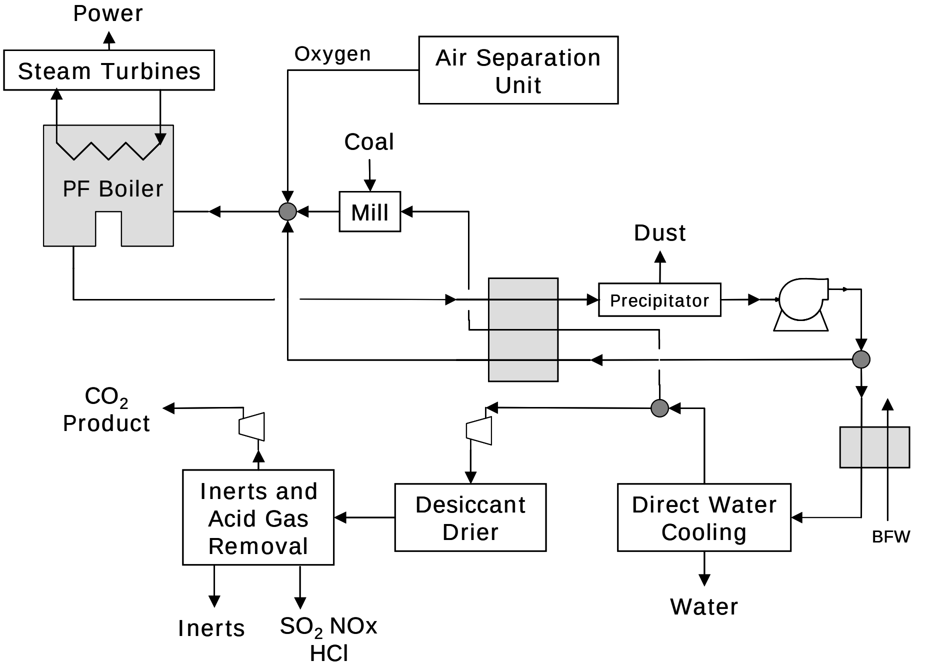

Figure 3.8. Schematic of an oxy-fuel, pulverized coal fired power plant.

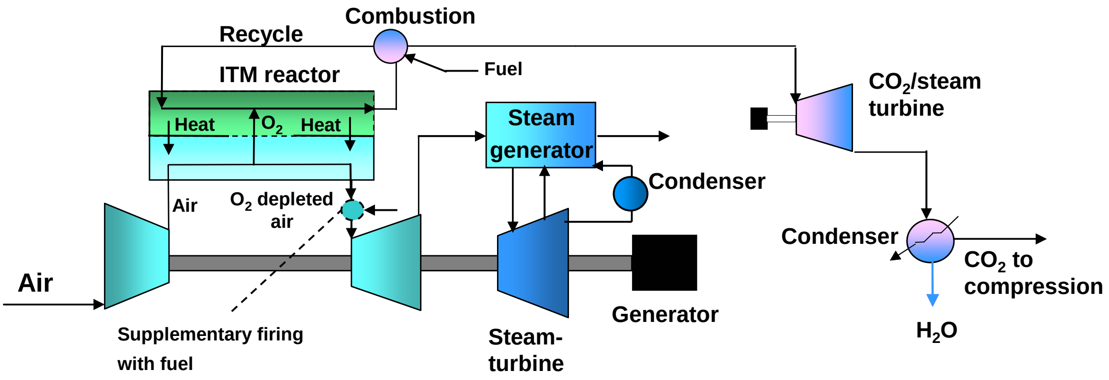

Figure 3.9. Principle flow scheme of the advanced zero emission power plant cycle.

