CO_2 CAPTURE FROM MEDIUM SCALE COMBUSTION INSTALLATIONS

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CO_2 CAPTURE FROM MEDIUM SCALE COMBUSTION INSTALLATIONS

Background

Large CO_2 sources such as power stations and large industrial plants are expected to provide the main opportunities for CO_2 capture and storage. Small and medium scale fixed sources account for a smaller but nevertheless substantial proportion of total global emissions of CO_2. In order to achieve the large reductions in CO_2 emissions that are expected to be necessary to avoid major climate change, major reductions in emissions from these sources will be required. Emissions from these sources could potentially be reduced by energy efficiency improvements, use of low-CO_2 energy carriers, particularly electricity and hydrogen, and by capture and storage of CO_2. Few studies have so far been carried out to assess capture and storage of CO_2 from medium scale sources and this was identified as a knowledge gap in the IPCC Special Report on CO_2 Capture and Storage, published in 2005. Although the costs of capture and storage of CO_2 are expected in general to be higher for medium scale sources than for large scale sources, the costs may be lower than those of alternative CO_2 abatement options.

IEA GHG has commissioned two studies to assess the feasibility and costs of capture and storage of CO_2 from medium scale fixed sources. This study focuses on CO_2 capture and a companion study focuses on collection of CO_2. Once the CO_2 from medium scale sources has been collected and combined into large capacity pipelines, further transportation and storage of the CO_2 will be same as for large scale sources.

This study was carried out for IEA GHG by Ecofys and ECN in the Netherlands.

Study Approach

This study focuses on 'medium scale' combustion installations in the industrial and commercial sectors. The commercial sector includes offices, retail and wholesale premises, hotels, hospitals and educational institutions etc and supply of energy to the residential sector through district heating schemes. CO_2 emissions from chemical processes and cement kilns are outside the scope of the study but IEA GHG is carrying out a separate study on CO_2 capture in the cement industry. 'Medium scale' is notionally defined as installations with a fuel input of between 1 and 100 MWth (a 100 MWth coal fired plant operating for 7,500 hours/year would produce about 250,000 tonnes/year of CO_2 and a natural gas fired plant would produce about 150,000 t/y). The study reviews publicly available information on the capacities of medium scale sources of CO_2, particularly from the USA, the Netherlands and China.

CO_2 is already captured from medium scale natural gas fired boilers, for use in the food, drinks and chemicals industries and standardised post-combustion amine scrubbing plants are commercially available. However, the main applications of medium scale CO_2 capture for greenhouse gas abatement, if any, are likely to be beyond 2020. The study therefore includes some novel capture technologies which are under development and which may become commercially available by that time.

A 'long list' of potential CO_2 capture technologies was identified and their suitability for each of the main types of medium scale CO_2 source was reviewed. A 'short list' of five combinations of fuel use and capture technology was selected for preliminary evaluation and one of these cases was then selected and evaluated in more detail. A case based on the current technology of amine scrubbing was also evaluated for comparison.

This report does not attempt to determine whether or not capture of CO_2 from medium scale installations would be commercially viable. That would depend on the costs of alternative abatement options (efficiency improvement, low-CO_2 energy carriers etc), which is beyond the scope of this study.

Results and Discussion

Medium scale CO_2 sources

The main types of medium scale energy utilisation equipment are reciprocating engines, gas turbines, boilers and furnaces and, possibly in the long term, fuel cells.

Reciprocating engines

Reciprocating engines are typically up to about 20 MWe, although most are less than 10 MWe. Their electricity generation efficiency is around 40%. Hot water or low pressure steam can be produced from the exhaust gas and engine cooling water. The main fuels are light fuel oil, natural gas and biogas. Many reciprocating engines, particularly smaller ones, are used only a fraction of the time, e.g. for stand-by or peak load generation. These installations would be unsuitable for CO_2 capture because it would be difficult to recover the investment cost over the small number of operating hours. However, the bulk of engines above 3.5 MWe are for continuous-duty operation. Reciprocating engines are used mainly in buildings with relatively high and coincident power and hot water demand, particularly universities, hospitals, office buildings and various other commercial buildings. They are also used in a variety of industrial applications where hot water or low pressure steam is required for process needs or space heating. The market for reciprocating engines has increased rapidly in recent years.

Gas turbines

Gas turbines cover the full range of sizes considered in this study, and beyond. About 15% of the total gas turbine capacity ordered in 2005 falls in the range 1-40 MWe. The largest market in terms of capacity is for units of >20MWe. The electricity generation efficiency of simple cycle gas turbines is between about 25% and 40%. The hot exhaust gas can be used in industrial processes or it can be used to generate steam, at high pressure if necessary, or hot water. The steam can be fed to a back-pressure or condensing steam turbine or it can be used directly in industrial processes. The main applications of gas turbines are in industry, particularly the oil, gas, chemicals, paper and food processing industries, and for heating at large sites such as universities.

Boilers and furnaces

The most significant medium scale energy users are boilers and furnaces. Data for the USA, the Netherlands and China are provided in this report. The main fuel for medium-sized boilers in the EU and USA is natural gas. In contrast, coal is the main fuel in China.

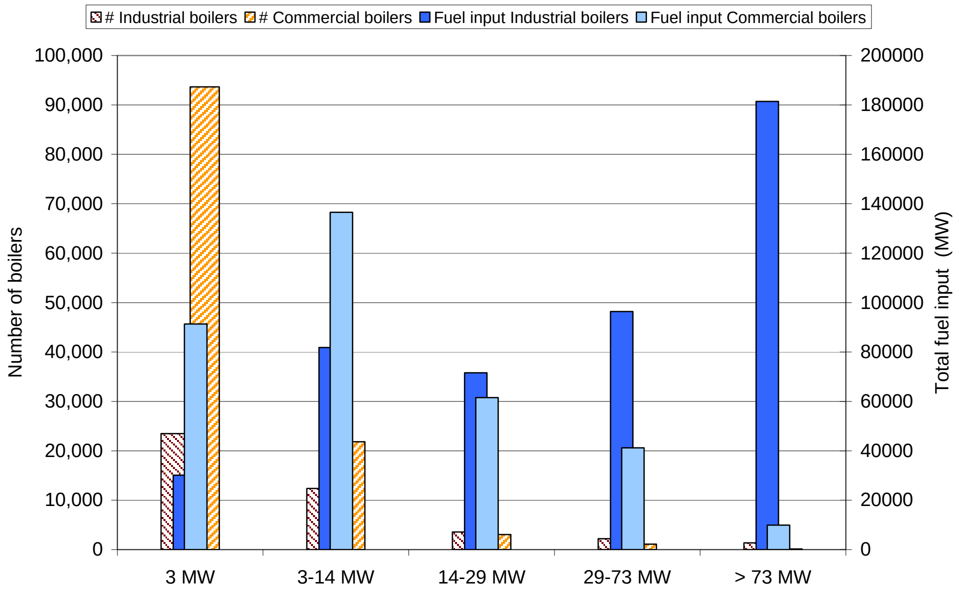

Figure 1 Fuel input capacity and number of boilers in the USA

For illustration, figure 1 shows the distribution of the numbers of boilers and installed capacity for different size ranges in the USA, although it should be noted that the size ranges extend beyond those defined in this study. It can be seen that commercial boilers account for the largest proportion of installed capacity for smaller sizes (<14 MWth) and industrial boilers are most significant for the larger sizes. In terms of capacity, the four most significant commercial sectors are, in descending order, health care, offices, lodgings and education and the most significant industrial sectors are the chemical and pulp and paper industries. Excluding the electric power industry there are 43,000 industrial boilers greater than 0.3 MW in the USA, with a total fuel input capacity of 480 GW. There are also 120,000 commercial boilers with a capacity of 330 GW. 75% of the overall capacity is for boilers smaller than 73 MWth.

In China in 2003 there were over 500,000 ‘industrial’ boilers with a total capacity of 1250 GW. 55% of the boiler capacity is used to provide energy for residential purposes. Analysis by some industrial experts indicates that in 2001 20% of boiler capacity is for boilers larger than 14 MWth, with only 1% for larger than 25 MWth. 75% is for boilers between 1.4 and 7 MWth and 5% is for less than 0.7 MWth. Most of the capacity is therefore within the ‘medium scale’ capacity defined in this study. In China, the average load factor of industrial boiler is reported to be between 50% and 70%. Generally speaking, the smaller the size of boiler, the lower the average load factor. It is debatable whether CO_2 capture would be realistic for many of the smaller scale coal fired boilers in China, which have relatively low efficiencies and high emissions, but it is possible that these may be replaced in future by larger scale installations which would be more amenable to CO_2 capture. The feasibility and costs of CO_2 capture is one of many factors which may influence the extent of distributed energy generation.

Fuel cells

Fuel cells are currently insignificant in terms of capacity installed but their role in the medium scale energy use sector may increase substantially in future. Fuel cells offer the prospect of high power generation efficiencies but they currently have high costs. IEA GHG is carrying out a further study to assess the prospects for fuel cells and opportunities for CO_2 abatement.

CO_2 emissions

Few data are available on CO_2 emissions from medium scale installations. Emissions data are reported in some countries, for example to environmental regulators, but they are usually for whole plants, which may include several different individual installations, some of which may not be ‘medium scale’. Data on emissions from plants and installations in the Netherlands are included in the report to illustrate this point. Ecofys estimates CO_2 emissions from medium scale installations to be 15-20% of total global emissions. The largest quantities of emissions are from boilers and furnaces, particularly those that are fired by coal.

Preliminary evaluation of CO_2 capture technologies

In principle a wide range of technologies could be used to capture CO_2 from medium scale sources. The report reviews the various capture options for each of the types of medium scale combustion installation. The most suitable technology will depend on site-specific conditions, including the size of the installation, its operating load factor and opportunities to utilise low grade heat. Modular technologies tend to be favoured for smaller scale installations because their economies of scale are relatively low. The following combinations of fuel use and capture technology, which cover a wide range of cases, were selected for preliminary evaluation in this study. It is emphasised that these are not necessarily the optimum technologies in all cases. The evaluation is based on new-build installations, rather than retrofits.

1. Natural gas and diesel oil fired reciprocating engines (1.5MW_e) with post-combustion membrane assisted liquid absorption for CO_2 capture
2. Natural gas fired gas turbine (5 MW_e) with pre-combustion PSA capture of CO_2
3. Oxy-combustion coal boiler (50 MW_{th}) with membrane oxygen production
4. Oxy-combustion natural gas boiler (5 MW_{th}) with membrane oxygen production
5. Solid oxide fuel cell (0.5 MW_e) with oxygen conducting membrane afterburner

Installations which are only operated seasonally or to meet peak demands will have relatively high CO_2 emissions avoidance costs, as shown in figure 2. The technical feasibility of operating CO_2 capture technologies at varying load factors needs to be assessed further. IEA GHG is planning a study on this subject, related to large scale power plants.

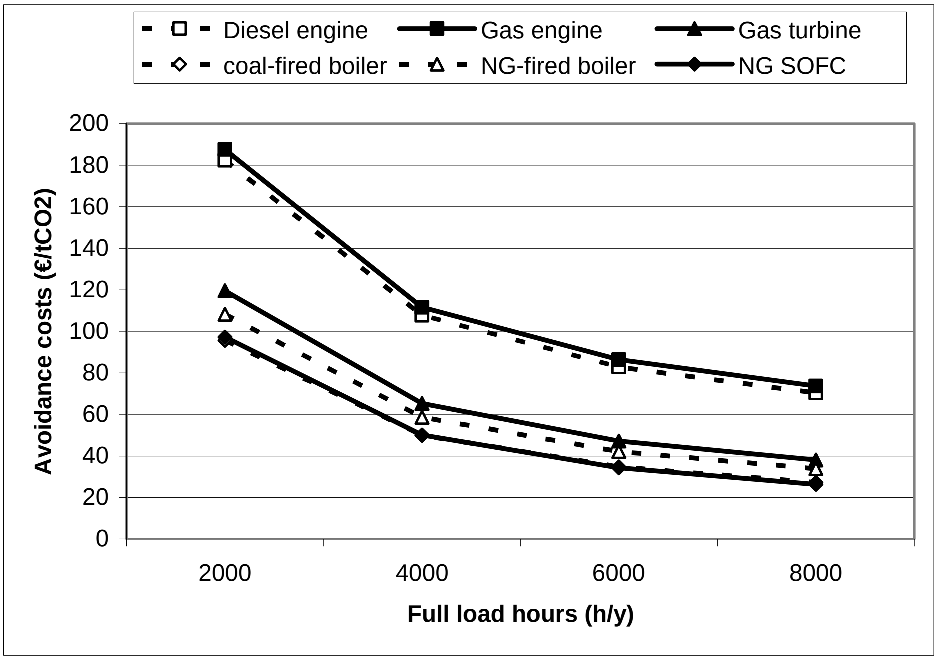

Figure 2 Cost of CO_2 avoidance for a range of annual full load operating hours
(Note: The data points for coal fired boilers and natural gas solid oxide fuel cells are almost the same)

The highest costs are for post combustion capture in the relatively small reciprocating engine cases. The fuel cells case has relatively low costs despite its small unit size.

The costs in figure 2 include the costs of CO_2 compression but in some circumstances CO_2 would be collected from medium scale capture plants in pipelines operating at low, or even sub-atmospheric, pressure. Compression would take place at nodes in a collection grid, as discussed in an IEA GHG study on CO_2 collection, which will be published shortly. If CO_2 compression is excluded, the cost of capture decreases by about €10-20/t CO_2 for plants operating at high load factors.

Evaluation of a selected case

A more detailed evaluation of CO_2 capture at 50 MWth coal fired circulating fluidised bed boilers was undertaken. Coal fired boilers were selected because they account for a large proportion of the emissions from medium scale sources. Two variants of an oxy-combustion boiler with oxygen conducting membranes (OCMs) were assessed: one that is solely coal-fired and one in which some natural gas fuel is used to heat the air input to the membrane module. Flowsheets of these plants are included in Appendices A and B of the report. A more conventional post combustion MEA scrubbing process was also assessed. The costs of this plant were derived by scaling down cost data for a larger plant.

The costs of CO_2 capture and compression are summarised in table 1. The costs are based on plants operating at full load for 7500 hours per year. Coal fired power plants are assumed to be used to generate the electricity which is imported by the plants.

Table 1 Performance and costs of plants with CO_2 capture

	Without CO_2 capture	With CO_2 capture		
	Coal fired boiler	Coal/natural gas OCM boiler	Coal-only OCM boiler	Coal boiler with MEA scrubbing
Heat output, MW	50.0	50.0	50.0	50.0
Coal feed, MW, LHV	57.5	47.3	57.7	79.2
Natural gas feed, MW, LHV		9.8		
Electricity consumption, MW	0.5	2.3	2.6	3.6
CO_2 captured, kg/s		4.30	5.10	6.04
CO_2 emissions, kg/s	5.16	0.68	0.09	1.07
CO_2 emissions due to electricity use, kg/s	0.10	0.48	0.55	0.76
CO_2 emissions avoided, kg/s		4.10	4.61	3.43
Capital cost, €M	18.7	25.3	30.8	45.1
Cost of CO_2 avoidance, €/tonne CO_2		22	21	70

The cost of post combustion capture using MEA solvent scrubbing, per tonne of CO_2 emissions avoided, is approximately twice as high as for a large scale (670MW_e) power plant. The estimated costs for the membrane oxy-combustion cases are substantially lower than for the post combustion capture case, because of a lower energy penalty and lower capital and operating costs. However, there are greater uncertainties regarding the performance and costs of the membrane cases. Significant technical issues need to be addressed, in particular the design of the membrane air separation unit and its sensitivity to impurities and fouling, integration of the equipment and, for the coal-only case, the high temperature heat exchanger which is needed to heat the air input to the membrane unit. (In the partially gas fired case heating of the membrane feed gas is achieved by in-line combustion of natural gas). Research, development and demonstration work would need to be successfully completed before the oxy-combustion technology could be widely adopted.

An important issue for medium scale installations is the transportation of the captured CO_2 to a storage site. The cost of CO_2 transport will depend on the proximity to other CO_2 sources, because there are large economies of scale in CO_2 pipeline systems. IEA GHG will shortly publish another study which assesses the feasibility and costs of collecting CO_2 from medium scale installations.

Expert Reviewers' Comments

The draft study report was reviewed by various external experts. IEA GHG is very grateful to all of those who contributed to this review. The comments from reviewers provided some significant information and suggestions which helped to improve the final report. In particular some significant issues were raised about the initial design of the oxy-combustion boilers. In response to reviewers' comments a conventional post combustion amine scrubbing case was included, in addition to the more novel oxygen membrane cases.

Some reviewers commented that the costs of the cases evaluated in this study appeared too low compared to published cost estimates for large scale CO_2 capture plants. Following the expert review, IEA GHG obtained some indicative cost and performance data from a company which supplies standardised MEA scrubbing plants for the food industry. The data were for a plant capable of capturing 24t/d of CO_2 from a natural gas fired boiler. Insufficient data were available for a detailed case evaluation but the capital cost per tonne of CO_2 appeared to be not significantly greater than published costs for large scale plants, possibly due to use of standardised, simplified construction. The steam consumption for solvent regeneration was higher than for large scale plants, so it would be necessary to be able to utilise much of the heat rejected from the capture unit, for example to provide hot water. Further information on this aspect of the plant design is needed. It may be worthwhile carrying out a more detailed site specific evaluation of a natural gas fired boiler with MEA scrubbing, in conjunction with an equipment supplier, to reduce uncertainties.

Major Conclusions

Medium scale combustion installations (1-100MWth) account for substantial emissions of CO_2, of the order of 15-20% of total global emissions. There is therefore considerable technical potential for CO_2 capture. Natural gas fired boilers are the most significant medium scale combustion installations in the EU and USA but coal fired boilers are most significant in China.

Plants for capture of CO_2 from medium scale natural gas-fired boilers using amine scrubbing are already commercially available, for example for supply of CO_2 to the food and drinks industries.

Various alternative CO_2 capture technologies may be suitable for medium scale installations.

The costs of CO_2 capture depend strongly on local conditions, including the size of the installation, the operating load factor and the ability to make use of any low grade heat that is rejected by the capture process. Installations which are only operated seasonally or for peak load duties will have relatively high costs of avoiding CO_2 emissions.

A boiler combined with membrane oxygen production offers the prospect of a significantly lower cost of capture than post combustion amine scrubbing for a medium scale installation. However, development of such a system presents considerable challenges and research, development and demonstration is needed.

The cost of capture and storage of CO_2 from medium scale installations depends on the proximity of other CO_2 sources, to give economies of scale in CO_2 pipeline systems. IEA GHG will shortly publish another study which assesses the feasibility and costs of collecting CO_2 from medium scale installations.

The economic feasibility of capturing CO_2 from medium scale installations will depend on the carbon price, which will depend on the extent to which global emissions of CO_2 need to be reduced and the costs of alternative ways of reducing emissions. Further work in this area is needed.

Recommendations

A study should be carried out to compare the alternative ways of reducing CO_2 emissions from medium scale energy users, including energy efficiency improvements and use of energy carriers (hydrogen and electricity) produced from renewable energy sources and large scale fossil fuel plants with CO_2 capture and storage.

If such a study shows that CO_2 capture is an attractive option for some medium scale energy users, further work should be carried out to assess capture technologies in more detail for specific sites. This could include commercially available MEA scrubbing and novel technologies. Routes to development and demonstration of novel technologies should be identified.

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CARBON DIOXIDE CAPTURE FROM MEDIUM SCALE COMBUSTION INSTALLATIONS

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1 Introduction

Up until today, the main focus of IEA GHG Programme's work has been on capture and storage of CO_2 from fossil fuel fired power plants and other large point sources. However, it is anticipated that small and medium-scale sources of CO_2 account for a substantial proportion of total emissions.

In order to achieve the reductions in CO_2 emissions that are necessary to comply with (inter)national targets, substantial reductions in emissions from small- and medium-scale sources may be required. There is significant scope for energy efficiency improvements to reduce CO_2 emissions from these sources but deeper reductions will probably be needed. Energy carriers such as hydrogen and electricity have usually been proposed to achieve such deep CO_2 reductions. CO_2 capture from decentralised plants has usually been neglected because it is expected to have high costs and energy consumptions. Furthermore, there are concerns about the safety of CO_2 transmission from sources in populated areas. However, the new energy carriers such as hydrogen would need major new infrastructure. It is therefore possible that small- and medium-scale capture of CO_2 may be competitive in some circumstances.

CO_2 could potentially be captured at any scale ranging from major power stations and industrial plants down to individual houses. IEA GHG R&D Programme commissioned Ecofys and ECN to carry out a study to the applicability of capture of carbon dioxide from medium-scale installations; at the scale of for example district heating plants, factories and large commercial buildings. In this study, we consider medium-scale plants to have a fuel input between 1 and 100 MW.

The study assesses the costs and performance of medium-scale CO_2 capture technologies The focus is on CO_2 capture from medium-scale power generation systems, from medium-scale systems for industrial heat production and from medium-scale combined heat and power systems. Other industrial processes which produce CO_2, such as cement, metallurgical and petrochemical processes are not considered in this study With respect to the size, we focus on novel technologies which will become available on the longer term, because it is expected that CO_2 is only likely to be captured from medium-scale sources when most large-scale capture opportunities have been exhausted. An exception is the MEA scrubbing technology which is currently used in industrial applications at a scale of 50 to 400 kt per year. The study only considers new systems and does not consider retrofit systems.

2 Medium-scale energy conversion installations

2.1 Introduction

This chapter gives a characterisation of medium-scale combustion installations for the production of power and heat. It will focus on different aspects like typical size, fuel use, efficiency and flue gas characteristics. The characterization forms the basis for analysis of the installations with capture of CO_2, which are described in chapter 2.
Power and heat in industry and the commercial sector is supplied by various types of energy conversion devices. A commonly used categorisation is:

- Furnaces (including boilers)
- Steam turbines
- Reciprocating engines
- Gas turbines
    - Large gas turbines (0.5 MW_e to 330 MW_e)
    - Micro turbines (30 to 500 kWe)
- Fuel cells

The installations are either used as ‘centralised’ production units, mainly large power production units with incidentally a district heating system attached to it, or ‘distributed’ production units supplying on site power and/or heat. As this study targets medium-scale applications, we focus at distributed production units. Figure 2-1 shows typical capacity ranges for some of the most common types of distributed power generation units.

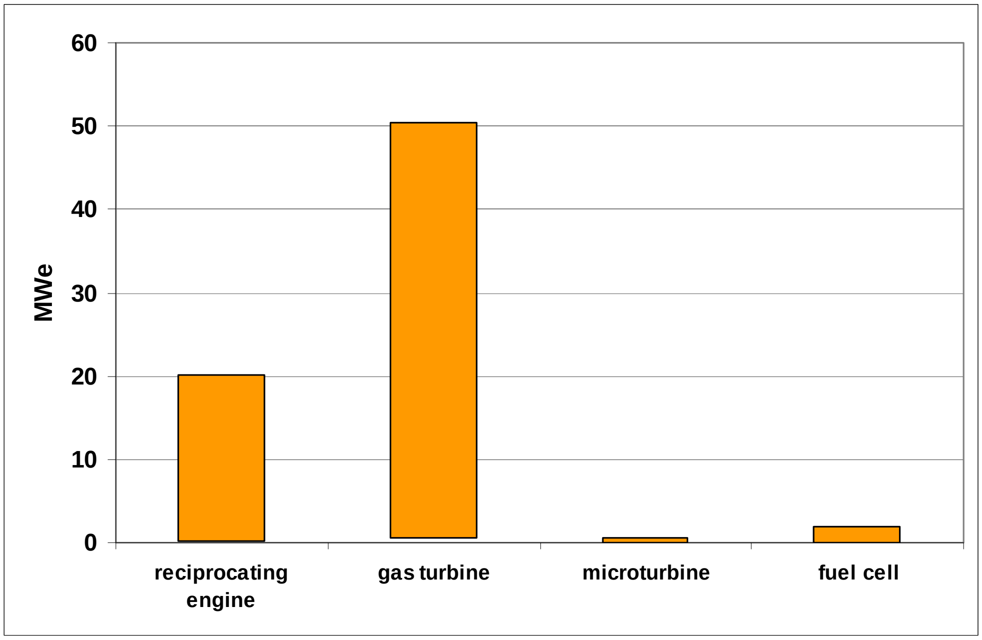


Figure 2-1. Power generators and general capacity range for distributed production

Table 2-1 summarizes the applicability and presents some typical characteristics for the distributed generation technologies.

Table 2-1 Characteristics of distributed generation technologies

	Steam Turbine	Reciprocating Engine	Gas turbine		Fuel cell
			Large gas turbine	Micro turbine	
Applicable market sectors	Institutional buildings, Industry, Waste fuels	Commercial buildings, Light industry, Grid, Waste fuels	Large commercial, Industry, Grid, Waste fuels	Commercial buildings, Light industry, Waste fuels	Residential, Commercial buildings, Light industry
Technology status	Commercial	Commercial	Commercial	Early entry	Early entry, development
Size (MW)	0.5 - 100	0.05- 20	0.5 - 50	0.03 – 0.25	0.005 – 2
Fuels	All	Natural gas, biogas, liquid fuels	Natural gas, biogas, distillate oil	Natural gas, biogas	Hydrogen, natural gas
Uses for heat recovery	LP and HP steam, district heating	Hot water, LP-steam, district heating	Direct heat, hot water, LP and HP steam, district heating	Direct heat, hot water, LP steam	Hot water, LP steam
Typical CO_2 flue gas concentration	8-15% (boiler) 3-4% (gas turbine HRSG)	9- 14%	3-4%	2-4%	8-10%

2.2 Furnaces (including boilers)

Furnaces either directly supply heat to processes or indirectly through the generation of hot water and steam; called boilers. Furnaces supplying direct heat are mainly used to produce high temperature heat, i.e. above 400 °C. Typical applications of furnaces are in the iron and steel industry, ceramic industry, cement and limestone industry and glass industry.
Steam from boilers is either used directly for industrial purposes (e.g. in the paper and pulp industry) or used to generate electricity in a steam turbine. If both heat and power are generated, we speak about a co-generation unit or combined heat and power installation (CHP).

Furnaces are applied in a wide range of scales. The size of furnaces ranges from small domestic heaters via industrial ovens to large boilers for coal-fired power plants (see Figure 2-2). Furnaces can be fuelled with all kind of fuels, like natural gas, oil, industrial off-gases, coal, biomass and waste. Furnaces are typically operated with a small excess of air of about 5-10% for natural gas-fired furnaces, 5-20% for oil-fired furnaces, and 15-60% for coal-fired furnaces. Concentrations of CO_2 in the flue gas vary depend on the fuel type used. Natural gas fuelled furnaces have a CO_2 concentration in the flue gases of about 9%, coal-fired furnaces of about 14%. The lifetime of furnaces is very long, and may exceed 50 years.

Boilers produce hot water or steam mostly for process heat up to 300 °C. They play a major role in manufacturing, heating and electricity generation. Boilers are used in many (energy-intensive) industries and commercial buildings and comprise a significant part of the energy consumption. Boilers can be differentiated by their configuration, size and quality of steam or hot water. Boilers are not only used for steam or hot water production, but are also used for combined heat and power production using a steam cycle.

Size
The size of boilers ranges from domestic heaters, typically smaller than 30 kW, via boilers for the commercial sector (10 MW) and industrial boilers (30 to 300 MW) to pulverised coal-fired boilers of large-scale power plants (over 1000 MW).

Quality of steam/hot water
Boilers can also be differentiated by the temperature and pressure of the hot water or steam that is required. Hot water systems are smaller than steam systems and many hot water boilers are installed for space heating applications. Hot water boilers generally deliver hot water below 120 °C and at pressures less than 20 bar.

Configuration
Boilers can furthermore be characterized by the configuration of the heat transfer surfaces – either firetube or watertube – and by the fuel-burning system. The appropriate

configuration is determined by the fuel, steam conditions and capacity. Figure 2-2 shows the typical capacity ranges for some of the common boiler types on a heat input basis. Firetube units are typically the smallest boilers, with most units less than 3 MW_{th} in capacity. Watertube boilers can be separated into two classes by size. Most operational watertube boilers have a capacity of less than 3 MW_{th}. A smaller number of watertube boilers are between 3 and 3,000 MW_{th}, including almost all large industrial and power generation boilers. Although fewer in number than the firetube and small watertube boilers, the large watertube boilers account for most steam production. Large capacity watertube boilers are most likely to use solid fuel and there are several different solid fuel combustion technologies in use at these larger size ranges, including stoker, fluidized bed and pulverized coal boilers.

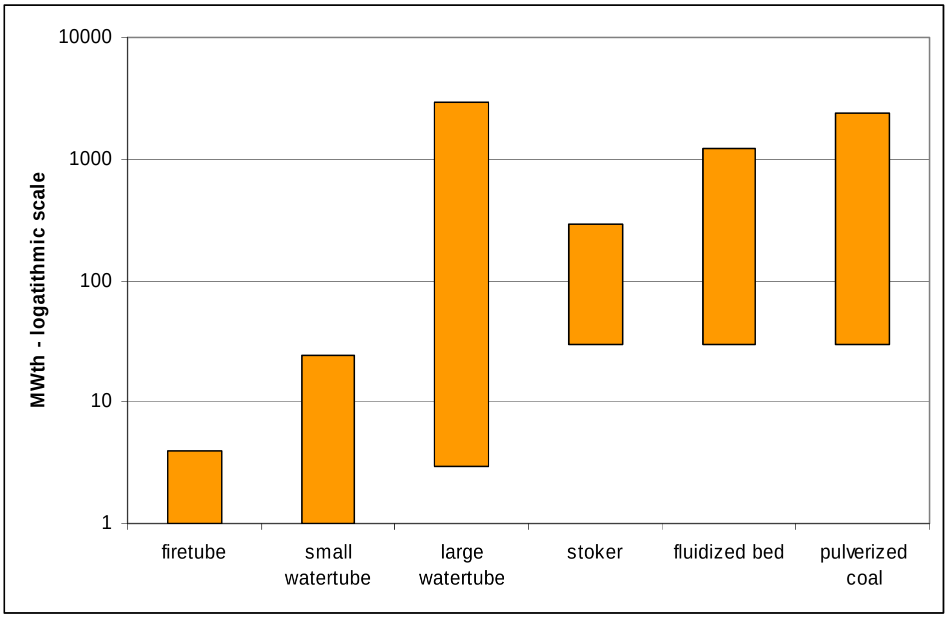


Figure 2-2. Boiler types and general capacity ranges

Efficiencies and costs of boilers

Boiler efficiency is determined by the fuel type used and the temperature and pressure of hot water or steam produced. Typical full load efficiencies for coal-fired boilers are in the range of 85% and for gas-fired boilers in the range of 80 to 90%. Boiler costs are typically in the range from 100 €/kWth-input to 50 €/kWth-input for larger units.

2.3 Steam turbines

In a steam cycle, the first step is to pump water to medium or high pressure. It is then heated to the boiling temperature corresponding to the pressure, boiled (heated from liquid to vapour), and then most frequently superheated (heated to a temperature above that of boiling). To generate power, a multistage turbine expands the pressurized steam to lower pressure. The steam is then exhausted either to a condenser at vacuum conditions or into an intermediate temperature steam distribution system that delivers the steam to the industrial or commercial application. The condensate from the condenser or from the steam utilization system returns to the feed water pump for continuation of the cycle.

Size

Steam turbines are one of the most versatile and oldest prime mover technologies still in general production. Power generation by means of steam turbines has been in use for about 100 years, when they replaced reciprocating steam engines due to their higher efficiencies and lower costs. In industrialised countries, like the United States, conventional steam turbine power plants generate most of the electricity produced. The capacity of steam turbines ranges from 50 kW to several hundred MWs for large utility power plants.

Configuration

Unlike gas turbine and reciprocating engine CHP systems where heat is a by-product of power generation, steam turbines are normally optimized for steam production and generate electricity as by-product of heat (steam). A steam turbine is captive to a separate heat source and does not directly convert fuel to electric energy. The energy is transferred from the boiler to the turbine through high-pressure steam that in turn powers the turbine and generator. This separation of functions enables steam turbines to operate with an enormous variety of fuels, from natural gas to solid waste, including all types of coal, wood, wood waste, and agricultural by products (sugar cane bagasse, fruit pits, and rice hulls). In CHP applications, steam at lower pressure is extracted from the steam turbine and used directly or is converted to other forms of thermal energy (Figure 2-3).

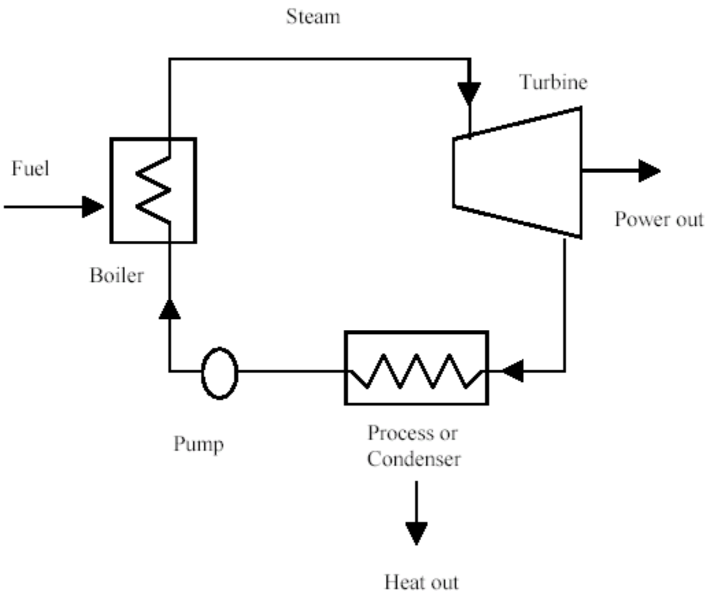


Figure 2-3. Schematic view of a combined heat and power system

Efficiencies and costs boiler/steam turbine combination
The electrical generating efficiency of standard steam turbine power plants varies from a high efficiency of 40% on LHV basis for large, electric utility plants designed for the highest practical annual capacity factor, to under 10% on LHV basis for small, simple plants which make electricity as a by-product of delivering steam to processes or district heating systems.

Steam turbine thermodynamic efficiency (isentropic efficiency) refers to the ratio of power actually generated from the turbine to what would be generated by a perfect turbine with no internal losses using steam at the same inlet conditions and discharging to the same downstream pressure (actual enthalpy drop divided by the isentropic enthalpy drop). Turbine thermodynamic efficiency is not to be confused with electrical generating efficiency, which is the ratio of net power generated to total fuel input to the cycle. Steam turbine thermodynamic efficiency measures how efficiently the turbine extracts power from the steam itself. Multistage (moderate to high-pressure ratio) steam turbines have thermodynamic efficiencies that vary from 65% for small (under 1,000 kW) units to over 90% for large industrial and utility sized units. Small, single stage steam turbines can have efficiencies as low as 50%. When a steam turbine exhausts to a CHP application, the turbine efficiency is not as critical as in a power only condensing mode. The majority of the energy not extracted by the steam turbine satisfies the thermal load. Power only applications waste the exhaust turbine steam energy in condensers.

2.4 Reciprocating engines

Reciprocating internal combustion engines represent a widespread and mature technology for power and heat generation. Reciprocating engines are used for all types of power generation, from small portable gensets to larger industrial engines of several megawatts. The engines can be fired with all kinds of fuel like natural gas, diesel and syngas, e.g. from wood or coal gasification. Spark ignition engines for power generation use natural gas as the preferred fuel – although they can be set up to run on propane or gasoline. Diesel-cycle, compression ignition engines operate on diesel fuel or heavy oil, or can be set up in a dual-fuel configuration that can burn primarily natural gas with a small amount of diesel pilot fuel.

Size and configuration
Reciprocating engines offer low first cost, easy start-up, proven reliability when properly maintained, and good load-following characteristics. Gas-fired reciprocating engines are well-suited for packaged CHP in commercial and light industrial applications of less than 7 MW. Smaller engine systems produce hot water. Larger systems can be designed to produce low-pressure steam. Figure 2-4 shows an example of a reciprocating engine.

Figure 2-4. Reciprocating engine > 1 MW_e (Jenbacher J620)

Efficiencies of reciprocating engines

The efficiencies of reciprocating engines in the range from 100 kW_e to 10 MW_e are depicted in Figure 2-5. Using the power output P (in kW), the average electrical efficiency (η_{el}) can be calculated as:

$$\eta_{el} = 0.24 \times P^{0.066} \times 100\%$$

The average total efficiency as:

$$\eta_{tot} = 0.90 \times P^{-0.006} \times 100\%$$

Investment costs of reciprocating engines

Figure 2-6 depicts the scale dependence of the investment costs of reciprocating engines. In this figure, the specific investment cost is shown, and a regression function is derived. Based on these data a scale factor for the total investment costs for reciprocating engines of 0.67 is derived.

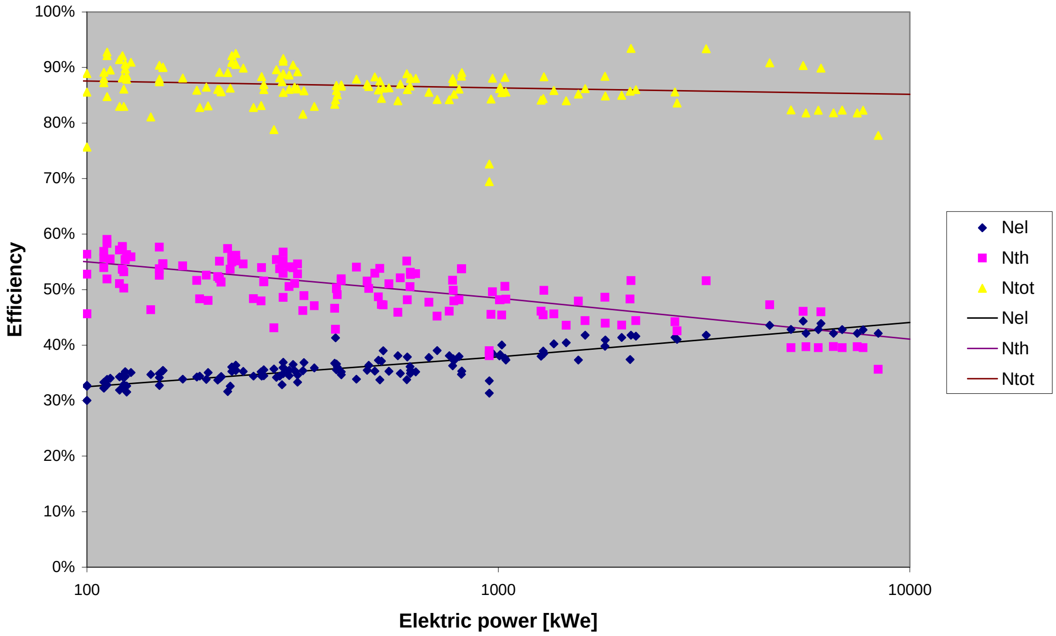


Figure 2-5. Electrical, thermal and total efficiency of reciprocating gas engines (ASUE, 2001)

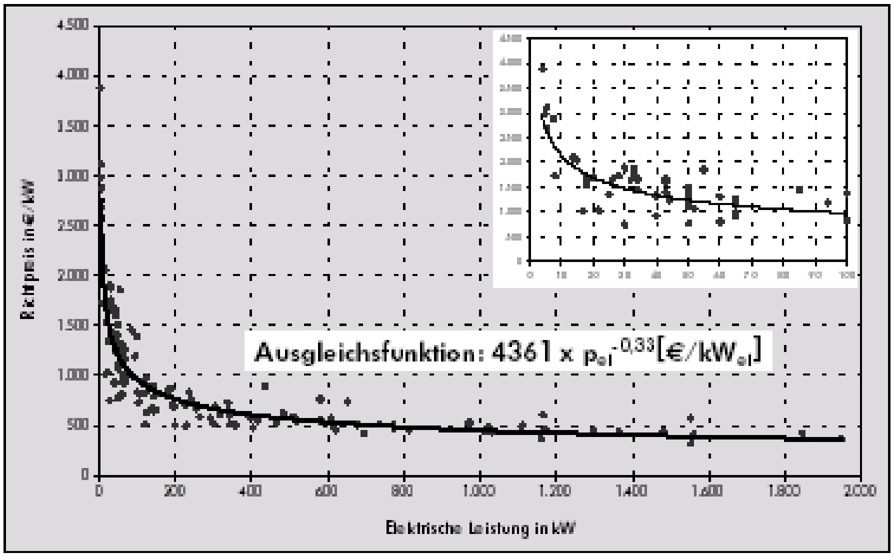


Figure 2-6. Specific investment costs of reciprocating natural gas engines (ASUE, 2005)

2.5 Gas turbines

Gas turbines for distributed generation applications are an established technology. A typical distinction is large gas turbines, ranging from 0.5 MW_e to 50 MW_e and micro turbines, ranging from 30 to 500 kW.

2.5.1 Large gas turbines

Gas turbines produce high quality heat that can be used to generate steam for on-site use. From capacities of 7 MW_e and higher some gas turbines are equipped with a heat recovery steam generator (HRSG) and a (multi stage) steam turbine for additional power generation (combined-cycle configuration). The largest units with one gas turbine go up to 480 MW_e (Gas Turbine World, 2005). Gas turbines can be set up to burn natural gas, a variety of petroleum fuels, oil or syngas from e.g. coal gasification. Gas turbines can also have a dual-fuel configuration. Maintenance costs per unit of power output are among the lowest of distributed generation technology options. Low maintenance and high-quality waste heat make gas turbines an excellent match for industrial or commercial CHP applications larger than 5 MW. Technical and economic improvements in small gas turbine technology are pushing the economic range into smaller sizes as well.

