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# Separation and Compression of $CO_2$ in an $O_2/CO_2$-fired Power Plant

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Thesis for the Degree of Master of Science

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# Abstract

The objective of this study is to suggest a process scheme for the treatment and compression of the flue gas from an $O_2/CO_2$-fired power plant, to be used for carbon dioxide sequestration for large scale power production. $O_2/CO_2$ combustion involves burning the fuel with pure oxygen, instead of air, in an atmosphere of recycled flue gas. $O_2/CO_2$ combustion can be designed to be a near zero emission concept. The carbon dioxide and sulphur dioxide are eliminated, except for some leak flows. Low emissions of nitrous gases are another advantage.

A 2x933 MW lignite-fired power plant, located in Lippendorf, Germany, has been selected as a reference plant to ensure that the selected gas treatment scheme can be implemented on such a large scale. The plant data was provided by the plant owner (VEAG).

The flue gas treatment scheme suggested involves units for dehydration, non-condensable gas separation and pressure increase. The sulphur dioxide content in the flue gas is treated and deposited together with the carbon dioxide, hence conventional desulphurising is not necessary. To meet the pipeline transport requirements, regarding hydrate formation and corrosion, the dehydration is carried out both in traditional flue gas condensers and in an active glycol dehydration unit. After the dehydration the flue gas is compressed to 58 bar and cooled to 15°C, which implies that carbon dioxide and sulphur dioxide are transformed into a liquid state, while the non-condensable gases are still in their gaseous state and easily separated. Finally, before transportation, the pressure of the carbon dioxide is increased to 80 to 100 bar in a high pressure pump.

The total investment cost for the $O_2/CO_2$ plant is slightly lower than for the original reference plant, mainly because of the removal of the desulphurising unit. The internal power consumption is, however, increased and consequently there is a decrease in the electrical efficiency. The emissions are significantly lower compared with the original plant; carbon dioxide and sulphur dioxide are limited to leak flows (1,470 tonnes/h $CO_2$ and 19.4 tonnes/h $SO_2$ are removed) and only 0.5 wt% of the gases is emitted to the atmosphere.

Keywords: O2/CO2, Oxyfuel, CO2 free, Zero Emission, Coal Combustion, Compression, Carbon Dioxide, Flue Gas Treatment, Desulphurising, Sequestration

# Preface

This work has been carried out at the department of Energy conversion at Chalmers University of Technology in corporation with Vattenfall AB.

I would like to take this opportunity to thank all my commercial contacts, which have been very helpful in the hard work of estimating the costs of special equipment that does not presently exist. A special thanks is dedicated to Mr. Kahlert, Mr. Haunschild and Mr. Ullrich at VEAG for their helpfulness and for an enjoyable and extremely beneficial visit at the Lippendorf power plant.

Thanks to my supervisor, Professor Lars Strömberg, who has been very helpful with contacts, advice and the Lippendorf visit, and to my examiner and supervisor, Professor Filip Johnsson for discussion support, good advice in the report writing phase and a tremendous proof reading. Thanks to Professor Lennart Vamling for his patient support with chemical process simulations.

Special thanks to my roommates at the department of energy conversion Klas Andersson and Peter Maksinen, with whom I have been discussing different problems, and thereby also making this report better and the work very pleasant.

Finally, I would like to thank my fiancée, Elin, for her great ideas and support.

Göteborg, January 2002.

Henrik Birkestad

# Table of Contents

# 1. Introduction

Today most scientists agree on the fact that there is a strong connection between emissions of green house gases and the global climate change. A large part, almost 85%, of these emissions come from energy production. Approximately 80% of the world energy production originates from fossil fuels and mostly from some sort of coal (Holloway *et al.*, p. 4, 1996). The amount of coal and fossil fuels used in energy production is expected to increase even more during the next few years due to the large coal reserves and the expected growth in world economy, especially in the third world. There are a number of reports which in more detail discuss the green house effect and its relations to energy consumption, *e.g.* Greenhouse Gas Control Technologies (1999).

Carbon dioxide sequestration is an effective and drastic way to make major cuts in emissions of carbon dioxide. At present, underground storage (in depleted oil wells, aquifers etc.) is the only realistic storage alternative that has general acceptance. In Norway, one million tonnes of carbon dioxide per year are already being separated and stored in the Utsira aquifer.

The interest in recovering carbon dioxide from the flue gas is not only limited to reducing the emission of green house gases. There are also commercial factors, such as the $CO_2$ merchant market and enhanced oil recovery (EOR). That is, carbon dioxide dissolved in water is more efficient of driving out oil than water alone is; oil wells earlier considered as depleted can again be of interest for extraction.

To be able to store the carbon dioxide or use it for another application the gas must be at high concentration, but from normal combustion the concentration of carbon dioxide is diluted. There are, however, a number of techniques for increasing the concentration of carbon dioxide in the flue gas. $O_2/CO_2$ combustion and absorption with an amine, for example MonoEthanolAmine (MEA), are interesting techniques since they are to a large extent based on commercially available technology. Of these two, $O_2/CO_2$ is more efficient, it produces no amine waste and it also has a lower investment cost than MEA (Singh *et al.*, 1999). $O_2/CO_2$ combustion involves burning the fuel in an atmosphere of oxygen and recycled flue gas instead of air combustion.