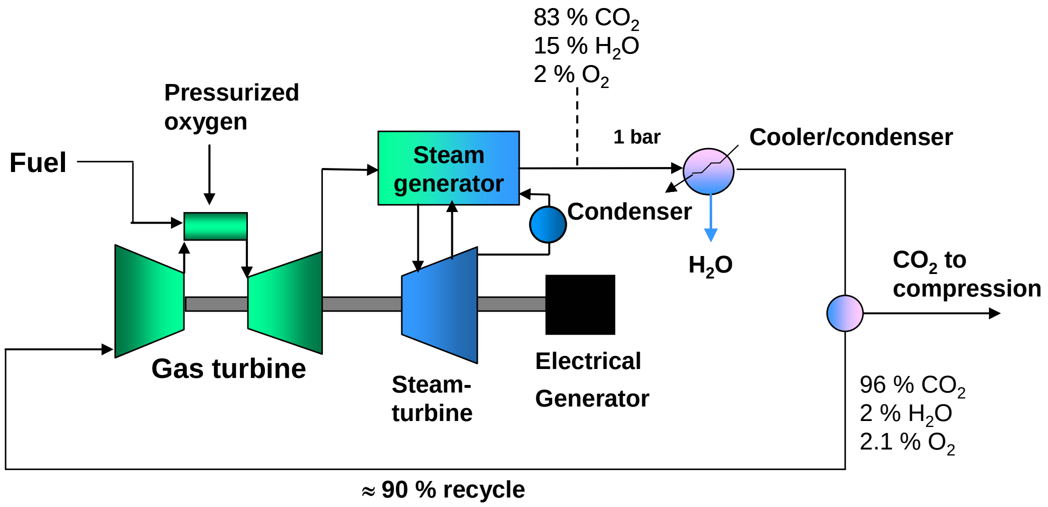

Figure 3.10. Principle of the oxy-fuel gas turbine combined cycle. Exhaust gas is recycled, compressed and used in the combustion chamber to control the temperature entering the turbine.

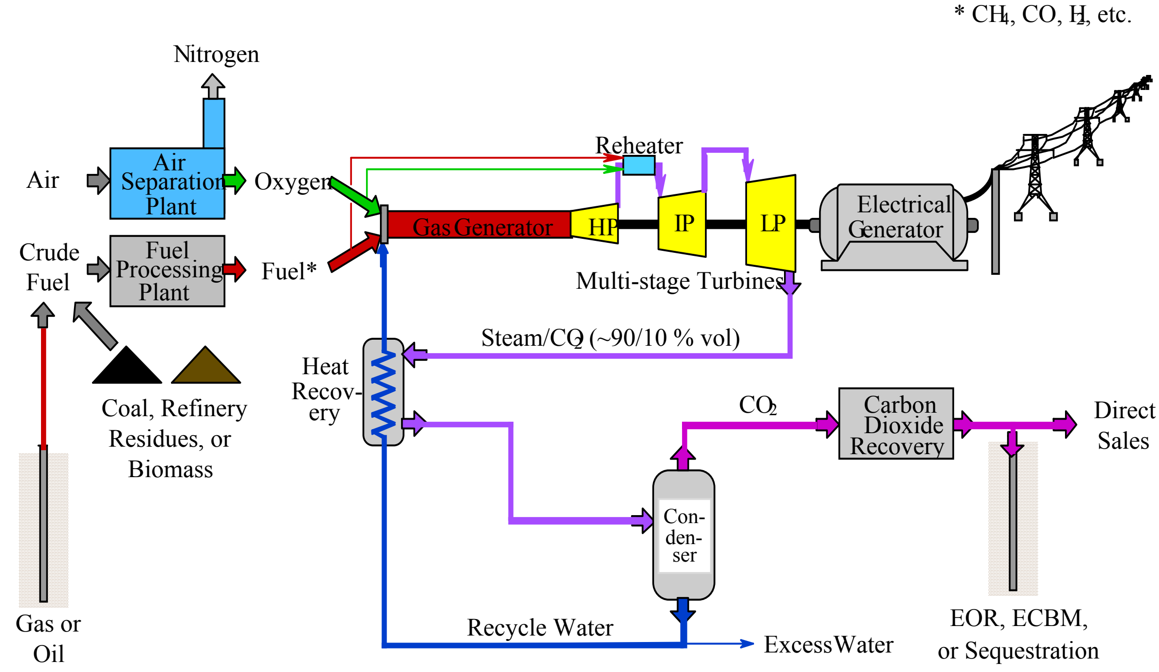

Figure 3.11. Principle of the Clean Energy Systems cycle. The combustion of the fuel and oxygen is cooled by injection of liquid-water, which is recycled in the process.