An important advantage of CHP using gas turbines is the high-quality waste heat available in the exhaust gas. The high-temperature exhaust gas is suitable for generating high-pressure steam, making gas turbines a preferred CHP technology for many industrial processes. In simple cycle gas turbines, hot exhaust gases can be used directly in a process or the exhaust heat can be used to generate steam or hot water by adding a heat-recovery steam generator (HRSG). Because gas turbine exhaust gas is oxygen-rich, it can support additional combustion through supplementary firing. A duct burner can be fitted within the HRSG to increase the steam production at lower-heating value efficiencies of 90% and greater.

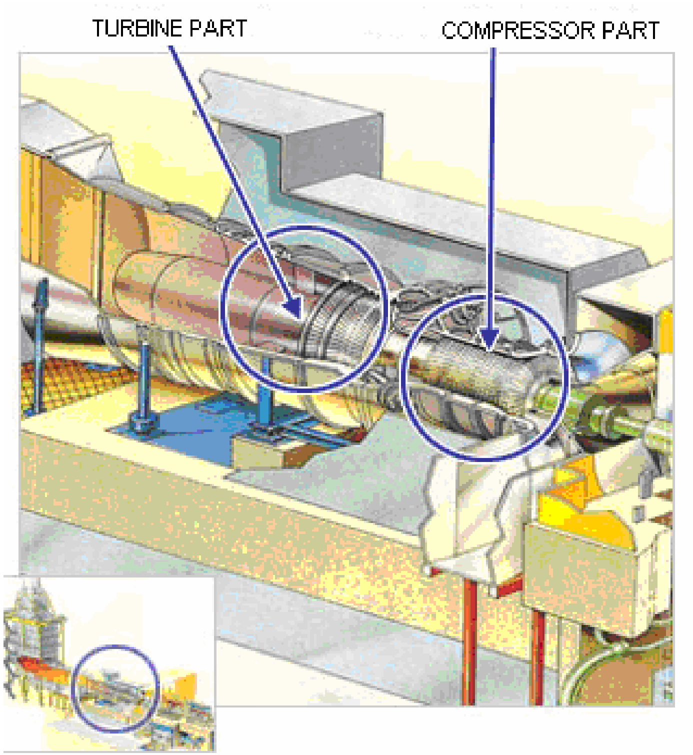


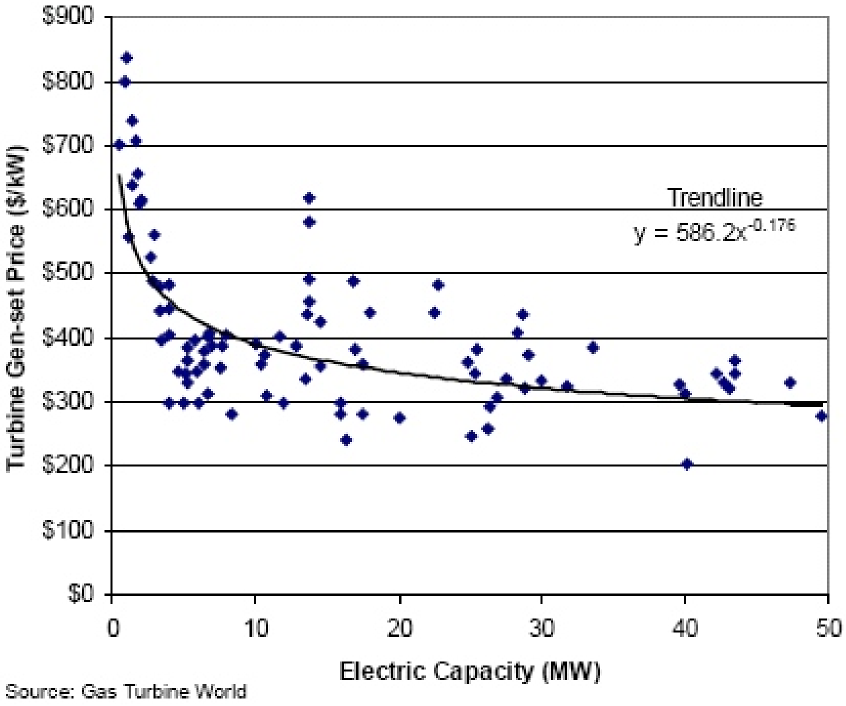
Figure 2-7. Industrial gas turbine as part of a combined cycle (Eems power plant, the Netherlands)

Efficiencies

Electrical efficiencies start from about 25% for micro gas turbines (from 30 kWe) to 40% for the largest simple cycles (334 MW_e). Combined cycles range from about 40% electrical efficiency up to 60% (Gas Turbine World, 2005). Thermal efficiencies are dependent on the temperature of the recovered heat in the heat recovery exchanger. In contrast to e.g. condensing boilers, thermal efficiencies will never exceed 100% LHV, because the dew point is 40 ºC. This temperature level is too low to extract useful heat. The low dew point is caused by the excess air, which is supplied to prevent high temperatures in the first turbine stage.

Investment costs

The scale dependence of the investment costs of gas turbines is depicted in Figure 2-8. In this figure, the specific investment cost is shown, and a regression function is derived. The scale factor for the total investment is calculated at 0.82.



Source: Gas Turbine World

Figure 2-8. Specific investment costs of gas turbines

2.5.2 *Micro turbines*

Micro turbines are very small gas turbines that are currently offered in a size range of 30 kW to 500 kW. Micro turbine technology has evolved from the technology used in automotive and truck turbochargers and auxiliary power units for airplanes and tanks. A common feature of micro turbines is that radial compressor and/or turbines are used whereas in the large gas turbines axial compressor and turbines are used.

Several companies have developed commercial micro turbine products, which are in early stages of market entry. Electrical efficiencies of 23-26% are achieved by employing a recuperator that transfers heat energy from the exhaust stream back into the combustion air stream. Micro turbines are compact and lightweight, with few moving parts. Many designs are air-cooled and some even use air bearings, thereby eliminating the cooling water and lube oil systems.

Micro turbines' potential for low emissions, reduced maintenance, and simplicity promises to make on-site generation much more competitive in the 30 to 500 kW size range characterized by commercial buildings or light industrial applications. Micro turbines for CHP duty are typically designed to recover hot water or low-pressure steam.

As the size range of micro turbines is out of the scope of this study, we will not deal with this technology in the remainder of the report.

2.6 Fuel Cells

Fuel cells produce power electrochemically. In fuel cells hydrogen fuel is delivered to the cathode of the cell, and oxygen in air is delivered to the anode. The fuel and oxygen (air) are separated by an electrolyte, through which ions can pass, but electrons cannot. By doing this, the electrons are forced to flow outside of the cell, producing electricity. The cell voltage is about 0.5-0.8 Volt. To achieve higher voltage and power output, multiple cells are built together in a stack. Figure 2-9 depicts the working principle of a fuel cell; Figure 2-10 shows the principle of a fuel cell stack.

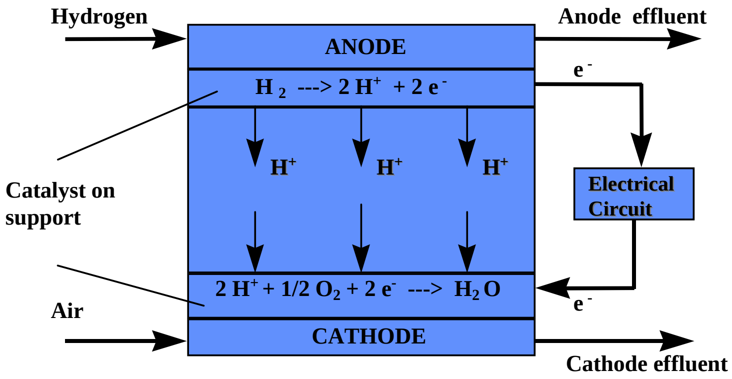


Figure 2-9. Working principle of a proton exchange membrane (PEM) fuel cell operating on hydrogen (Huijsmans, Jansen & Ruijg, 2002)

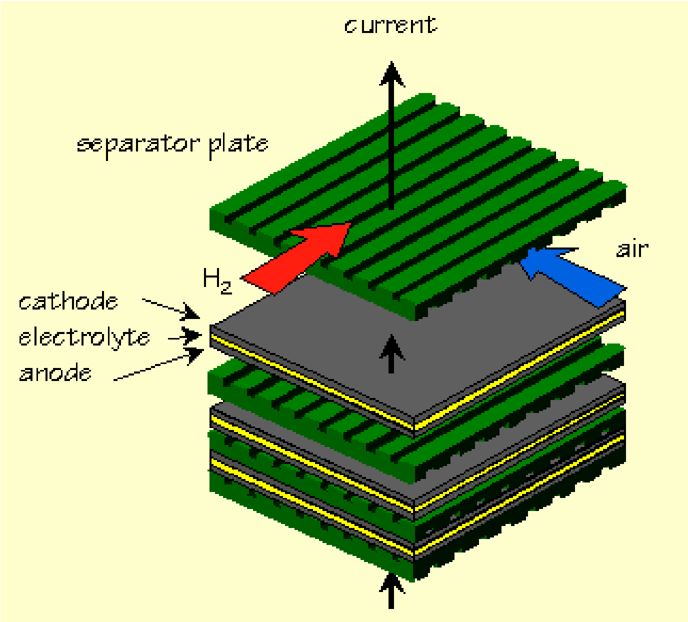


Figure 2-10. Principle of a fuel cell stack (ECN, 2005)

Hydrogen is an energy carrier, which implies that it has to be produced from other fuel sources. Hydrogen fuel can come from a variety of sources, but the most economic is steam reforming of natural gas – a chemical process that strips the hydrogen from both the fuel and the steam. Several different liquid and solid media can be used inside fuel cells – phosphoric acid (PAFC), molten carbonate (MCFC), solid oxide (SOFC), and proton exchange membrane (PEMFC). Each of these media comprises a distinct fuel cell technology with its own performance characteristics (see Table 2-2).

Table 2-2. Characteristics of different types of fuel cells (ECN, 2005)

	PEMFC	**PAFC**	**MCFC**	**SOFC**
Operation temp. [ºC]	40-80	200	650	650-1000
Fuel*	H_2 (/CO_2)	H_2(/CO_2)	H_2, CO	H_2, CO
Electrolyte	Polymer	Phosphoric acid	Molten Carbonate	Solid Oxide
Application	Transportation, Distributed generation, CHP**, Portable, Backup	Distributed generation, CHP**	Distributed generation, CHP**	Distributed generation, CHP**

*Fuel Cell types which can utilize a mixture of H_2 and CO_2 can also utilize higher hydrocarbons if combined with a fuel reformer.

**CHP: Combined heat and power

The various fuel cell technologies are in different stages of development. PAFCs are in early commercial market development now, with 200 kW units delivered to more than 120 customers worldwide. The PEMFC and MCFC technologies are now in early market introduction and demonstration. SOFC units are in development and testing. Fuel cells promise higher efficiency than generation technologies based on heat engine prime movers. In addition, fuel cells are inherently quiet and their running is extremely clean. Many fuel cell technologies are modular and capable of application in small commercial and even residential markets. Some fuel cell types operate at high temperatures in larger sized systems that would be well suited to industrial CHP applications.

Fuel cells are envisioned to serve a variety of distributed generation applications and markets. However, at this stage of development and early commercial use, fuel cells have high capital cost and high project risk due to unproven durability and reliability.

3 Market analysis

3.1 Introduction

This chapter gives an overview of the market share (by size and fuel use) of medium-scale combustion technologies described in chapter 2.

The overview of medium-scale combustion installations is valuable for several reasons. First, the overview will be used to determine the applicability of various existing and novel capture technologies that will be identified in chapter 4. The overview also gives an indication of the future potential market for carbon dioxide capture equipment and the potential emission reduction that can be obtained by applying the various technologies.

In this chapter we present the role and market share of reciprocating engines, gas turbines and boilers - currently the most important medium-sized power and heat generating installations. The information is gathered from in-house knowledge, and public literature. It should be noted that information on installation level is scarcely available, in contrast to information on plant level for which there are several databases.

The Diesel & Gas Turbine Worldwide magazine gives an overview of units of gas engines and gas turbines ordered annually. A comprehensive dataset on worldwide ordered boilers was not available. Nevertheless, for some regions in the world some information on installed boiler units is available, notably the United States, Europe and China. Another source of information is databases, like the European National Allocation Plans (NAPs), the IEA CO_2 emissions database compiled by Ecofys and the European Pollutant Emission Register (EPER). However, these databases have only aggregated information on company level rather than on installation level.

The medium-sized installations often operate in the CHP-mode, e.g. by attaching a steam turbine to a boiler or by adding a heat recovery steam generator with steam turbine to a gas turbine. As microturbines and fuel cells are just a very small portion of the market, we only dedicate a small paragraph to this type of installation. Fuel cells, however, might become interesting in the medium to long-term future.

As information is sometimes given in MW_{th} and sometimes in kt (or Mt) of CO_2 Table 3-1 provides a conversion table from MW_{th} to $MtCO_2$. The corresponding amount of CO_2 produced per installation capacity differs according to the fuel used. Table 3-1 shows the installation size (in MW_{th}) for five installation capacities and three fuels (coal, natural gas or oil). For each installation an operating time of 7500 hours per year is assumed. Coal has the highest carbon content and therefore a coal-fired installation with the same capacity as a natural gas or oil-fired installation produces a larger stream of CO_2. Medium-scale installations can therefore be qualified as installations producing annually up to 250 kt of carbon dioxide, corresponding up to about 93 MW coal fired-installations or up to 165 MW natural gas-fired installations. It should be noted that many medium-scale installations are used only a fraction of the time (e.g. as stand-by, peak load, or partly during the year).

Table 3-1. Combustion installation capacity (MW_{th}) based on annual emission and fuel type

Annual CO_2 emissions	**50 kt/y**	**100 kt/y**	**150 kt/y**	**200 kt/y**	**250 kt/y**
Coal (EF=100kg/GJ)	**19 MW_{th}**	37 MW_{th}	56 MW_{th}	74 MW_{th}	93 MW_{th}
Oil (EF=76 kg/GJ)	24 MW_{th}	49 MW_{th}	73 MW_{th}	97 MW_{th}	122 MW_{th}
Natural gas (EF=56 kg/GJ)	33 MW_{th}	66 MW_{th}	99 MW_{th}	132 MW_{th}	**165 MW_{th}**

3.1.1 Plant size versus size of installation

Information on installation level is scarcely available and often relatively outdated. This is mainly due to the liberalisation of the market, which makes companies reluctant to provide detailed information on the operational activities. In this section we provide an overview of the share in total CO_2 emissions for each defined emission category for both installation level and plant level. This shows to what extent the share of CO_2 emissions from medium-scale plants change according to the level of analysis (plant or installation) Furthermore, this might give a rough indication of the break-down in other industrialised countries.

In this section emission data on plant level and installation level are compared to see the effect of level of analysis on the spread of installations/plant in the various emission categories. The results of this analysis are presented in Figure 3-1.

The information on installations is taken from the Dutch emission database ERL - “Emissie Registratie Lucht” (1986). Although updates have been made for this database, confidentiality agreements make it impossible to access the required information. As we use the data just for a breakdown of the size of installation, we regard the issue that the information may be somewhat outdated as a minor problem.

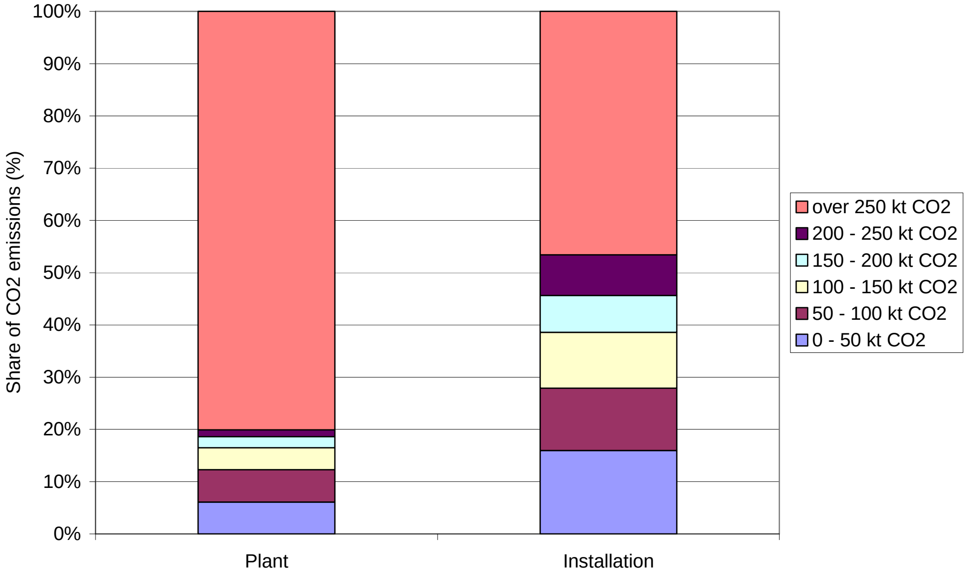


Figure 3-1. Share of CO_2 emissions in six emission classes for plants and individual installations covering all industries in the Netherlands (ERL, 1986)

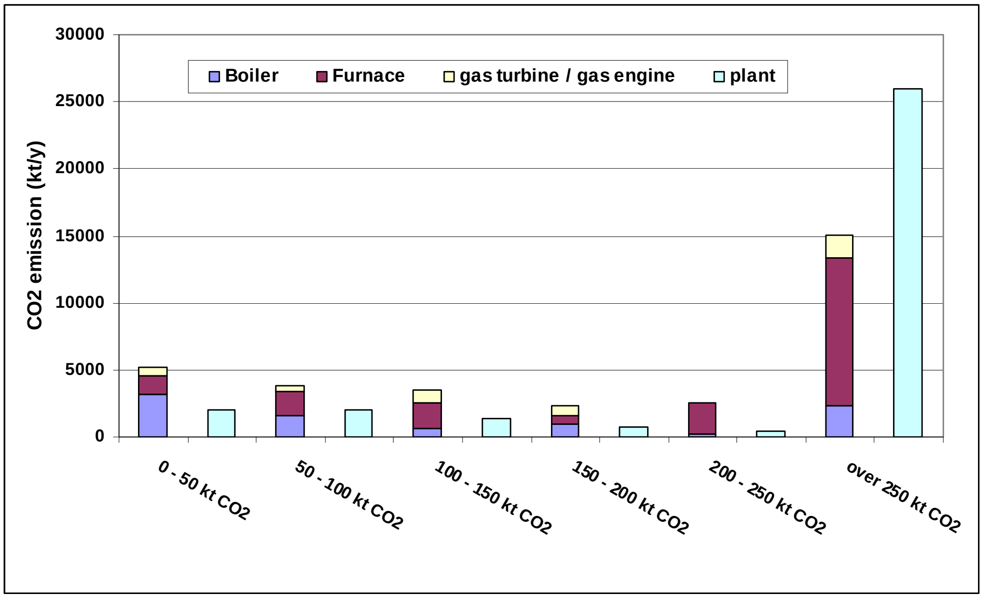


Figure 3-2. Division of emissions per emission size category for installations and plants

The left bar of Figure 3-1 shows the share of CO_2 emissions per emission category for *plants*. Around eighty percent of total CO_2 emissions are emitted by plants with emissions over 250 kt per year. Because in one plant often more combustion installations as gas turbines, furnaces and boilers are present, the division over emission categories by installation shows a different picture. From the right column in Figure 3-1 it can be seen that about 50% of the CO_2 emissions stems from installations emitting over 250 ktCO_2 per year. Installations that emit less than 50 kt per year are responsible for about 15% of total CO_2 emissions. On plant level, small plants (less than 50 kt CO_2) account for only 5% of total CO_2 emissions.

Figure 3-2 shows the same division as Figure 3-1 but in absolute emissions and subdivided per type of installation. It should be noted that the group ‘gas turbines’ comprises gas turbines as well as gas engines (especially in the lower capacity classes). The figure also shows that the greatest part of the (smaller) installations is furnaces/boilers. This distinction is somewhat more addressed in the next section.

From Figure 3-1 and Figure 3-2 it can be concluded that a number of medium-scale installations undeservedly might be excluded from the analysis when using information on plant level. It must be noted that it is highly probable that the amount of CO_2 that can be captured from medium-scale combustion installations is rather conservative when using information on plant level. Actual potential of carbon dioxide capture from medium-scale installations could be significantly larger based on above results for installations in the Netherlands.

3.2 Sector

The power and heat generation technologies identified in chapter 2 can be found in various economic sectors. Generally, the following sector division is made:

Commercial sector: The commercial sector is generally defined as non-manufacturing business establishments, including hotels, motels, restaurants, wholesale businesses, retail stores, and health, social, and educational institutions. Commercial refers to any building that is neither residential (used as a dwelling for one or more households), manufacturing/industrial (used for processing or procurement of goods, merchandise raw materials or food), nor agricultural (used for the production, processing, sale, storage, or housing of agricultural products, including livestock).

Industrial sector: The industrial sector is generally defined as manufacturing, construction, mining, agriculture, fishing and forestry establishments; sometimes *agriculture* is regarded as a separated sector.

In the commercial sector, some of the prime movers function as standby power or for peak shaving. The standby power only provides power when the primary source is out of service or is not working properly. Peak shaving equipment operates typically between a few hundred hours and a maximum of a few thousand hours per year. Both categories, standby and peak shaving, are not appropriate installations for capturing CO_2, as technologies need to be simple (standby), flexible (short start-up period) and run a significant part of the year in order to recover the extra investment costs for capture.

The National Allocation Plans (NAPs) for the 25 member states of the European Union provide information on the economic sectors in which to find combustion installations. The NAPs database covers all plants that are under the EU Emission Trading Scheme (EU-ETS); including about 10512 emission sources around Europe. It should be noted that combustion installations with an installed capacity of less than 20 MW_{th} do not fall under the Emissions Trading Scheme and therefore are not included in the NAPs. According to the categorisation used in the NAPs, the installations either fall in the category 'Combustion >20 MW_{th}' or in a specific industry sector. Unfortunately, the categorization that is used is not always clear and consequently is not always used throughout the Member States of the European Union. Normally, boilers are included in the category 'Combustion >20 MW_{th}'.

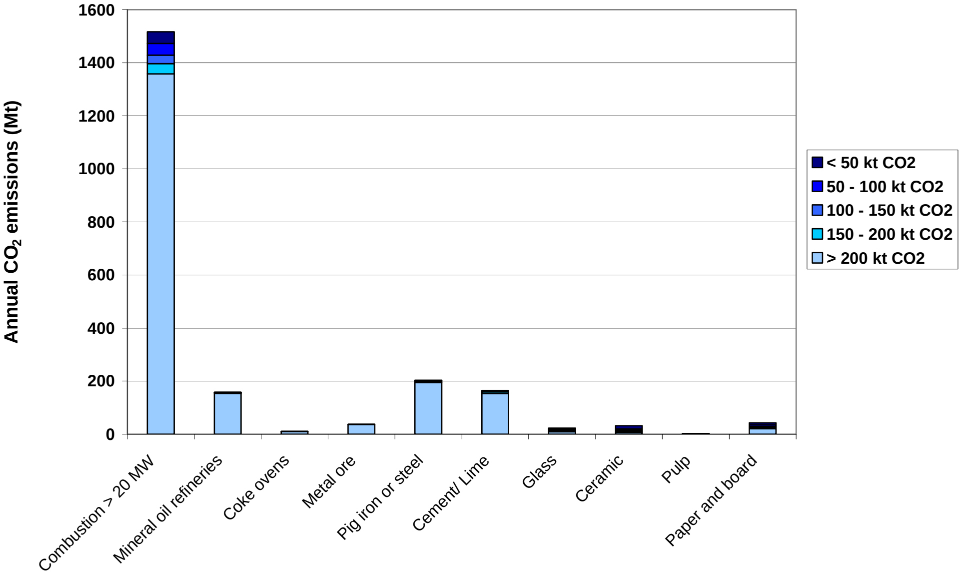


Figure 3-3. Annual emissions from combustion installations under the EU emission trading scheme (source: Ecofys, 2005)

From Figure 3-3 it can be concluded that about 90% of the emissions originate from plants participating in the NAPs that emit over 200 kt per year.

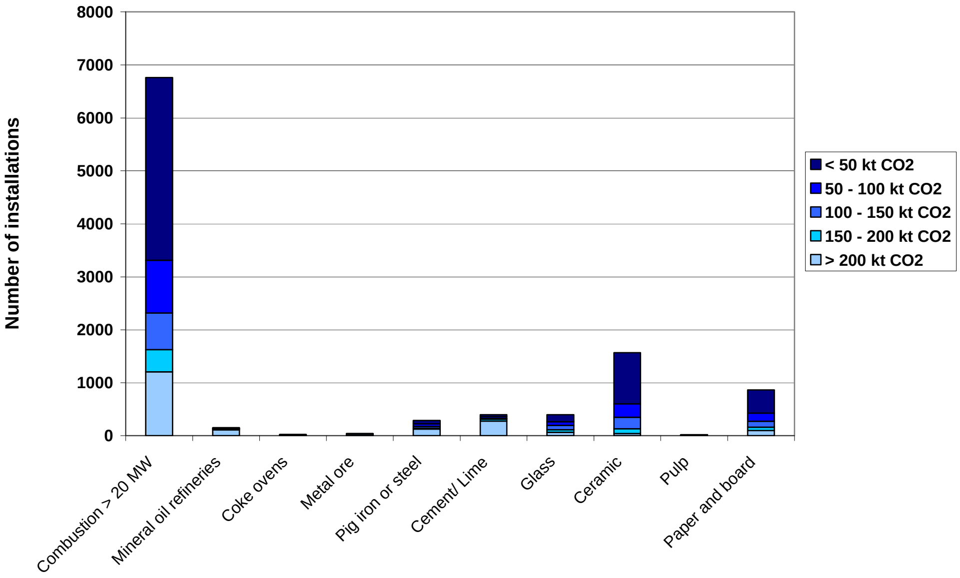


Figure 3-4. Number of plants under the EU emission trading scheme (source: Ecofys, 2005)

Figure 3-4 shows that 80% of the plants under the NAPs are in the range of 50 kt to 200 kt CO_2. Also it can be concluded that 10% of the plants emit 80% of the CO_2. However, NAPs do not give information on individual installations. In paragraph 3.1.1. it is explained that the number of medium-scale installations and the potential of CO_2 that could be captured might be underestimated when using information on plant level.

3.3 Type of installation

In this section we analyse the type of installation, its application and market share. We mainly do this based on information from installations worldwide and more specifically based on information from the United States, China, Europe and the Netherlands.

3.3.1 Boilers

Industrial and commercial boilers are available in a wide range of capacities, ranging from less than 3 MW to over hundreds of megawatts. A recent survey in the United States included about 163,000 boilers with a total capacity of 810,000 MW. The inventory included 43,000 industrial boilers[1] with a total capacity of 480,000 MW and 120,000

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[1] Not included in this figure are electric boilers (22,000) and boiler smaller than 0.3 MW_{th} input (387,000).

commercial boilers with a total capacity of 330,000 MW. Boilers consume about 40% of the energy in the industrial and commercial sector, of which 75% in boilers smaller than 73 MW_{th} (see Figure 3-5).

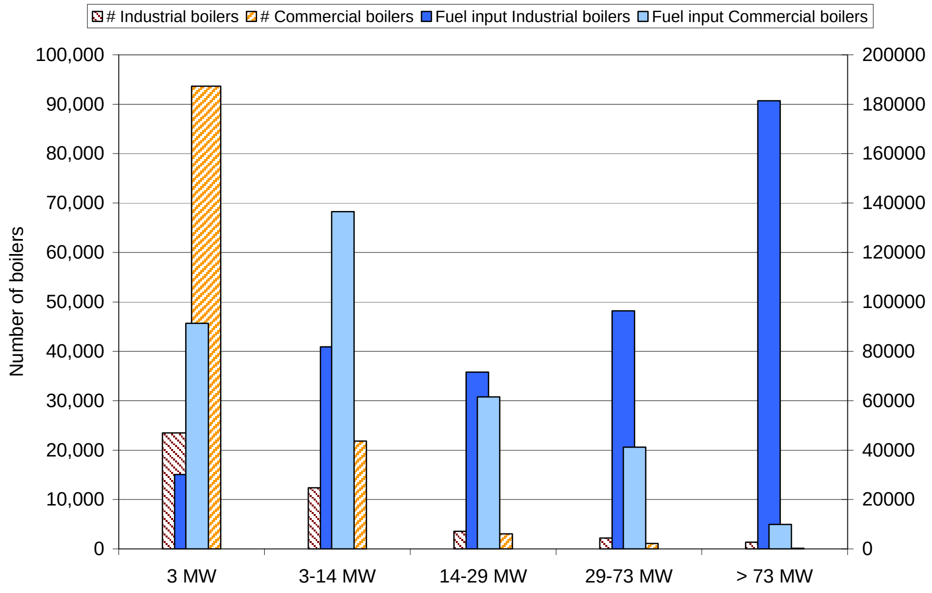

Figure 3-5. Number of boilers (shaded bars) and fuel input capacity (blue bars) of various boiler capacity classes (industrial boilers and commercial boilers) in the United States (ORNL, 2005)

More than 95% of the industrial and commercial boilers fall in the capacity range of less than 73 MW (see Figure 3-5). Commercial boilers at office buildings, health care facilities and educational establishments account for almost two-third of the commercial boiler units and capacity (see Figure 3-6). The on average largest boiler capacity is in the health care facilities (over 6 MW).

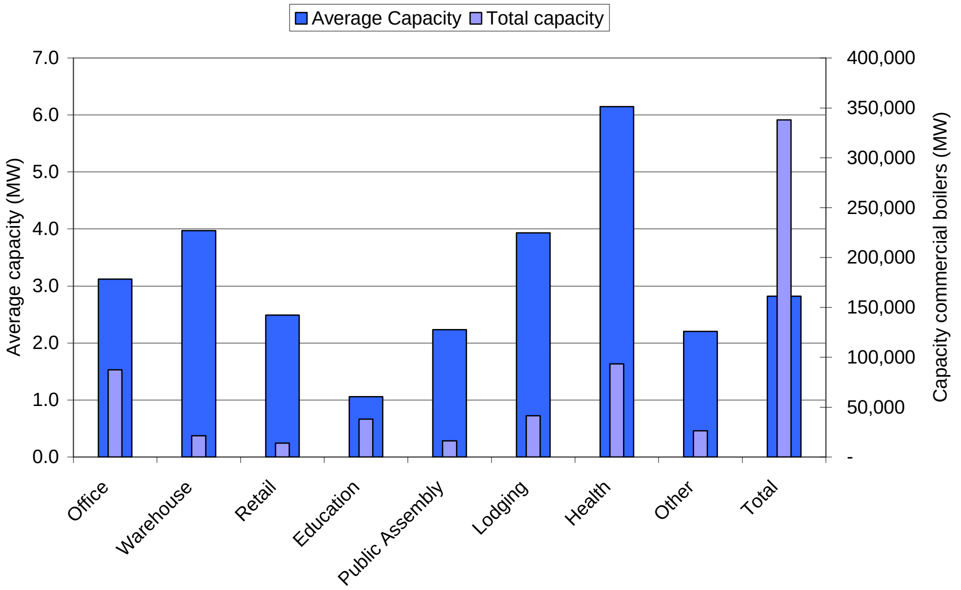

Figure 3-6. Total capacity and average capacity for commercial boilers classified by commercial sector

The most important steam-intensive industries are paper and pulp, chemical industry, refineries and primary metals (see Figure 3-7). In the USA these industries have installed 82 percent of all boiler capacity. Large capacity boilers are typically found in the paper and pulp industry and refineries.

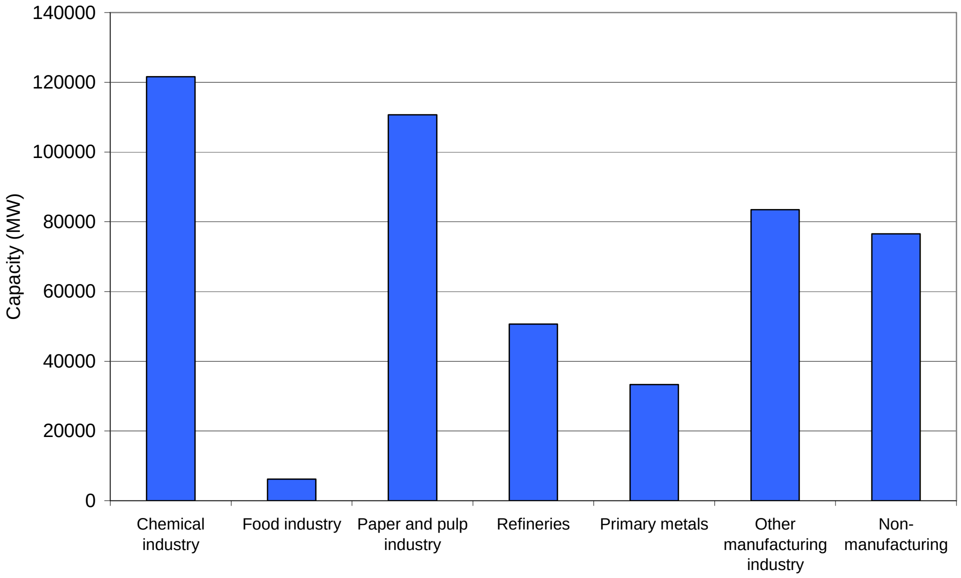

Figure 3-7. Total boiler capacity per industrial sector (ORNL, 2005)

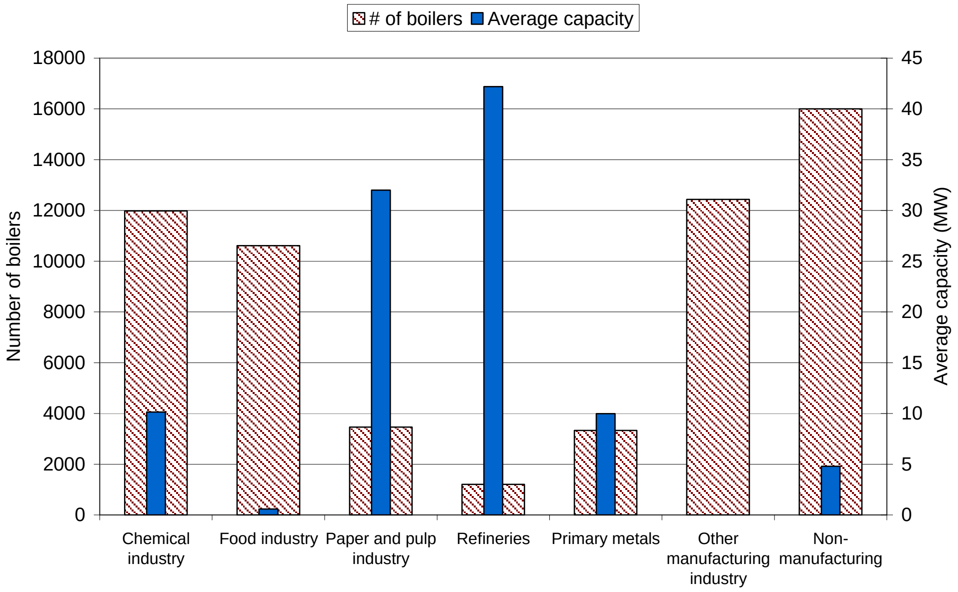

Figure 3-8. Number of boilers in industry (blue bar) and average capacity (red bar) in the United States (ORNL, 2005)

Figure 3-8 shows the average capacity of boilers in the major steam-intensive industries in the United States. It shows that the number of boilers in the food and chemical industry is large, but capacities are relatively small. Probably most of the boilers installed in the food industry are smaller than 1 MW and fall out of the scope of this study. Considerably less in number, but much larger boilers, can be found in the pulp and paper and refinery industry.

Besides the manufacturing industries, boilers are also used in the agricultural sector, mining and engineering. The boiler capacity in these sectors is estimated at about 14 percent of the manufacturing boiler capacity. The average capacity is about 5 MW.

Natural gas is the most common fuel in the USA for industrial boilers and is identified as the primary fuel for 78 percent of the boiler units and 56 percent of boiler capacity. Some industries use by-product fuel in a large portion of their boiler capacity. On a United States national scale, the coal, oil and wood fuel only count for a small portion of the boiler capacity. This may be different in other regions in the world, e.g. in China, where the share of coal in boilers is considerably higher.

<u>China</u>

By the end of 2003, over 525,000 industrial boilers were in use in China, with a total capacity of 1254 GW. Sixty percent of the total capacity is steam boilers, and the rest (40%) is hot water boilers. Forty-five percent of the total capacity is used for industrial purposes and 55% for residential purposes. Table 3-2 presents an overview of the development in both boiler capacity and number of boilers in industry.

The average operating efficiency of the industrial boilers in China is around 65%, which is 15 to 20 percentage points lower than the international advanced level. More than 80% of the boilers are layer-combustion boilers with low efficiency and high pollutant emission. The realization of high-efficiency and clean coal combustion for medium and small-sized coal-fired boilers is the emphasis of cleaner energy technologies in China [Du Minghua, 2001].

Table 3-2 Industrial Boilers in use in China (WANG Shanwu, 2005)

Year	Total number (1000)	Capacity (GW)				
		Steam	**Hot water**	**Total**	**Production**	**Life**
1990	491	589	273	862	483	378
2000	510.5	591	284	875	482	393
2001	509.3	550	271	821	470	350
2002	469.7	734	570	1304	513	791
2003	527.4	756	497	1254	562	691

Among the industrial boilers in use, about 90% (in terms of number) of the hot water boilers are situated in Northeast China, Northwest China, and North China. In recent years pollution control requirements have become more stringent and China strives to a more balanced energy supply diversification. Therefore the share of oil-fired boilers is increasing, partly accomplished by converting existing coal-fired boilers into oil-fired boilers. According to the Boiler Sub-committee of China Electrical Appliance Industrial Association, in Beijing, Nanjing, and some other cities, the share of oil-fired boilers has already reached over 50%.

Considering the size of industrial boilers, analysis by some industrial experts indicates that 99% of the industrial boilers in use in 2001 have a capacity of less than 35 t/h (approximately 25 MW_{th}). More specifically, the share of industrial boilers with a capacity of more than 20t/h (14 MW_{th}) is less than 20%; and 75% of the industrial boilers are of a capacity between 2 and 10 t/h (7 MW_{th}), and 5% are less than 1t/h (0.7 MW_{th}). However, as far as the number of boilers is concerned, around 35% still has a capacity of less than 2 t/h (1.4 MW_{th}), about 60% are of a capacity between 2 and 10 t/h (1.4-7 MW_{th}), and approximately 5% of the boilers are of a capacity above 10t/h (7 MW_{th}). Total heat capacity amounts to about 880 GWth in various industries and social sectors.

The average life of industrial boilers is 15 to 20 years. Currently, most of the industrial boilers in use in China were made and put into use in the 1980s. A small share of the boilers, around 10%, is already more than 20 years in use. The majority of the industrial boilers older than 15 years are coal-fired. Industrial boilers still take coal as the most important fuel, with some also using oil and gas. It is estimated that the 480,000 coal-fired boilers for industrial and residential purposes each year consume around 12 EJ. In China, the average load rate of industrial boilers is between 50% and 70%. Generally speaking, the smaller the size of boilers, the lower the average load rate.

3.3.2 Reciprocating engines

The smaller capacity range of combustion technologies is mainly covered by gas engines. Below 3.5 MW_e, diesel and gas-fired reciprocating engines are the dominant segment players. Above 25 MW_e the dominant technology is gas turbines. The mid-range is equally shared between the technologies although the trend is that smaller gas turbines, despite their somewhat poorer performance, become more available in the lower range classes as they have better environmental performances (e.g. lower methane slip and other emissions).

The most prevalent on-site generation application for reciprocating engines is CHP. The economics of such applications are enhanced by effective use of the thermal energy contained in the exhaust gas and cooling systems, which is normally equal to 60 to 70% of the input fuel energy. Medium pressure steam can be generated from the engine's high-temperature exhaust gas, but the hot exhaust gas contains only about one-half of the available thermal energy from a reciprocating engine. Sometimes the hot exhaust gases are directly used for process drying.

The market for gas engines is growing rapidly (see Figure 3-9). The number of units ordered over 1 MW_e increased by 41% in 2005 compared to 2004, while the total output increased by roughly 29% to almost 17,000 MW. Orders for the 7.5 to 10 MW category increased by more than 100% from 64 units in 2004 to 137 units in 2005. The bulk of the engines ordered above 3.5 MW were destined for continuous-duty operation. Geographically, both North America and the Far East, which includes China, reported large increases in ordered engines.
Figure 3-10 shows the number of engines ordered in 2004 subdivided into size category, fuel type and region. The overall picture is that the power generation market for reciprocating engines is robust. Although natural gas engine orders are increasing, the impact of much higher gas prices will have an impact on that area.