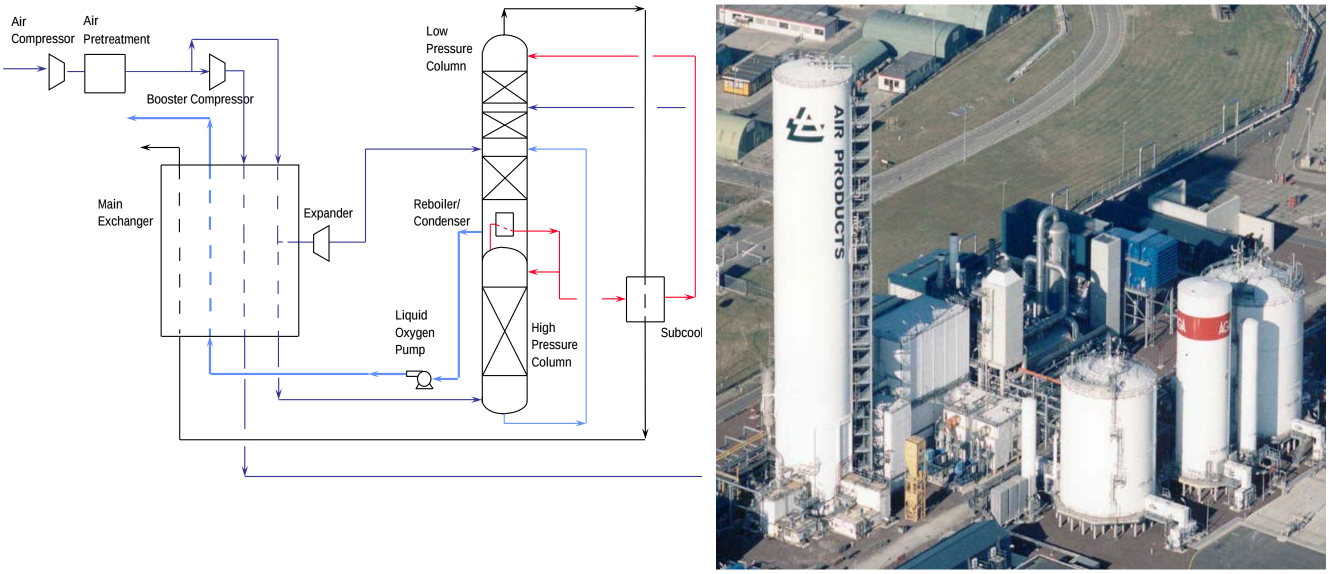

Figure 3.12. Oxygen production by distillation of liquid air. The illustration shows a 3000 tonnes d^{-1} oxygen plant (courtesy of Air Products).

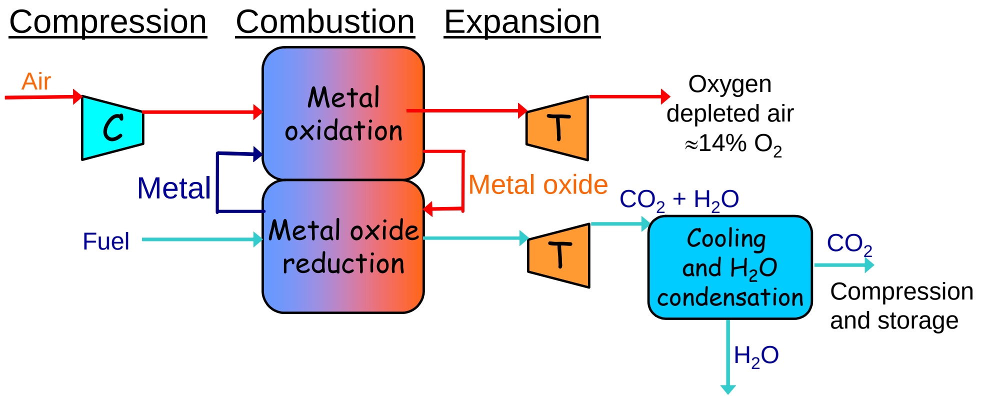

Figure 3.13. The chemical looping combustion principle − in a gas turbine cycle.

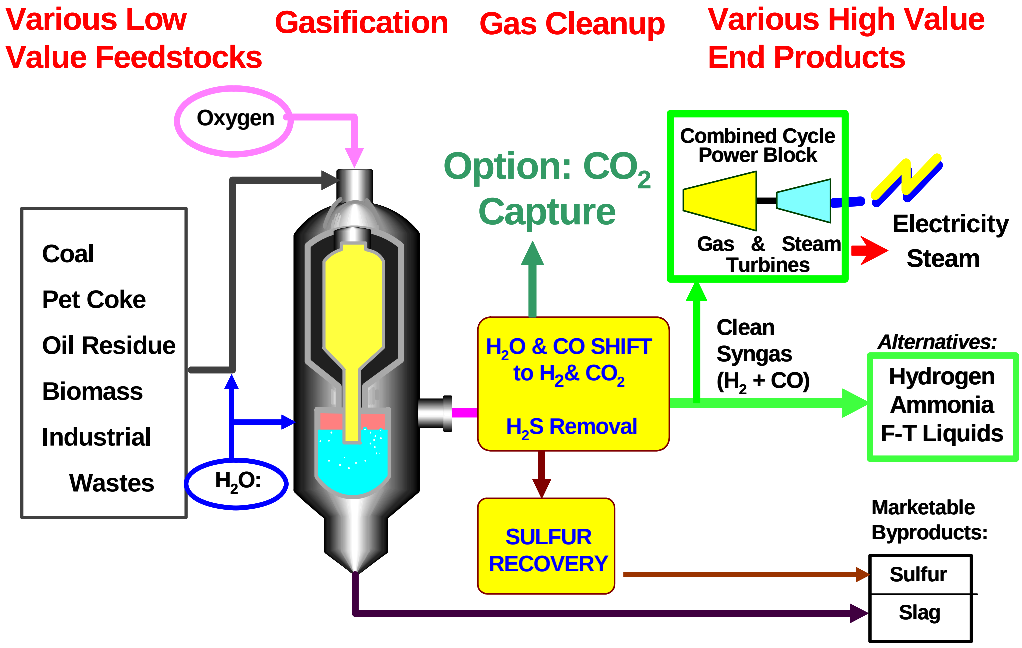

Figure 3.14. Simplified schematic of a gasification process showing options with CO_2 capture and electricity, hydrogen or chemical production.

Figure 3.15. North Dakota coal gasification plant with 3.3 $MtCO_2\ yr^{-1}$ capture using a cold methanol, physical solvent process (cluster of 4 tall columns in the middle of the picture represent the H_2S and CO_2 capture processes; part of the captured stream is used for EOR with CO_2 storage in Weyburn, Saskatchewan, Canada).

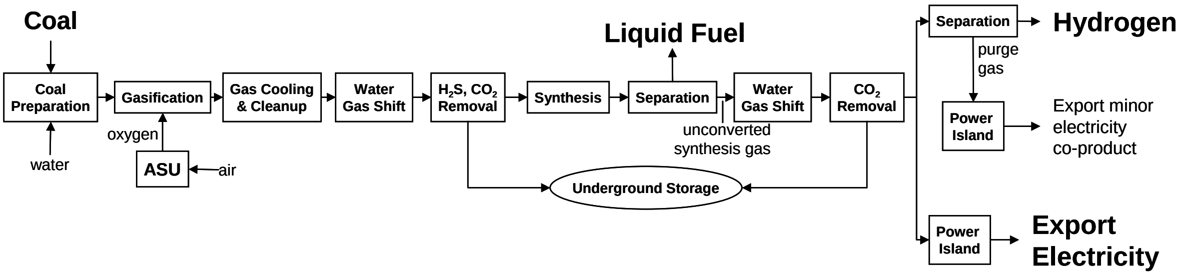

Figure 3.16. Making liquid fuel, electricity and hydrogen from coal via gasification, with CO_2 capture and storage.