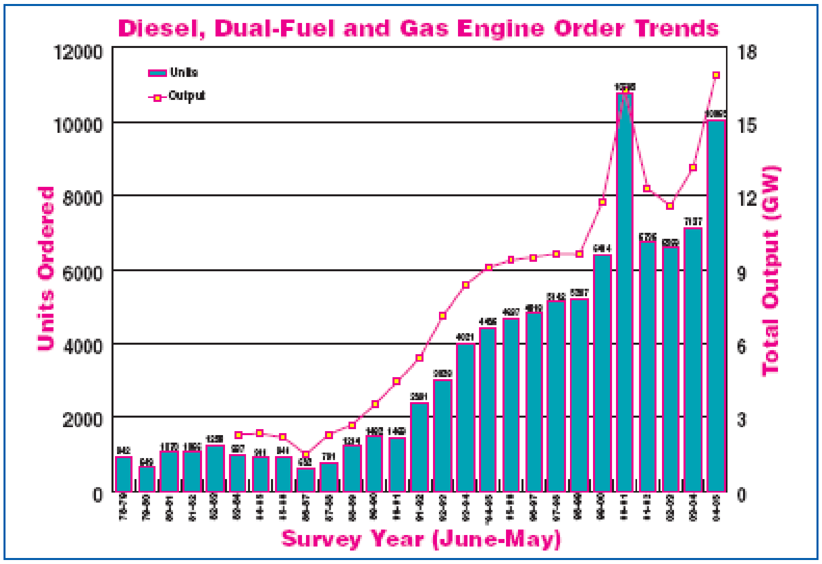

Figure 3-9. Trend in units ordered and total output of reciprocating engines (Diesel & Gas turbine worldwide, 2005)

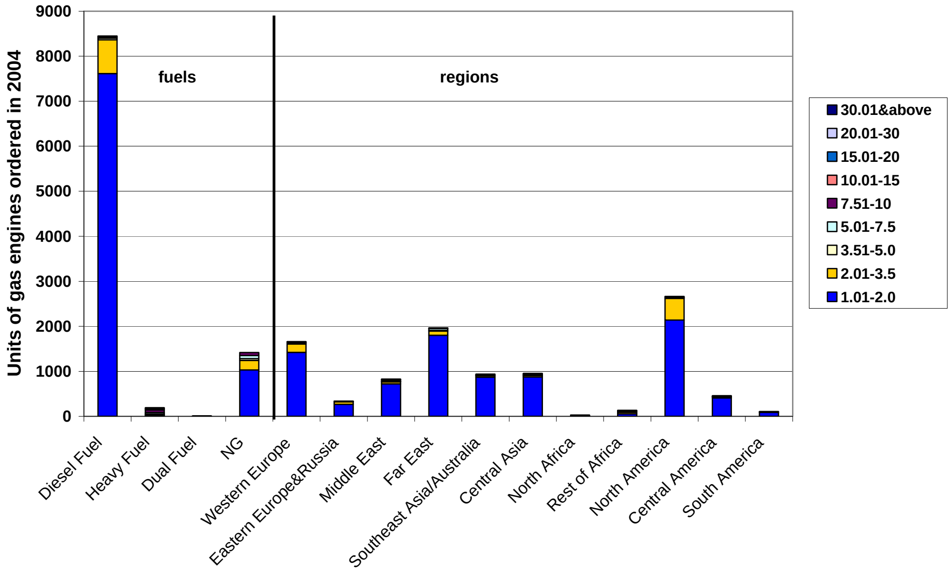

Figure 3-10. Engines ordered per size category per fuel (left side of the graph) and per region (right side of the graph) (Diesel & Gas Turbine Worldwide, 2005)

Where are gas engines placed – an example from the United States

In 2000, over 1000 engine-based CHP systems were operating in the United States, representing more than 800 MW of electric power capacity. Facility capacities range from 30 kW to more than 40 MW, with many larger facilities comprising multiple units. This is, however, a small fraction of the total power capacity in the United States, which amounts to over 750,000 MW_e (including 150,000 MW_e owned by non-utility).

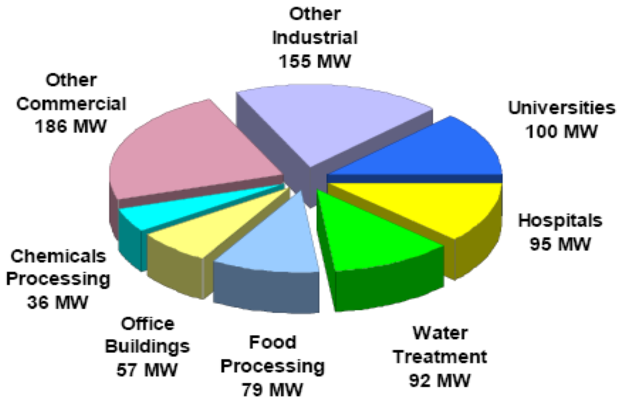

Figure 3-11. Existing reciprocating engine CHP in the United States (801 MW at 1055 sites)

From Figure 3-11 it can be seen that applications for engine CHP are mainly in buildings with relatively high and coincident power and hot water demand, such as colleges and universities, hospitals and nursing homes. Engine driven CHPs are also used in a variety of industrial applications where hot water or low-pressure steam is required for process needs or space heating.

3.3.3 Gas turbines

Gas turbines are available in sizes ranging from 0.5 MW to 330 MW. Figure 3-9 shows the trend in sales since 1978/1979. In 2004 almost 700 units were ordered with a total output of almost 35,000 MW. Total units ordered remained essentially unchanged over 2004, but the total output decreased by approximately 10%, reflecting the fact that fewer large machines were ordered. Figure 3-12 shows the trend of units ordered since the end of the seventies.

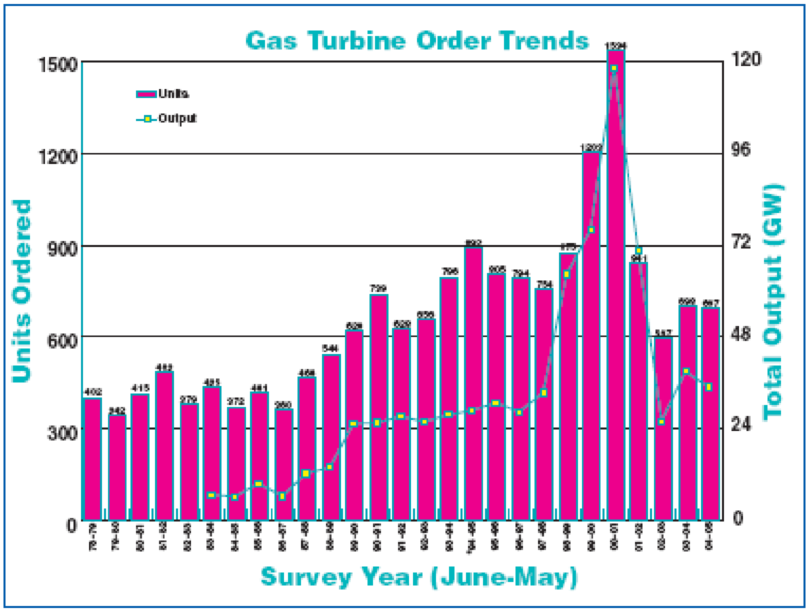

Figure 3-12. Trend in units ordered and total output of gas turbines (Diesel & Gas Turbine Worldwide, 2005)

In this analysis we focus on small gas turbine systems in the size range of 1 to about 40 MW_e. About 15% of total capacity ordered falls in this range The largest growth in orders was in the output range of 20 to 30 MW, where orders increased by almost 60%, to almost 100 units in 2005. In a geographical perspective, orders went up especially in the Far East, North America, Central America and Africa. Figure 3-13 shows the units ordered in 2005 by fuel and by regions for various size categories. Units ordered running on diesel fuel and heavy fuel oil both increased, 21% and 12% respectively, while units on natural gas fuel decreased by 23%.

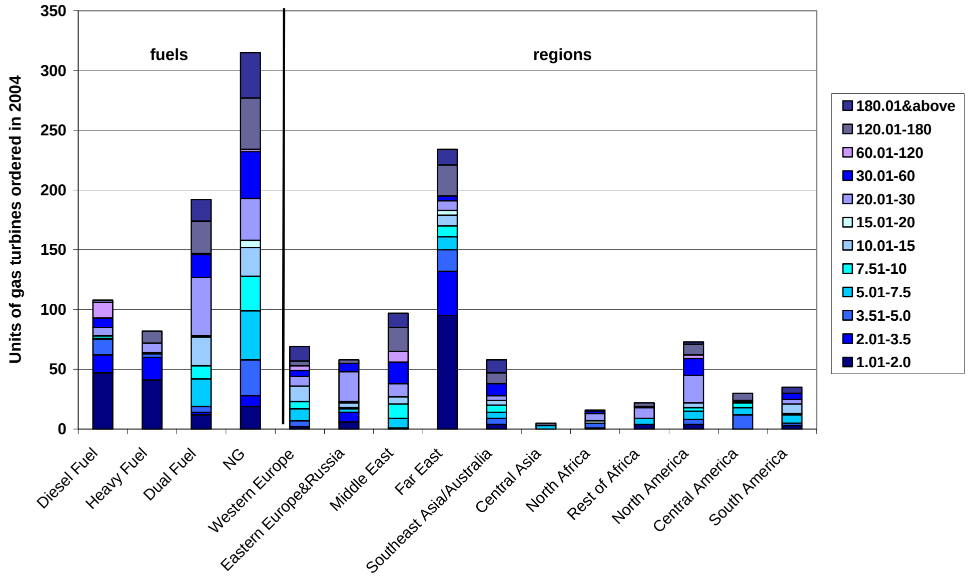

Figure 3-13. Gas turbines ordered per size category (in MW$_e$) and fuel (left side of the graph) and per region (right side of the graph)

The efficiency and reliability of small gas turbines (1 to 40 MW) has progressed sufficiently to be an attractive choice for industrial and large institutional users for CHP applications. Gas turbines are commonly used in the oil and gas industry to drive pumps and compressors, in process industries to drive compressors and other large mechanical equipment, and in many industrial and institutional facilities to generate electricity for use on-site.
The most efficient commercial technology is the gas turbine combined cycle, with efficiencies approaching 60% (LHV). Simple-cycle gas turbines are available for on-site generation with efficiencies approaching 40% (LHV). The general trend is that gas turbines are increasingly used for base-load power.

Like reciprocating engines, the economics of gas turbines in on-site generation applications often depend on effective use of the thermal energy contained in the exhaust gas, which generally represents 60% to 70% of the input fuel energy. The most common use of this energy is for steam generation in unfired or supplementary-fired heat recovery steam generators (HRSGs). However, gas turbine exhaust gases can also be used as a source of direct process energy, for unfired or supplementary process fluid heaters, or as preheated combustion air for boilers.

According to Diesel & Gas Turbine Worldwide, the market for gas turbines is likely to stay stable for the next years, but is highly dependent on external factors like the natural gas price. Several years ago, one high level discussion predicted the gas turbine power generation market at roughly 60,000 MW yearly. From the previous three years seen, it is likely that in the near term gas turbine orders will total something closer to 40,000 MW on a yearly basis.

Where are gas turbines placed: example from the United States

In 2000, a total capacity of 40,000 MW of gas turbine-based CHP systems was in operation in the United States, located at more than 575 industrial and institutional sites. This amount comprises about 7% of the total installed capacity (750,000 MW) in the United States. About 10,000 MW was installed as simple-cycle gas turbine-based CHP system, typically in units less than 40 MW in size. Figure 3-14 shows that the largest capacity is installed in the oil and chemical industry.

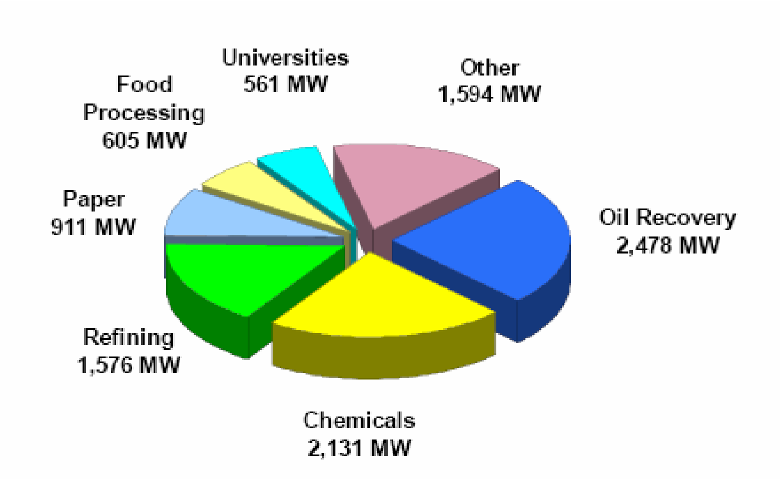

Figure 3-14. Existing small gas turbines in the United States (9,854 MW at 359 sites)

Figure 3-15 shows the share of gas engines and gas turbines for twelve capacity classes. It is clear that gas engines serve the lower capacity ranges and gas turbines the higher capacity ranges. Worldwide the annual average new capacity in the period 2001 to 2010 is estimated at about 150 GW. For installations smaller than 30 MW this amounts to about 25 to 30 GW per year.

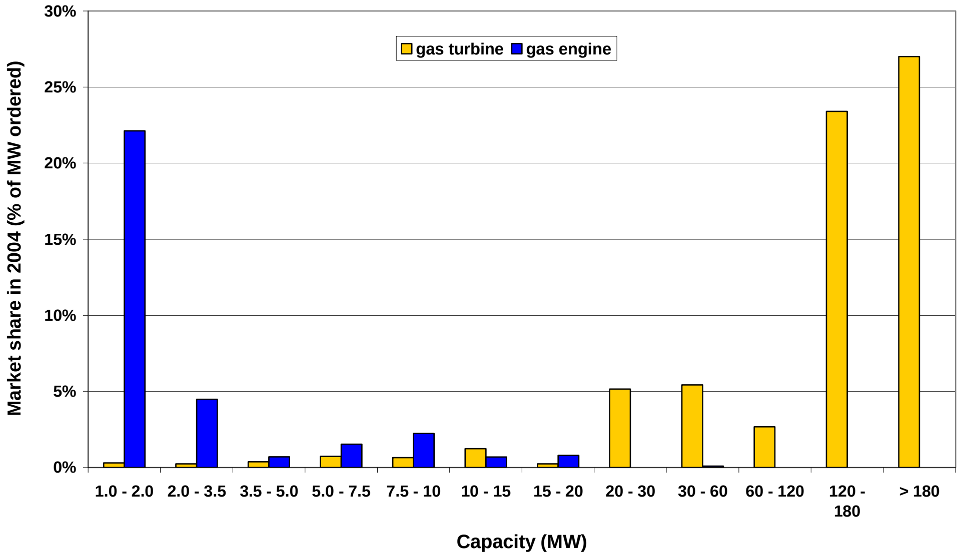

Figure 3-15 Share of ordered gas engines and gas turbines in 2004

The main fuel for gas engines and gas turbines is natural gas. A small share of the machines is fired by coal (e.g. through gasification) or oil.

Figure 3-16 depicts the capacity of gas turbine and gas engines sold in 2004 with the estimated associated power production and related CO_2 emissions of these machines. The total CO_2 emissions are estimated at 178 Mt. Assuming a lifetime of 20 years, the total emissions will be 3.6 $GtCO_2$.

Figure 3-17 presents an estimate of the economics of the machines sold in 2004. The emissions are subdivided into type of fuel; diesel, heavy fuel, dual fuel and natural gas for both types of machine; gas engines and gas turbines.

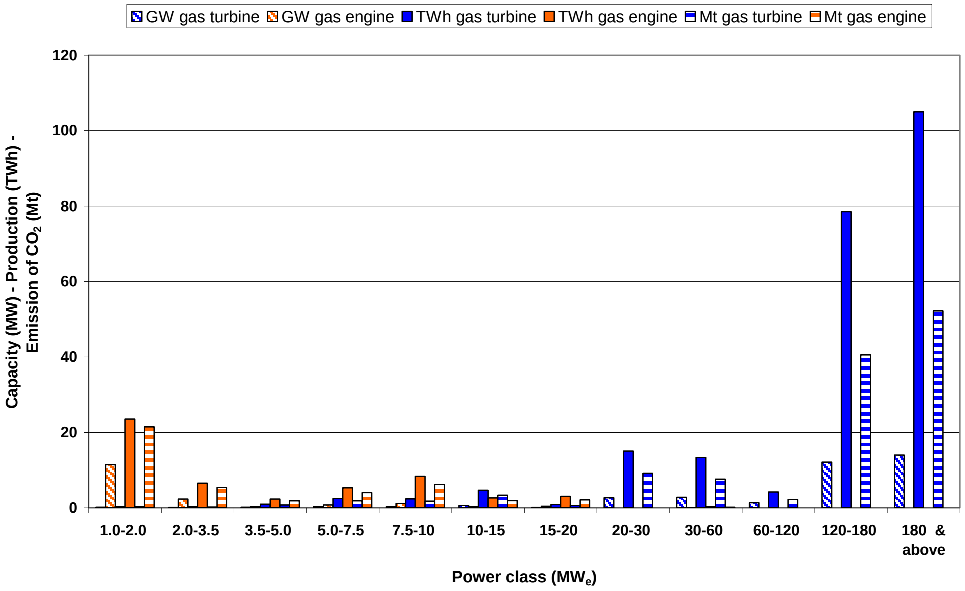

Figure 3-16. Capacity sold in 2004, associated power production (assuming 70% load factor) and related CO_2 emissions

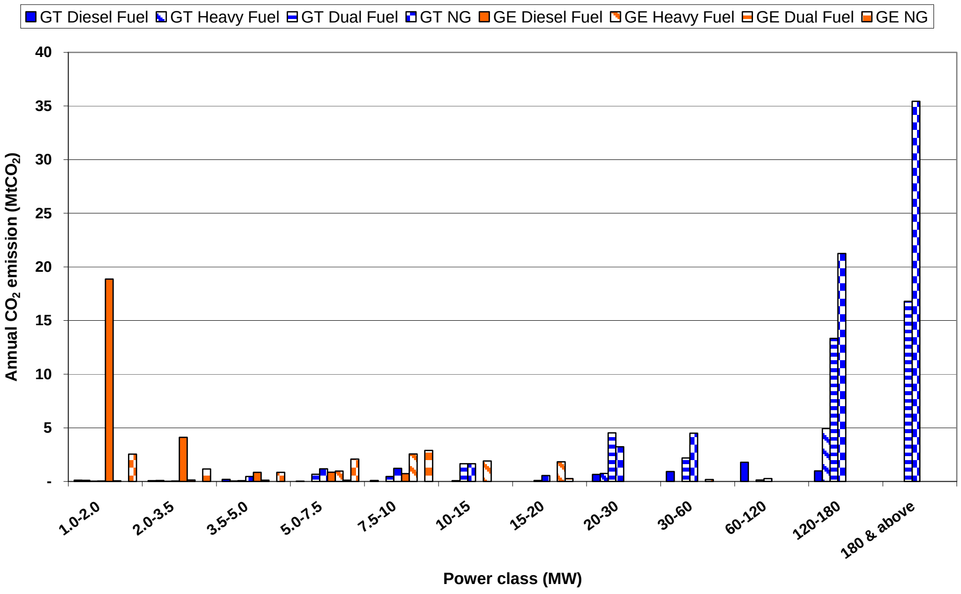

Figure 3-17. Emission of CO_2 subdivided by type of machine and fuel

3.3.4 Fuel cells

Fuel cells are emerging installations mainly addressing the capacity range below 5 MW_e. Unlike large turbine systems competing against higher priced coal-fired power plants, these emerging products are competing against products that feature lower capital costs.

Stationary applications comprise:

- Small-scale, on-site, non-utility power generation (3kW-1MW).
- Commercial CHP (up to 1 MW).
- Distributed power generation (1-30 MW).
- Centralised power generation (>100 MW).

The use of fuel cells for centralised electricity generation is unlikely within the next ten years, although by 2015 Europe, Japan and the USA could all have significant installed capacity. This market looks particularly promising in Japan where power generation costs are high.

Fuel cells, particularly high temperature fuel cells, are potentially well suited to the distributed power generation and CHP markets, both of which are growing. However, their market share will depend on long-term performance and cost-competitiveness.

3.4 Type of fuel

The fuels used in combustion processes vary between oil, gas, coal or biomass and waste. The main fuel used in OECD countries is natural gas. Fuels like coal, oil and biomass/waste are used to a lesser extent in medium-scale combustion installations. Coal boilers (and also waste/by-product boilers) are generally much larger than average size. In the United States they contribute to just 3% of the total number of boilers, but have a capacity share of 14%. In contrast to OECD countries, coal is the pre-dominant fuel in boilers in China. For gas engines, the share of capacity in diesel is high, but these installations are more often used as standby units.
Figure 3-18 shows the fuel use for the number and capacity of industrial boilers in the United States. A similar division can be found for European boilers. The use of natural gas in boilers in the commercial sector is even more dominant and constitutes about 85% of the fuel used. When restrictions on emissions become more stringent, fuels like biomass might be used more extensively in the future.

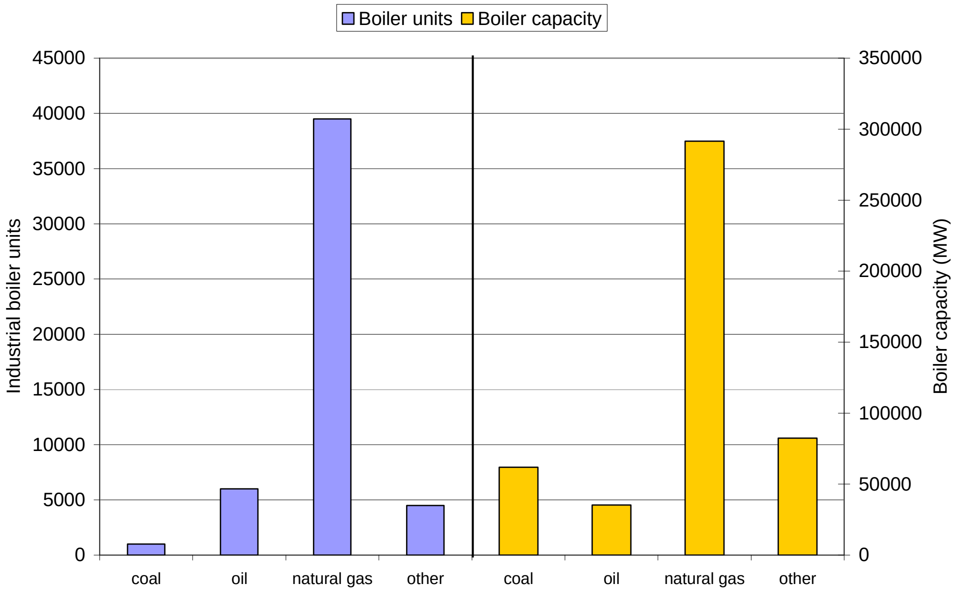

Figure 3-18. Industrial boilers and boiler capacity by primary fuel use in the USA

3.5 Flue gas composition

The analysis of the composition of the flue gases of the plant concentrates on the concentration of carbon dioxide in the flue gas. The amount of CO_2 per unit of volume of flue gas is depending on the carbon content of the fuel used and the excess amount of air used in the combustion. Data on large-scale CO_2 emission sources are used, derived from the IEA CO_2 emissions database to address the composition of the flue gas. The CO_2 concentrations in flue gases resulting from the power sector vary from 3% to 15% depending on the fuel and technology used. Concentrations of CO_2 in flue gases produced with steam turbine technology are 8% for oil and gas and 15% for coal. CO_2 streams with a 15% concentration of CO_2 make up 44% of the total CO_2 emissions in Europe (IEA GHG, 2002).

3.6 Conclusion

Information on combustion installation level is only fragmented and scarcely available. The main types of small and medium-sized combustion installations are gas engines, gas turbines, furnaces – including boilers, micro turbines and fuel cells.

From the available information it can be seen that reciprocating engines dominate in the smaller capacity classes (up to 5 MW). Above this level gas turbines are more dominantly in use. The trend is that both gas engines and gas turbines become more and more available in the capacity classes which originally were not their domain.
The number of gas turbines, and especially gas engines, is large but the share in total installed capacity and CO_2 emissions is limited and does not account for more than a few

percent of the total emissions from combustion installations. Furnaces and boilers are considerably more important in terms of CO_2 emission volumes. Furnaces are the dominant sources for heat and CHP installations, especially in the smaller capacity range. In the United States, for instance, practically all boilers in the commercial sector are smaller than 73 MW_{th}. In industrial boilers, about 60% of the fuel is used in boilers smaller than 73 MW_{th}.
Most boilers between 1 – 100 MW_{th} can be found in the chemical industry, primary metals industry and non-manufacturing industry. Boilers in the paper and pulp industry and refineries typically have larger sizes.

Fuel cells are currently insignificant in terms of capacity installed. However, in ten to twenty years from now, the role of fuel cells in some industrial sectors may substantially increase. Micro turbines are typically much smaller than 1 MW, and therefore do not fit in the capacity range (1 to 100 MW) set in this study.

The main fuel in the medium-sized installations in the USA and Europe is natural gas, with a share of more than 80%. The main fuel in China is coal.

4 Capture technologies for medium-scale combustion plants

4.1 Introduction

In this chapter we present an extensive inventory of existing and emerging CO_2 capture technologies, which may be best suited for medium-scale combustion installations for (combined) generation of heat and power in the range from 1 to 100 MW fuel input. We start with the identification of a 'long list' of capture processes for power and heat generation, which is based on a working scheme with capture principles. These capture processes are discussed with respect to their applicability for the power and heat generation installations as described in chapter 2.

A structured approach is used, based on working scheme Table 4-1, a long list of CO_2 capture possibilities is created (Table 4-2). Subsequently, applications for medium-scale heat and power generation are listed (paragraph 4.3). For each generation option the application of technologies from the long list is analysed.

For each of the separation technologies a fact sheet is drawn up in which the following items are addressed (Appendix F.1 to F.18):

- Block flow diagram and key characteristics: data are gathered from IEA GHG studies, supplemented with results from in-house studies and public literature.
- Characterisation of separation technology: the working principle is described and for each of the main separation operations the key thermodynamic and physical/chemical properties, which determine the separation efficiency, are tabulated and described. This may include properties like operating condition limits, CO_2 solubility in solvents, chemical reaction characteristics, membrane selectivity, sorbent adsorption characteristics and technological development potential.
- Practicable lower scale limit of the separation technology: the technically feasible lower limits of scale on which the separation technologies could be used are assessed. Furthermore, the main implications for cost, energy efficiency and other emissions when pushed to lower capacity limit are described and quantified.

Coal, oil as well as gas fuelled processes are considered. Carbon extraction has not been considered. A carbon extraction process is an alternative to CO_2 capture. This kind of process recovers the hydrogen from the fuel and rejects the carbon content. The key reaction involved is the thermal decomposition of a fossil fuel. In case of methane decomposition, hydrogen and 'carbon black' are the main products and 50% of the energy remains in the extracted carbon.

4.2 Long list of processes

A long list of possible processes for CO_2 capture can be produced in a structured way using the working scheme of Table 4-1. Processes can be distinguished by the nature of the capture process (post-combustion, pre-combustion and oxy-fuel, based on separation of CO_2/N_2, CO_2/H_2 or O_2/N_2 respectively), or by the materials used to capture CO_2 (liquids, solids, membranes or cryogenic).

Table 4-1. Working scheme for long-list of capture methods

Separation technology	**Capture method**		
	Post-combustion decarbonisation (CO_2/N_2-separation)	Pre-combustion decarbonisation (CO_2/H_2-separation)	Oxy-fuel conversion (O_2/N_2-separation)
Liquid phase absorption	Chemical solvent scrubbing	Scrubbing using chemical and physical solvents	Absorbents for O_2/N_2 separation
Solid sorption	Post combustion CO_2 sorbents	CO_2 sorbents, including sorption enhanced reaction.	Adsorbents for O_2/N_2 separation Oxygen chemical looping
Membranes	Membrane gas assisted gas absorption Carbon molecular sieve membranes	CO_2/H_2 separation based on hydrogen and CO_2 separating membranes	O_2-separating membranes
Cryogenic	CO_2 separation by liquefaction	Cryogenic CO_2/H_2 separations	Cryogenic air separation

It should be noted that pre-combustion capture is essentially preceded by or integrated with a fuel processor and conversion of carbon monoxide to carbon dioxide. This fuel processor can for example be a hydrocarbon steam reformer, partial oxidiser or autothermal reformer, or a solid fuel gasifier (coal or biomass). Water-gas shift reactors often perform the CO conversion. Figure 4-1 gives an example of an IGCC with CO_2 capture.

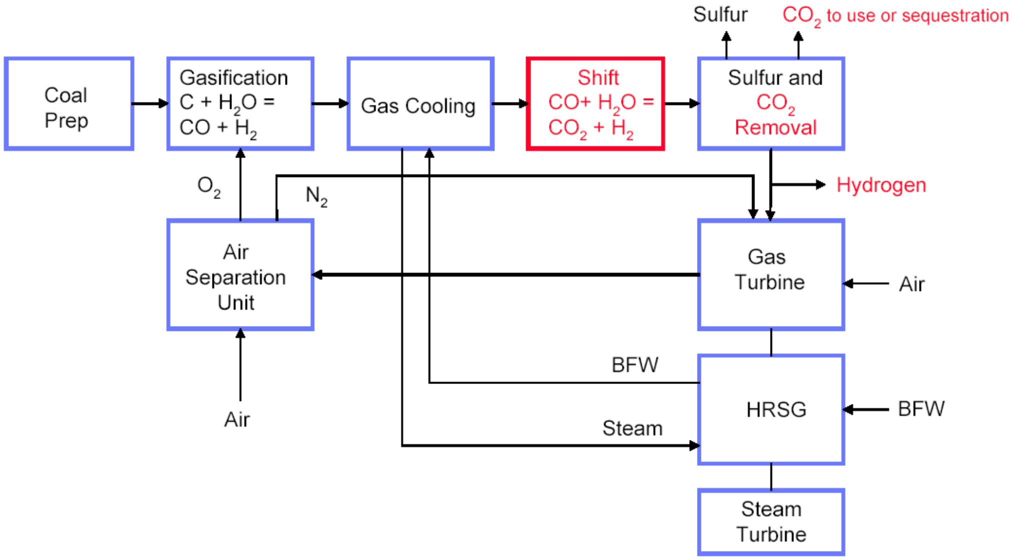

Figure 4-1. Example of an IGCC power plant with CO_2 capture

The resulting long list of processes with CO_2 capture from power generation is presented in Table 4-2. Emphasis has been put on processes, which are thought to have good prospects for medium-scale applications. For descriptions of the options see Appendix F.1 to F.19.

Other processes that were considered but that are not selected for the long list are:

- *Cryogenic separation of CO_2 from flue gases (in post combustion decarbonisation) or syngas (in pre-combustion decarbonisation).*
  Cryogenic separation of CO_2 from the flue gas would involve refrigeration of the flue gas to solidify CO_2. The solid CO_2 is subsequently liquefied.
  Cryogenic separation of CO_2 is mainly applied to gas streams with a relatively high CO_2 concentration. For flue gas from a coal-fired boiler or from a gas turbine it is economically not attractive, because of the very low CO_2 concentration in the flue gas. The total flue gas volume has to be cooled to very low temperature resulting in a high energy demand. In addition water vapour in the flue gas may interfere, because of potential blockage in heat exchange equipment due to freeze out and may need removal prior to refrigeration.
- *Pressurized PC boiler with CO_2 expansion*

Table 4-2. Long list of processes for power generation with CO_2 capture

Separation technology	**Capture method**		
	Post-combustion decarbonisation *(CO_2/N_2-separation)*	Pre-combustion decarbonisation *(CO_2/H_2-separation)*	Oxy-fuel conversion *(O_2/N_2-separation)*
Liquid phase absorption	❑ MEA absorber	❑ Selexol CO_2 absorber.	
Solid sorption	❑ Flue gas CO_2 adsorption (PSA or TSA)	❑ Sorption enhanced reforming ❑ Sorption enhanced shift ❑ In situ CO_2 separation ❑ PSA for CO_2 separation	❑ Chemical Looping Combustion (CLC)
Membranes	❑ Membrane assisted absorption	❑ Membrane reformer ❑ Shift membrane reactor	❑ OCM combustor (AZEP) ❑ SOFC (GT) with afterburner ❑ Boiler with integrated OCM ❑ Membrane oxygen production
Cryogenic			❑ Oxyfuel conversion with CO_2 recycle ❑ Oxy fuel boiler ❑ Matiant cycle ❑ Water cycle

* For abbreviations see list on page 100.

4.2.1 Characteristics of medium-scale installations

This study assesses CO_2 capture technologies, which can be applied to medium-scale sources of carbon dioxide. In general one can say that CO_2 capture from smaller energy conversion systems will have higher specific energy consumptions and higher costs both due to the economy of scale. Higher energy consumptions are the result of lower efficiencies of rotating equipment as pumps and compressors, and relatively higher heat losses.

Scale effects of compression

Small-scale compression has lower efficiency than large-scale compression, mainly because of relative longer length of the seals, and therefore higher leakage. Further research should determine at which scale compression losses become unacceptable. Lower compression efficiencies at small-scale favour the collection of captured CO_2 from several sources at low or intermediate pressure and compression centrally to higher pressures needed for transport and storage. In the order of magnitude analysis (see chapter 5) energy use for compression (0.108 kWh/kg CO_2) is taken into account assuming that the CO_2 is finally compressed to 110 bar before injection into the transfer pipeline.

Scale effects of heat loss

Small-scale thermal equipment will have higher heat losses, because the ratio between outer surface and volume is higher, which implies that the heat loss per unit of volume will be higher. To maintain temperatures at a sufficiently high level, heat must be supplied to the equipment. This leads to a lower availability of energy for heating purposes. These scale effects are neglected in the calculations of chapter 5.

Cryogenic air distillation

The main cause of efficiency loss in cryogenic equipment is heat leakage into the coldest parts, e.g. the cryogenic distillation column. There are two contributions:

- heat transport through thermal insulation
- heat transport along heat exchangers and supporting equipment.

This implies the main reasons why cryogenic equipment is preferably applied at large scale; the ratio between outer surface and the interior is the smallest, so heat transportation lengths become greater, which leads to relatively smaller heat transport. Another reason is that cryogenic distillation uses compression as the driving force to reach low temperatures. The leakage effect counts double then in small-scale cryogenic installations. More gas needs to be compressed because of the higher heat leak and the compression efficiency is lower. Because of the fast diminishing efficiency at smaller scale, it is often more efficient, both energetically and economically, to produce oxygen at large scale, and transport liquid oxygen to the customers. This may also be an option for small-scale oxyfuel capture.

4.3 Discussion of possible capture technologies

4.3.1 Possible capture technologies in relation with small furnaces/boilers

- MEA absorbers need sufficiently low levels of NO_2 and SO_X, to prevent MEA degradation. The fact that MEA absorption is widely considered as appropriate for CO_2 capture at coal fired power stations indicates that in furnaces, eventually with flue gas desulphurisation, sufficient low levels of these emissions can be realised. However, flue gas desulphurisation units for smaller systems are relative expensive.

- In principle furnaces can be equipped with all pre-combustion options. However, if the excess heat of the fuel processor can not be utilised (for example because its temperature is too low), there is a severe efficiency loss. It is expected that this often occurs. Moreover, furnace combustion of hydrogen is widely considered to be less appropriate.
- Vattenfall, one of the largest utilities of Europe, considers oxyfuel coal fired furnaces as an appropriate technology. The Vattenfall pilot oxyfuel lignite plant has 30 MW thermal output (Blackaby, 2005) and will be operational in 2008.

It can be concluded that for coal-fired furnaces oxyfuel using a cryogenic air separation unit is a suitable option. For natural gas and oil, post combustion MEA, PSA or TSA seems more appropriate, because of the relatively high ratio of the element hydrogen in the fuel, leading to relative high oxygen demand for the capture of one tonne of CO_2. This drawback may partly be circumvented when using an OCM to supply the oxygen.

4.3.2 Possible capture technologies in relation with reciprocating engines

- No technical problems are foreseen when applying post combustion options to reciprocating engine exhaust gases.
- To prevent excessive MEA degradation, concentrations of SO_X and NO_2 need to be below 1-10 ppm(v) and 20 ppm(v) respectively. Other forms of NO_X do not pose problems (Fluor, 2004). In natural gas-fired engines, sulphur levels are lower. Normal syngas desulphurisation of gasification equipment reaches also sufficiently low SO_X levels. Total NO_X emissions of reciprocating engines amount up to 180 ppm, which can be reduced by selective catalytic reduction (SCR) to 20 ppm (Pieterse, 2005). After passing the catalyst, only a small part of the NO_X is present as NO_2. Therefore it can be assumed that MEA degradation by NO_2 will not be a problem.
- To capture 85% of the CO_2 from natural gas fired engines with MEA absorption, 19% of the LHV of the natural gas is required for regeneration of the solvent, at temperatures ranging from 150°C up to 170°C, depending on the heat exchangers (Appendix F.1). It is expected that this heat is sufficiently available in the flue gas, as almost two third of all generated heat is in the flue gas. This corresponds to roughly 45% of the energy content of the natural gas. Therefore electrical efficiency loss is expected for mainly solvent pumping and CO_2 compression. In case the condenser heat of the stripper is high enough in temperature for a useful application (e.g. space heating), a minor loss in thermal efficiency is expected.
- The pre-combustion capture technologies mentioned above are all applicable to reciprocating engines. The fuel processor can be a natural gas reformer as well as a coal gasifier. However, the latter might suffer from economy of scale problems, especially concerning cleaning of the fuel gas. Small-scale natural gas reformers

are already available for fuel cell applications. It is expected that reciprocating engine operation on hydrogen does not differ much from operation on natural gas.[2] A high compression ratio can be applied, and the engines have low NO_x formation, due to the fast combustion (Schock, 1996). A few modifications must be made to valve and ignition timing (Fokker, 2005). Because the efficiency of small-scale fuel processors is not expected to exceed 80% on LHV basis, the electrical efficiency of the engine will decrease by at least 20%. If the waste heat of the reformer is still useful, the total efficiency will be less affected.

- Not many problems are foreseen for the oxyfuel option. In the past many oxyfuel applications for submarines have been developed.[3] The compression ratio of the engine might need some modifications. However, the general remark that oxyfuel is less preferred in case of use of natural gas applies here as well. The oxyfuel option seems good if applied to the combination of a coal gasifier and a reciprocating engine. The oxyfuel option is already tested in the past with a diesel engine for use in submarines. Here argon is added to the closed cycle, to avoid the need for compression ratio adjustment. CO_2 is removed from the closed cycle by scrubbing with sea water (Dutchsubmarines.com, 2005).
- For solid fuels, a gasifier with pre-combustion capture or oxyfuel combustion is the preferred option. Because the engine intake is atmospheric, and gasifiers often operate under pressure, the use of water gas shift membrane reactors delivering the hydrogen at atmospheric pressure (i.e. maximal partial pressure difference) might be the preferable pre-combustion option. Prior to shift, the syngas must be desulphurised, which can be done with zinc oxide, which is a preferred sulphur capture technology for smaller systems. In case of atmospheric gasification, oxyfuel operation of the reciprocating engine is also an option, because the shift section can be avoided then.
- For diesel oil, no reformers at a sufficient large scale are available today but partial oxidation (gasification) could be used. Because of the higher hydrogen content of the fuel compared to coal, although less than natural gas, oxyfuel combustion might be less preferable.

4.3.3 Possible capture technologies in relation with gas turbines

- In principle all capture options identified in Table 4-2 can be applied to gas turbines.
- For 85% capture of CO_2 with MEA absorption, 19% of the LHV of the natural gas is required for regeneration of the solvent, at temperatures ranging from 150 °C up to 170 °C, dependent on the heat exchangers. For coal this is 32% of the LHV. This heat can be completely derived from the flue gas. Therefore only electrical efficiency loss is expected for solvent pumping and CO_2 compression.

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[2] This is expected because BMW introduced a bi-fuel engine able to operate on both hydrogen and gasoline in their first series produced car with hydrogen fuelled reciprocating engines. This indicates that hydrogen fuelled spark ignition engines are ready for market introduction.
[3] see for instance www.dutchsubmarines.com

In case the condenser heat of the stripper is high enough in temperature for a useful application (e.g. space heating), a minor loss in thermal efficiency is expected. In case of a combined cycle, there is a loss in electrical efficiency due to the extraction of low pressure steam for the MEA regeneration, and/or a loss in thermal efficiency, depending on whether the condenser heat of the stripper is of high enough temperature for a useful application.

- From all studies published so far it can be concluded that SO_X and NO_2 levels are low enough for the application of MEA absorption.
- General Electric reports that their Frame 5 and Frame 6 gas turbines can be modified easily for hydrogen as fuel, and that the power output and efficiency increase with the hydrogen content of the fuel (Wotzak, 2005). This means that gas turbines can be made suitable for all pre-combustion capture options. It should be noted that the permeate pressure needs to be at least 5 bars above the pressure of the combustor of the gas turbine.
- In case of oxyfuel with oxygen from cryogenic distillation or membrane separation, a recycle of cooled flue gas is necessary, to prevent high turbine inlet temperatures. This flue gas has a high CO_2 concentration. The use of CO_2 instead of air as the working fluid for the gas turbine requires drastic modifications of the gas turbine. These modifications are necessary due to the difference in physical and chemical properties between air and CO_2. These modifications require massive upfront R&D investments.
- An innovative approach combining the oxyfuel principle with “gas turbines” is the water cycle developed by Clean Energy System (CES). This system will probably require less R&D investments than above mentioned modification to gas turbines.
- In case of chemical looping combustion, the gas turbine heat is generated by oxidation of metal in air. The disadvantage of oxyfuel options mentioned in the last two bullets above is therefore not present with CLC, because all fuel is used for the metal oxide reduction in a separate reactor. Thus CO_2 remains separated from the air cycle in the gas turbine. This is also the case for the OCM membrane boiler and for the AZEP (Advanced Zero Emissions Power Plant) concept.

4.3.4 Possible capture technologies in relation with fuel cells

- In principle all CO_2 capture options identified can be applied to fuel cells.
- PEMFC and PAFC are not capable of handling carbon monoxide. If they are not fuelled with pure hydrogen, they need a fuel processor. This makes them suitable for pre-combustion capture options.
- In SOFC the fuel remains completely separated from the air. Only oxygen passes through the electrolyte, which makes the SOFC in principle an oxyfuel option. The maximum fuel utilisation however is about 85%. To avoid large energy losses, this makes after-combustion necessary. If some form of oxyfuel combustion is applied here, e.g. by supplying liquid oxygen from a storage facility, or

through the use of an OCM, CO_2 can be captured by cooling with water knock-out.

- The working principle of MCFC is conduction of CO_3^{2-} ions through the electrolyte, from the oxygen side to the fuel. Therefore CO_2 containing flue gas from the afterburner has to be passed to the oxygen side. This makes pre-combustion capture difficult and oxyfuel the favourite capture option for MCFC. Then the cathode is fed with a mixture of flue gas and pure oxygen. By doing this, the concentration of O_2 as well as CO_2 is higher than in the case of air operation. The cell voltage benefits from this, and herewith the efficiency. Detailed analysis must determine how far this offsets the efficiency loss due to capture.

4.4 Summary of possible capture technologies

In the summary Table 4-3 an overview is given of possible combinations of energy conversion technologies and capture technologies. Further information on the different technologies can be found in Appendix E.1 and E.2 and Appendix F.1 to F.19.

Table 4-3. Possible combinations of capture technologies and energy conversion technologies

	Reciprocating engines	Gas turbines	Furnaces & boilers	Fuel cells
Post combustion				
1. MEA Absorber	•	•	•	•
2. Solid adsorption of CO_2 from flue gas (PSA and TSA)	•	•	•	•
3. Post combustion membrane absorber	•	•	•	•
Pre combustion, in combination with a fuel processor				
4. Selexol pre combustion CO_2 capture	•	•	•	•
5. Pre-combustion solid adsorption of CO_2 from syngas	•	•	•	•
6. Sorption enhanced reforming	•	•	•	•
7. Sorption enhanced shift (SE-WGS)	•	•	•	•
8. Reforming with CaO		•	•	•
9. Membrane reformer	•	•	•	•
10. Shift membrane reactor	•	•	•	•
Oxyfuel combustion				
11. Oxyfuel conversion with CO_2 recycle	•	•	•	•
12. Matiant Cycle		•		•
13. Water Cycle		•		•
14. Chemical Looping		•	•	•
15. Membrane oxygen production	•	•	•	•
16. GT/CC with oxygen conducting membrane (AZEP)		•		•
17. Oxygen Conducting Membrane boiler		•	•	•

5 Selection and evaluation of typical plants potentially suitable for CO_2 capture

5.1 Introduction

An inventory of combustion installations was prepared in chapter 3, whereas potential capture technologies were identified in chapter 4. The combination of prime movers with various fuels covers a range of capture technologies, applications (heat, power or combined) and plant sizes in order to form a ‘long list’ of cases. Five cases are selected for in-depth evaluation (short list), in close cooperation with the IEA GHG R&D programme. These cases cover a broad range of options, whilst complying with reasonable economic prospects.

The evaluation is performed with respect to the following aspects:

- Brief assessment of the main advantages and disadvantages of each case.
- The potential for application of the cases (i.e. amount of installations in operation and CO_2 mitigation potential).
- Indication of the energy and cost performance as well as an indication of cost reduction potential by large-scale implementation of the cases (Order of Magnitude Analysis, OMA).

5.2 Long list of cases

Based on the conclusions in chapter 4 on the application of capture technologies to medium-scale combustion installations, a list has been prepared that comprises combinations of selected prime movers and possible fuels. The CO_2 capture options that are expected to be commercially viable in the near future were identified for each of the combinations of prime mover and fuel type. These options are listed in Table 5-1.

Table 5-1. Long list of prime movers and fuels with potential CO_2 capture options

	Natural gas	**Oil**	**Solid Fuels**
Reciprocating engines	Post-C membrane MEA absorption, Pre-C membrane options	Post-C membrane MEA absorption	Pre-C shift membrane reactors, oxyfuel with CO_2 recycle
Gas Turbines	Pre- and Post-C options, CLC	Pre- and Post-C options, CLC	Pre-combustion SE-WGS, membrane shift
Furnaces, ovens and boilers	Post-C (membrane) MEA, PSA, TSA	Post-C (membrane) MEA, PSA, TSA	Oxyfuel
PEMFC, PAFC	Membrane reforming, SERP	Pre-C SE-WGS, membrane shift	Pre-C SE-WGS, membrane shift
MCFC	Post-C MEA, PSA, TSA	Post-C MEA, PSA, TSA	Oxyfuel
SOFC	OCM oxyfuel	OCM oxyfuel	OCM oxyfuel

5.3 Short list of cases

A selection of the most viable cases was prepared in order to reduce the number of cases in the long list. The emphasis of this short list is on the application of natural gas as fuel. The selection has been made in close cooperation with the IEA GHG R&D Programme and is based on following considerations:

- Worldwide total installed capacity and number of installations
- Gas/diesel motors are used a lot in CHP mode in Europe, United States etc.
- Gas turbines are one of the prime movers for small- to medium-scale CHP
- SOFC is one of the potential future competitors for gas engines and small gas turbines in CHP applications. However, only when investment costs of SOFC systems are competitive. This means that costs should be below €2000 /kWe
- Natural gas boilers for heat and CHP are selected on the basis of installed capacity
- Small to medium size coal boilers have a significant market share that will increase in the near future.

The five selected combinations of prime mover, fuel and capture method for the short list are:

1. Reciprocating engine with membrane assisted liquid absorption
2. Natural gas turbine with pre-combustion PSA capture
3. Oxyfuel coal boiler with oxygen conducting membranes

4. Oxyfuel natural gas boiler with oxygen conducting membranes
5. Solid oxide fuel cell with oxygen conducting membrane afterburner

These combinations will be discussed and evaluated in the next paragraphs, which results in the selection of one case for detailed assessment. The quoted costs numbers are total installed costs, including site preparation, fuel supply system, owner's costs, and contingency.

5.3.1 Reciprocating engine with membrane assisted liquid absorption

Description of the option

This option is selected, because reciprocating engines (gas and diesel fuelled) are one of the designated prime movers in the segment below 10 MW_e. The other frequently used prime mover in this market segment below 10 MW_e is the gas turbine. A detailed description of reciprocating engines can be obtained in section 2.4.

The concentration of CO_2 in the exhaust gas of reciprocating engines is relatively high compared to gas turbines, and therefore beneficial for post-combustion capture. This is induced by the reduced air excess rate, for example by application of stoichiometric engines with catalytic NO_x conversion.

The membrane assisted liquid absorption option is selected from the three possible capture technologies identified in the long list. This option is expected to be more economically attractive at small scale compared with conventional MEA absorbers, since the specific investments for MEA absorbers are relatively high for small units (see Appendix F.1). On top of that, membrane systems are very compact, which is a significant advantage for such systems.

Potential for application

Reciprocating engines are applied in various market sectors, such as commercial buildings, light industry, utilities and the agricultural sector. Reciprocating engines are applied to serve a wide variety of purposes, e.g. base load power generation, utility support as well as demand responding peak- and standby-power. Furthermore, reciprocating engines are applied within Combined Heat and Power (CHP) installations.

The installed capacity of stationary reciprocating engines ranges from several kilowatts up to 10 MW, where the global market displays a tendency towards procurement of larger units. The annual global market comprises procurement of 25 GW_e, where the installed capacity between 1 and 2 MW_e amounts to 45% of the market. The segment between 2 and 20 MW_e comprises 25% of the market. The majority of the units, as well as the units with the largest capacity are commissioned in the Far East, including China. The worldwide market displays a steady growth and doubled during the past ten years. Approximately 85% of the engines are fired by diesel fuel, whilst the remainder is natural gas-fired. Less than one percent of the installed fossil-fuelled power capacity in the United States utilises engines as prime mover. However, the share in capacity in the segment up to 40 MW_e is 18%, and moreover the number of machines covers 70% of all prime movers for power generation. These numbers confirm that the majority of the installed diesel-fired gas engines are applied within Uninterruptible Power Supply (UPS) systems.

Based on the global market sales statistics we made an estimate of the total emissions of carbon dioxide from reciprocating engines in the capacity class of 1 to 20 MW_e. This is estimated at about 600 to 700 Mt, which equals about 6% of total carbon dioxide emissions related to electricity and heat production (world wide approximately 11 Gt). Figure 5-1 shows a breakdown of capacity classes. The main share of the reciprocating engine related emissions is in the category of 1 to 2 MW_e.

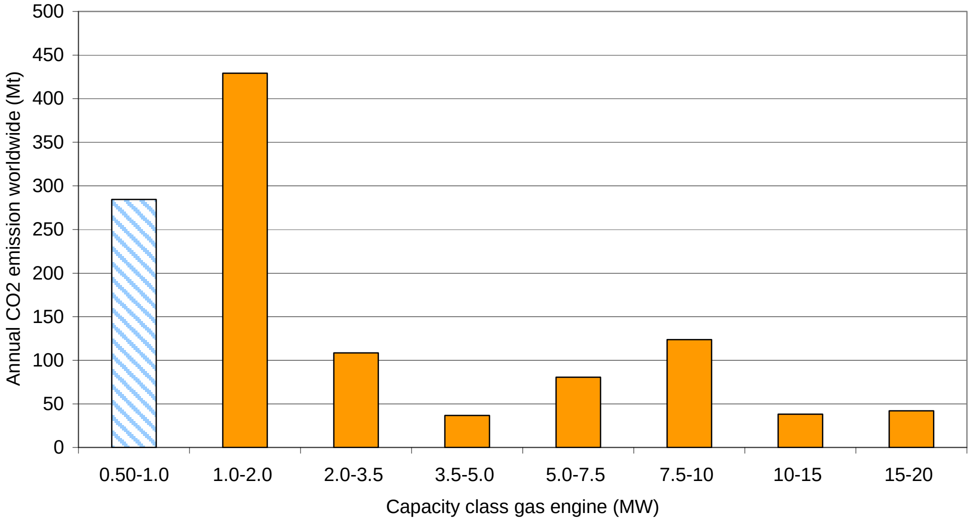

Figure 5-1. Estimated emissions from gas engines worldwide. Category 0.5-1.0 (bar in shaded blue) falls outside the scope of this study

Indication of performance and cost reduction potential

Reciprocating engines with an electrical power output of 1.5 MW_e are selected as a characteristic case for small-scale power generation. The capture costs are calculated for both diesel and natural gas-fired engines. The specifications for the diesel and natural gas-fired engines without CO_2 capture are obtained from (ASUE, 2005) and are summarised in Table 5-2.

Table 5-2. Specifications of 1.5 MW_e reciprocating engines (ASUE, 2005)

	Diesel	**Natural Gas**
Fuel input	3.3 MW	3.9 MW
Electrical power output	1.5 MW_e	1.5 MW_e
Thermal output	1.4 MW_{th}	2.0 MW_{th}
Electrical efficiency LHV	46 %	39 %
Thermal efficiency	44 %	52 %
Operation time, full load	4000 h/y	4000 h/y
Total efficiency	90 %	91 %
CO_2 emission	0.24 kg/s	0.22 kg/s
CO_2 emission	3.5 kt/y	3.2 kt/y
Genset costs	1.4 M€	0.6 M€
O&M costs	0.008 €/kWh_e	0.008 €/kWh_e
Overhaul costs (every 25.000 h)	0.003 €/kWh_e	0.003 €/kWh_e

Most of the data for diesel-fired CHP installations are available in literature. However, the operation and maintenance costs, as well as the overhaul costs (e.g. replacement of cylinders, pistons, bearings and valves) are derived from data for natural gas-fired installations and are presumed to be representative.

The following assumptions were made for the assessment of similar engines equipped with membrane assisted MEA sorption:

1. Carbon capture ratio of 90%.
2. Heat of sufficient temperature is available from the exhaust gas for MEA regeneration (being 0.62 MW_{th} (NG) and 0.70 MW_{th} (diesel) for 90% capture).
3. 80% of total amount of heat needed for the MEA regeneration still can be used for the original purpose i.e. heating.
4. Electrical output is reduced by power consumption to overcome pressure drop in the system (0.02 kWh/kg) and CO_2 compression to 11 MPa (0.108 kWh/kg)[4].

The results of the assessment are summarised in Table 5-3.

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[4] The compression energy is assumed to be 390 kJe/kg CO_2 (Wildenborg, 1999; Hendriks, 2001). This value is relatively high, however it is considered reasonable in case CO_2 is distributed from small capture units at low pressures and compressed at nodes in a CO_2 distribution network.

Table 5-3. Specifications of 1.5 MW_e combustion engines with post-combustion CO_2 capture

	Diesel	**Natural Gas**
Fuel input	3.3 MW	3.9 MW
Capture	90 %	90 %
Electrical power output	1.4 MW_e	1.4 MW_e
Thermal power output	1.3 MW_{th}	1.9 MW_{th}
Electrical efficiency LHV	43 %	37 %
Thermal efficiency	40 %	49 %
Total efficiency	83 %	85 %
Operation time, full load	4000 h/y	4000 h/y
CO_2 emission	0.024 kg/s	0.022 kg/s
CO_2 emission	0.35 kt/y	0.32 kt/y
Additional capture investments[5]	2.1 M€	1.9 M€
Additional O&M costs	0.015 €/kWh_e	0.014 €/kWh_e

The additional capture investments are assessed by combination of literature values for the investment costs of large-scale membrane contactors including CO_2 compression (Grönvold, 2005), with a scale exponent amounting to 0.8 regarding conventional MEA scrubbers obtained from (Hamelinck, 2001; Hendriks, 1994). The relatively large scale exponent is expected to suit application due to the enhanced modularity of membrane equipment (Grönvold, 2005).

The CO_2 compressor investment costs are estimated using two sources (Kreutz, 2005) and (Lokurlu, 2004). The first reference states that a 13 MW_e CO_2 compressor costs approximately 14.8 M$ (2004). From the second reference, it is derived that a 0.7 MW_e CO_2 compressor costs approximately 1.8 M€. Application of both literature sources results in a scale exponent of 0.65, which corresponds with the scale exponent 0.67 reported by (Kreutz, 2005).

5.3.2 Natural gas turbine with pre-combustion PSA capture

Description of the option

This option is selected since natural gas-fired gas turbines are together with reciprocating engines the most applied prime mover in the segment of 1 MW_e and larger. A detailed description of gas turbines can be obtained from Paragraph 2.5.

The concentration of CO_2 in the exhaust gas of gas turbines is relatively low, due to the high air excess ratio. This large amount of excess air is applied to comply with construction material constraints, by reduction of the entrance temperature at the first turbine stage. The elevated pressure at which fuel is combusted in gas turbines is advantageous

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[5] The compression costs may be estimated somewhat conservative, because the possibility of low pressure collection is neglected here.

for pre-combustion PSA capture. Carbon dioxide can be adsorbed at combustion pressure, and be released at lower pressures, e.g. at ambient pressure.

Potential for application

Compared to reciprocating engines, gas turbines are more expensive and are seldom applied for standby (only in the 1-2 MW_e classes) or demand response peak purposes. Gas turbines are frequently applied for peak shaving and base load purposes, also in CHP configuration. The main application lies within industry, the large commercial sector and the power sector. Gas turbines comprise 9% of the installed fossil fuel based power capacity in the United States. The share of gas turbine based power generation increases significantly, when the output of steam turbines within combined cycle plants is also taken into account. In the category up to 40 MW_e (about 5% of total capacity) the share of gas turbines amounts to almost 50% (2.5% of total capacity installed). Independent power production spurs the procurement of distributed power generation units with a capacity smaller than 50 MW_e. The global annual market consists of approximately 600 gas turbines (1 MW_e and larger), where approximately 500 units are smaller than 40 MW_e. The gas turbine market is stable at present; however increasing gas prices may decrease annual sales. The majority of the units are sold in the Far East, including China. The annual global market comprises procurement of 35 GW_e (reference year 2004), where the installed capacity up to 40 MW_e amounts to 15% of the market.

Based on the global market sales statistics we made an estimate of the total emissions of carbon dioxide from gas turbines in the capacity class of 1 to 60 MW_e. This is estimated at about 250 to 300 Mt, which equals to about 2.5% of total carbon dioxide emissions related to electricity and heat production (world wide approximately 11 Gt). Figure 5-2 shows a breakdown according to capacity classes. The major share of the gas turbine related emissions is in the category above 120 MW_e.

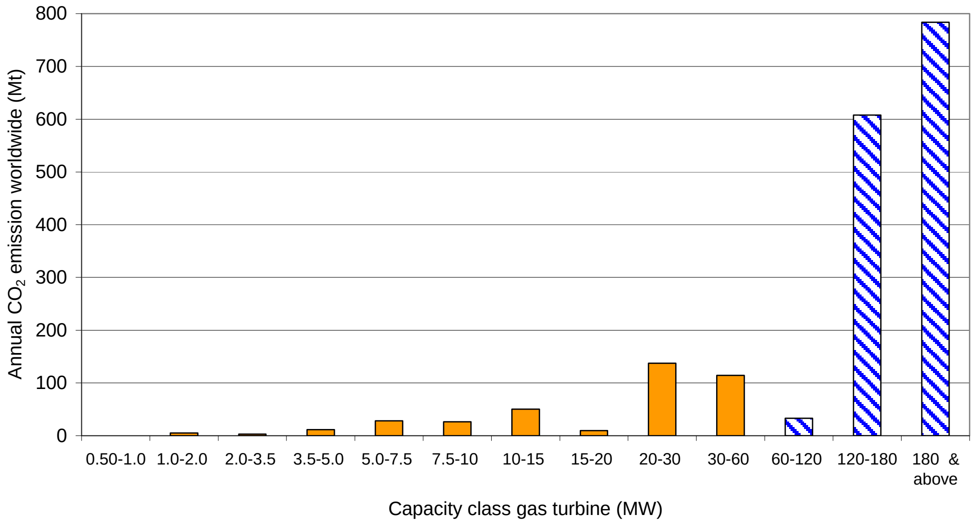

Figure 5-2. Emissions from gas turbines world wide. Categories 60 to 120 MW, 120-180 MW and 180 MW and above (bars in shaded blue) fall outside the scope of this study

Indication of performance and cost reduction potential

A gas turbine with an electrical power output of 5 MW_e is selected as a characteristic case for medium-scale power generation. The Siemens SGT 100 gas turbine genset matches the output. The specifications for this gas turbine are summarised in Table 5-4.

Table 5-4. Specification of Siemens SGT 100 gas turbine (Gas Turbine World, 2005)

Fuel input	16.7 MW
Electrical power output	5.05 MW_e
Thermal output (100 °C flue gas)	9.71 MW_{th}
Electrical efficiency LHV	30.2 %
Thermal efficiency	58.1 %
Total efficiency	88.3 %
Operation time, full load	6000 h/y
CO_2 emission	0.94 kg/s
CO_2 emission	20.3 kt/y
Genset costs, including heat recovery boiler	2.8 M€
Gas turbine O&M	4 %/y
Pressure ratio	14.8
Air flow	21 kg/s
Exhaust temperature	546 °C

The thermal efficiency is estimated using the specified outlet temperature of the gas turbine and an estimated flue gas exit temperature of 100 °C. The genset costs are estimated using the trend line found in Chapter 3, while an additional specific investment of 100 €/kW$_{th}$ is assumed for the heat recovery boiler.

The following assumptions were made during the assessment of a similar turbine within a CHP system, equipped with pre-combustion carbon capture through pressure swing adsorption:

1. Carbon capture ratio of 85%.
2. Natural gas is converted in an auto thermal reformer (ATR), shifted in two sequential water-gas shift reactors, and subsequently CO_2 is captured in a Pressure Swing Adsorber (PSA).
3. Heat of combustion is transferred in the ATR from natural gas to hydrogen with an efficiency of 75% (on LHV basis).
4. The syngas cooler provides an additional efficiency of 15% of the heat of combustion of natural gas (on LHV basis).
5. CO_2 is captured at the operation pressure of the shift section. This pressure is 2 to 5 bars above the gas turbine combustor pressure. We assumed that natural gas is available at pressures well above the minimum pressure needed. Pipeline pressure normally is at least 25 bars.

The results of the assessment are summarised in Table 5-5.

Table 5-5. Specifications of Siemens SGT 100 gas turbine with pre-combustion CO_2 capture

Fuel input	22.3 MW
Capture	85%
H_2 generating efficiency	75%
Useful heat from ATR	15%
Electrical power output	4.6 MW$_e$
Thermal power output	13.1 MW$_{th}$
Electrical efficiency LHV	20%
Thermal efficiency (100 °C flue gas)	59%
Total efficiency	79%
Operation time, full load	6000 h/y
CO_2 emission	0.19 kg/s
CO_2 emission	4.1 kt/y
Additional capture investment	5.7 M€
Additional capture O&M	2%/y

The results display a reduced electrical efficiency from 30.2% to 20.0%. The thermal efficiency remains constant and the thermal output increases.

Estimates of the investment for capture equipment (including compressor) are based on investments for natural gas combined cycles (NGCC) which are approximately equal to the investment for the capture part of the system. If the investment costs for the steam cycle (about 25% of an NGCC) are subtracted from the NGCC investment, the relative capture investment amounts to about 133% of the investment of a gas turbine. Furthermore, the efficiency of a small gas turbine is about 33% lower than for a large turbine (30% compared to 40%) and the efficiency of a small-scale auto-thermal reactor (ATR) is expected to be somewhat lower than the larger variant. All these effects are estimated to add up to a capture investment, which resulted in an investment twice as high as for a gas turbine and heat recovery boiler together. The investment for the additional capture equipment is therefore estimated at 5.7 M€. Because there are only few moving parts, the O&M costs are assumed at 2% of the capture investment (IEA GHG, 2000).

5.3.3 Oxyfuel coal boiler with oxygen conducting membranes

Description of the option

Natural gas-fired boilers are more abundant than coal-fired boilers, however an increase in implementation of coal-fired boilers is foreseen, particularly when CO_2 can be captured. A coal-fired boiler with Oxygen Conducting Membranes (OCM) is selected in order to maintain diversity amongst the fuels used in the cases in this study, combined with carbon capture through oxyfuel combustion. The selected boiler type is water tube. Live steam pressure and temperatures of the boiler are rather low compared to the steam pressures and temperature in modern pulverised coal power generation boilers.

Integration of both burner and OCM is undesirable in this case, due to the nature of coal combustion. OCM appears to be particularly vulnerable for sulphurous components as well as fouling by ash. Air will be compressed and preheated upstream of the OCM, expansion of the lean air downstream of the OCM provides the required work for the air compressor. It could also be used to drive the CO_2 compressor. Oxygen permeated through the OCM is guided to the coal-fired burner where flue gas recycling is applied to reduce the temperature in the water tube boiler where steam is produced. The principle is depicted in Figure 5-3; additional information concerning this concept can be obtained from Appendices F.9 and F.19.

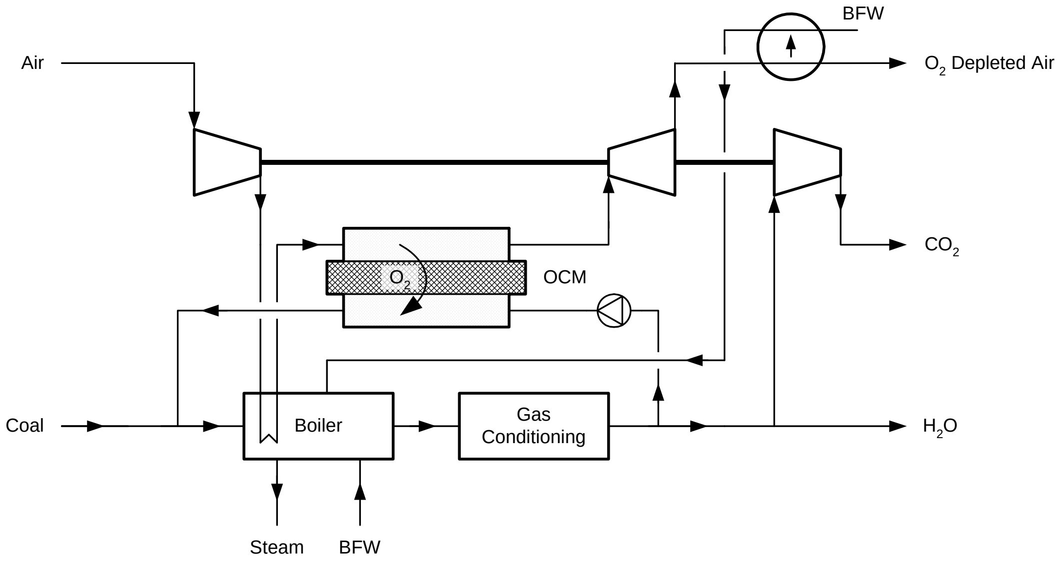

Figure 5-3. Oxyfuel coal-fired boiler with oxygen conducting membranes (OCM)

Potential for application

Boilers are widely applied in the power, manufacturing, commercial and agricultural sector. The majority of the boiler capacity is installed for electricity generation by expansion of steam in a steam turbine. Nevertheless, a substantial part is employed to generate steam for industrial processes. The dominant fuel for boilers in the power sector in the United States is coal (over 50%). In units connected to steam turbines that are smaller than 40 MW_e, coal use amounts to approximately 33%. Natural gas is the dominant fuel for industrial boilers, while coal is only applied in approximately 15% of the boilers. Coal is hardly applied in boilers that are operated within the commercial sector.

Emissions from boilers are substantially larger than from gas engines and gas turbines in the same capacity class, i.e. up to about 100 MW of fuel input. Emissions from industrial boilers and commercial boilers in the United States are estimated at about 500 Mt and 110 Mt respectively. Coal-related emissions of both sectors together amount to about 100 Mt. Emissions from oil and ‘other’ fuels (except natural gas) amount to 300 Mt. Together they make up approximately 16% of the emissions in the United States and about 4% of the emissions worldwide related to electricity and heat. ‘Other fuels’ include all kinds of waste gases (e.g. refinery gases or blast furnace gases) and waste fuel like black liquor. A part of this fuel can also be decarbonised in the same way as described here for the coal-fired boilers. Figure 5-4 shows an emission breakdown for the United States by fuel type. Emissions from industrial and commercial coal-fired boilers in China amount to about 1200 Mt. This equals 45% of all emissions from electricity, heat, manufacturing and construction in that country.

For the calculation of worldwide emissions from boilers fired with coal, oil or ‘other fuels’ the United States serves as a typical example for the rest of the world except China.

Seven percent of the CO_2 emissions in the United States stems from boilers fired with coal, oil and 'other fuel'. If this is typical for the rest of the World (except China), about 1500 Mt of CO_2 is emitted worldwide (except China) by boilers with this type of fuel. If we add the emissions from China of 1200 Mt to this number, the total emissions from boilers amount to about 2700 Mt (about 11% of worldwide CO_2 emissions).

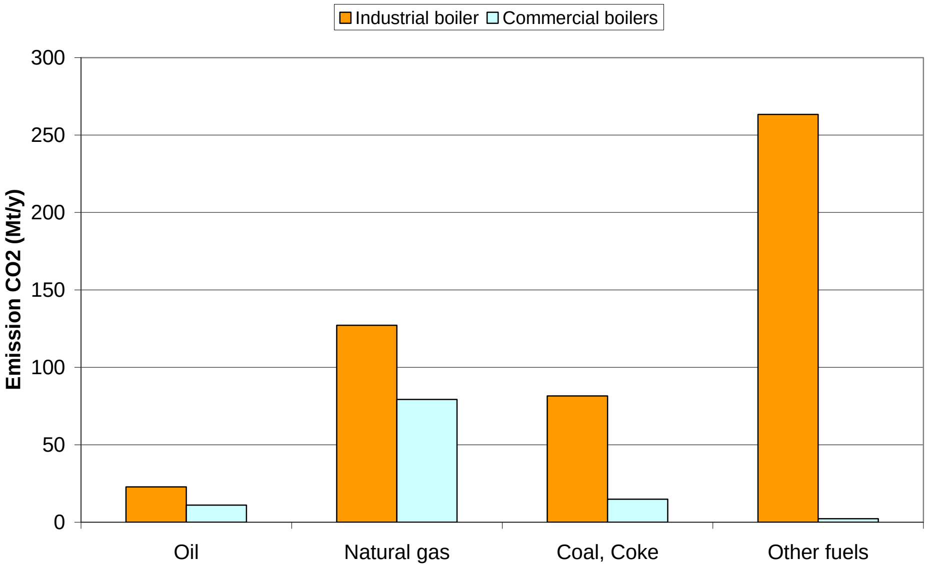

Figure 5-4. Annual CO_2 emissions from boilers in the U.S (derived from (ORNL, 2005))

Indication of performance and cost reduction potential

A coal-fired steam boiler with a thermal output of 50 MW_{th} is selected as a characteristic case for medium-scale steam production. The specifications for the steam boiler without CO_2 capture are summarised in Table 5-6.

Table 5-6 Specifications of a 50 MW coal-fired steam boiler

Fuel input	58.8 MW
Thermal output (steam)	50 MW_{th}
Thermal efficiency	85%
Operation time full load	8000 h/y
CO_2 emission	5.65 kg/s
CO_2 emission	163 kt/y
Boiler specific costs	250 €/kW_{th}
Boiler O&M	4%/y

The following assumptions were made during the assessment of a similar coal-fired boiler with oxyfuel firing through oxygen separation in an OCM.

1. Carbon capture ratio of 95%.
2. A stoichiometric (air) factor of 101% is applied to obtain full conversion while minimising the oxygen content of the mitigated CO_2 stream. For a conventional boiler this is equivalent to an excess air factor of 105%.
3. The oxygen permeation ratio through the OCM is assumed at 50%, in case of a permeate pressure of 18 bar and operation at a temperature of 800°C.

The results are summarised in Table 5-7.

Table 5-7. Specifications of a 50 MW coal-fired steam boiler with oxyfuel CO_2 capture

Fuel input		65 MW
Capture ratio		95%
Thermal output (steam)		50 MW_{th}
Thermal efficiency		77.3%
Operation time full load		8000 h/y
Net CO_2 emission		0.37 kg/s
Net CO_2 emission		10.7 kt/year
Additional capture costs:		
	Oxygen conducting membrane	6.8 M€
	Heat exchangers	6.2 M€
	Compressor & turbine	5.0 M€
	CO_2 compressor	3.9 M€
Capture O&M		1.3 M€/y

The thermal output is maintained at 50 MW_{th}, which results in an increase in boiler size (from 58.8 to 65 MW) to overcome the efficiency penalty induced by carbon capture. The capital expenditure for the OCM is based on specific costs that amount to 1,500 $/m^2 (Yantovski, 2004; Allam, 2004). The oxygen permeation rate is assumed at 1 g/m^{2}s. The capital expenditure for the air compressor and turbine are determined with data for mechanical drive turbines as mentioned in the Gas Turbine World Handbook (2005).

The CO_2 compressor investment costs are estimated using two literature sources (Kreutz, 2005) and (Lokurlu, 2004). The first reference states that a 13 MW_e CO_2 compressor costs approximately 14.8 M$ (2004). From the second reference, it is derived that a 0.7 MW_e CO_2 compressor costs approximately 1.8 M€. Application of both literature sources results in a scale exponent of 0.65, which corresponds with the scale exponent 0.67 reported by (Kreutz, 2005).

The CO_2 compressor power for the coal-fired boiler is estimated at 2.2 MW_e, which leads to an investment of 3.8 M€. The annual O&M costs are estimated at 4% of the investment costs.

5.3.4 Oxyfuel natural gas boiler with oxygen conducting membranes

Description

Natural gas-fired boilers are predominantly applied for central heating purposes, which implies that the water outlet temperature is relatively low (typically 100 °C or less).
In accordance with the coal-fired boiler case, oxyfuel combustion with oxygen conducting membranes is selected as a CO_2 capture option. Due to the absence of fouling or wear by ashes, natural gas can be fired in direct contact with the membranes. Therefore the OCM is integrated with the natural gas-fired boiler. The working principle is depicted in Figure 5-5.

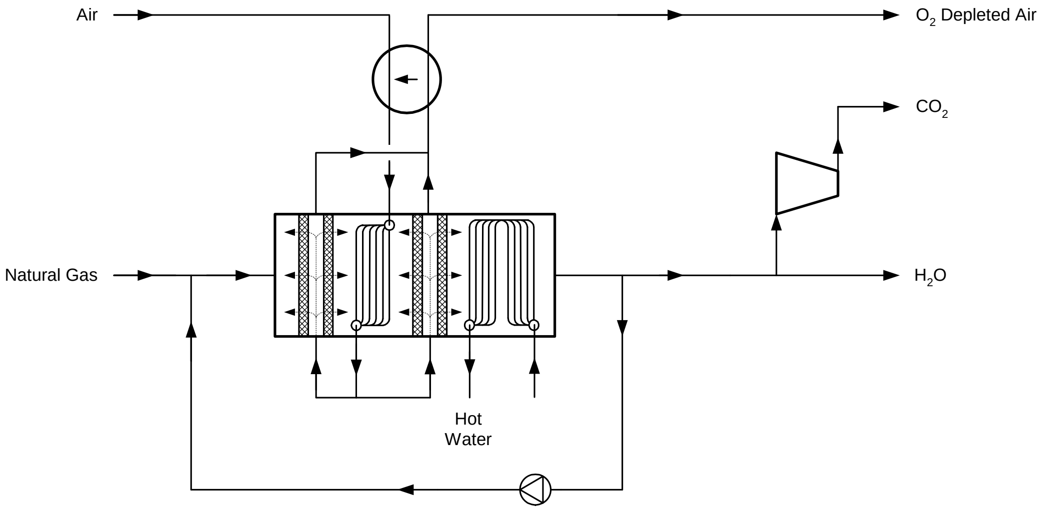

Figure 5-5. Oxyfuel natural gas-fired boiler with oxygen conducting membranes (OCM)

Potential for application

Paragraph 3.3.1 displays that the dominant fuel for small-scale boilers is natural gas. In the United States alone, over 50% of the industrial boilers and over 85% of the commercial boilers are fired with natural gas. Moreover, approximately 45% of the boilers in the power sector (connected to steam turbines smaller than 40 MW_e) are fired with natural gas.
Carbon dioxide emissions from industrial boilers and commercial boilers in the United States are estimated at about 500 Mt and 110 Mt respectively. Natural gas-related emissions amount to about 200 Mt, approximately 6% of all emissions in the United States related to electricity, heat, manufacturing and construction. It should be noted that ‘other fuels’ have a high share in total fuel use in boilers. ‘Other fuels’ includes all kinds of waste gases (e.g. refinery gases or blast furnace gases) and waste fuels like black liquor. A part of this fuel can also be decarbonised in the same way as described here for the natural-fired boilers.

The United States contributes 3.2 $GtCO_2$ to about 22% of the world wide carbon dioxide emissions of 15 $GtCO_2$ related to electricity, heat, manufacturing and construction. Assuming the same emission division for the rest of the world, industrial and commercial natural gas-fired boilers worldwide emit about 800 - 1000 Mt each year. This equals about 10% of total emissions from heat and power production. This percentage might be an overestimate as it may be expected that less natural gas is used on average in other countries than is used in the United States

Indication of performance and cost reduction potential

A natural gas-fired boiler with a thermal output of 5 MW_{th} is selected as a characteristic case for hot water production. The specifications of this boiler are summarised in Table 5-8.

Table 5-8. Specifications of a 5 MW_{th} natural gas-fired central heating boiler

Fuel input	5.4 MW
Thermal output (hot water)	5.0 MW_{th}
Thermal efficiency	92.5%
Operation time full load	2000 h/y
CO_2 emission	0.3 kg/s
CO_2 emission	2.18 kt/y
Specific costs boiler	100 €/kW_{th}
Boiler O&M	4%/y

The following assumptions were made during the assessment of a similar natural gas-fired boiler equipped with carbon capture through oxyfuel combustion with oxygen conducting membranes:

1. Capture ratio of 95%.
2. A stoichiometric air factor of 101% is applied to obtain full conversion while minimising the oxygen content of the mitigated CO_2 stream.
3. The oxygen permeation rate through the OCM is 75% at 800 °C. This is higher than in the coal-fired boiler case due to negligible oxygen partial pressure at the combustion side.
4. Air compression and expansion can be abandoned, since the oxygen permeation is already at an acceptable level under atmospheric pressure, i.e. the oxygen partial pressure on the permeate side is 3 to 4 bars, whereas the partial pressure at the retentate side is 0.2 bars.

The results of the assessment are summarised in Table 5-9.

Table 5-9. Specifications of a 5 MW_{th} natural gas-fired boiler with oxyfuel CO_2 capture

Fuel input	5.6 MW
Capture ratio	95%
Thermal output (hot water)	5.0 MW_{th}
Thermal efficiency	89.6%
Efficiency penalty	2.9%
Operation time full load	2000 h/y
Net CO_2 emission	0.034 kg/s
Net CO_2 emission	0.25 kt/y
Additional costs	
Oxygen conducting membrane	0.58 M€
Heat exchangers	0.13 M€
CO_2 compressor	0.57 M€
Capture O&M	0.07 M€/y

The thermal output is maintained at 5 MW_{th}, which results in an increase in boiler size to overcome the efficiency penalty induced by carbon capture. Furthermore, an additional electricity demand of 140 kW_e is introduced for CO_2 compression. The additional capture investments and O&M costs match the estimations made for the coal-fired boilers in Paragraph 5.3.3.

5.3.5 Solid oxide fuel cell with oxygen conducting membrane afterburner

Description
A Solid Oxide Fuel Cell (SOFC) with an electrical output of 500 kWe is selected as a characteristic case for a small-scale CHP installation that incorporates a high-temperature fuel cell. Application of an SOFC is considered an oxyfuel option in correspondence with chapter 4.3.4 and Appendix F.17., since it incorporates a non-electron conductive oxygen conducting membrane as electrolyte. Downstream of the SOFC 15% of the fuel remains unconverted; therefore the application of an oxygen conducting membrane as afterburner is desirable.

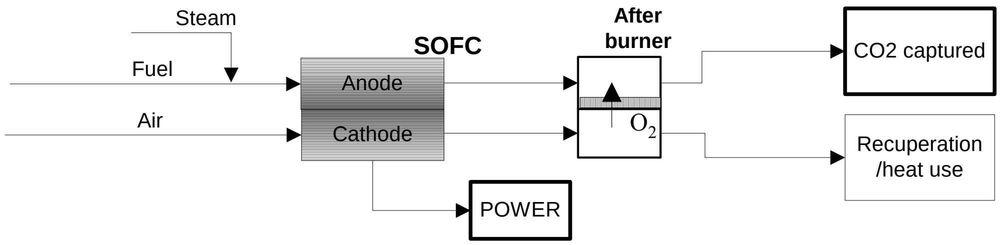

Figure 5-6. SOFC with OCM afterburner and CO_2 capture

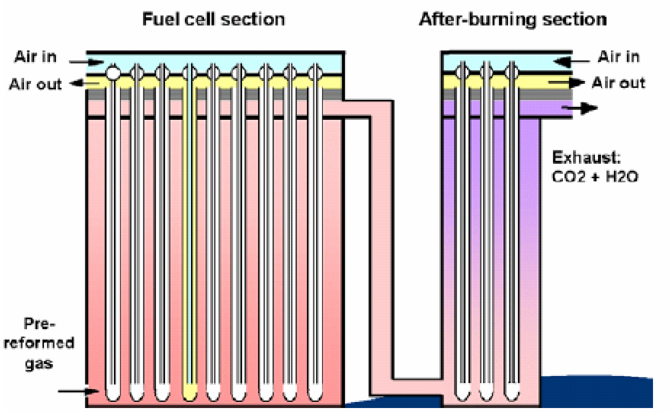

Figure 5-7. SOFC with OCM afterburner (Veyo et al., 2002)

Potential for application

Operating temperatures for SOFCs are among the highest of the available fuel cell technologies (up to 1,000°C). The available waste heat makes SOFCs suitable for combined heat and power (CHP) applications. The ratio of power plus heat to the lower heating value of the fuel can range up to 70 to 75% within the SOFC only. Combined with the afterburner, total energy conversion efficiency can go up to 85% or higher. The second major benefit of the high operating temperature is reduced reforming process requirements, because the high temperatures required for the reforming process are already available within the system.

The SOFC can be applied for grid-connected power systems as well as for the (commercial) building sector. At present, development proceeds relatively slowly and costs continue to be considerable. The broad implementation of SOFC in the midterm future is unknown, since the present market is virtually non-existent. Future driving forces for implementation can be high energy prices and an increased tendency to distributed energy supply. However, the high specific costs of SOFC that range from 10,000 to 20,000 €/kW presently, should be reduced to 1,000 €/kW or less to ensure sufficient market penetration. It is not expected that the market share of SOFC and therefore the contribution of this technology to the reduction of CO_2 emissions will be in the order of mega tonnes before the year 2020-2030.

Indication of performance and cost reduction potential

An SOFC with an electrical power output of 500 kW_e is selected as a characteristic case for small-scale CHP. The specifications for the SOFC without CO_2 capture are summarised in Table 5-10.

Table 5-10. Specifications of a SOFC system

Fuel input	1000 kW
Electrical power output	500 kW_e
Thermal output	350 kW_{th}
Electrical efficiency LHV	50%
Thermal efficiency LHV	35%
Total efficiency	85%
Operation time full load	6000 h/y
CO_2 emission	0.056 kg/s
CO_2 emission	1.2 kt/y
Genset costs	1000 k€
O&M costs	4%/y
Operating pressure	1 bar
Fuel utilisation	85%
Air factor (λ)	5

The following assumptions were made during the assessment of a similar SOFC system equipped with CO_2 capture:

1. The fuel conversion rate of the fuel cell amounts to 85%, the remaining fuel is oxidised in the OCM afterburner.
2. A stoichiometric air factor of 101% is applied to obtain full conversion while minimising the oxygen content of the mitigated CO_2 stream.
3. Cathode feed air is applied to feed the OCM module, since the high temperature suits permeation in the OCM.
4. The oxygen permeation rate through the OCM is 75% at 800 °C at atmospheric pressure, due to negligible oxygen partial pressure at the combustion side.
5. Air compression and expansion for the OCM can be abandoned, since the oxygen permeation is already at an acceptable level under atmospheric pressure.
6. The electrical efficiency of the SOFC system is not affected by the addition of the OCM afterburner.
7. Costs for SOFC systems are reduced to 2000 €/kWe. This is a *target cost* that is needed to enable SOFCs to enter the combined heat and power market. Yet it is not clear whether this target will be achieved.