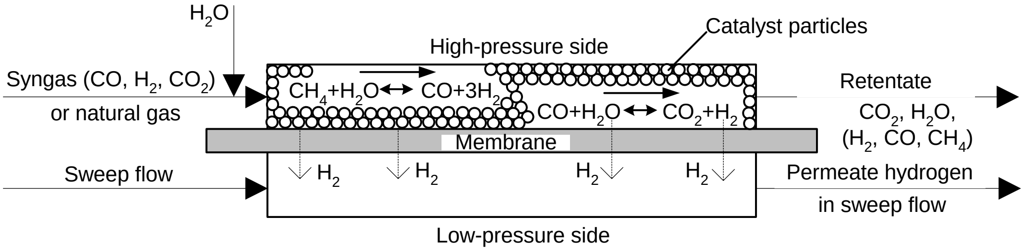

Figure 3.17. Operating principle of a membrane reactor.

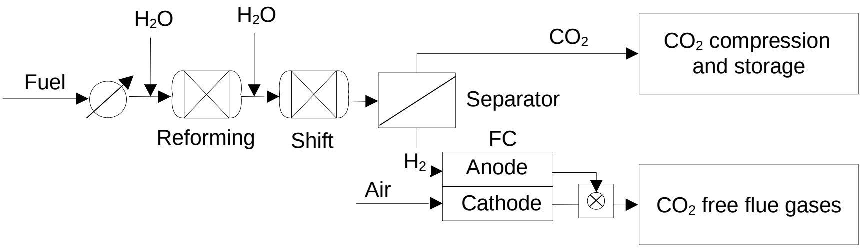

Figure 3.18a. Fuel cell system with pre-fuel cell CO_2 capture. The carbon-containing fuel is first completely converted into a mixture of hydrogen and CO_2. Hydrogen and CO_2 are then separated and the H_2-rich fuel is oxidized in the fuel cell to produce electricity. The CO_2 stream is dried and compressed for transport and storage.

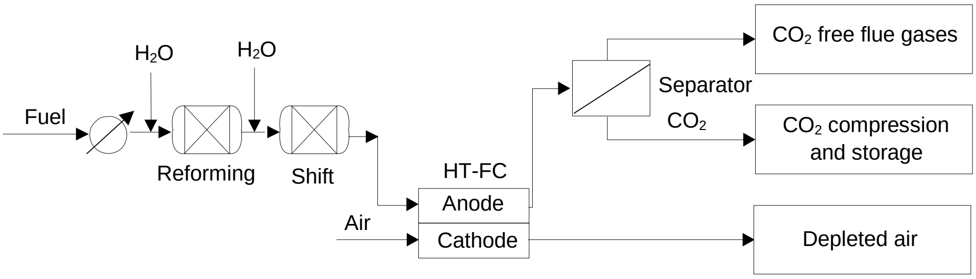

Figure 3.18b. Fuel cell system with post-fuel cell CO_2 capture. The carbon-containing fuel is first converted into a syngas. The syngas is oxidized in the fuel cell to produce electricity. At the outlet of the fuel cell CO_2 is separated from the flue gas, dried and compressed for transport and storage.

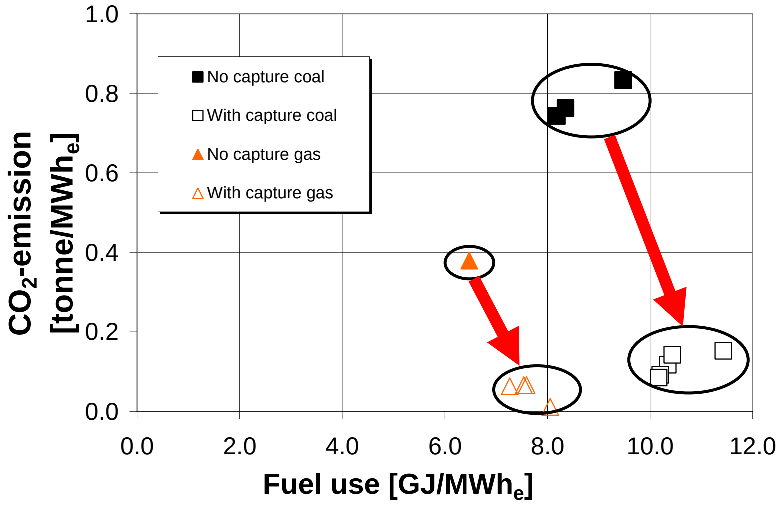

Figure 3.19. Fuel use for a reduction of CO_2 emissions from capture plants (data presented from design studies for power plants with and without capture shown in Figures 3.6 and 3.7).