The thermal efficiency of the SOFC system with OCM afterburner is higher then the efficiency of an independent SOFC. The unconverted fuel is combusted with the entire cathode exhaust air stream in the stand-alone SOFC configuration, which results in a high dilution of the flue gas with water vapour. The dew point of the flue gas is low enough to avoid the occurrence of condensation. Application of an oxyfuel afterburner prevents flue gas dilution and allows the ability to condense water vapour from the flue gas, which results in a higher thermal efficiency. However, the gain in thermal efficiency is assumed to be smaller than the uncertainties in the estimates, and is therefore not accounted for. An additional heat exchanger to quench the anode outlet stream is required, as a consequence of the separation of the streams. The specifications for the SOFC without CO_2 capture are summarised in Table 5-11.

Table 5-11. Specifications of a SOFC system with OCM afterburner and CO_2 capture

Fuel input	1000 kW
Capture ratio	100%
Electrical power output	470 kW_e
Thermal output	350 kW_{th}
Electrical efficiency LHV	47%
Thermal efficiency LHV	35%
Total efficiency	82%
Operation time full load	6000 h/y
CO_2 emission	-
CO_2 permeation	1 $g/m^2.s$
OCM surface	6.2 m^2
Additional capture costs	
Oxygen conducting membrane	8 k€
Heat exchanger	8 k€
CO_2 compressor	190 k€
Capture O&M	8 k€/y

It can be observed that CO_2 compression on a small scale is relatively expensive, whereas oxygen conducting membranes are inexpensive. The required OCM surface is small due to the high fuel conversion in the SOFC; the high oxygen permeation also works beneficially.

5.4 Costs per tonne of CO_2 avoided

5.4.1 Method

The economics of medium-scale combustion installations with carbon capture are calculated as the costs per tonne of CO_2 avoided. The net reduction of CO_2 emissions to the atmosphere by applying carbon capture depends on several factors, with the fraction of CO_2 captured and increased CO_2 emissions resulting from loss in overall efficiency being the most important. The costs (expressed as €/tCO_2 avoided) can only be defined against a pre-defined reference situation. For the economic analysis a reference situation needs to be defined for each case: the plant equipped with CCS is similar to the reference plant. In the calculations we assume that the heat and power co-generation installations with capture have the same electric output as the reference case without capture. For the heat-only installation we assume a similar heat output. In the co-generation variants, the difference in heat output is assumed to be compensated by an additional installation, i.e. the emissions and costs of the loss or gain of heat are taken into account. The same approach is taken for the power required for the compressors for CO_2; emission and capital, O&M and power costs are separately taken into account.[6] For each case we use the characteristics and costs given in the previous paragraph to calculate the avoidance costs (€/tCO_2 avoided).

The annual costs are calculated based on the estimated operation hours per year and a capital charge rate of 10%. The capture costs per tonne of CO_2 avoided are obtained with the formula below:

$$CostPerTonAvoided = \frac{(Annual\cos ts_{cap} - Annual\cos ts_{no\_cap})}{(Annual_{Avoided-CO_2})} \quad [€/ton]$$

5.4.2 Results

The results of the calculation of the costs per tonne of CO_2 avoided are summarised in Table 5-12 and Table 5-13. In the second table, the *compression costs are not* included, but the CO_2 is delivered at low/atmospheric pressure to the collection system. It might be more economically attractive to install central compression nodes in the system instead of compressing the CO_2 at each individual capture location. The costs and associated emissions for central compression systems are not included in the economic evaluation. The avoidance costs for reciprocating engines with post-combustion capture and natural gas-fired boilers with oxyfuel capture are relatively high and amount to above 100 €/tCO_2 when assuming a typical annual operational time for such installations of 4000 hours. The cheapest options are coal-fired boilers with oxyfuel capture (around 25 €/t) and natural gas-fired Solid Oxide fuel cells (around 34 €/t). The economic advantages of the coal-fired boiler are maintained even if fuel prices would double or triple.

The natural gas-fired SOFC also offers an inexpensive option with respect to costs per tonne of CO_2 avoided. The SOFC, however, suffers from the effects of its small scale (500 kW_e) as considered in this study. In case SOFC systems become commercially competitive in the future, they are likely to offer an economically viable method to capture CO_2. The latter can be attributed to the fact that 85% of the fuel is already converted under oxygen-rich conditions in the fuel cell, and oxygen conducting membranes are expected to become very cost effective (due to the reasonably high flux and inexpensive materials).

Table 5-12 also shows a limited dependency of the avoidance costs to the fuel price.

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[6] The assumed costs for power are 0.05 €/kWh_e. The assumed costs for heat varies from 0.015 to 0.035 €/kWh for fuel price of 2 and 8 €/GJ respectively.

Table 5-12. Costs per tonne of CO_2 avoided at various fuel prices, CO_2 emissions from plant without capture and typical annual operation time. Avoidance costs for 'realistic' fuel prices are highlighted

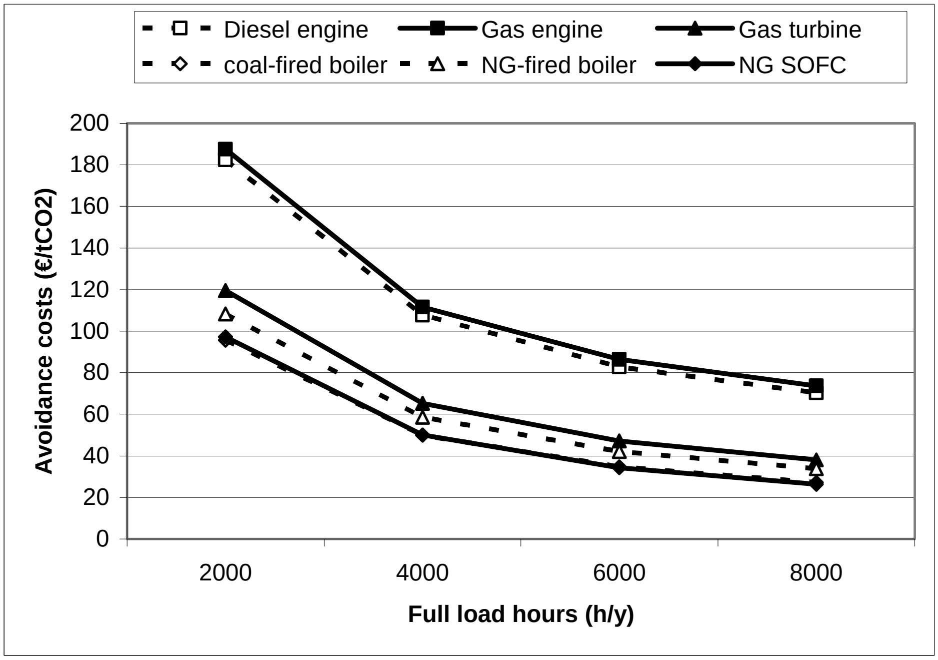
Fuel price [€/GJ]	**2**	**4**	**6**	**8**	**CO_2 emission [t/h]**	**Operation Time [h]**
Diesel Engine 1.5 MW_e	106	**108**	**110**	112	0.9	4000
Gas Engine 1.5 MW_e	109	**112**	**114**	117	0.8	4000
Gas turbine 5 MW_e	42	**47**	**53**	58	3.4	6000
Coal-fired boiler @18bar, 50 MW_{th}	**25**	**27**	29	32	20.3	8000
NG boiler @1bar, 5 MW_{th}	107	**108**	**110**	111	1.1	2000
Natural Gas SOFC, 500 kW_e	33	**34**	**36**	37	0.2	6000

Table 5-13. Same as table Table 5-12, but without costs to compress CO_2. Avoidance costs for 'realistic' fuel prices are highlighted

Fuel price [€/GJ]	**2**	**4**	**6**	**8**	**CO_2 emission [t/h]**	**Operation Time [h]**
Diesel Engine 1.5 MW_e	84	**86**	**87**	88	0.9	4000
Gas Engine 1.5 MW_e	91	**92**	**93**	95	0.8	4000
Gas turbine 5 MW_e	30	**34**	**39**	43	3.4	6000
Coal-fired boiler @18bar, 50 MW_{th}	**15**	**18**	20	22	20.3	8000
NG boiler @1bar, 5 MW_{th}	79	**81**	**82**	84	1.1	2000
Natural Gas SOFC, 500 kW_e	9	**9**	**9**	9	0.2	6000

Figure 5-8 depicts the sensitivity CO_2 avoidance costs to operation time. The graph demonstrates that the avoidance costs are strongly influenced by the annual time the installation is in operation. Comparing Table 5-12, Figure 5-8 and Table 5-14 demonstrates that cases with elongated annual operation time are favourable with respect to the lowest costs per tonne of CO_2 avoided. Therefore efforts on CO_2 capture should best be concentrated on continuous operating processes.

Comparing the natural gas fired boiler and the coal-fired boiler with oxyfuel capture shows that the avoidance costs of a natural gas fired boiler are substantial higher than the avoidance costs of the coal-fired version. The costs of heat for a natural gas fired boiler are also considerably larger, as they are rather used for peak load than for base load heat supply. Medium-scale natural gas-fired boilers are, for example, applied for central heating applications, where they typically operate up to 2000 hours on an annual basis.



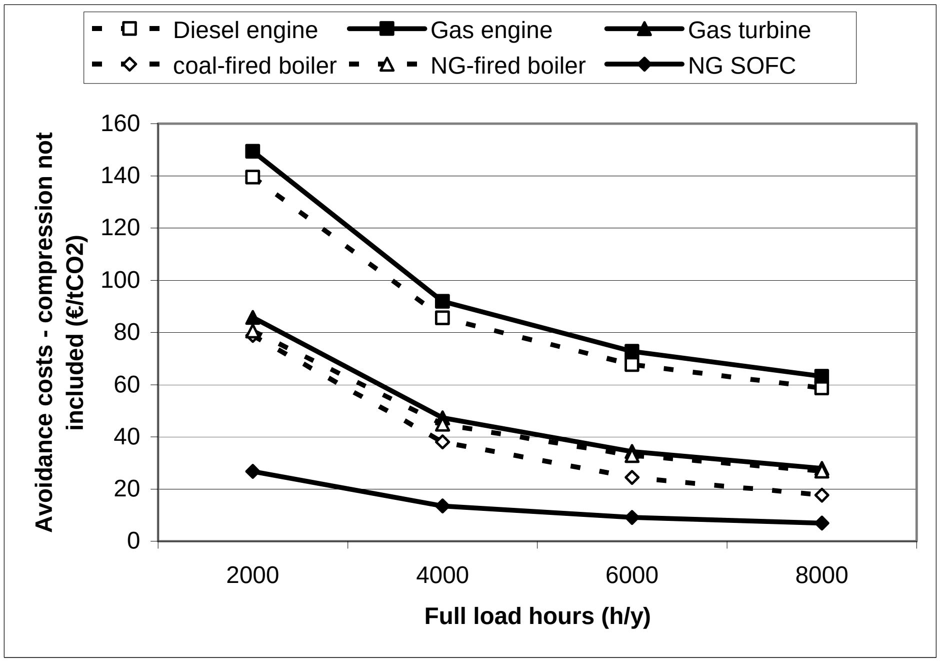


Figure 5-8. Avoidance costs of CO_2 for six different medium-scale installations as function of the annual full load hours. First graph (on top) shows costs of systems including compression. Second graph (below) shows costs of systems without compression. The fuel price is assumed to be 4€/GJ in all cases

Table 5-14 and Table 5-15 display the avoidance costs for various fuel prices when the annual operation time is fixed at 6000 hours. It should be noted that the annual operation time solely depends on the application of the equipment. The application of 6000 operation hours per year is for some applications therefore fairly unrealistic.

Table 5-14. Costs per tonne of CO_2 avoided at various fuel prices, normalised for 6000 annual operating hours. Avoidance costs for 'realistic' fuel prices are highlighted

Fuel price [€/GJ]	**2**	**4**	**6**	**8**	**CO_2 emission [t/h]**	**Operation time [h]**
Diesel Engine 1.5 MW_e	81	**83**	**85**	87	0.9	6000
Gas Engine 1.5 MW_e	83	**86**	**89**	91	0.8	6000
Gas turbine 5 MW_e	42	**47**	**53**	58	3.4	6000
Coal-fired boiler @18bar, 50 MW_{th}	**33**	**35**	37	39	20.3	6000
Natural Gas boiler @1bar, 5 MW_{th}	41	**42**	**44**	45	1.1	6000
Natural Gas SOFC, 500 kW_e	33	**34**	**36**	37	0.20	6000

Table 5-15. Same as Table 5-14, but without costs to compress CO_2. Avoidance costs for 'realistic' fuel prices are highlighted

Fuel price [€/GJ]	**2**	**4**	**6**	**8**	**CO_2 emission [t/h]**	**Operation time**
Diesel Engine 1.5 MW_e	66	**68**	**69**	70	0.9	6000
Gas Engine 1.5 MW_e	71	**73**	**74**	76	0.8	6000
Gas turbine 5 MW_e	30	**34**	**39**	43	3.4	6000
Coal-fired boiler @18bar, 50 MW_{th}	**26**	**28**	30	33	20.3	6000
Natural Gas boiler @1bar, 5 MW_{th}	31	**33**	**34**	36	1.1	6000
Natural Gas SOFC, 500 kW_e	9	**9**	**9**	9	0.20	6000

From Figure 5-8, Table 5-12 and Table 5-14 it can be concluded that coal-fired boilers offer the cheapest medium-scale capture option.

For industrial natural gas-fired boilers with OCM capture, costs per tonne of CO_2 avoided are in the same range as the avoided cost for industrial gas turbines, when assuming equal operation time, but more expensive assuming typical operation time. It is expected, however, that the PSA pre-combustion CO_2 capture for gas turbine applications will be available earlier than oxyfuel capture of CO_2 in boilers on the basis of the OCM technology.

5.5 Conclusions

The following conclusions are derived, based on the observations in paragraph 5.4.2.

1. Operation time of the installation has a large influence on the avoidance costs. Therefore, installations which are only operated seasonally or in peak load, for example, have considerably higher avoidance costs.
2. The coal-fired boiler with OCM capture offers the lowest avoidance costs while simultaneously producing heat at low costs. It should be noted that the OCM technology is still in development.
3. Pre-combustion capture in combination with industrial gas turbines under elongated annual operating hours is expected to be promising within the near future, due to the existing availability of the technology.
4. Capture from reciprocating engines tends to be the most expensive solution for medium-scale CO_2 capture, due to the small scale and the low number of operating hours.
5. Capture from SOFC appears promising in the midterm future if the SOFC market develops and fuel cell capital expenditures can be reduced significantly.

For each of the small-scale applications discussed in above paragraphs the potential for application has been addressed. Table 5-16 gives an overview of the CO_2 emissions that potentially can be avoided by each of the applications individually. Coal-fired boilers have by far the largest CO_2 reduction potential.

Table 5-16. Overview of emission reduction potentials per medium-scale application

Medium-scale application	**CO_2 emission [Mt per year]**
Diesel & Gas engines	600 - 700
Gas turbines	250-300
Coal-fired boilers	2500 - 3000
Natural Gas boilers	800 - 1000
Natural Gas SOFC	< 20

5.6 Selection of case for detailed analysis

The coal-fired boiler with oxygen conducting membrane capture technology has been selected for an in-depth analysis on performance and costs. The in-depth analysis is presented in chapter 6.

The coal-fired boiler is selected for further analysis based on the following arguments:

- The indicative avoidance costs are the lowest of the studied cases;

- The market share, and the related share in emissions, of (coal-fired) boilers is relatively large;
- The capture technology, although modified, might also be applied to boilers fired with other types of fuel;
- Coal might be the dominant fuel for the longer-term, i.e. at the moment that CCS for medium-scale applications might become relevant;
- The expectation is that oxygen conducting membrane technology is a promising technology, although still in a pre-stage of development;
- Solid Oxide Fuel Cells are cost-wise attractive but the market share of SOFC is virtual non-existent and very likely not very large in the medium term future. It is still quite uncertain whether the SOFC technology will have a break-through at all.

6 Oxyfuel Coal Boiler with Oxygen Conducting Membranes

6.1 Introduction

This chapter describes the detailed assessment of a new-build oxyfuel coal boiler with Oxygen Conducting Membranes (OCM). This alternative has been selected for further evaluation (see chapter 5). It comprises a 50 MW_{th} industrial Circulating Fluidized Bed (CFB) boiler for generation of low-pressure steam (10 bar, 200 ºC). Important aspects discussed during this detailed assessment are:

- Plant description (case 1: “natural gas and coal”, case 2: “coal only”)
- Operational flexibility (start-up, shut down and part load operation)
- Simulation results (stream summary, efficiency penalties)
- CO_2 purity
- Equipment sizing
- Economic evaluation
- Sensitivity analysis

6.2 Plant Description

Two cases of the oxyfuel coal boiler with OCM are evaluated. The flowsheets of both processes are presented in Appendix A and B. In case 1 “natural gas and coal”, natural gas is combusted to preheat air that is fed to the OCM. Case 2 “coal only” involves preheating of air fed to the OCM, by mounting a heat exchanger inside the boiler.
An important modification with respect to the description of the coal-fired oxyfuel boiler in section 5.3.3 comprises the stand-alone installation of the CO_2 compressor instead of extracting work from the drive shaft of the turbine. This enhances the operational flexbility, however it requires power from the electricity grid that results in additional CO_2 emissions beyond the boundary limits of the plant.

Air must be heated to 800 °C to facilitate oxygen permeation through the membrane, and moreover sufficient difference in oxygen partial pressure on the feed and permeate side of the membrane is required (Yantovski et al., 2004). It must be noted that OCM are still subject to research, therefore little is known about the effect of constituents such as SO_x, chlorines and mercury on its performance. Moreover, the chemical and mechanical stability of the membranes as well as the membrane sealing form important issues in pursuit of commercial implementation.

The flowsheets of both cases (Appendix A and B) show that the air inlet stream <**1**> is compressed to 18 bar <**2**> and further heated by the OCM retentate outlet stream <**6**>. For case 1, the targeted temperature of 800 °C is reached by addition of natural gas <**4**>

in the combustion chamber. This results in a lower carbon capture ratio since flue gases obtained from natural gas combustion are vented into the atmosphere. Case 2 comprises further preheating of air by a gas-to-gas heat exchanger mounted inside the coal boiler.

The oxygen-lean air **<7>** downstream of the OCM is expanded **<8>** and applied to pre-heat Boiler Feed Water (BFW) **<23>**. The oxygen-lean air **<9>** is subsequently released to the atmosphere at 105 ºC. Coal **<10>** is fed to the oxyfuel CFB boiler together with limestone to capture SO_x in-situ. The boiler generates steam (10 bar, 200 ºC) **<25>** from Boiler Feed Water (BFW) **<24>**. The flue gas from the boiler **<11>** is led to a cyclone and a dry Electro-Static Precipitator (ESP) where ash **<13>** is removed. The LP steam conditions allow a flue gas temperature of 150 ºC, which is above the acid dew point temperature due to in-situ desulphurisation. The CO_2-rich primary recycle **<15>** maintains the boiler temperature at 900 ºC, by recycling approximately 75% of the particulate-free flue gas **<12>** to the inlet of the boiler. Approximately half of the remaining flue gas **<14>** is fed to the knockout drum <**16**>, whereas the other half **<19>** is preheated by the permeate outlet stream **<21>** of the OCM. Stream **<20>** is enriched with oxygen that permeated through the OCM, subsequently cooled and applied as secondary recycle **<22>** to the boiler.

With respect to the second case it must be noted that the implementation of the air preheater (HX4) inside the boiler is highly challenging. Gas-to-gas heat exchangers typically require large surface areas due to the low heat transfer coefficient, and moreover state-of-the-art heat exchangers are suitable for operation at relatively low temperatures. Research on heat exchangers capable of heating pressurised air up to temperatures of approximately 1000 ºC is ongoing, pursuing application in externally fired combustion cycles (Aquaro, Pieve, 2007). These heat exchangers are to be positioned in a pulverised coal boiler, after which the air is further heated by natural gas addition and subsequently expanded in a turbine. Two prototypes are under development: a finned-tube convective arrangement and a radiant panel where tubes a covered with refractory liner. Experiments demonstrated that the first arrangement suffers from particulate deposition on the finned-tubes whereas the refractory liner in the second arrangement is particularly vulnerable with respect to alkali slag corrosion, resulting in refractory replacement at regular intervals (Oakey et al., 2003). Moreover, the application within a CFB boiler operated at 900 ºC results in a relatively large heat exchanger surface area compared with a pulverised coal boiler operated at more elevated temperatures.

Start-up and shut down

Coal boilers require significant start-up and shut down periods to prevent excessive thermal stresses in the refractory liner, which could result in failure. A typical start-up and shut down period of 8 hours is appropriate for a boiler of the size specified in these cases. If the heating rate of the OCM amounts to two Kelvin per minute, the start-up and shut down period matches with the boiler. This seems practically feasible for materials applied for the OCM. The flow of the secondary recycle can be gradually increased if both temperature and pressure at the feed side of the membrane allow oxygen permeation.

The combustion of methane in the first case allows start-up of the OCM prior to start-up of the boiler. This implies that oxygen is also produced and applied as oxidation agent during start-up of the boiler. Moreover, the primary recycle is operated to recycle CO_2 to the boiler during start-up, which results in reduced size of the AGR/SCR section since both throughput and NO_x formation are significantly reduced. Moreover, CO_2 can be captured from the flue gases of the coal boiler during start-up.

In the second case coal is combusted with air during start-up; hence the primary recycle is not applied because oxygen is already sufficiently diluted by nitrogen. This results in an increased NO_x formation, since air is applied as oxidation agent. Application of the secondary recycle is omitted during start-up, since both temperature and pressure on the feed side of the OCM do not result in oxygen permeation through the membrane until the preheated air approaches a temperature of 800 ºC.

In the first case, the upper part of the flow sheet (above the OCM) resembles the gas generator part of a gas turbine. The application of methane to increase the air inlet temperature of the OCM allows enhanced adaptation of the oxygen separation section of the plant upon thermal load fluctuations. The flexibility upon thermal load fluctuations is slightly hampered by the plant configuration in the second case, due to the integration of air preheating within the CFB boiler.

6.3 Modelling

The flow sheets displayed in Appendix A and B, have been thermodynamically assessed with Aspen Plus (version 13.1, AspenOne). The dedicated membrane model developed by ECN (Pex, Delft, 2005) was applied during the simulations in Aspen Plus. This Fortran based model allows selection of logarithmic transport behaviour across membranes, amongst other transport mechanisms. The most important assumptions are displayed in Table 6-1.

Table 6-1. Assumptions for modelling

Section	Parameter	Units	Value
OCM	Temperature	°C	800
	Air inlet pressure	bar	18
	Average oxygen permeation	$g\ m^{-2}\ s^{-1}$	2.50
	Secondary recycle (stream <19> & <20>)	mol/s	120.0
Gas generator	Isentropic efficiency compressor	-	0.88
	Isentropic efficiency turbine	-	0.88
	Mechanical efficiency	-	0.98
Boiler	Adiabatic temperature	°C	900
	Outlet temperature	°C	150
	Carbon content in ash[1]	wt%	2.0
	Heat loss	MW_{th}	2.0
ESP	Separation efficiency ash	wt%	100.0
Limestone injection	Total separation efficiency SO_x	wt%	95.0
	Limestone/Sulphur molar ratio		2.5
CO_2 compressor (five-stage)	Isentropic efficiency per stage	-	0.85
	Mechanical efficiency per stage	-	0.90

[1] CFB coal boiler, allows application of ash in construction materials

The natural gas composition is based on the standard composition of Norwegian North Sea natural gas (LHV = 46.88 MJ/kg); the applied coal composition appertains to Eastern Australian coal (LHV = 25.87 MJ/kg), as described in IEA-GHG report no. 2005/9 (Dillon et al., 2005). The delivered prices of Norwegian North Sea natural gas and Eastern Australian coal were assumed to be 5 and 2 €/GJ respectively. Addition of limestone to reduce SO_x emissions was not included during modelling to allow rapid convergence of the flowsheet, moreover the amount added is rather small. Moreover, the costs related to ash, limestone and gypsum disposal are omitted since application in construction materials is foreseen.

During modelling a CO_2 purity of 95.0 mol% is pursued, however this purity is governed by the stoichiometric ratio over the boiler, being 1.2 (20% O_2 excess). The secondary recycle is fixed at 120 mol/s to allow rapid convergence of the Fortran-based membrane model in Aspen Plus. The average oxygen permeation through the OCM amounts to 2.5 $g\ m^{-2}\ s^{-1}$ (10.5 $nml\ cm^{-2}\ min^{-1}$), which corresponds to recent experimental results by (Vente, Haije & Rak, 2006) upon correction for higher oxygen partial pressures at the feed side of the membrane in this particular case. An oxygen permeation of 10 $nml\ cm^{-2}\ min^{-1}$ is generally accepted as the lower boundary limit required for cost effective OCM operation. Heat transfer through the membrane is neglected.

The presence of limited amounts of nitrogen and oxygen in captured CO_2 is not expected to impose any difficulties during transport and storage. However if the CO_2 is used for enhanced oil recovery, the allowed oxygen content remains subject to discussion (De Koeier 2006, personal communication). Moreover, the presence of impurities increases the compression duty to some extent (Aspelund, Jordal, 2006).

The development of boilers that are specifically adapted for oxyfuel combustion in combination with OCM has a significant potential. A sensitivity analysis for Case 1 provides the resulting CO_2, H_2O, O_2 and N_2 concentration in the CO_2 product stream as a function of the air in-leakage, which are displayed in Figure 6-1. The build-up of nitrogen in both recycles appears to have a very large influence on the purity of the CO_2 product stream. Increasing the air in-leakage from 0 to 10%, results in an increase of the NO_x concentration in the CO_2 product stream from 6 to 65 ppm, due to thermal NO_x formation. However, the SO_x concentration decreases from 27 to 22 ppm upon increasing the air in-leakage, due to nitrogen dilution. Air in-leakage is neglected in the assessment.

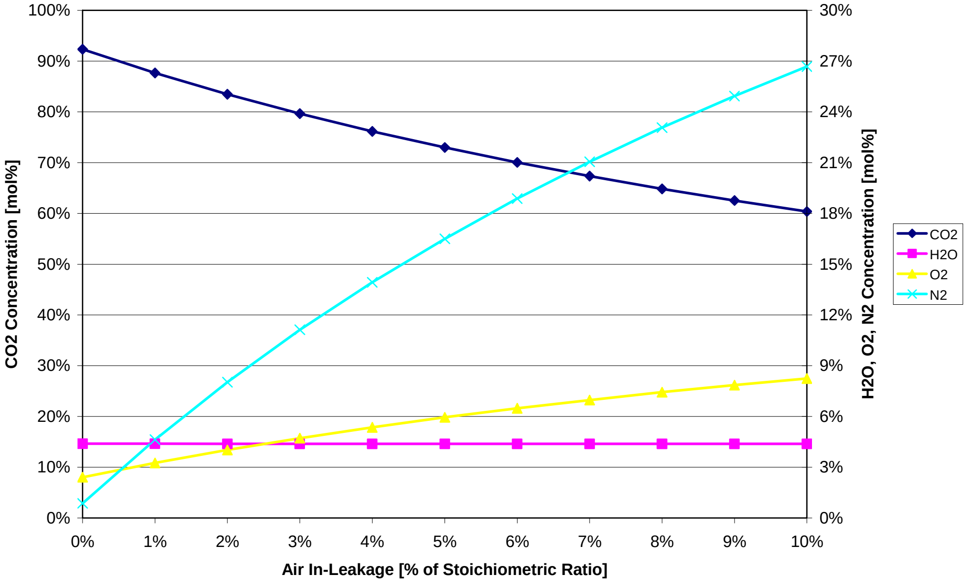


Figure 6-1 Air in-leakage versus CO_2, H_2O, O_2 and N_2 concentrations downstream of knockout drum

CO_2 obtained from oxyfuel combustion can be further purified by cryogenic distillation. A small fraction of compressed CO_2 is expanded to facilitate sufficient cooling duty. The feed conditions of the mixture approach the triple point of CO_2, when cooled CO_2 condenses whereas incondensables such as nitrogen and oxygen remain in the vapour phase. Expansion of 1.8 wt% of the captured CO_2 is reported to result in a CO_2 purity increase from 95.90 to 99.97%, moreover the increase in power requirement for CO_2 liquefaction amounts to 5% (White, Allam & Miller, 2006). The economic evaluation for each case was prepared both with (Table 6-3) and without (Table 6-4) compression to 110 bar and purification by cryogenic CO_2 distillation.

6.4 Results

The stream summary obtained during the flowsheet calculations for both cases are displayed in Appendix C and D. Based on the thermal input of coal and natural gas, the thermal efficiency of the oxyfuel boiler amounts to 87.6% on LHV basis for Case 1. This relatively high efficiency is attributable to application of gas for air pre-heating, as well as the higher extent of heat recovery from the oxygen-lean air. The electricity required for the CO_2 compressor is not taken into account with respect to this efficiency, but solely to determine the operating costs and the overall amount of avoided CO_2 emissions. The stoichiometric oxygen ratio in the boiler amounts to 1.20, which results in a CO_2 purity of 92.3 mol% for Case 1 without cryogenic CO_2 distillation. Total moisture removal would result in a CO_2 purity of 96.6 mol%.

The CO_2, H_2O, O_2 and N_2 concentrations as a function of the stoichiometric ratio are displayed in Figure 6-2, these are similar for both cases. This chart demonstrates that relatively large stoichiometric ratios still allow high CO_2 purities, which is mainly ascribed to the absence of nitrogen permeation during oxygen production in the OCM (nitrogen present at the boiler outlet solely originates from coal-bound nitrogen). Moreover, relatively low boiler temperatures require a larger amount of inert gas at the boiler inlet; this is obtained by a larger primary recycle and subsequently results in an elevated CO_2 concentration at the boiler outlet. The latter poses an advantage for application of CFB boilers in comparison with pulverised coal boilers.

Additionally, oxygen production in an Air Separation Unit (ASU) typically contains 3-5% nitrogen, which predominantly builds up in the primary recycle over the boiler. This gives rise to research in coal combustion at low stoichiometric ratios: total combustion is reported at stoichiometric ratios of 1.01 to 1.03 (IPCC, 2005).

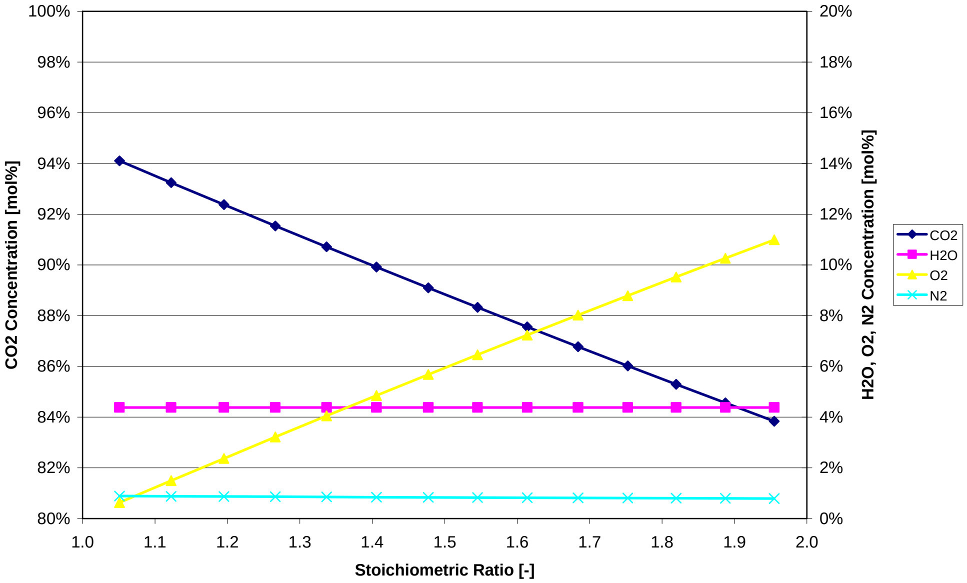


Figure 6-2. Stoichiometric oxygen ratio versus CO_2, H_2O, O_2 and N_2 concentrations downstream of the knockout drum

Table 6-2 displays the equipment sizing resulting from the flowsheet calculations. The heat transfer coefficients were obtained from (Peters, Timmerhaus & West, 2004), and are characteristic values for the heat transfer between the involved phases in the various heat exchangers.

Table 6-2. Sizing of additional equipment

Section	Parameter	Units	Case 1	Case 2
OCM	Surface area	m^2	1478	1806
Gas generator	Work compressor	MW	15.00	16.59
	Work turbine	MW	15.30	16.92
HX1	Heat duty	MW_{th}	6.01	5.55
	Heat transfer coefficient	$W\ m^{-2}\ K^{-1}$	30	30
	Surface area	m^2	888	743
HX2	Heat duty	MW_{th}	3.25	3.25
	Heat transfer coefficient	$W\ m^{-2}\ K^{-1}$	20	20
	Surface area	m^2	1609	1453
HX3	Heat duty	MW_{th}	3.56	4.17
	Heat transfer coefficient	$W\ m^{-2}\ K^{-1}$	20	20
	Surface area	m^2	3439	3951
HX4[1]	Heat duty	MW_{th}	-	11.33
	Heat transfer coefficient	$W\ m^{-2}\ K^{-1}$	-	50
	Surface area	m^2	-	1128
Knockout drum	Heat duty	MW_{th}	2.77	3.38
	Heat transfer coefficient	$W\ m^{-2}\ K^{-1}$	20	20
	Surface area	m^2	2482	3028
CO_2 compressor	Work compressors	MW_e	1.72	2.10
(5-stage, 110 bar)	Heat duty coolers	MW_{th}	2.45	2.99

[1] *Costs estimated to be twice as high as stainless steel heat exchanger with equal surface area*

The results of the economic evaluation are displayed in Table 6-3. It demonstrates that operation of an oxyfuel coal boiler with OCM is economically viable at market prices above €22.22 per tonne CO_2 avoided for case 1. The costs related to coal consumption decrease slightly in comparison with the reference case (no CO_2 capture), however the costs for natural gas and electricity that are introduced upon carbon capture are more significant.

The results for the second case demonstrate that an oxyfuel coal boiler with OCM is economically viable at market prices above €21.46 per tonne CO_2 avoided. A third case is introduced to compare both cases with conventional MEA scrubbing, downstream of an air-fired coal CFB boiler. The MEA case proves to be economically viable at €70.31 per tonne CO_2 avoided.

Table 6-3. Economic evaluation oxyfuel coal boiler with OCM with CO_2 purification and compression

	Case 1: **Coal/Nat. gas**	**Case 2:** **Coal only**	**Case 3:** **MEA**	**Reference** **w/o capture**	**Unit**
Coal (Eastern Australian)	1.827	2.231	3.062	2.221	kg/s
LHV Coal (Eastern Australian)	25870	25870	25870	25870	kJ/kg
Natural Gas (Norwegian)	0.209				kg/s
LHV Natural Gas (Norwegian)	46880				kJ/kg
Load	50000	50000	50000	50000	kW_{th}
Annual operation time at 100% capacity	7500	7500	7500	7500	h
Overall thermal efficiency	87.6%	86.6%	63.1%	87.0%	
Fuel input	57078	57704	79226	57471	kW_{th}
Carbon capture ratio (from coal)	98.2%	98.2%	85.0%		-
CO_2 for storage (from coal)	4.304	5.104	6.043		kg/s
CO_2 emission combustion coal & NG	0.679	0.094	1.066	5.156	kg/s
CO_2 emission electricity consumption	0.483	0.551	0.761	0.104	kg/s
CO_2 avoided	4.099	4.615	3.433		kg/s
Auxiliary power consumption	0.571	0.577	1.188	0.489	MW_e
CO_2 compressor duty	1.689	2.003	2.371		MW_e
Specific power CO_2 compressor	0.109	0.109	0.109		kWh_e/kg
Investment OCM					
Specific costs[1]	€ 1,500	€ 1,500			€/m^2
Estimated oxygen permeation	2.50	2.50			$g\ m^{-2}\ s^{-1}$
Oxygen required	3.695	4.516			kg/s
Subtotal OCM	€ 2,217,109	€ 2,709,605			
Investment boiler					
Specific costs[2]	€ 325,000	€ 325,000	€ 325,000	€ 325,000	€/MW_{th}
Subtotal coal boiler	€ 15,362,111	€ 18,753,730	€ 25,748,483	€ 18,678,161	
Investment heat exchangers					
HX1	€ 217,266	€ 199,473			
HX2	€ 288,882	€ 275,101			
HX3	€ 415,478	€ 444,345			
HX4		€ 487,327			
Knockout drum	€ 355,588	€ 391,144			
Subtotal heat exchangers	€ 1,277,214	€ 1,797,389			
Investment compressor & turbine	€ 3,000,000	€ 3,663,783			
Investment CO_2 Compressor	€ 3,397,168	€ 3,841,379	€ 4,338,931		
Investment MEA Plant[3]			€ 15,000,000		
Total capital costs	**€ 25,253,602**	**€ 30,764,438**	**€ 45,087,413**	**€ 18,678,161**	

Table 6-3 (Continued)

		Case 1: Coal/Nat. gas	Case 2 : Coal only	Case 3: MEA	w/o capture	Unit
Operational costs						
Annuity	11%	€ 2,777,896	€ 3,384,088	€ 4,959,615	€ 2,054,598	€/a
O&M	4%	€ 1,010,144	€ 1,230,578	€ 1,803,497	€ 747,126	€/a
Electricity	0.05 (€/kWh)	€ 880,917	€ 1,030,493	€ 1,563,395	€ 183,190	€/a
Coal	2.00 (€/GJ)	€ 2,552,474	€ 3,116,004	€ 4,278,209	€ 3,102,508	€/a
Natural Gas	5.00 (€/GJ)	€ 1,324,403	€ 0	€ 0	€ 0	€/a
Costs per tonne CO_2 avoided		**€ 22.22**	**€ 21.46**	**€ 70.31**		€/tonne
Costs deducted		-€ 2,458,412	-€ 2,673,742	-€ 6,517,294		€/a
Total annual operational costs		**€ 6,087,422**	**€ 6,087,422**	**€ 6,087,422**	**€ 6,087,422**	**€/a**

[1] *Derived from (Allam et al., 2004)*

[2] *Derived from (Nsakala et al., 2004) upon downscaling*

[3] *Derived from (Singh et al., 2003) upon downscaling; assumed thermal loss 3.6 MJ/kg CO_2 captured*

The costs per tonne CO_2 avoided for Case 1 and 2 are more or less equal, however Case 3 proves that coal-fired OCM boilers (case 1 and 2) offer a large potential over commercially available technology. The cases presented in Table 6-3 are also calculated without CO_2 compression and purification; this allows application of the obtained results in the assessment of central CO_2 collection nodes, where further compression occurs. The results are displayed in Table 6-4.

Table 6-4. Economic evaluation of oxyfuel coal boiler with OCM without CO_2 purification and compression

Input	**Case 1: Coal/Nat. gas**	**Case 2: Coal only**	**Case 3: MEA**	**Reference w/o capture**	**Unit**
Coal (Eastern Australian)	1.827	2.231	3.062	2.221	kg/s
LHV Coal (Eastern Australian)	25870	25870	25870	25870	kJ/kg
Natural Gas (Norwegian)	0.209				kg/s
LHV Natural Gas (Norwegian)	46880				kJ/kg
Load	50000	50000	50000	50000	kW_{th}
Annual operation time at 100% capacity	7500	7500	7500	7500	h
Overall thermal efficiency	87.6%	86.6%	63.1%	87.0%	
Fuel input	57078	57704	79226	57471	kW_{th}
Carbon capture ratio (from coal)	100.0%	100.0%	85.0%		-
CO_2 for storage (from coal)	4.383	5.197	6.043		kg/s
CO_2 emission comb. coal & NG	0.600	0.000	1.066	5.156	kg/s
CO_2 emission electricity consumption	0.122	0.123	0.254	0.104	kg/s
CO_2 avoided	4.538	5.137	3.940		kg/s
Auxiliary power consumption	0.571	0.577	1.188	0.489	MW_e
CO_2 compressor duty					MW_e
Specific power CO_2 compressor					kWh_e/kg
Investment OCM					
Specific costs[1]	€ 1,500	€ 1,500			€/m2
Estimated oxygen permeation	2.50	2.50			$g\ m^{-2}\ s^{-1}$
Oxygen required	3.695	4.516			kg/s
Subtotal OCM	€ 2,217,109	€ 2,709,605			
Investment boiler					
Specific costs[2]	€ 325,000	€ 325,000	€ 325,000	€ 325,000	$€/MW_{th}$
Subtotal coal boiler	€ 15,362,111	€ 18,753,730	€ 25,748,481	€ 18,678,161	
Investment heat exchangers					
HX1	€ 217,266	€ 199,473			
HX2	€ 288,882	€ 275,101			
HX3	€ 415,478	€ 444,345			
HX4		€ 487,327			
Knockout drum					
Subtotal heat exchangers	€ 921,626	€ 1,406,245			
Investment compressor & turbine	€ 3,000,000	€ 3,662,335			
Investment CO_2 Compressor					
Investment MEA Plant[3]			€ 15,000,000		
Total capital costs	**€ 21,500,846**	**€ 26,531,915**	**€ 40,748,481**	**€ 18,678,161**	

Table 6-4 (Continued)

	Input	Case 1: Coal/Nat. gas	Case 2: Coal only	Case 3:MEA	w/o capture	Unit
Operational costs						
Annuity	11%	€ 2,365,093	€ 2,918,511	€ 4,482,333	€ 2,054,598	€/a
O&M	4%	€ 860,034	€ 1,061,277	€ 1,629,939	€ 747,126	€/a
Electricity	0.05 (€/kWh)	€ 214,044	€ 216,389	€ 445,647	€ 183,190	€/a
Coal	2.00 (€/GJ)	€ 2,552,474	€ 3,116,004	€ 4,278,209	€ 3,102,508	€/a
Natural Gas	5.00 (€/GJ)	€ 1,324,403	€ 0	€ 0	€ 0	€/a
Costs per tonne CO_2 avoided		**€ 10.26**	**€ 9.01**	**€ 45.58**		€/tonne
Costs deducted		-€ 1,228,626	-€ 1,224,759	-€ 4,748,707		€/a
Total annual operational costs		**€ 6,087,422**	**€ 6,087,422**	**€ 6,087,422**	**€ 6,087,422**	**€/a**

[1,2,3] *See* Table 6-3

Footprint

The footprint of case 1 was estimated as displayed in Table 6-5. The footprint of the OCM boiler is significantly larger than the footprint of the reference case, being 236 m^2 versus 48 m^2 respectively. The required surface area is estimated at 150% of the calculated surface area, to allow inspection and maintenance of the unit operations.

Table 6-5. Footprint OCM Boiler (Case 1)

	Surface Area [m^2]	Height [m]
Boiler[1]	16	14
Coal bunker & HX3	16	10
OCM[2]	90	2
Compressor & turbine[3]	30	4
CO_2 compressor[4]	5	2
HX 1, HX 2 & HX 3[5]	15	2
Overall	236	14

[1] Based on downscaled coal-fired CFB boiler of 67.0 MW_{th} (Basu, Fraser, 1991)

[2] Based on tubular configuration: 12 modules with membrane surface area of 125 m^2 each (Vente, Haije, IJpelaan & Rusting, 2006)

[3] Based on Siemens SGT-400 mechanical drive (Gas Turbine World, 2005)

[4] Educated guess based on on-site compression to 30 bara

[5] Based on heat exchanger surface area to volume ratio of 150 m^2/m^3 (Bejan, 1993)

6.5 Sensitivity analysis

In section 4 the results of the economic evaluation of two oxyfuel boiler cases and the MEA scrubbing case are presented (see Table 6-3 and Table 6-4). A sensitivity analysis is performed for the cases with CO_2 purification and compression.

The input parameters used for the calculation of the economic performance of each of these processes are subject to uncertainties. In this section we present the results of the sensitivity analysis, which aims to investigate the influence of the variations of input parameters on the CO_2 avoidance costs of the different capture processes studied. The input parameters (e.g. component costs, discount rate) are characterized by variability and/or uncertainty. A parameter's variability refers to the range of that specific parameter and is determined by external conditions (e.g. discount rate). The uncertainty of a parameter refers to the limited knowledge of the system under investigation (e.g. oxygen flux). For the purpose of the analysis, the economic evaluation of all three cases has been implemented in a stand-alone Excel workbook. The Excel add-in @risk module performs the analysis using Monte Carlo simulation. Distributions are assigned to the most relevant input parameters using the @risk module.

The input parameters with significant uncertainty and/or variability have been identified and displayed in Table 6-6. For each input parameter a probability distribution is defined that best approaches the variability or uncertainty of the parameter. The type of probability distribution that is chosen (normal, uniform, exponential) depends on the expected uncertainty or variability in the parameter values. To give an example, a normal distribution (also called Gaussian distribution) is the most common way to describe the uncertainty within a parameter. While the mechanisms underlying these phenomena are often unknown, the use of the normal model can be theoretically justified if one assumes many small (independent) effects contribute to each observation in an additive fashion. A uniform distribution, sometimes also known as a rectangular distribution, is a distribution that has constant probability. This type of distribution is used when no extra information is available or when all values within a range have equal probability (like for the discount rate). The first step of the sensitivity analysis is to define the type of distribution for each input parameter. The second step is to define the minimum and maximum values for each parameter. The basis for the specification of these minimum and maximum values is a general understanding of the status of technologies and components. For example, the range applied to the probability distribution of costs for the coal boiler is assumed to be smaller than the range of the probability distribution for the innovative (and unproven) concept of the OCM. The probability distribution reflects that there is much uncertainty regarding the cost development of the OCM. Table 6-6 presents the parameters, distribution description and ranges. For each cell value of an input parameter the range of values within the distribution is defined (90% confidence interval).

Table 6-6 Parameters, distribution function and range taken into account for sensitivity analysis

Parameter	Type of distribution	Case 1 natural gas/ coal	Minimum-Maximum	Case 2 all coal	Minimum - Maximum	Case 3 MEA	Minimum - Maximum
Investment costs							
Specific costs OCM	Uniform	1,500	min -50% max +50%	1,500	min -50% max +50%	-	-
Specific costs boiler	Normal	€ 325,000	min -5% max +5%	€ 325,000	min -5% max +5%	€ 325,000	min -5% max +5%
Costs heat exchanger 1	Normal	€ 217,286	min -10% max +10%	€ 199,473	min -10% max +10%	-	-
Costs heat exchanger 2	Normal	€ 288,882	min -10% max +10%	€ 275,101	min -10% max +10%	-	-
Costs heat exchanger 3	Normal	€ 415,478	min -10% max +10%	€ 444,345	min -10% max +10%	-	-
Costs heat exchanger 4 (inside boiler)	Normal		min -40% max +40%	€ 487,327	min -40% max +40%	-	-
Costs knock-out drum	Normal	€ 355,588	min -5% max +5%	€ 391,144	min -5% max +5%	-	-
Costs Compressor, Turbine & Comb. Chamber	Normal	€ 3,000,000	min -5% max +5%	€ 3.662.335	min -5% max +5%	-	-
Specific costs CO_2 Compressor	Normal	€ 2,328,000	min -25% max +25%	€ 2,328,000	min -25% max +25%	€ 4,338,931	min -25% max +25%
Costs MEA plant	Normal	-	-	-	-	€ 15,000,000	min -25% max +25%
Technical parameters							
Estimated Oxygen Flux	Exponential	2.50	min -5% max +50%	2.50	min -5% max +50%	-	-
Economic parameters							
Annual operation time	Normal	7,500	min -20% max +10%	7,500	min -20% max +10%	7,500	min -20% max +10%
Discount rate	Uniform	0.08	min 0.04 max 0.12		min 0.04 max 0.12	0.08	min 0.04 max 0.12
Depreciation or lifetime	Uniform	20	min 15 max 30	20	min 15 max 30	20	min 15 max 30
Price Electricity	Exponential	€ 0.05	min 0% max 50%	€ 0.05	min 0% max 50%	€ 0.05	min 0% max 50%
Price Coal	Normal	€ 2.00	min -25% max 10%	€ 2.00	min -25% max 10%	€ 2.00	min -25% max 10%
Price Natural Gas	Normal	€ 5.00	min -25% max 50%	€ 5.00	min -25% max 50%	€ 5.00	min -25% max 50%

Results

Figure 6-3 to Figure 6-5 show the distribution of the CO_2 avoidance costs for the examined cases. As reported in the previous section the costs for case 1 (coal and natural gas) amount to 22.19 €/tCO_2 avoided. Based on the assumed distributions of the input parameters, 90% percent of the outcomes of the economic evaluation (costs per tonne CO_2 avoided) is between 16.12 and 29.40 €/tCO_2. The economic evaluation of case 2 (all coal) gives slightly lower costs of 21.41 €/tCO_2. The distribution of the costs for case 2 is a little smaller than case 1: costs range from 17.28 to 27.67 €/tCO_2. This shows that the costs per tonne avoided CO_2 are more sensitive to changes in natural gas price than to changes in costs of heat exchanger 4. The CO_2 avoidance costs for case 3 are 70.17 €/tCO_2. The distribution of the costs is 54.78 to 90.57 €/tCO_2 avoided (see Figure 6-5).

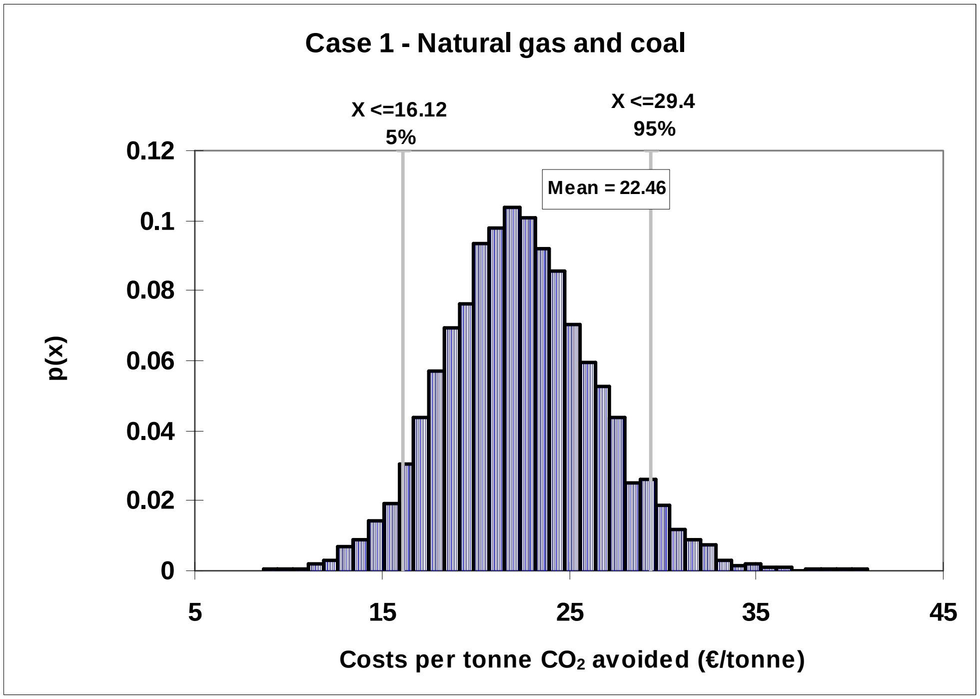


Figure 6-3 Distribution for avoidance costs for case 1: “natural gas and coal”.

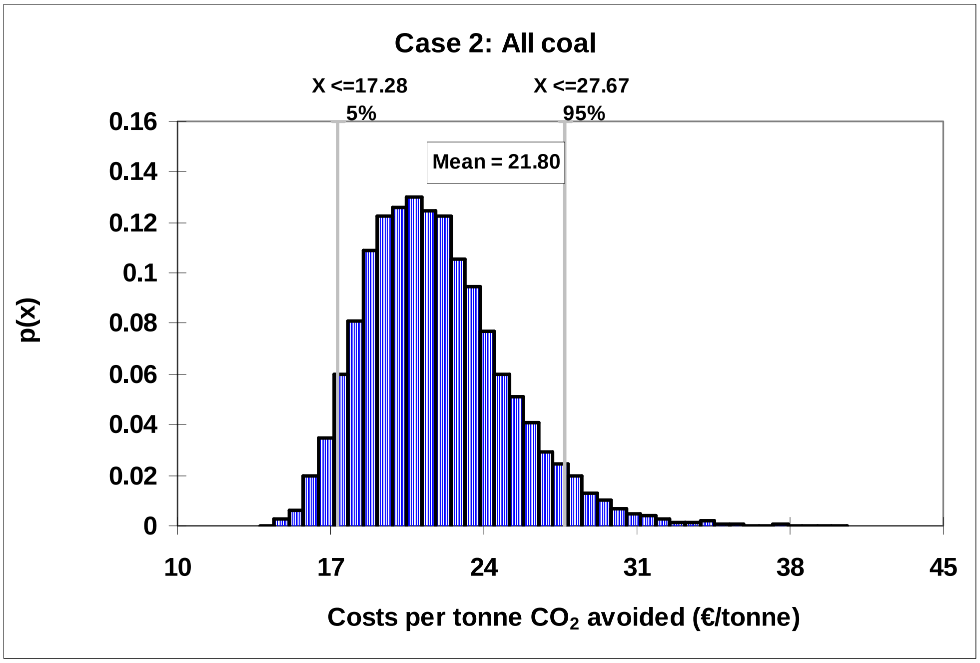


Figure 6-4 Distribution for avoidance costs for case 2: “all coal”.

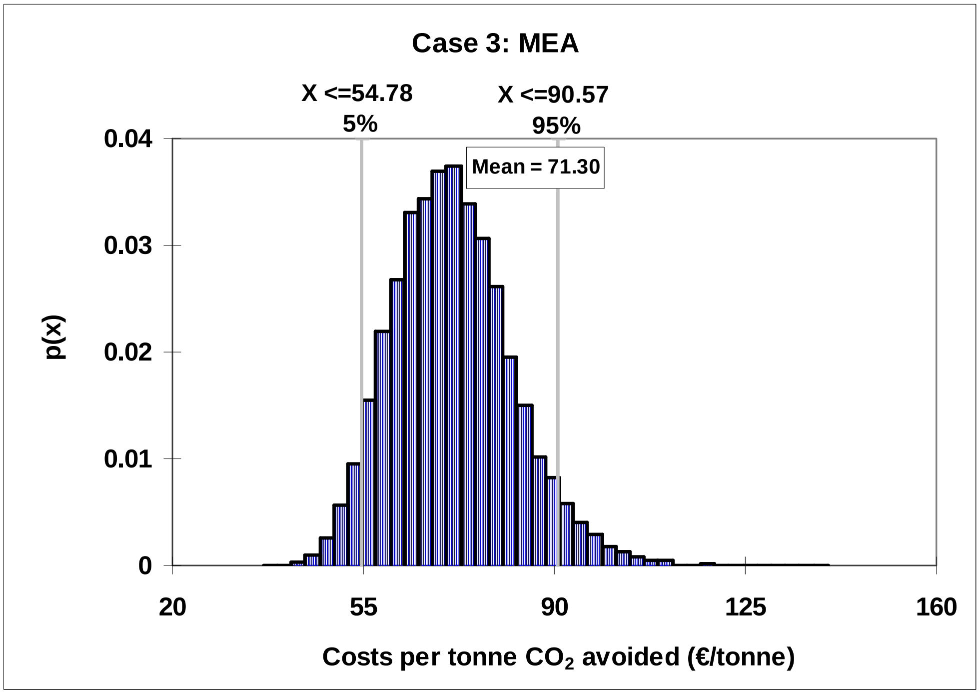


Figure 6-5 Distribution for avoidance costs for case 3: “MEA”

The sensitivity of the costs per tonne CO_2 avoided (€/ton) for selected parameters (see Table 6-6) is analysed and presented by means of so-called Tornado graphs. The input parameters are ranked in terms of their impact on the output parameter ‘costs per tonne CO_2 avoided’. The graph shows which input distributions are ‘significant’ in determining the value of the output variable (in this case costs per CO_2 avoided). The ranking of the parameters has been done by applying regression analysis. This is an appropriate method to investigate the effect of several X variables on Y. The Monte Carlo simulation drops a variable if the impact on the output variable is close to zero (or zero).

The regression analysis of case 1 “coal and natural gas” shows that CO_2 avoidance costs are most sensitive to the price of natural gas (see Figure 6-6). It has been assumed that the gas price varies between 3 and 7 €/GJ. The second largest impact on costs results from fluctuations in electricity price, the range of which has been set to 0.036 to 0.08 €/kWh. The regression coefficient of coal price and annual operation time is negative for both parameters, which implies that higher coal prices and increasing annual operation hours results in lower costs per tonne CO_2 avoided.

Figure 6-6 also shows that changes in costs of system components costs have a moderate effect on the avoidance costs. According to Figure 6-6 the avoidance costs are most sensitive to the (specific) investment costs of the OCM and to a somewhat lesser extent to the CO_2 compressor. These results reflect the high uncertainty in OCM costs (plus or minus 50%), because of the immaturity of this system component. The investment costs of other system components are of minor importance, especially those of the heat exchangers, because of the relatively small contribution to the total costs. Because the avoidance costs (output parameter) have very small sensitivity to the variability of the costs of heat exchanger 1, this parameter is dropped from the regression analysis.

The price of coal and the costs per tonne CO_2 avoided are negatively correlated (correlation coefficient of -0.30). This negative correlation implies that higher coal prices affect the economics of the oxyfuel coal boiler in a positive way, since coal consumption and therefore the costs of coal are higher in the reference case without CO_2 capture. In the case 1, part of the coal consumption is replaced by natural gas[7]. Higher coal prices also result in lower prices per tonne of CO_2 avoided for case 2, although the effect is much smaller. With increasing coal prices case 2 becomes competitive to the reference case at lower prices per tonne CO_2 avoided. The same argument holds for the impact of annual operation time on the avoidance costs for CO_2.

When only coal is used to fuel the system (represented by case 2) electricity prices have the strongest positive correlation with the cost per tonne of CO_2 avoided. Natural gas prices are not included since no natural gas is consumed in case 2. Whereas coal prices have a negative correlation with CO_2 avoidance cost in case 1, in case 2 the correlation of coal prices with the output parameter is positive (0.006). For case 1, the effect of higher coal prices is small compared to the reference case, because part of the fuel consumption is fulfilled by natural gas. In case 2 all coal is used and since capture also consumes en-

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[7] It should be noted that in this analysis the coal and natural gas prices are assumed to be independent of each other. In practice, there will be (some) interdependency.

ergy coal consumption is slightly higher (57704 kWth) than in the reference case without capture (57471 kWth).

Several input parameters of the “all coal” case show a negative correlation to the output parameter costs per tonne of CO_2 avoided: annual operation time, estimated oxygen flux and coal price. Increasing values for these parameters result in lower costs per tonne of CO_2 avoided. For case 3 “MEA”, the parameters “lifetime” and “annual operation time” increase are negatively correlated to the output parameter “costs per tonne CO_2”.

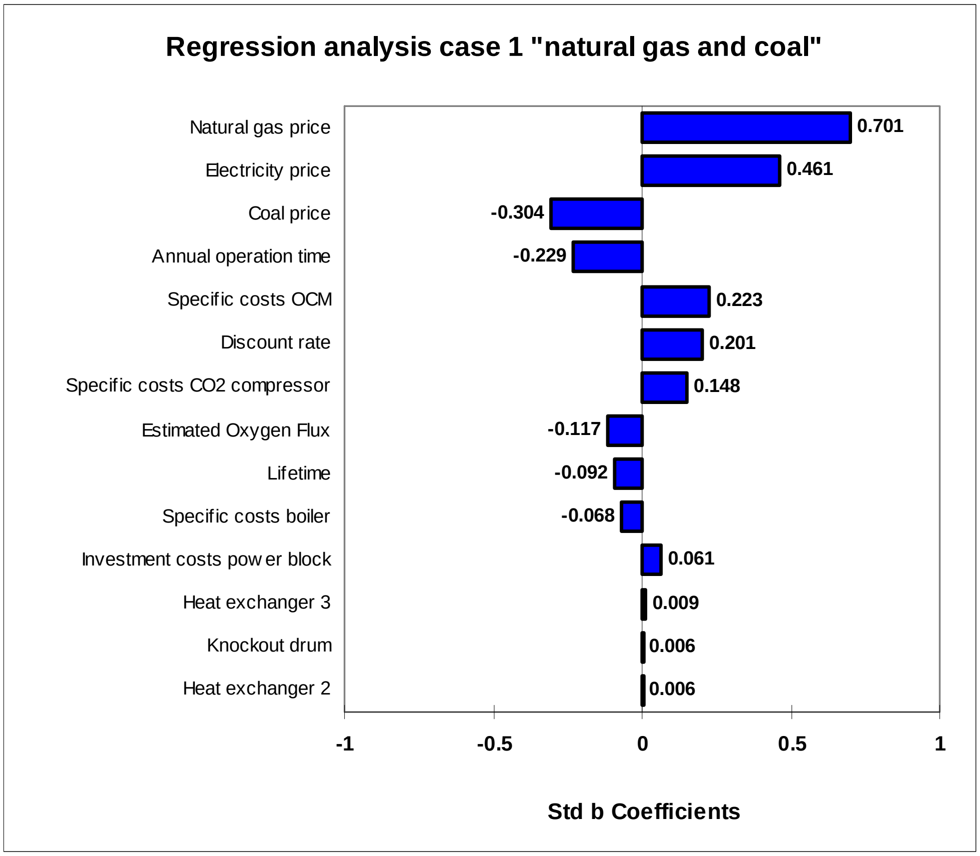

Figure 6-6 Regression analysis for coal-fired boiler with OCM for case 1: “natural gas and coal”

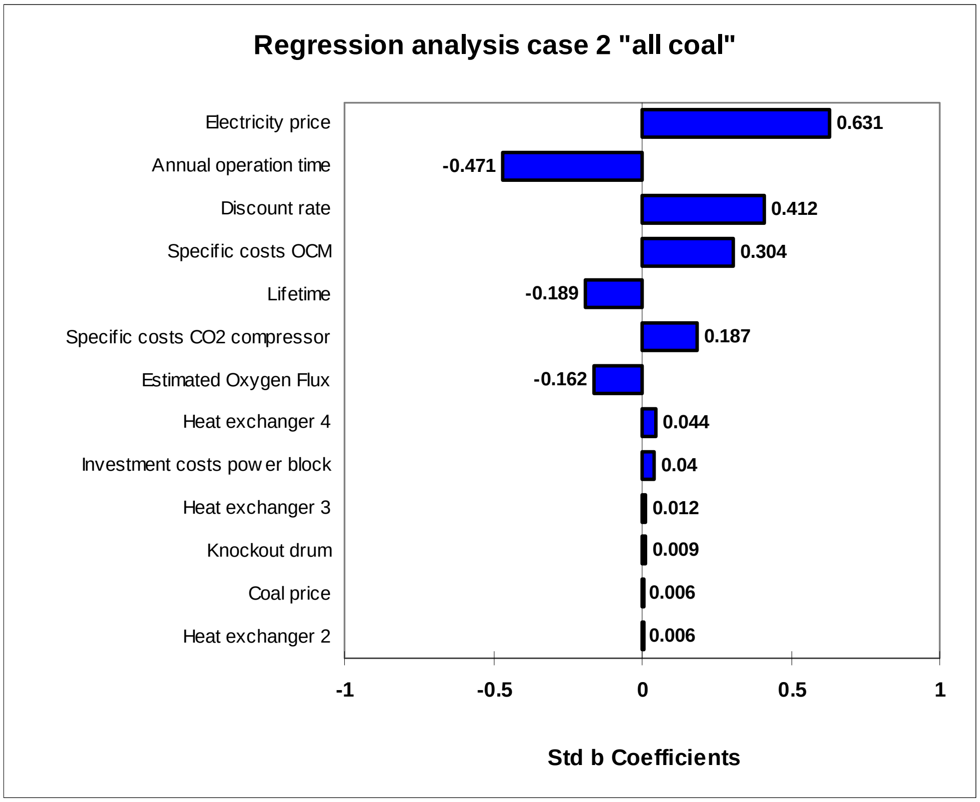

Figure 6-7 Regression analysis for coal-fired boiler with OCM for case 2 “all coal”

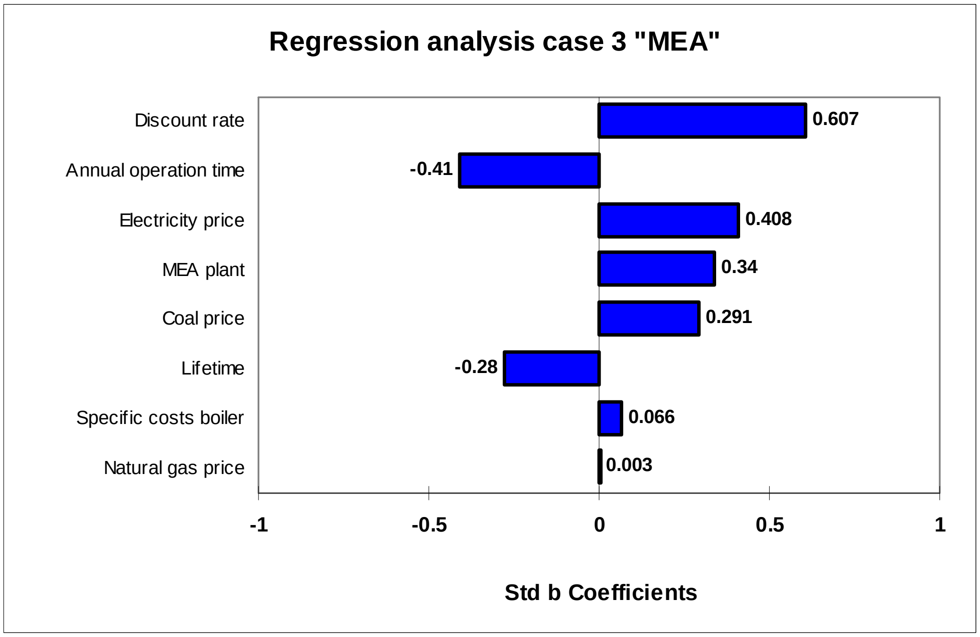

Figure 6-8 Regression analysis for coal-fired boiler for case 3 “MEA”

6.6 Conclusions

Coal-fired boilers with capture of CO_2 based on Oxygen Conducting Membranes may provide an effective way to reduce emissions of CO_2 to the atmosphere. In this detailed analysis two cases were assessed; case 1 “natural gas and coal” using additional natural gas to preheat air and case 2 “all coal” using an additional heat exchanger inside the coal-fired boiler to preheat air. A third case is introduced to compare both cases with conventional MEA scrubbing (case 3).

The results of the economic evaluation of all cases show that case 1 and 2 are almost equally attractive from an economic point of view. The difference in costs per tonne CO_2 avoided is less than 1 Euro. However, the capture ratio is significantly larger for case 2 (78% for case 1 versus 88% for case 2) and the uncertainty in costs is significantly smaller. For case 1, the costs are most strongly dependent on the price of natural gas. The variability of electricity prices has the largest effect on CO_2 avoidance costs in case 2. The large uncertainty in the required surface area of the OCM means that CO_2 avoidance costs are quite sensitive to changes of this parameter.

The capital investment of the OCM boiler is significantly lower for case 1 compared with case 2, being 25 M€ versus 31 M€ respectively. Moreover, the development of the gas-to-gas heat exchanger to be positioned inside the coal-fired boiler as applied in case 2 is still in a premature phase. Particulate fouling (CFB-boiler) and slag corrosion (PC-boiler) of these heat exchangers remain substantial challenges to overcome.

Furthermore it should be noted that the Oxygen Conducting Membranes with elevated oxygen permeation (2.5 g m^{-2} s^{-1}) only exist on lab-scale at present, whereas scale-up and increased membrane stability is required to facilitate technical and economic viability in advanced energy conversion systems. It might require several years before these membranes can be applied on the described scale.

However, the application of Oxygen Conducting Membranes for oxygen production in oxyfuel coal boilers results in elevated CO_2 purities compared with conventional oxygen production in Air Separation Units. This is mainly attributable to the high purity of the produced oxygen, which significantly reduces the nitrogen build-up in the primary recycle over the coal boiler. As a consequence a coal boiler with OCM can be operated at higher stoichiometric ratios than a coal boiler with an ASU, both resulting in equal CO_2 purities and under the assumption that air leakage into the boiler can be neglected.

7 Overall conclusions

The main focus on capture of carbon dioxide has been on large combustion plants, notably fossil fuel power plants. However, to achieve reductions in CO_2 emissions that are expected to be necessary, substantial reductions in emissions from medium-scale sources may be required. This study assesses the potential, performance and costs of capture from medium-scale CO_2 sources. A medium-scale installation is considered to consume between 1 and 100 MW of fuel input.

Carbon dioxide emissions of medium-scale installations aggregate to an estimated share of global emissions between 15 and 20%. A subdivision of emissions and shares of different types of combustion installations is presented in Table 7-1.

Table 7-1. Overview of annual worldwide CO_2 emissions and share in total emission for medium-scale applications

Medium-scale application	CO_2 emission [Mton per year]	Share in CO_2 emissions worldwide
Diesel & Gas reciprocating engines	600 - 700	2.5%-3.0%
Gas turbines	250-300	1%
Coal-fired boilers	2500 - 3000	10-12%
Natural gas boilers	800 - 1000	3-5%
Natural gas SOFC (in 2020)	< 20	<0.1%

There exists a broad range of applicable capture options for medium-scale combustion installations. In general it can be stated that small-scale thermal installations will have higher heat losses and higher specific investment costs than large-scale installations. Moreover, small-scale compressors demonstrate lower efficiency. The latter can partly be avoided by collecting CO_2 from various capture units by a low pressure suction collection system, prior to a large central compression system.
Conversely, medium-scale capture may offer certain benefits over large-scale application. An example is the application of waste heat in gas turbine flue gases for the desorption process of MEA absorption systems.

Five potential combinations of combustion systems with capture have been selected for a performance and cost assessment, based on criteria such as applicability of capture systems and market share. The selected systems are:

1. Reciprocating engine with membrane assisted liquid absorption
2. Natural gas turbine with pre-combustion PSA capture
3. Oxyfuel coal boiler with oxygen conducting membranes

4. Oxyfuel natural gas boiler with oxygen conducting membranes
5. Solid oxide fuel cell with oxygen conducting membrane afterburner

The diesel and gas engine demonstrate the highest costs which exceed 100 euro per tonne of carbon dioxide avoided (€/t), despite the fact that sufficient low-temperature heat is available in the flue gases. Gas turbines show a much better cost performance: approximately 40 to 60 €/t CO_2 avoided. Gas turbines are larger in size and are generally in operation for more hours than reciprocating engines. The lowest costs are obtained for coal-fired boilers with integrated capture using oxygen conducting membranes: around 25 €/t CO_2 avoided. However, this option is not yet commercial available. The development of oxygen conducting membranes is still in the research phase.

Comparable systems for natural gas fired boilers appear less favourable. This is mainly attributable to the higher hydrogen content of the natural gas in comparison with coal. Solid oxide fuel cells also offer an attractive method for low-costs emission reduction. However, it is not expected that the SOFC will obtain a substantial market share within the next 20 years.

Because of the high potential to reduce carbon dioxide emissions and estimated low costs, a more detailed techno-economic analysis of the oxyfuel coal boiler with oxygen conducting membranes (OCM) is performed. The analysis confirms that costs are competitive, and show that capture costs can be as low as 21 €/t CO_2 avoided. Without compression, the costs approximately amount to 9 €/t CO_2 avoided. This indicates that the integration of oxygen conducting membranes is very attractive, even in medium-scale coal-fired boilers. Oxygen separation is relatively cheap due to the level of integration between the OCM and the industrial boiler. Moreover, the CO_2 purity that can be obtained is relatively high in comparison with application of an air separation unit.

Two possible layouts for the oxyfuel coal boiler with OCM have been studied: case 1 with a separate air preheating system using natural gas combustion and case 2 with an integrated high-temperature gas-to-gas air preheater inside the boiler. The estimated avoidance costs are approximately equal for case 1 and case 2, both with and without CO_2 purification and compression. Application of case 2 offers a higher amount of CO_2 avoided, however the capital expenditure is 20% higher than case 1. Moreover, development and integration of high-temperature heat exchangers for application inside boilers remains highly challenging. Research dedicated to this type of heat exchangers is in progress.

Capture of CO_2 from medium-scale boilers using state-of-the art MEA separation systems has also been considered and the costs are estimated at approximately 70 €/t CO_2 avoided. Without CO_2 purification and compression the costs approximately amount 46 €/t CO_2 avoided.

Overall it can be concluded that CO_2 capture from medium-scale sources appears attractive for coal-fired boilers with OCM. The potential of these boilers is relatively large, i.e.

emissions represent about 10% of global emissions of CO_2 while capture costs are relatively low. The proposed technology requires considerable technological developments and full-scale application is not foreseen within the next 10 to 15 years. Furthermore, capture from dispersed medium-scale sources requires an extensive distribution and transport network which was not part of the scope during the study presented in this report.

Abbreviations

ASU	Air separation unit
ATR	Autothermal reformer
AZEP	Advanced zero emission power plant
BAT	Best Available Technology
BFW	Boiler feed water
$CaCO_3$	Calcium Carbonate
CaO	Calcium Oxide
CC	Combined cycle
CCR	Carbon Capture ratio, see page 11
CCS	Carbon capture and storage
CF	Coal fired furnace
CG	Coal Gasifier
CLC	Chemical looping combustion
CO	Carbon monoxide
CO_2	Carbon dioxide
C-partic	Carbon particulate matter
CRYO	Cryogenic
CuO	Copper Oxide
DeNOx	NO_x removal
ECN	Energy research Centre of the Netherlands
Efficiency	LHV electrical efficiency
ESP	Electrostatic precipitator
Eur	Euro
FGD	Flue gas desulphurisation
GT	Gas turbine
H_2	Hydrogen
H_2O	Water/steam
H_2S	Hydrogen Sulphide
Hg	Mercury
HHV	Higher heating value
HP steam	High pressure steam
HRSG	Heat recovery steam generator
HT shift	High temperature shift
HTS	High temperature shift
HX	Heat exchanger
IEA	International Energy Agency
IEA GHG	International Energy Agency Greenhouse gas R&D programme
IGCC	Integrated gasification combined cycle

JC	Jacobs Consultancy
kWhF	kWh fuel
LT shift	Low temperature shift
LTS	Low temperature shift
LHV	Lower heating value
LP steam	Low pressure steam
Me	A metal atom
MCM	Mixed conducting membrane
MEA	MonoEthanol Amine (absorber), a liquid phase chemical CO_2 absorbent
MeO	A metal oxide
N_2	Nitrogen
NG	Natural gas
NGCC	Natural gas fired combined cycle
NOx	Nitrogen Oxides
O_2	Oxygen
OCM	Oxygen conducting membranes
OOM	Order of Magnitude
PC	Pulverized coal boiler
PM10	Small particulate matter
Post-C	Post combustion
Pre-C	Pre-combustion
PSA	Pressure Swing Absorption
R&D	Research and Development
RE	Reciprocating (piston) Engine
SCR	Selective catalytic reduction
Selexol	A liquid phase physical CO_2 absorption agent
SERP	Sorption enhanced reaction process
SE-WGS	Sorption enhanced water gas shift
SOFC	Solid Oxide Fuel Cell
SOFC-GT	Solid Oxide Fuel Cell - Gas Turbine combination
SOx	Sulphur Oxides
TSA	Temperature swing adsorption
USD	US dollars
VOC	Volatile organic substances
WGS	Water gas shift

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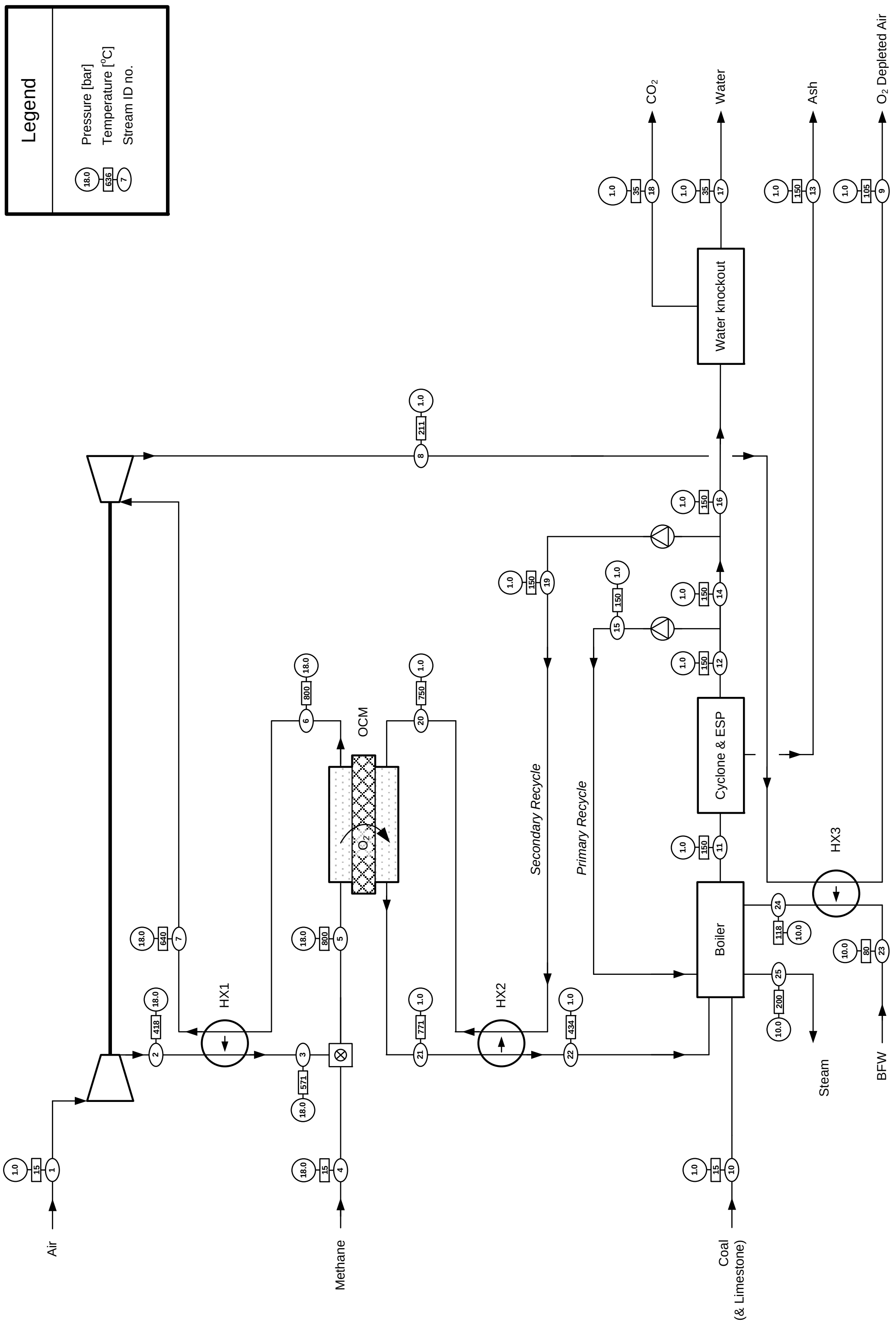
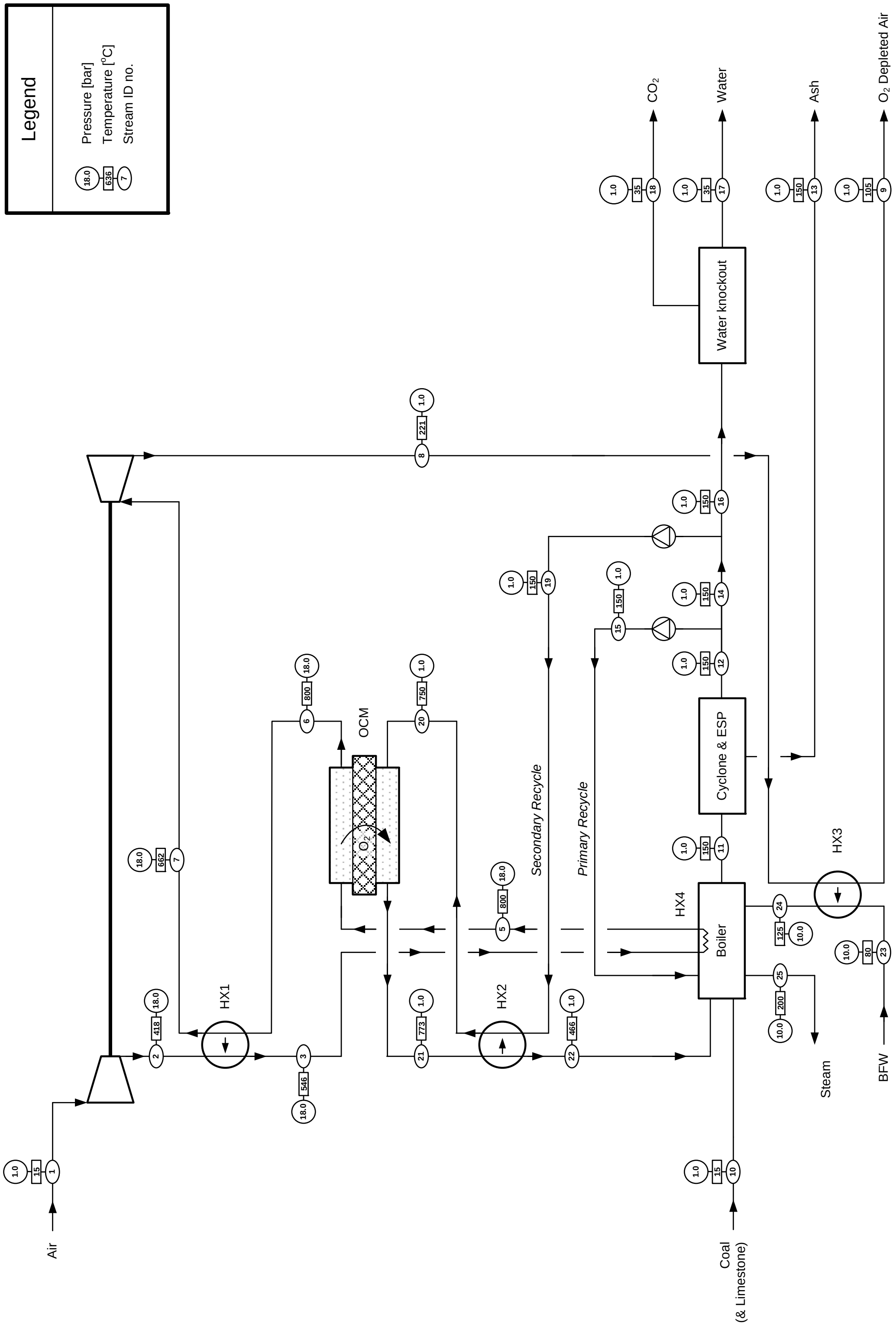
Appendix A Flowsheet OCM Boiler Case 1