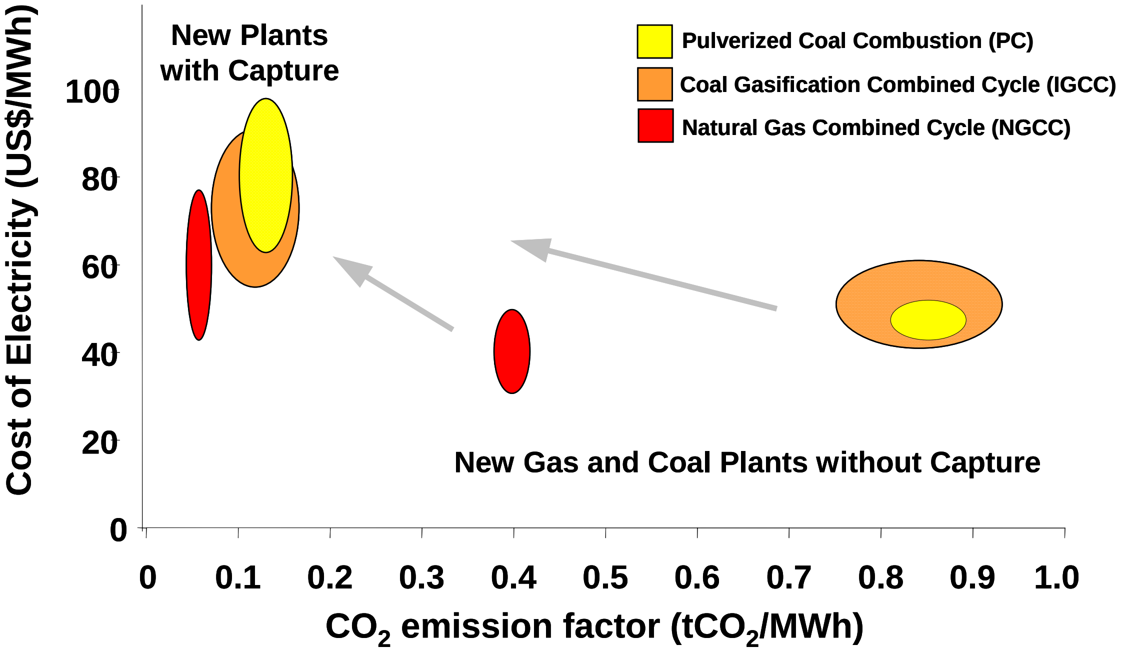

Figure 3.20. Cost of electricity (excluding transport and storage costs) compared to CO_2 emission rate for different reference and capture plants based on current technology. The shaded areas show the Table 3.15 ranges of CO_2 emission rates and levelized cost of electricity (COE) for new PC, IGCC and NGCC plants with and without CO_2 capture. All coal plant data are for bituminous coals only. PC plants are supercritical units only. (See Tables 3.7, 3.9, 3.10 and 3.15 for additional assumptions.) The cost of CO_2 avoided corresponds to the slope of a line connecting a plant with capture and a reference plant without capture (i.e., the change in electricity cost divided by the change in emission rate). Avoidance costs for the same type of plant with and without capture plant are given in Table 3.15. When comparing different plant types, the reference plant represents the least-cost plant that would 'normally' be built at a particular location in the absence of a carbon constraint. In many regions today, this would be either a PC plant or an NGCC plant. The cost per tonne of CO_2 avoided can be highly variable and depends strongly on the costs and emissions of new plants being considered in a particular situation. See Chapter 8 for the full COE and full cost of CO_2 avoided for different plant types.