Appendix B Flowsheet OCM Boiler Case 2



Appendix C Stream Summary OCM Boiler Case 1

Stream ID	1	2	3	4	5	6	7	8	9	10	11	12
Temperature C	15.0	418.3	571.4		800.0	800.0	639.7	210.9	105.0	15.0	150.0	150.0
Pressure bar	1.0	18.0	18.0		18.0	18.0	18.0	1.0	1.0	1.0	1.0	1.0
Vapor Frac	1	1	1	1	1	1	1	1	1		0.995	1
Liquid Frac												
Solid Frac										1	0.005	
LIQUID & VAPOUR PHASE												
Mole Flow kmol/sec												
N2	0.951	0.951	0.951	4.31E-05	0.951	0.951	0.951	0.951	0.951		0.008	0.008
O2	0.255	0.255	0.255		0.231	0.115	0.115	0.115	0.115		0.022	0.022
AR	0.011	0.011	0.011		0.011	0.011	0.011	0.011	0.011			
CO2	0.000	0.000	0.000	1.94E-04	0.013	0.013	0.013	0.013	0.013		0.859	0.859
H2O	0.012	0.012	0.012		0.036	0.036	0.036	0.036	0.036		0.443	0.443
CH4				9.05E-03								
C2H6				9.92E-04								
C3H8				3.56E-04								
C4H10				1.51E-04								
CO											6.16E-08	6.16E-08
H2											2.41E-08	2.41E-08
NO											5.84E-06	5.84E-06
NO2											3.03E-08	3.03E-08
SO2											4.88E-04	2.44E-05
SO3											1.96E-05	9.78E-07
Mole Frac												
N2	0.773	0.773	0.773	0.004	0.766	0.844	0.844	0.844	0.844		0.006	0.006
O2	0.208	0.208	0.208		0.186	0.102	0.102	0.102	0.102		0.017	0.017
AR	0.009	0.009	0.009		0.009	0.010	0.010	0.010	0.010			
CO2	0.000	0.000	0.000	0.018	0.011	0.012	0.012	0.012	0.012		0.644	0.645
H2O	0.010	0.010	0.010		0.029	0.032	0.032	0.032	0.032		0.333	0.333
CH4				0.839								
C2H6				0.092								
C3H8				0.033								
C4H10				0.014								
CO											4.62E-08	4.62E-08
H2											1.81E-08	1.81E-08
NO2											2.27E-08	2.27E-08
NO											4.38E-06	4.38E-06
SO2											3.66E-04	1.83E-05
SO3											1.47E-05	7.34E-07
Total Flow kmol/sec	1.230	1.230	1.230	0.011	1.242	1.127	1.127	1.127	1.127		1.334	1.333
Total Flow kg/sec	35.500	35.500	35.500	0.209	35.709	32.014	32.014	32.014	32.014		46.777	46.746
Total Flow cum/sec	29.471	3.960	4.830	0.014	6.189	5.614	4.780	45.366	35.436		46.803	46.786
Enthalpy kJ/kmol	-2.86E+03	9.33E+03	1.42E+04	-8.37E+04	1.34E+04	1.21E+04	6.79E+03	-6.79E+03	-9.94E+03		-3.29E+05	-3.29E+05
Enthalpy kJ/kg	-9.91E+01	3.23E+02	4.93E+02	-4.31E+03	4.64E+02	4.27E+02	2.39E+02	-2.39E+02	-3.50E+02		-9.39E+03	-9.39E+03
Entropy kJ/kmol-K	3.777	5.967	12.345	-118.769	20.518	19.224	13.843	17.935	10.575		6.696	6.663
Entropy kJ/kg-K	0.131	0.207	0.428	-6.122	0.714	0.677	0.487	0.631	0.372		0.191	0.190
Density kmol/cum	0.042	0.311	0.255	0.791	0.201	0.201	0.236	0.025	0.032		0.028	0.028
Density kg/cum	1.205	8.966	7.349	15.350	5.770	5.702	6.698	0.706	0.903		0.999	0.999
Average MW	28.854	28.854	28.854		28.747	28.414	28.414	28.414	28.414		35.076	35.065
Solid Phase												
Mass Flow kg/sec												
COAL										1.827		
ASH											0.220	

Stream ID	13	14	15	16	17	18	19	20	21	22	23	24	25
Temperature C	150.0	150.0	150.0	150.0	35.0	35.0	150.0	750.0	770.5	433.8	80.0	118.3	200.0
Pressure bar	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	10.0	10.0	10.0
Vapor Frac	0.124	1	1	1		1	1	1	1	1			1
Liquid Frac					1						1	1	
Solid Frac	0.876												
LIQUID & VAPOUR PHASE													
Mole Flow kmol/sec													
N2		0.002	0.006	0.001	0.000	0.001	0.001	0.001	0.001	0.001			
O2		0.005	0.018	0.003	0.000	0.003	0.002	0.002	0.117	0.117			
AR													
CO2		0.174	0.685	0.097	0.000	0.097	0.077	0.077	0.077	0.077			
H2O		0.090	0.354	0.050	0.045	0.005	0.040	0.040	0.040	0.040	1.058	1.058	1.058
CH4													
C2H6													
C3H8													
C4H10													
CO		1.25E-08	4.91E-08	6.93E-09		6.93E-09	5.54E-09	5.54E-09	5.54E-09	5.54E-09			
H2		4.89E-09	1.92E-08	2.72E-09		2.72E-09	2.17E-09	2.17E-09	2.17E-09	2.17E-09			
NO		1.18E-06	4.65E-06	6.57E-07	2.73E-13	6.57E-07	5.25E-07	5.25E-07	5.25E-07	5.25E-07			
NO2		6.14E-09	2.42E-08	3.41E-09	9.81E-11	3.32E-09	2.73E-09	2.73E-09	2.73E-09	2.73E-09			
SO2	4.64E-04	4.95E-06	1.95E-05	2.75E-06	1.70E-09	2.75E-06	2.20E-06	2.20E-06	2.20E-06	2.20E-06			
SO3	1.86E-05	1.98E-07	7.80E-07	1.10E-07	1.15E-09	1.09E-07	8.80E-08	8.80E-08	8.80E-08	8.80E-08			
Mole Frac													
N2		0.006	0.006	0.006	0.000	0.009	0.006	0.006	0.003	0.003			
O2		0.017	0.017	0.017	0.000	0.024	0.017	0.017	0.499	0.499			
AR													
CO2		0.645	0.645	0.645	0.000	0.923	0.645	0.645	0.328	0.328			
H2O		0.333	0.333	0.333	1.000	0.044	0.333	0.333	0.170	0.170	1.000	1.000	1.000
CH4													
C2H6													
C3H8													
C4H10													
CO		4.62E-08	4.62E-08	4.62E-08		6.62E-08	4.62E-08	4.62E-08	2.35E-08	2.35E-08			
H2		1.81E-08	1.81E-08	1.81E-08		2.59E-08	1.81E-08	1.81E-08	9.23E-09	9.23E-09			
NO2		2.27E-08	2.27E-08	2.27E-08	2.16E-09	3.16E-08	2.27E-08	2.27E-08	1.16E-08	1.16E-08			
NO		4.38E-06	4.38E-06	4.38E-06	6.02E-12	6.27E-06	4.38E-06	4.38E-06	2.23E-06	2.23E-06			
SO2	0.961	1.83E-05	1.83E-05	1.83E-05	3.75E-08	2.62E-05	1.83E-05	1.83E-05	9.34E-06	9.34E-06			
SO3	0.039	7.34E-07	7.34E-07	7.34E-07	2.54E-08	1.04E-06	7.34E-07	7.34E-07	3.74E-07	3.74E-07			
Total Flow kmol/sec	0.000	0.270	1.063	0.150	0.045	0.105	0.120	0.120	0.235	0.235	1.058	1.058	1.058
Total Flow kg/sec	0.031	9.474	37.272	5.266	0.817	4.449	4.208	4.208	7.903	7.903	19.056	19.056	19.056
Total Flow cum/sec	0.017	9.482	37.304	5.270	0.001	2.673	4.211	10.210	20.439	13.843	0.026	0.027	3.924
Enthalpy kJ/kmol	-2.95E+05	-3.29E+05	-3.29E+05	-3.29E+05	-2.88E+05	-3.74E+05	-3.29E+05	-3.02E+05	-1.42E+05	-1.55E+05	-2.84E+05	-2.81E+05	-2.37E+05
Enthalpy kJ/kg	-4.57E+03	-9.39E+03	-9.39E+03	-9.39E+03	-1.60E+04	-8.80E+03	-9.39E+03	-8.62E+03	-4.22E+03	-4.63E+03	-1.58E+04	-1.56E+04	-1.31E+04
Entropy kJ/kmol-K	23.961	6.663	6.663	6.663	-167.975	4.898	6.663	45.856	48.926	32.997	-156.161	-147.117	-49.216
Entropy kJ/kg-K	0.370	0.190	0.190	0.190	-9.323	0.115	0.190	1.308	1.458	0.983	-8.668	-8.166	-2.732
Density kmol/cum	0.029	0.028	0.028	0.028	41.630	0.039	0.028	0.012	0.012	0.017	40.161	38.698	0.270
Density kg/cum	1.847	0.999	0.999	0.999	750.047	1.664	0.999	0.412	0.387	0.571	723.514	697.164	4.856
Average MW	64.681	35.065	35.065	35.065	18.017	42.444	35.065	35.065	33.562	33.562	18.015	18.015	18.015
Solid Phase													
Mass Flow kg/sec													
COAL													
ASH	0.220												

Note: SO_2 and SO_3 in stream <13> are captured by limestone and exist in the solid phase as gypsum (result of omitting the capture reaction in AspenPlus)

Appendix D Stream Summary OCM Boiler Case 2

Stream ID	1	2	3	4	5	6	7	8	9	10	11	12
Temperature C	15.0	418.3	546.3		800.0	800.0	661.9	220.5	105.0	15.0	150.0	150.0
Pressure bar	1.0	18.0	18.0		18.0	18.0	18.0	1.0	1.0	1.0	1.0	1.0
Vapor Frac	1	1	1		1	1	1	1	1		0.995	1
Liquid Frac												
Solid Frac										1	0.005	
LIQUID & VAPOUR PHASE												
Mole Flow kmol/sec												
N2	1.051	1.051	1.051		1.051	1.051	1.051	1.051	1.051		0.010	0.010
O2	0.282	0.282	0.282		0.282	0.141	0.141	0.141	0.141		0.029	0.029
AR	0.013	0.013	0.013		0.013	0.013	0.013	0.013	0.013			
CO2	0.000	0.000	0.000		0.000	0.000	0.000	0.000	0.000		1.057	1.057
H2O	0.014	0.014	0.014		0.014	0.014	0.014	0.014	0.014		0.546	0.546
CH4												
C2H6												
C3H8												
C4H10												
CO											7.38E-08	7.38E-08
H2											2.90E-08	2.90E-08
NO											7.35E-06	7.35E-06
NO2											3.91E-08	3.91E-08
SO2											5.96E-04	2.98E-05
SO3											2.45E-05	1.22E-06
Mole Frac												
N2	0.773	0.773	0.773		0.773	0.862	0.862	0.862	0.862		0.006	0.006
O2	0.208	0.208	0.208		0.208	0.116	0.116	0.116	0.116		0.018	0.018
AR	0.009	0.009	0.009		0.009	0.010	0.010	0.010	0.010			
CO2	0.000	0.000	0.000		0.000	0.000	0.000	0.000	0.000		0.644	0.644
H2O	0.010	0.010	0.010		0.010	0.011	0.011	0.011	0.011		0.332	0.333
CH4												
C2H6												
C3H8												
C4H10												
CO											4.49E-08	4.49E-08
H2											1.76E-08	1.76E-08
NO2											2.38E-08	2.38E-08
NO											4.48E-06	4.48E-06
SO2											3.63E-04	1.81E-05
SO3											1.49E-05	7.45E-07
Total Flow kmol/sec	1.360	1.360	1.360		1.360	1.219	1.219	1.219	1.219		1.642	1.641
Total Flow kg/sec	39.250	39.250	39.250		39.250	34.734	34.734	34.734	34.734		57.579	57.540
Total Flow cum/sec	32.584	4.378	5.183		6.778	6.075	5.297	50.068	38.346		57.622	57.606
Enthalpy kJ/kmol	-2.86E+03	9.33E+03	1.34E+04		2.17E+04	2.13E+04	1.68E+04	2.90E+03	-5.24E+02		-3.29E+05	-3.29E+05
Enthalpy kJ/kg	-9.91E+01	3.23E+02	4.65E+02		7.53E+02	7.48E+02	5.89E+02	1.02E+02	-1.84E+01		-9.38E+03	-9.39E+03
Entropy kJ/kmol-K	3.777	5.967	11.374		20.220	18.834	14.298	18.400	10.508		6.719	6.689
Entropy kJ/kg-K	0.131	0.207	0.394		0.701	0.661	0.502	0.646	0.369		0.192	0.191
Density kmol/cum	0.042	0.311	0.262		0.201	0.201	0.230	0.024	0.032		0.028	0.028
Density kg/cum	1.205	8.966	7.572		5.791	5.717	6.557	0.694	0.906		0.999	0.999
Average MW	28.854	28.854	28.854		28.854	28.490	28.490	28.490	28.490		35.068	35.057
Solid Phase												
Mass Flow kg/sec												
COAL										2.231		
ASH											0.269	

Stream ID	13	14	15	16	17	18	19	20	21	22	23	24	25
Temperature C	150.0	150.0	150.0	150.0	35.0	35.0	150.0	750.0	773.0	465.5	80.0	124.9	200.0
Pressure bar	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.0	10.0	10.0	10.0
Vapor Frac	0.124	1	1	1		1	1	1	1	1			1
Liquid Frac					1						1	1	
Solid Frac	0.876												
LIQUID & VAPOUR PHASE													
Mole Flow kmol/sec													
N2		0.002	0.008	0.001	0.000	0.001	0.001	0.001	0.001	0.001			
O2		0.005	0.024	0.003	0.000	0.003	0.002	0.002	0.143	0.143			
AR													
CO2		0.195	0.861	0.118	0.000	0.118	0.077	0.077	0.077	0.077			
H2O		0.101	0.445	0.061	0.055	0.006	0.040	0.040	0.040	0.040	1.058	1.058	1.058
CH4													
C2H6													
C3H8													
C4H10													
CO		1.36E-08	6.01E-08	8.24E-09		8.24E-09	5.39E-09	5.39E-09	5.39E-09	5.39E-09			
H2		5.35E-09	2.36E-08	3.24E-09		3.24E-09	2.12E-09	2.12E-09	2.12E-09	2.12E-09			
NO		1.36E-06	5.99E-06	8.22E-07	3.41E-13	8.22E-07	5.38E-07	5.38E-07	5.38E-07	5.38E-07			
NO2		7.23E-09	3.19E-08	4.37E-09	1.26E-10	4.25E-09	2.86E-09	2.86E-09	2.86E-09	2.86E-09			
SO2	5.66E-04	5.51E-06	2.43E-05	3.33E-06	2.06E-09	3.33E-06	2.18E-06	2.18E-06	2.18E-06	2.18E-06			
SO3	2.32E-05	2.26E-07	9.97E-07	1.37E-07	1.43E-09	1.35E-07	8.94E-08	8.94E-08	8.94E-08	8.94E-08			
Mole Frac													
N2		0.006	0.006	0.006	0.000	0.009	0.006	0.006	0.003	0.003			
O2		0.018	0.018	0.018	0.000	0.025	0.018	0.018	0.549	0.549			
AR													
CO2		0.644	0.644	0.644	0.000	0.922	0.644	0.644	0.296	0.296			
H2O		0.333	0.333	0.333	1.000	0.044	0.333	0.333	0.153	0.153	1.000	1.000	1.000
CH4													
C2H6													
C3H8													
C4H10													
CO		4.49E-08	4.49E-08	4.49E-08		6.44E-08	4.49E-08	4.49E-08	2.06E-08	2.06E-08			
H2		1.76E-08	1.76E-08	1.76E-08		2.53E-08	1.76E-08	1.76E-08	8.11E-09	8.11E-09			
NO2		2.38E-08	2.38E-08	2.38E-08	2.27E-09	3.32E-08	2.38E-08	2.38E-08	1.10E-08	1.10E-08			
NO		4.48E-06	4.48E-06	4.48E-06	6.16E-12	6.42E-06	4.48E-06	4.48E-06	2.06E-06	2.06E-06			
SO2	0.961	1.81E-05	1.81E-05	1.81E-05	3.71E-08	2.60E-05	1.81E-05	1.81E-05	8.34E-06	8.34E-06			
SO3	0.039	7.45E-07	7.45E-07	7.45E-07	2.58E-08	1.06E-06	7.45E-07	7.45E-07	3.42E-07	3.42E-07			
Total Flow kmol/sec	0.001	0.303	1.338	0.183	0.055	0.128	0.120	0.120	0.261	0.261	1.058	1.058	1.058
Total Flow kg/sec	0.038	10.640	46.901	6.433	0.998	5.434	4.207	4.207	8.723	8.723	19.055	19.055	19.055
Total Flow cum/sec	0.021	10.652	46.955	6.440	0.001	3.267	4.212	10.210	22.720	16.041	0.026	0.028	3.924
Enthalpy kJ/kmol	-2.95E+05	-3.29E+05	-3.29E+05	-3.29E+05	-2.88E+05	-3.73E+05	-3.29E+05	-3.02E+05	-1.25E+05	-1.38E+05	-2.84E+05	-2.80E+05	-2.37E+05
Enthalpy kJ/kg	-4.57E+03	-9.39E+03	-9.39E+03	-9.39E+03	-1.60E+04	-8.80E+03	-9.39E+03	-8.61E+03	-3.75E+03	-4.12E+03	-1.58E+04	-1.55E+04	-1.31E+04
Entropy kJ/kmol-K	23.905	6.689	6.689	6.689	-167.975	4.930	6.689	45.868	48.676	34.618	-156.161	-145.648	-49.216
Entropy kJ/kg-K	0.369	0.191	0.191	0.191	-9.323	0.116	0.191	1.308	1.457	1.036	-8.668	-8.085	-2.732
Density kmol/cum	0.029	0.028	0.028	0.028	41.630	0.039	0.028	0.012	0.011	0.016	40.161	38.427	0.270
Density kg/cum	1.847	0.999	0.999	0.999	750.047	1.664	0.999	0.412	0.384	0.544	723.514	692.278	4.856
Average MW	64.696	35.057	35.057	35.057	18.017	42.429	35.057	35.057	33.404	33.404	18.015	18.015	18.015
Solid Phase													
Mass Flow kg/sec													
COAL													
ASH	0.269												

Note: SO_2 and SO_3 in stream <13> are captured by limestone and exist in the solid phase as gypsum (result of omitting the capture reaction in AspenPlus)

Appendix E - Fuel processors

In this appendix, two options for fuel processing for pre-combustion capture are described.

E.1. Hydrogen production facilities

At present, most hydrogen in the world is produced from natural gas. Two working principles are applied:

1. Steam reforming, in which methane, or higher hydrocarbons, react with steam, to produce hydrogen and carbon monoxide, using the reformer reaction:
   $CH_4 + H_2O \rightarrow CO + 3\ H_2$
2. Partial oxidation, in which methane, or higher hydrocarbons are partly oxidised. The reaction:
   $CH_4 + ½\ O_2 \rightarrow CO + 2\ H_2$.

The syngas (mixture of basically CO and H_2) is commonly applied in several industrial processes. If the purpose is hydrogen production, the reformer or partial oxidiser is followed by one or more water-gas shift reactors, where the carbon monoxide is converted to hydrogen and carbon dioxide:

$$CO + H_2O \rightarrow CO_2 + H_2.$$

After the shift reactors feed impurities, unconverted fuel and CO_2 are most commonly removed with PSA, and up to 99.99% pure hydrogen remains.

The application of these hydrogen production facilities ranges from fuel processors for domestic fuel cell cogenerators (typical 3 kW_{LHV}) to industrial scale hydrogen plants, over 1 GW_{LHV}.

Discussion of possible capture technologies

- Which capture technologies can be applied on small and medium scale depends mainly on the types of energy converter applied after the hydrogen generator. As can be seen when comparing this list with the long list, in principle all the options identified can be applied. Reformers and partial oxidisers are however best suited for pre-combustion capture.
- When the hydrogen is used elsewhere, capture is restricted to the pre-combustion options. Removal of the CO_2 before the hydrogen purification enhances this process, because less impurities remain in the feed of the PSA.
- The use of oxyfuel is restricted to after combustion of the tail gas remaining after hydrogen separation with PSA. This tail gas is used for heating the reformer reaction.
- Oxyfuelling of the total process including the power generation is less preferable because of the high hydrogen content of the fuel

E.2. Coal gasifiers

In coal gasifiers, coal is partial oxidised to produce a synthesis gas. Pulverised coal is fed to a reactor, together with pure oxygen, about 25% to 30% of the stoichiometric amount. The reaction takes place at temperatures above 1600 °C, at a pressure of about 27 bar (Dijkstra, 2005).

Nowadays coal gasifiers are large-scale devices, from about 500 MW_{LHV}. However, developments are going on towards biomass gasifiers, on a much smaller scale. These gasifiers are expected to be able to handle coal as well, and can possibly be equipped with CO_2 capture devices. This even opens the possibility to capture CO_2 from biomass.

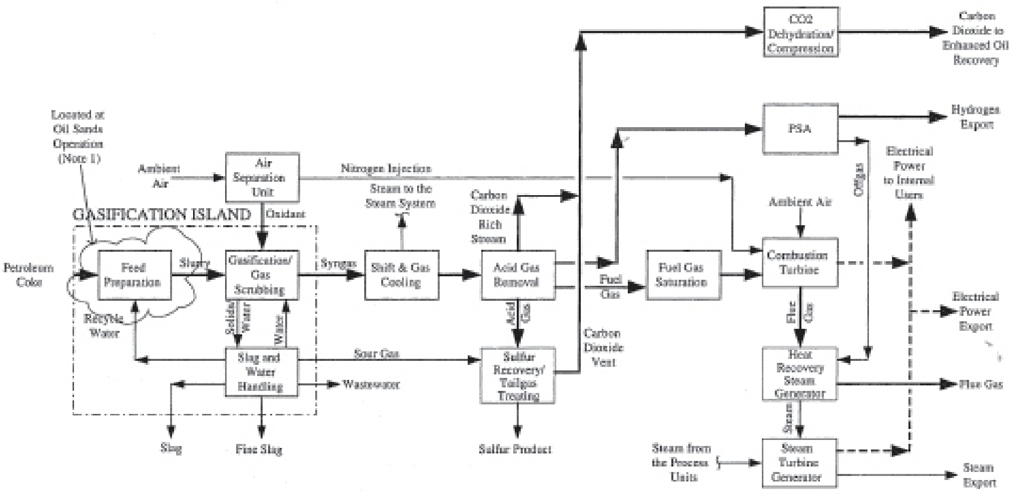


Figure E-1 Example of a gasifier with CO2 capture, in this case petcoke (Holysh, 2005)

Discussion of possible capture technologies

- In principle all the capture options identified can be applied to gasifiers. But this is also depending on the type of energy conversion applied after the gasifier.
- Coal gasifiers can be equipped with high temperature syngas cleaning, followed by one or more shift reactors. Because the coal is gasified with pure oxygen, after hydrogen separation with PSA only steam and CO_2 remain. The steam is easily removed by cooling and water knock-out, leaving a pure CO_2 stream.

Appendix F Process descriptions: processes with CO_2 capture

In this appendix a qualitative description is provided for electricity production processes with CO_2 capture. The focus is on the scale of CO_2 capture, possibilities for downscaling and the effects of CO_2 capture on other emissions. The descriptions are based on the descriptions of the Best Available Technologies without CO_2 capture.

The qualitative description of power generation processes with CO_2 capture includes:

- Block diagram
- Principle of separation
- Practicable lower limits to the scale of the separation technology

The following processes are discussed:

Post combustion

F.1.	MEA Absorber
F.2.	Solid adsorption of CO_2 from flue gas (PSA and TSA)
F.3.	Post combustion membrane absorber

Pre combustion, in combination with a fuel processor

F.4.	Selexol pre-combustion CO_2 capture
F.5.	Pre-combustion solid adsorption of CO_2 from syngas
F.6.	Sorption enhanced reforming
F.7.	Sorption enhanced shift (SE-WGS)
F.8.	$CaO-CaCO_3$ capture
F.9.	GT/CC-NG with oxygen conducting membrane (AZEP)
F.10.	Membrane reformer
F.11.	Shift membrane reactor

Oxyfuel combustion

F.12.	Oxyfuel conversion with CO_2 recycle
F.13.	Membrane oxygen production
F.14.	Matiant Cycle
F.15.	Water Cycle
F.16.	Chemical Looping Combustion
F.17.	SOFC-GT
F.18.	CG SOFC
F.19.	OCM boiler

F.1. MEA absorber

In principle, post combustion CO_2 capture can be applied on all natural gas and coal conversion equipment. Of these, NG and coal fired furnaces (eventually equipped with steam cycles), NG reciprocating engines and gas turbines are the most common.

Block scheme

The block scheme of this option is depicted in Figure F-1. Flue gas from a furnace, gas turbine or reciprocating engine is cooled (for example in an HRSG), cleaned in case of coal firing, and treated in the CO_2 capture unit. In this unit, CO_2 is removed from the flue gas by absorption in a solvent. The solvent used for absorption is monoethanolamine (MEA). CO_2 is recovered from the solvent by heating with low pressure (LP) steam from the flue gas cooler.

Due to the steam use for CO_2 recovery, the steam turbine of an NGCC loses about 1/3 of its original power output. Similarly, the output of high temperature heat of a heat recovery exchanger is diminished by 1/3, and this heat comes available again at low temperature.

Recycling of half the flue gas from the HRSG, shown in Figure F-1 as "optional", can reduce the volume of gas to be treated in the CO_2 absorber by 50% and has the effect of doubling the concentration of CO_2. The overall result is a small reduction of the overall investment, rather than an improvement of the overall CO_2 capture ratio.

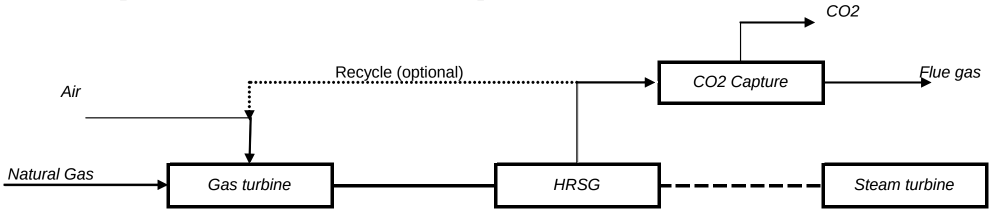


Figure F-1 System configureation for NGCC + CO_2 capture (Dijkstra, 2005)

Characterisation of separation technology

A basic process scheme of the system is shown in Figure F-2. Flue gas from the gas turbine or engine is first cooled to near ambient temperature, before it enters the absorber. In the absorber, a conventional packed bed column, CO_2 in the pre-cooled flue gas bonds with the solvent. CO_2-lean off-gas is discharged from the absorber to the atmosphere. The CO_2-rich solvent is routed to the regenerator (packed column) where CO_2 is recovered from the solvent by the addition of heat (provided by the heat recovery exchanger, or by condensation of LP steam extracted from an HRSG). The lean solvent from the regenerator is cooled in the lean/rich heat exchanger and further cooled in a solvent after cooler before it is used again in the absorber. Some solvent is lost in both the stripper and regenerator. These losses are compensated for by solvent make-up.

Further main equipment items in the system (not shown) are a flue gas fan down stream of the flue gas cooler to overcome the pressure drop in the absorber and solvent pumps to circulate the rich and lean solvent.

The CO_2 concentration in the flue gas of a gas turbine is approximately 4 vol-%. This low concentration (low partial CO_2 pressure) favours a chemical solvent (MEA) instead of a physical solvent to achieve a high capture ratio. CO_2 is chemically bound to the MEA solvent.

The energy consumption for regeneration of the MEA is reported to be 3 to 5 MJ/kg CO_2 (Hamelinck, 2001), with 4 MJ/kg as average. As combustion of natural gas produces 56 kg CO_2 per GJ (Vreuls, 2004), the regeneration energy for capture of 85% of the CO_2 accounts for 19% of the lower heating value of natural gas. The CO_2 emission of coal is 94 kg/GJ, which leads to an energy consumption for regeneration of 32% of the lower heating value, if 85% CO_2 is captured. For MEA a typical regeneration temperature is reported of 139 ^{o}C (Hendriks, 1994). Dependent on heat exchanger design, the heat must be supplied at temperatures between 150 to 170 ^{o}C.

Instead of MEA, which is a primary amine, also secondary or tertiary amines can be used in this process. Secondary and tertiary amines are weaker CO_2 binders, but they can be regenerated at lower temperatures. For dimethanolamine (DEA), a regeneration temperature of 128 ^{o}C is reported (Hendriks, 1994).

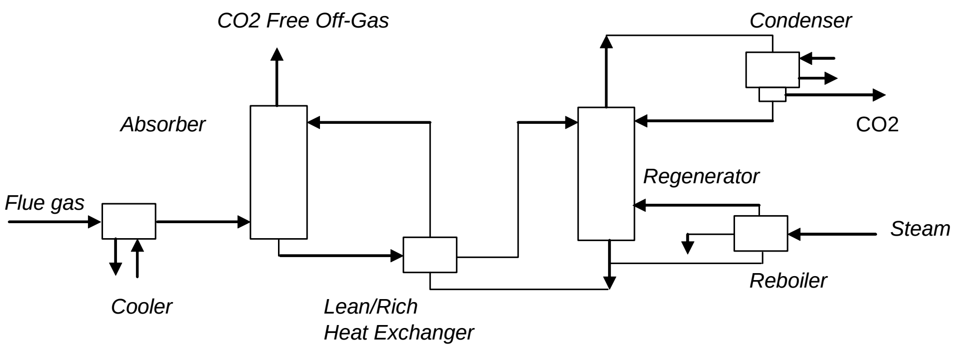


Figure F-2 Working principle for CO2 absorption in solvent

Practicable lower limit to the scale of the separation technology

Although scrubbers can be applied on any scale, the cost of MEA absorbers is scale sensitive. A scale factor of 0.8 is used in an equation for the investment costs (Hamelinck, 2001):

$$I = (I_{42tph,fix} + I_{42tph,\mathrm{var}} \times \frac{c_0}{c}) \times (\frac{m}{m_0})^R \times f$$

with I = investment in M€

$I_{42tph,fix}$ = fixed investment for 42 tonne CO_2/h = 13.2 US$_{1994}$

$I_{42tph,var}$ = CO_2 concentration dependent investment for 42 tonne CO_2/h = 8.8

US$_{1994}$

c_0 = base case CO_2 concentration = 8 %

c = actual CO_2 concentration in %

m_0 = base case CO_2 flow = 42 tonne/h

m = actual CO_2 flow

R = scale factor = 0.8

f = currency inflation factor from US$$_{1994}$ to €$_{2000}$ = 1.22

A CO_2 flow of 42 tonne/h applies to a 100 MW_e combined cycle. The value of 0.8 might even be fairly high. Values in the magnitude of 0.6 are also applied. As a result the MEA absorbers are regarded applicable in medium-scale installations, but less applicable in small installations.

F.2. Post combustion Solid adsorption

Block scheme

The block scheme is similar to that of a CF/GT-NG/GT-CG plant with post combustion amine absorption, see Figure F-3.

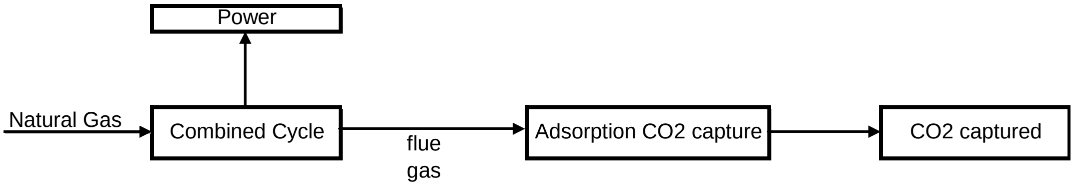


Figure F-3 System configuration for post combustion solid sorption.

Principle of separation

An adsorption process consists of two major steps: adsorption and desorption. Adsorption of CO_2 takes place on a solid material with a high surface area and strong affinity for CO_2, such as zeolites or activated carbon. After the sorbent has been loaded to its capacity, it is regenerated, where desorption takes place producing a concentrated CO_2 stream. The sorbent can be regenerated by increasing the temperature; temperature swing adsorption (TSA), or by decreasing the pressure; pressure swing adsorption (PSA). The stronger the affinity of the sorbent, the deeper the flue gas can be decarbonised, but also the more difficult it is to desorb the adsorbed material and the more energy is consumed in regenerating the adsorbent for reuse in the next cycle. An adsorption unit comprises a series of vessels containing sorbent in packed beds through which the gas flows. The process operates on a repeated cycle of adsorption and regeneration.

Practicable lower limit to the scale of the separation technology

This limit is very low. Nowadays already, Netherlands based fuel processor manufacturer HyGear B.V. (former Hexion) offers hydrogen production facilities down to 5 Nm^3 per hour, which are equipped with pressure swing adsorption (HyGear b.v., 2005; Adamson, Jollie, 2005).

F.3. Post combustion membrane absorption

Block scheme

The block scheme is identical to that of conventional post combustion NGCC Figure F-1 However, instead of a packed column a membrane contactor is used. In the membrane contactor, there is no direct contact between absorption liquid and gas phase. The absorption liquid flows trough polymer hollow fibre membranes with a high mass transfer per volume.

Principle of separation

The principle of separation is identical to the combined cycle with MEA absorption. The CO_2 is chemically bonded to the liquid absorbent. Regeneration takes place similar to MEA absorption.

Practicable lower limit to the scale of the separation technology

As membranes are considered as highly modular, a lower limit to the scale might be absent.

F.4. Selexol pre combustion decarbonisation

Block scheme

The block scheme of this option is depicted in Figure F-4. In this process, natural gas is converted to a synthesis gas, and subsequently shift converted to a mixture of hydrogen and CO_2. The CO_2 is captured in a physical solvent. The remaining hydrogen rich fuel gas is burned in a gas turbine or reciprocating engine.

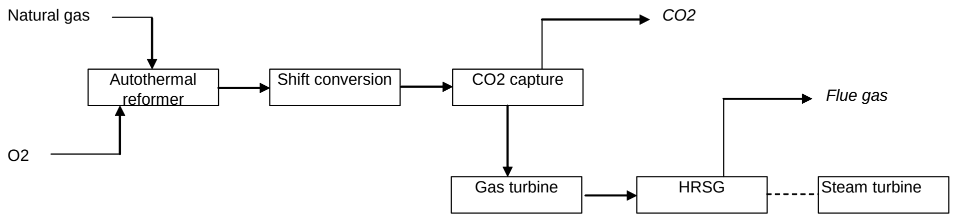


Figure F-4 Combined cycle with Selexol pre combustion decarbonisation

In case of coal use, the block scheme of this option is similar, and depicted in Figure F-5. The example is an integrated coal gasification combined cycle (IGCC) with pre-combustion CO_2 removal by means of shift conversion and subsequent CO_2 removal. The resulting hydrogen rich fuel is burned in the gas turbine.

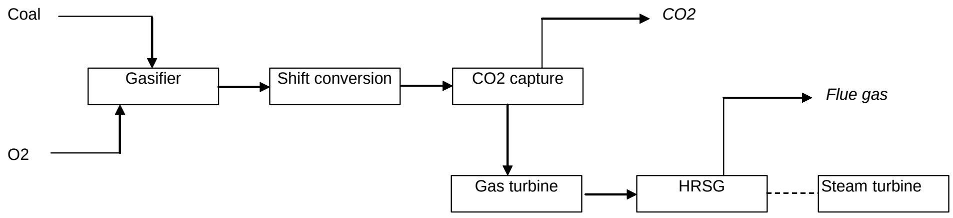


Figure F-5 IGCC + CO_2 capture

Principle of separation

A basic process scheme is shown in Figure F-6. In the autothermal reformer natural gas is converted into syngas (consisting of mainly CO and H_2) using a catalyst. Steam (from the HRSG) and oxygen (from an air separation unit or enriched air) are used to enable the conversion. The hot syngas supplies heat to the gas feed to the reformer, before it enters the two stage shift conversion section. In the two stage shift reactor CO in the syngas is converted with steam into a mixture of H_2 and CO_2. Steam for the shift reaction is added in the autothermal reformer. As the shift reaction is exothermic, heat can be recovered. The heat recovering system is integrated into the steam and syngas cycle. The H_2/CO_2 mixture then enters an absorption process with a physical solvent (Selexol) where CO_2 is removed from the gas mixture leaving a hydrogen rich fuel gas. The CO_2 is recovered from the solvent by a two stage pressure reduction step and a final regeneration with steam. The H_2 rich fuel gas is mixed with N_2 from the air separation unit and proceeds through the saturator where water is added to enhance gas turbine performance and reduce NO_x emissions.

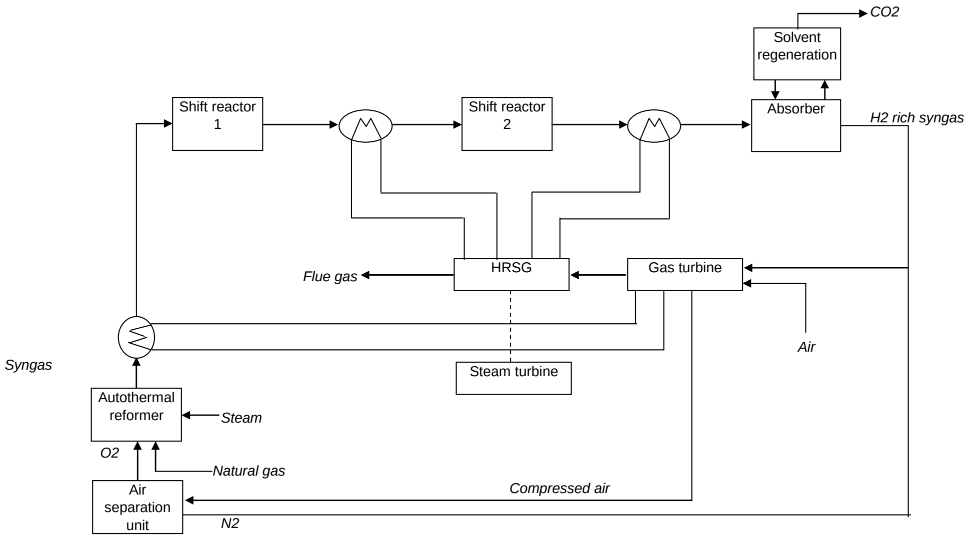


Figure F-6 Working principle of NGCC with Selexol pre-combustion decarbonisation

Oxygen for the gasifier is produced by the air separation unit (ASU). Air side integration is applied between the gas turbine and the ASU to reduce cost. The nitrogen that is produced by the ASU is added to the gas turbine.

Practicable lower limit to the scale of the separation technology

At small scale air separation is too expensive, which poses a lower limit to the scale. This problem might be surpassed by air-blown reforming or gasification, but that can result in too lean fuel to the gas turbine or engine.

F.5. Pre-combustion solid CO_2 sorption

Block scheme

The block scheme for this option is similar to that of liquid pre-combustion CO_2 sorption, Figure F-6. Natural gas is fed to an autothermal reformer together with steam and oxygen from an ASU. In the reactor autothermal reforming takes place. The product gas is then cooled and sent to a water gas shift reactor (WGS). Here the stream is further converted to CO_2 and H_2, followed by adsorption of CO_2 to a solid sorbent. The energy content of the natural gas is by this means transferred to the hydrogen. The hydrogen is sent to a conventional option for power generation. This stream will also contain some unconverted fuel.

If the sorbent is loaded with CO_2 the sorbent is switched to regeneration mode. In this mode a purge steam flow is used to desorb CO_2 from the sorbent. The result from this mode is a mixed

CO_2/H_2O stream. After water knock-out, drying and compression this CO_2 stream is ready for storage.

Principle of separation

An adsorption process consists of two major steps: adsorption and desorption. Adsorption of CO_2 takes place on a solid material with a high surface area and strong affinity for CO_2, such as zeolites or activated carbon. After the sorbent has been loaded to its capacity, it is regenerated, where desorption takes place producing a concentrated CO_2 stream. The sorbent can be regenerated by increasing the temperature, temperature swing adsorption (TSA), or by decreasing the pressure, pressure swing adsorption (PSA). The stronger the affinity of the sorbent, the deeper the fuel gas can be decarbonised, but also the more difficult it is to desorb the adsorbed material and the more energy is consumed in regenerating the adsorbent for reuse in the next cycle. An adsorption unit comprises a series of vessels containing sorbent in packed beds through which the gas flows. The process operates on a repeated cycle of adsorption and regeneration.

Practicable lower limit to the scale of the separation technology

Similar to post combustion solid CO_2 adsorption, the scale limit of pre combustion solid sorption is very low (Appendix F.5).

F.6/7 Pre-combustion CO_2 sorption enhanced reforming and Pre-combustion CO2 sorption enhanced shift

Block scheme

The block scheme for these options is depicted in Figure F-7. Natural gas is fed to an autothermal reformer together with steam and oxygen from an ASU. In the reactor autothermal reforming takes place. The product gas is then cooled and sent to a sorption enhanced water gas shift reactor (SE-WGS). Here the stream is further converted to CO_2 and H_2 parallel to adsorption of CO_2 to a solid sorbent. The energy content of the natural gas is by this means transferred to the hydrogen. The hydrogen is sent to a conventional combined cycle for power generation. This stream will also contain some unconverted fuel.

If the sorbent is loaded with CO_2 the sorbent enhanced reactor is switched to regeneration mode. In this mode a purge steam flow is used to desorb CO_2 from the sorbent. The result from this mode is a mixed CO_2/H_2O stream. After water knock-out, drying and compression this CO_2 stream is ready for storage.

Optionally the ATR and SE-WGS can be integrated. In this case, the autothermal reforming reactions as well as the shift reaction are carried out in the presence of a CO_2 sorbent. Integration of the ASU with the gas turbine may be applied but is not required. Another process variant is autothermal reforming with air instead of oxygen. All these options will not be treated further since the CO_2 capture ratio will not be fundamentally different.

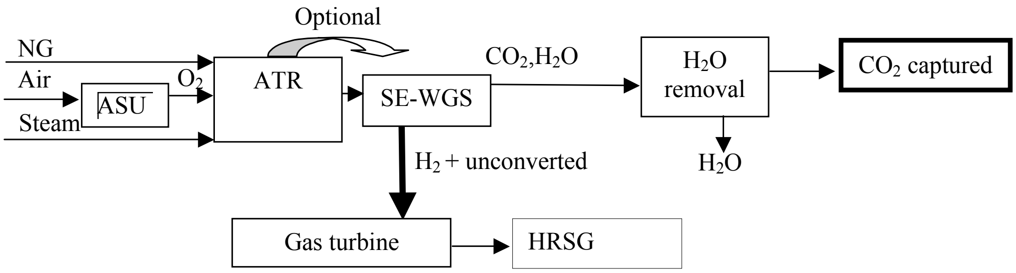


Figure F-7 System configuration for CC with SE-WGS reactor

Principle of separation
The key element of this option is the sorption enhanced reactor. A schematic depiction of a sorption enhanced reactor is shown in Figure F-8. Here the water gas shift reaction takes place parallel to absorption of the CO_2. The chemical equilibrium of the reaction is pulled to the right side as a result of the absorption of CO_2.

$$CO + H_2O \rightarrow CO_2\ (abs) + H_2$$

The conversion is therefore linked to the amount of CO_2 absorbed. By means of the combination of ATR and shift, the energy content of the natural gas is transferred to hydrogen. The by-product is mainly CO. The hydrogen, with possibly some unconverted fuel and excess H_2O leaves the reactor and goes to the combined cycle.

If the sorbent is loaded with CO_2 the reformer is switched to regeneration mode. Then the CO_2 is released by purging with steam, lowering the pressure, by increasing the temperature or a combination thereof. The result is a mixed CO_2/H_2O stream. After water knock-out, drying and compression this CO_2 stream is ready for storage. By having multiple reactors operating in alternating mode, a continuous process can be achieved.

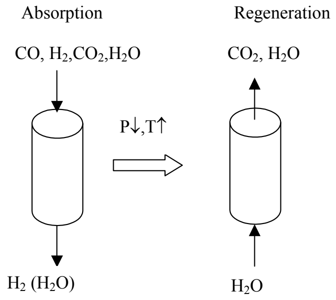


Figure F-8 Working principle of a sorption-enhanced reformer or shift reactor

Important reactor performance indicators are the CO conversion and the amount of CO_2 captured. The most important factor for the reactor performance is the chemical affinity of the sorbent with CO_2. A high affinity will mean a high CO conversion at relatively low temperatures (both will result in a lower efficiency penalty and a higher carbon capture ratio).

However, regeneration will become more difficult and will require lower pressures, higher temperature or higher amounts of purge steam. All of these factors will increase the efficiency penalty and lower the carbon capture ratio.

Practicable lower limit to the scale of the separation technology
The equipment of a sorption enhanced reactor is very similar to post- and pre-combustion solid sorption. As already concluded there, the scale limit is very low (Appendix F.2).

F.8. $CaO-CaCO_3$ Capture

Block scheme
The block scheme is depicted in Figure F-9. There is little information on the exact lay-out of the process scheme, so the description is limited to the fundamentals of the process. Coal is combusted with air, and the flue gasses are led to the calciner in which at moderate temperature (below 750°C) CO_2 is absorbed with the reaction

$$CaO + CO_2 \rightarrow CaCO_3$$

The $CaCO_3$ is transported to a calciner in which the reverse reaction takes place at high temperature (above 1050°C) producing a hot CO_2 stream. After use of the high temperature heat this CO_2 is available for storage. The heat for the calciner originates from the combustion. This heat flux is significant as a result of the large heat of reaction. Constant replacement of CaO is required because of the rapid deactivation of the sorbent.

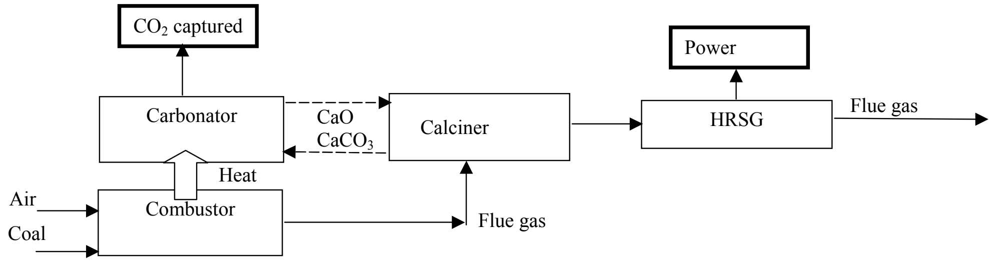


Figure F-9 Block scheme of PC boiler with CaO absorption

Principle of separation
The principle of separation is that of a gas-solid chemical reaction, as can be seen in Figure F-11. Here the example is a CaO enhanced reformer, but the NG/H_2O feed can be replaced with flue gas from a combustor. The level of CO_2 in the flue gas is limited by the chemical equilibrium composition at the given temperature and pressure. Furthermore the rate of reaction must be sufficiently fast to achieve this chemical equilibrium (Dijkstra, 2005).

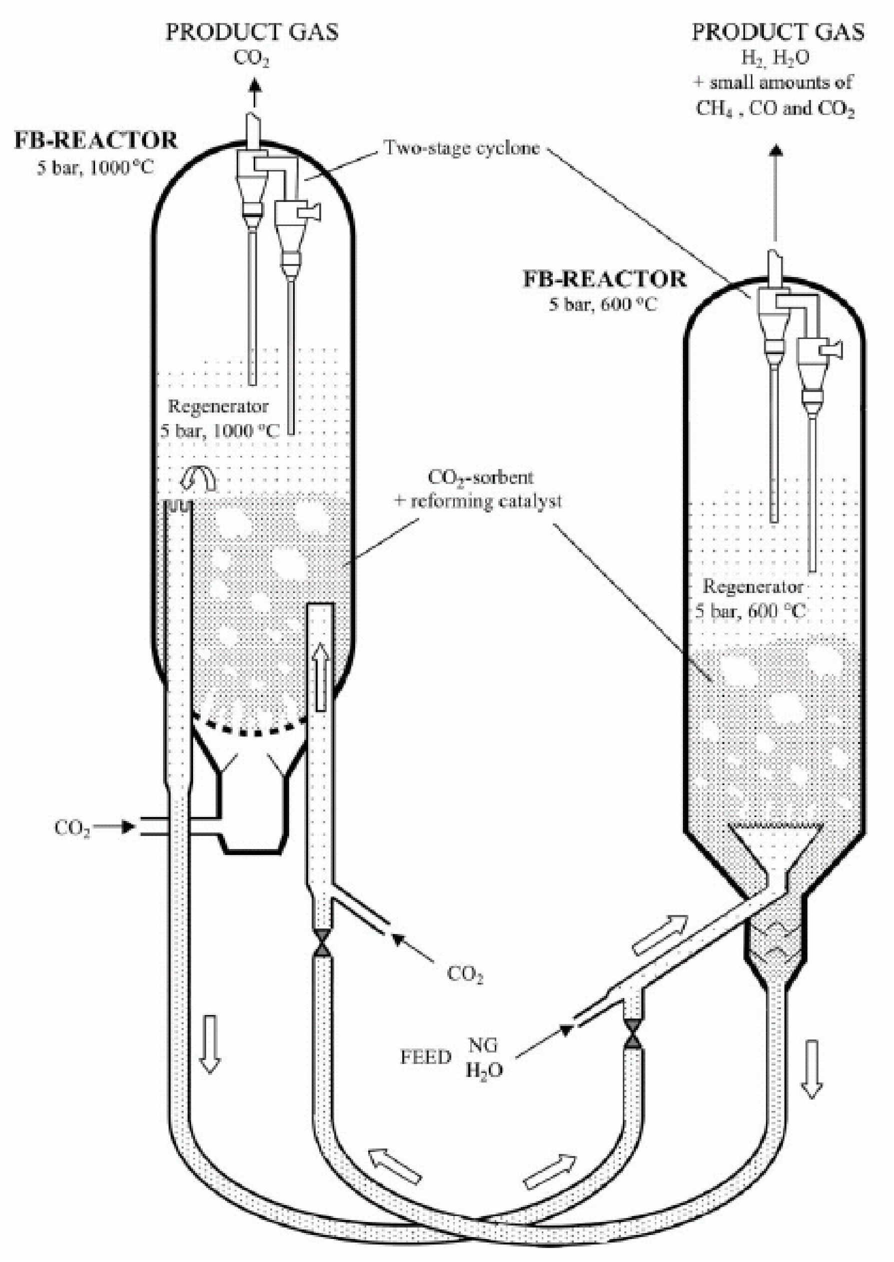


Figure F-10 CaO reforming process (Meyer, Aaberg& Andresen, 2005)

Practicable lower limit to the scale of the separation technology

Since there is little technical information about the process lower limits are hard to predict. It is however not expected that there would be more limitations than heat loss, which might mean that the equipment can become sufficiently small.

F.9. GT/CC-NG with oxygen conducting membrane (AZEP)

Block scheme

The block scheme is depicted in Figure F-11. Air is compressed by a gas turbine compressor. Then the air is heated further and oxygen withdrawn in a membrane module. The membrane module is equipped with an oxygen selective high-temperature membrane working at 800-900° C. The hot air exhaust is then expanded in the gas turbine expander. The off gasses from the expander pass through a HRSG.

Natural gas is added to the permeate of the membrane reactor and then fed to a combustor in which the natural gas is combusted. A stream consisting of mainly CO_2 and H_2O results. A large part of this stream is recycled to the permeate side of the membrane module, thus creating a sweep stream and transferring heat to the incoming air. The remaining off-gasses are used for heat exchange and steam production. After dewatering a concentrated CO_2 stream results for storage.

Additional firing of natural gas before the GT expander inlet gas may be applied to obtain a significant increase in system efficiency (3.8%-points at 15% additional firing). However, the resulting CO_2 is not captured.

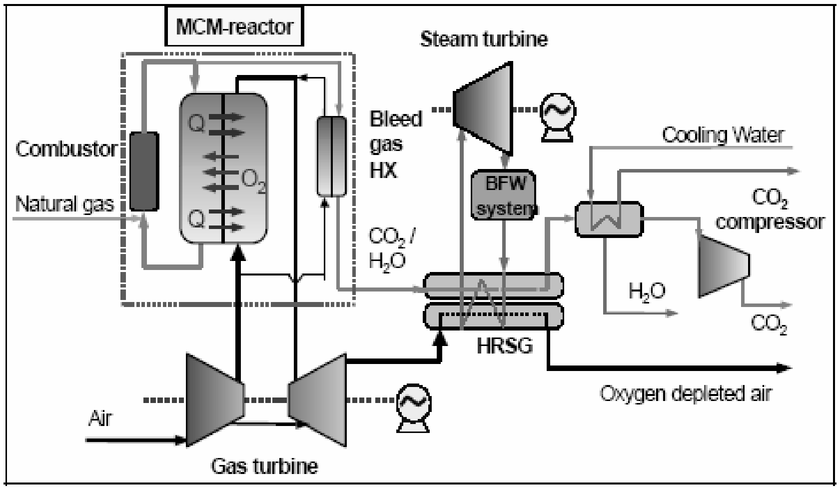


Figure F-11 System configuration for AZEP concept

Principle of separation

The principle of separation is the oxyfuel combustion principle. The natural gas is combusted in an $O_2/CO_2/H_2O$ atmosphere. After water-knock-out a concentrated CO_2 stream results. The membrane module has to perform two tasks: separating oxygen from the air and heat transfer from the permeate side to the feed side. The special combustor is located outside the membrane module, but is in integral part of the concept.

Practicable lower limit to the scale of the separation technology

As membranes are considered to be highly modular, a lower limit to the scale might be virtually absent.

F.10/11. Pre-combustion membrane reformer and Pre-combustion membrane shift reactor

Block scheme

The block scheme for this option is depicted in Figure F-12. The natural gas is fed to a membrane reformer together with steam. In the membrane reformer the steam reforming reaction takes place parallel to separation of hydrogen. The energy content of the natural gas is by this means transferred to the hydrogen. The hydrogen is sent to a conventional combined cycle for power generation. The other product stream of the membrane reformer is a CO_2 stream diluted with steam (of which the bulk content can be removed by condensation and knock-out) and unconverted CH_4, CO and H_2. In the clean-up section the CH_4 and unconverted fuel are separated (e.g. by cryogenic distillation during CO_2 compression or by Selexol absorption) or are converted by oxidation. The result is a concentrated CO_2 stream available for storage. The membrane reformer is fired with natural gas and/or a part of the hydrogen product to supply the heat for the steam reforming reaction.

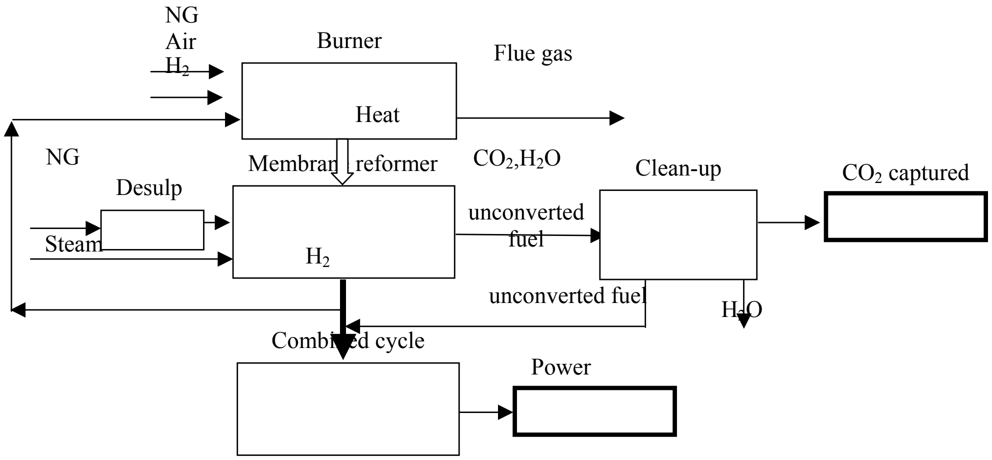


Figure F-12 System configuration for NGCC with membrane reformer

Principle of separation

The key element of this option is the membrane reformer. A schematic depiction of a membrane reformer is shown in Figure F-13. Natural gas and steam are fed to the membrane reformer. The natural gas originates from an adiabatic pre-reformer (not depicted) to convert the higher hydrocarbons and increase the H_2 content of the feed. At the feed side of the membrane reformer, the steam reforming reaction takes place. Here the energy content of the natural gas is transferred to hydrogen. The hydrogen permeates through a hydrogen selective membrane. The retentate (exit at the feed side) is CO_2, diluted with H_2O and unconverted CH_4, CO and H_2.
At the permeate side a sweep stream of steam (from the combined cycle) is introduced to lower the partial pressure of hydrogen. The permeate stream consists of a mixture of hydrogen and steam.
The removal of hydrogen parallel to reaction shifts the steam reforming equilibrium reaction to the product side. This allows operation at lower temperatures compared to conventional steam reforming with a high CH_4 conversion. The conversion at the feed side is therefore linked to the amount of hydrogen permeating.

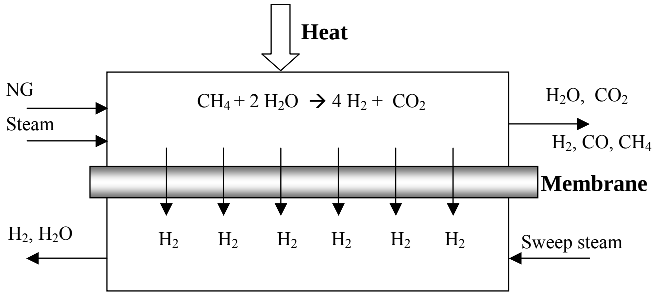


Figure F-13 Working principle of a membrane reformer

The important factors for the membrane reactor performance are:
1. The membrane surface area. Increasing the membrane surface area increases the hydrogen recovery (relative amount of hydrogen permeating) and the CH_4 conversion. It is not possible to achieve 100% conversion. This would require an infinite membrane surface area.
2. Membrane permeance. This is the amount of hydrogen permeating for a given surface area and pressure differential. A high membrane permeance has a similar effect as a large membrane surface area. It increases the hydrogen recovery and the CH_4 conversion.
3. Membrane selectivity. Depending on the membrane type, other components besides hydrogen could also permeate.
4. The sweep flow. Increasing the sweep flow increases the hydrogen recovery and thus the conversion in the membrane reactor for a given membrane surface area.
5. Permeate side hydrogen partial pressure. Decreasing the permeate side pressure increases the hydrogen recovery and thus the conversion for a given membrane surface area. The feed side pressure is fixed at the available natural gas pressure.

It is not possible to have full 100% conversion in the membrane reactor. In order to have full conversion, the membrane surface area would need to be infinitely large. In practice the membrane surface area will be determined by an economic optimum between membrane investments and system performance.

Practicable lower limit to the scale of the separation technology
As membranes are considered to be highly modular, a lower limit to the scale might be virtually absent.

F.12/13. Oxyfuel conversion with CO_2 recycle and Membrane oxyfuel conversion with CO_2 recycle

Block scheme
The block schemes of these options are depicted in Figure F-14 for natural gas, and in Figure F-15 for coal. Natural gas is combusted with pure oxygen under pressurized conditions, coal is combusted under atmospheric conditions. To limit the high temperatures resulting from combustion with pure oxygen, CO_2 rich flue gas is recycled to the compressor, furnace or engine

inlet and mixed with the oxygen prior to combustion. Oxygen is supplied by a cryogenic air separation unit (or indirectly from liquid storage), or by a membrane air separator. Flue gas (mainly CO_2 and H_2O) is cooled down and water is separated by knock-out.

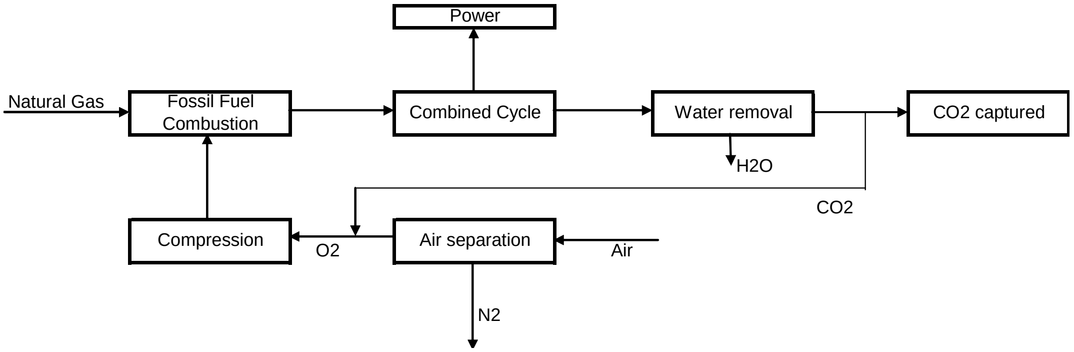


Figure F-14 System configuration for CC Oxyfuel conversion with CO_2 recycle

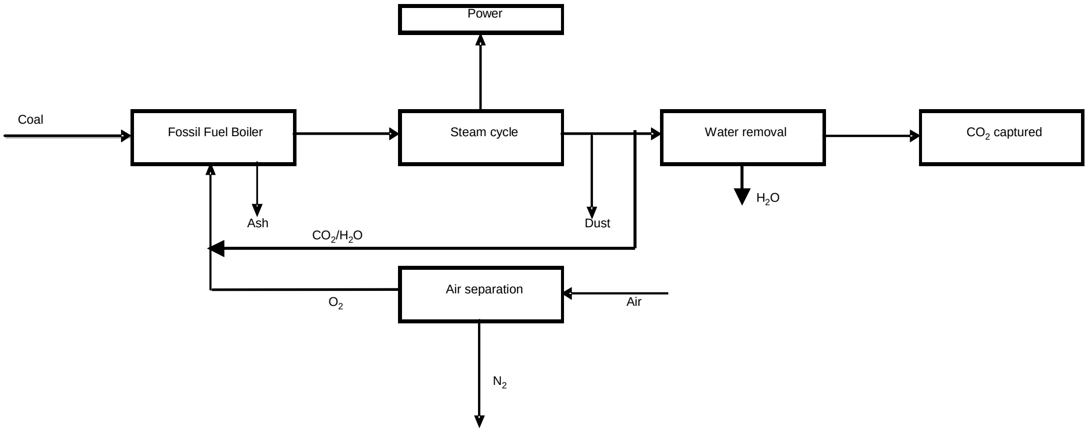


Figure F-15 System configuration for Oxyfuel PC conversion

Principle of separation

The key element of this option is combustion with pure (±95%) oxygen, resulting in a flue gas, which mainly consists of CO_2 and H_2O. The water can relatively easily be removed from the flue gas by cooling and knock-out.

Oxygen for the combustion is produced in an air separation unit (ASU). In a conventional cryogenic air separation unit the following process steps are taken:

- air compression to ±6 bar
- removal of CO_2, H_2O and hydrocarbons with temperature swing adsorption (TSA)
- cooling by heat integration with product flows and expansion
- cryogenic distillation

See also Figure F-16.

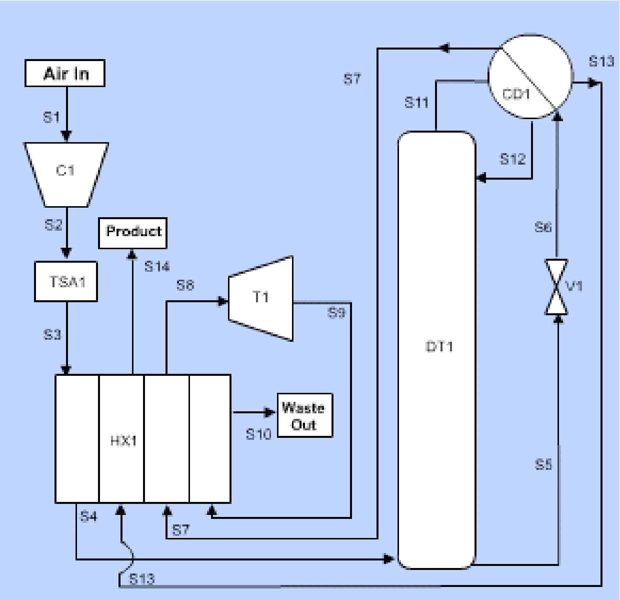


Figure F-16 Cryogenic Air separation unit

Alternatively, air separation can be done by oxygen selective membranes or pressure swing adsorption (PSA).

Direct combustion of natural gas with pure oxygen in a gas turbine or engine is not feasible because of too high combustion temperature. Therefore the O_2 is mixed with a large CO_2 recycle stream. Because CO_2 has a different expansion coefficient (κ-value) to air, the compressor and the turbine have to be redesigned. Gas engines probably need a modification of the compression ratio

The purity of oxygen is an important factor for oxyfuel conversion. The remaining nitrogen ends up in the CO_2 stream and gives an efficiency penalty when the CO_2 is compressed for storage. Reduction of this efficiency penalty is a trade off with the higher energy requirement for air separation.

Practicable lower limit to the scale of the separation technology

Depends on minimum scale for air separation. On-site cryogenic air separation is only applicable on (very) large scale. Membrane oxygen production is expected to be less scale sensitive, but it is not known if the technology is already available. If liquid oxygen from large-scale separation is delivered on-site in tanks, there is actually no lower scale limit.

F.14. Oxyfuel boiler without CO_2 recycle

Block scheme

The block scheme of this option is depicted in Figure F-17. Here the fuel is combusted with pure oxygen. To limit the high temperatures resulting from combustion with pure oxygen, water is recycled to the compressor, furnace or engine inlet and mixed with the oxygen prior to combustion. Oxygen is supplied by a cryogenic air separation unit (or indirectly from liquid storage), or by a membrane air separator. Flue gas (mainly CO_2 and H_2O) is cooled down and water is separated by knock-out (Simmonds, Walker, 2005).

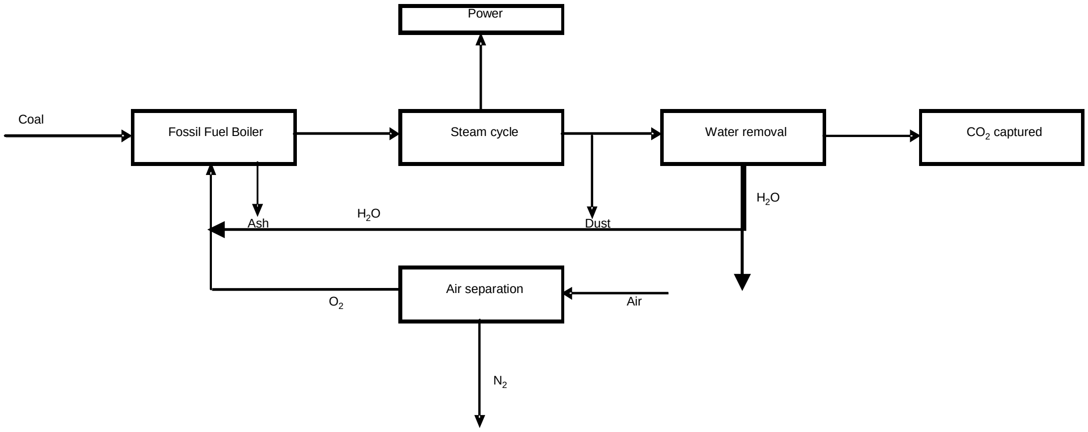


Figure F-17 System configuration for Oxyfuel PC conversion

Principle of separation
The key element of this option is combustion with pure (±95%) oxygen, resulting in a flue gas, which mainly consists of CO_2 and H_2O. The water can relatively easily be removed from the flue gas by cooling and knock-out.

The purity of oxygen is an important factor for oxyfuel conversion. The remaining nitrogen ends up in the CO_2 stream and gives an efficiency penalty when the CO_2 is compressed for storage. Reduction of this efficiency penalty is a trade off with the higher energy requirement for air separation.

Practicable lower limit to the scale of the separation technology
The minimum scale of the energy conversion depends on the minimum scale for air separation. On-site cryogenic air separation is only applicable on (very) large scale, as discussed in paragraph 4.2. Membrane oxygen production is expected to be less scale sensitive, but it is not known if the technology is already available. If liquid oxygen from large-scale separation is delivered on-site in tanks, there is actually no lower scale limit.

F.15. Matiant cycle

Block scheme
The block scheme of the Matiant cycle is depicted in Figure F-18. The Matiant cycle is a more sophisticated variation of the combined cycle with oxy-fuel combustion. Natural gas is combusted with oxygen in two stages in a gas turbine producing a flue gas, which consists mainly of CO_2 and H_2O. CO_2 and H_2O are recycled thus forming the working fluid of the gas turbine. The Matiant cycle has the potential of higher overall thermal efficiency than oxy-fuel combustion in a combined cycle.

Gas turbines with CO_2/H_2O as working fluid would be different to conventional gas turbines, because of different expansion coefficients, and are currently not available on the market. Potential application of the Matiant cycle is therefore dependent on gas turbine development.

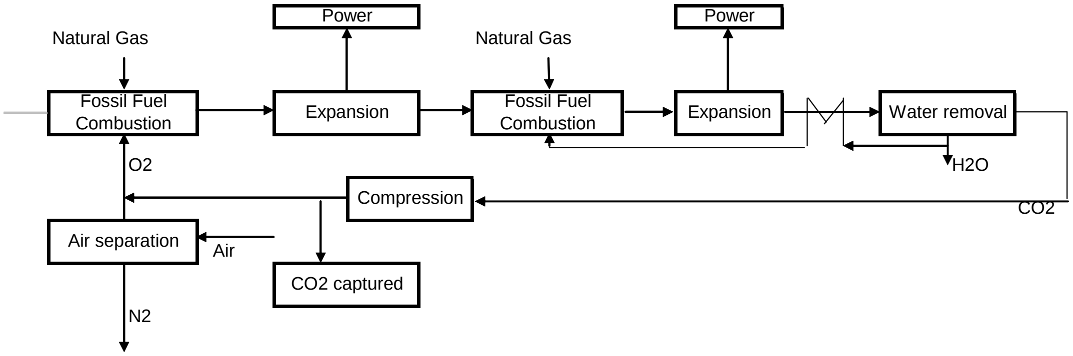


Figure F-18 System configuration for Matiant cycle

Principle of separation

The principle of separation is based on combustion with pure oxygen giving a CO_2/H_2O mixture as flue gas.

Practicable lower limit to the scale of the separation technology

Depends on the scale for air separation, as discussed in F.12.

F.16. Water cycle

Block scheme

The block scheme for this option is depicted in Figure F-19. The water cycle is an oxyfuel cycle using natural gas as a fuel. In a specially designed combustor the temperature is lowered by water injection. The resulting high-temperature gas is used in an expander. A steam topping cycle may be present (not depicted). Optionally there is a CO_2 cycle. After water-knock-out a pure CO_2 stream results.

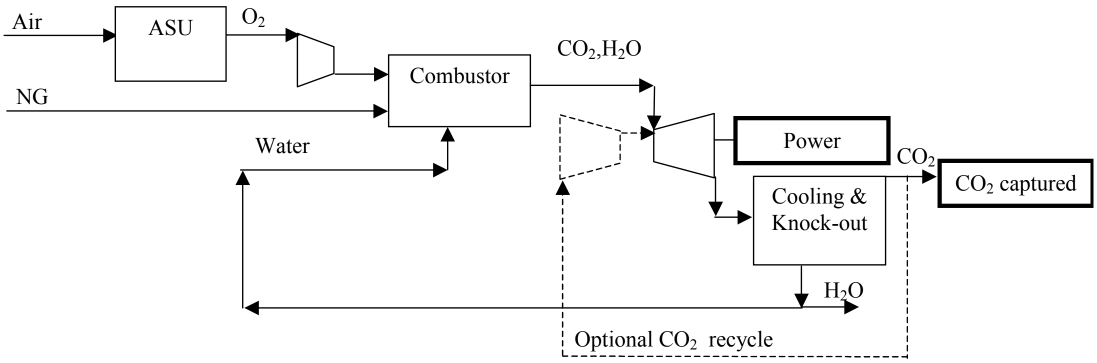


Figure F-19 System configuration for the water cycle

Principle of separation

The principle of separation is based on combustion with pure oxygen (in stoichiometric quantity) giving a CO_2/H_2O mixture as flue gas. Gas expanders with CO_2/H_2O as the working fluid would be different to expanders of conventional gas turbines (because of different expansion coefficient) and are currently not available on the market. Potential application of the Water Cycle is therefore dependent on gas expander development.

Practicable lower limit to the scale of the separation technology

Depends on the scale for air separation, as discussed in F.12.

F.17. GT/CC with chemical looping combustion

Block scheme

The block scheme is depicted in Figure F-20. The natural gas (fuel) is fed to a fluidised bed reactor where it reacts with a metal oxide to CO_2 and water.

$$CH_4 + 4\,MeO \rightarrow 4Me + CO_2 + H_2O$$

This reaction is moderately endothermic. Several metal oxides are being considered such as iron or nickel.

The resulting stream contains only CO_2 and water. After water removal a CO_2 stream is available for storage. The metal oxide is transported pneumatically to a second reactor, placed inside a gas turbine cycle. The metal is oxidized here to a metal oxide.

$$2Me + O_2 \rightarrow 2MeO$$

This reaction is highly exothermic. The resulting hot air flow drives the gas turbine and is then used to raise steam in a HRSG.

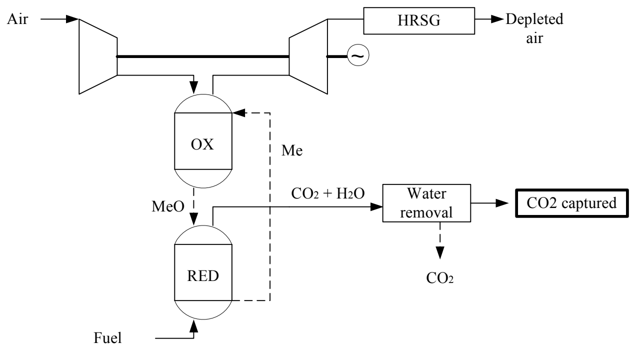


Figure F-20 System configuration for CC with chemical looping combustion

Principle of separation

The principle of separation is similar to an oxyfuel combustion principle. The metal oxide is used as a carrier for the oxygen. Dilution with nitrogen from air is prevented. Since the metal and metal oxide are in solid phase these can be easily separated from the gaseous phase consisting of a CO_2/H_2O mixture. Water can be removed by conventional means, leaving a pure CO_2 stream.

Practicable lower limit to the scale of the separation technology

As the process and the reactors are quite similar to those of the CaO-$CaCO_3$ cycle, the same assumptions may be applied. As discussed in F.8, the lower limit to the scale may be well under the desired size (1 to 100 MW_{LHV} fuel input).

F.18. SOFC-GT with afterburner CO_2 capture

Block scheme

The block scheme is depicted in Figure F-21. Natural gas is pre-heated, mixed with steam and fed to the SOFC anode. Air is pre-heated and fed to the cathode side of the SOFC. In the SOFC the fuel is converted into electricity. The anode and cathode streams remain separated. Oxygen is transferred through the SOFC membrane. The fuel utilisation in the SOFC has a maximum of about 85%. Larger amounts of fuel converted could damage the SOFC as a result of locally occurring oxidizing conditions at the anode side. The resulting anode off-gas consists of CO_2 and H_2O with unconverted CO and H_2. All CH_4 has been converted into CO an H_2. The unconverted CO and H_2 are converted in an afterburner so a CO_2/H_2O stream results. After water-knock-out a concentrated CO_2 stream results for storage.

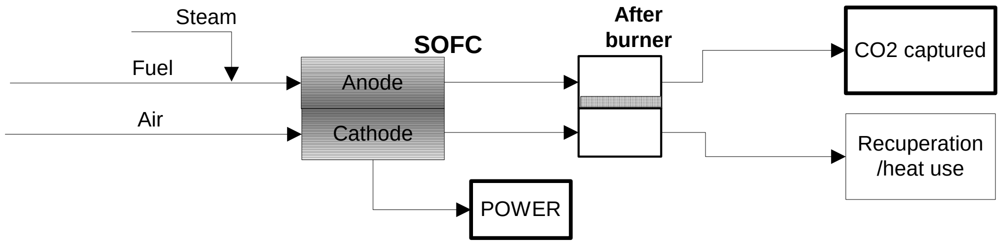


Figure F-21 SOFC with CO_2 capture

For the afterburner two options exist, see Figure F-22. The first type is the OCM afterburner, which uses an oxygen conducting membrane. Oxygen from the air permeates through the membrane and then oxidizes the unconverted fuel (CO_2 and H_2). The resulting stream is a CO_2/H_2O mixture.
The second type is the shift membrane burner. The CO is converted to CO_2 and H_2 catalytically, by means of the water gas shift reaction. The resulting H_2 permeates through a hydrogen selective membrane. After permeation, the hydrogen is combusted using oxygen from the air on the other side of the membrane.

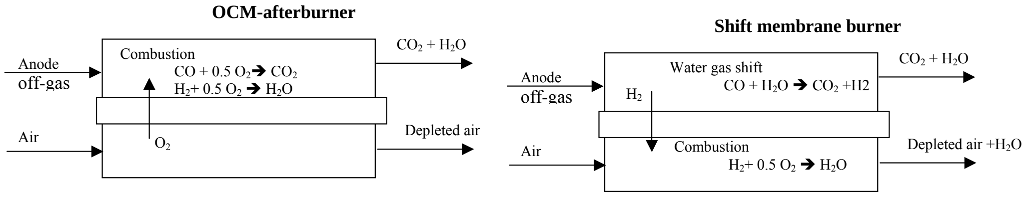


Figure F-22 Working principle of two afterburner types

The actual system may include integration with a gas turbine. In this case the air originates from a gas turbine compressor, and the cathode exhaust (depleted) air is fed to a gas turbine expander. As with the AZEP system, additional firing may be applied in an additional combustion chamber before the expander. Here natural gas (or anode off-gas) is combusted to raise the turbine inlet temperature and thus increase the system efficiency. The CO_2 resulting from this combustion is not captured.

Principle of separation

The principle of separation is the oxyfuel combustion principle. The natural gas is converted without dilution with N_2 in the SOFC. The resulting stream is a CO_2/H_2O stream. After water-knock-out a concentrated CO_2 stream results.

Practicable lower limit to the scale of the separation technology

Solid oxide fuel cells, and the types of afterburners described here, consist of oxygen conducting membranes. As stated in F.9, membranes are considered to be highly modular, a lower limit to the scale might be virtually absent. Indicative is the fact that in several places in the world SOFC micro CHP systems are developed in the range of 1 to 5 kW_e, e.g. Sulzer Hexis in Switzerland, Fuel Cell Technology in Canada and Australia based CFCL.

F.19. Boiler with integrated OCM

Block scheme

The block scheme is depicted in Figure F-23. Fuel and air remain separated through an oxygen conducting membrane (OCM).

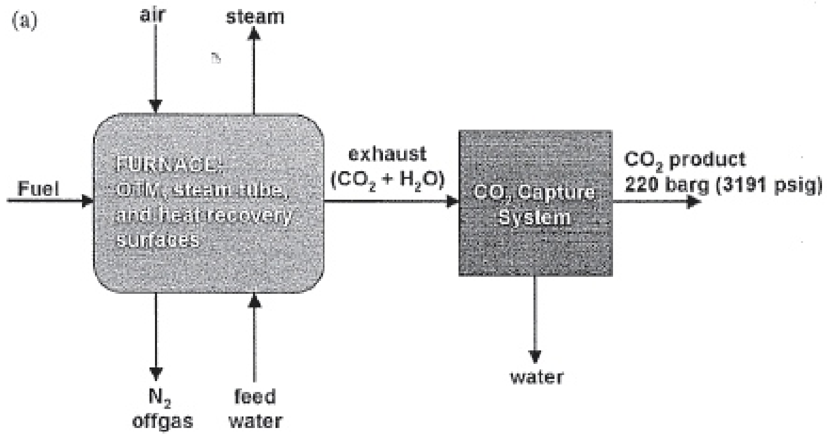


Figure F-23 Conceptual layout of a boiler system with CO_2 capture using a oxyfuel combustion via OCM (Switzer, 2005).

Principle of separation

See AZEP, Appendix F.9.

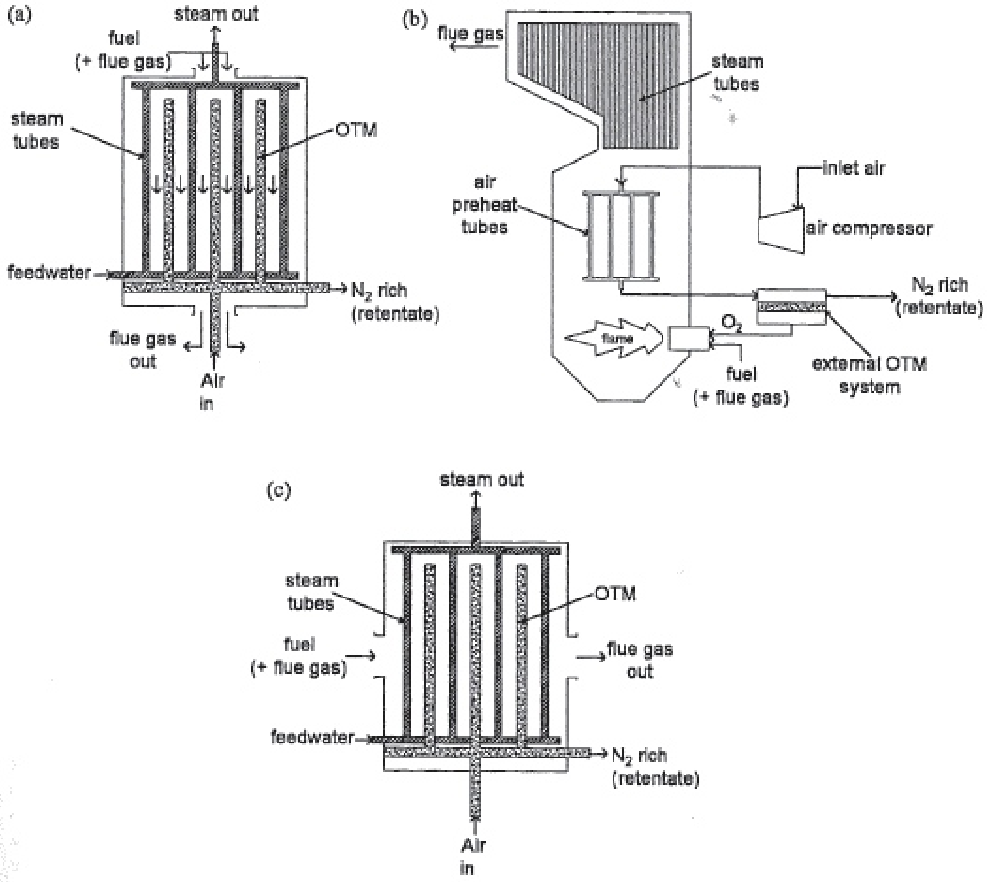


Figure F-24 The OCM boiler designs: (a) co/counter-flow design, (b)external membrane design, and the (c) cross-flow design (Switzer, 2005)

Practicable lower limit to the scale of the separation technology

As stated in F.9, membranes are considered to be highly modular, a lower limit to the scale might be virtually absent.

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