Almost all previous studies concerning $O_2/CO_2$ combustion have concentrated their objectives on either economical or environmental issues. For example Croiset *et al.* (2000) and Kimura *et al.* (1995) are studying emissions and combustion parameters. When the process is discussed as for example by Nsakala *et al.* (2001) or by Hult and Nilsson (2001) the flue gas treatment mainly consists of compression of the gas. In short, there is a need for a technical evaluation with respect to the flue gas conditioning and compression process.

The main objective of this study is to investigate the requirements of the flue gas treatment in a large $O_2/CO_2$ power plant. Identification of equipment and estimation of their size and investment costs are carried out. A large state-of-the-art lignite-fired power plant has been selected as a reference plant to ensure that selected techniques can be implemented in a large power plant.

Moreover, parallel to this work a study is conducted concerning process integration with respect to the air separation unit and the power plant itself (Andersson and Maksinen, 2002). Thus, air separation and calculations of energy and exergy improvements achieved in the $O_2/CO_2$ power plant are only briefly treated in this report. Transport and storage of carbon dioxide are not covered in this work.

# 2. $O_2/CO_2$ Combustion

The advantages and disadvantages of $O_2/CO_2$ combustion in a power plant are described in more detail in this chapter.

## 2.1 Overall picture

$O_2/CO_2$ combustion involves burning the fuel in an atmosphere of oxygen and recycled flue gas instead of in air. Figure 2.1 shows a schematic layout of an $O_2/CO_2$ power plant. The oxygen is produced in an air separation unit (ASU). There are several types of ASU. However, for the oxygen flows required for a large-scale power plant only a cryogenic air separation process is possible today (Andersson and Maksinen, 2002). The maximum oxygen purity which can be obtained from the cryogenic process is 99.7% $O_2$. To save energy in the air separation process the purity of the oxygen is reduced to 95% (Andersson and Maksinen, 2002), which is the lowest possible purity from which no nitrogen is obtained in the oxygen stream. The remaining 5% consists entirely of argon (Kjellström, 2001). This implies that 1 kg of the oxygen stream includes 66 g of argon.

Argon is an inert gas and does not negatively affect the combustion reactions, it only contributes to the mass flow. The argon will increase the compressor work in the flue gas treatment process since it is non-condensable and must therefore be separated before the last step of compression. The optimisation of the purity of the oxygen is further discussed by Andersson and Maksinen (2002).

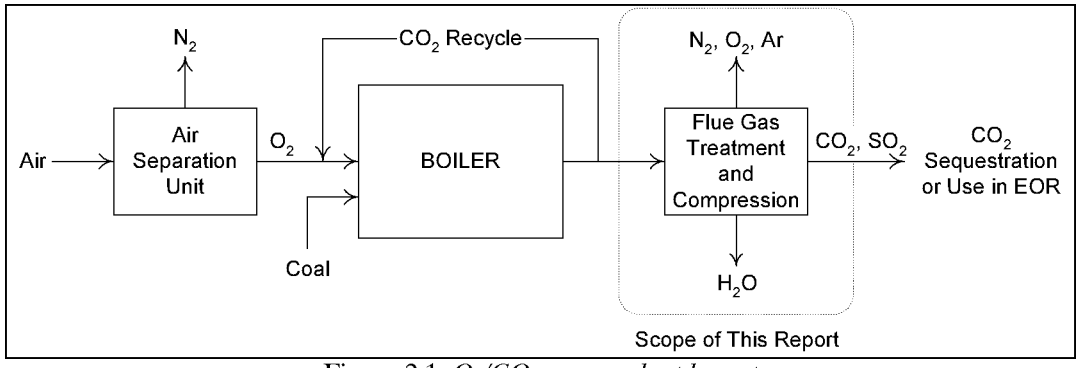


Figure 2.1: *$O_2/CO_2$ power plant layout*

The oxygen can be diluted with recycled flue gas in order to attain combustion conditions that are as normal as possible. Consequently, the concentration of oxygen in the feed gas can be varied from pure oxygen to lower concentrations. This means that it is possible repowering an existing boiler to $O_2/CO_2$ combustion, even if a design of a new power plant is more preferable, since it opens for optimisation of the oxygen concentration in the feed gas which should yield a slightly higher combustion efficiency.

The mixed flow of oxygen and recycled flue gas is fed to the boiler together with fuel and burned as in a conventional plant. A part of the flue gas is separated downstream of the economizer, recycled and mixed with new oxygen. The remaining part of the flue gas is treated, compressed and later transported to storage or to another application, Figure 2.1.

At normal combustion of lignite in air the concentration of carbon dioxide in the flue gas is approximately 14%. This means that an expensive process is necessary to increase the concentration of the carbon dioxide in order to gain the concentrated carbon dioxide flow required by compression. A post treatment to separate the carbon dioxide is not necessary in the case of $O_2/CO_2$ combustion because the major diluent, nitrogen, is removed in the air

separation unit and the concentration of carbon dioxide is relatively high already from combustion. Still, the overall efficiency of an $O_2/CO_2$ power plant is lower than a conventional plant. This is due to the energy required to compress the air in the air separation unit and the carbon dioxide in the flue gas stream. The scale advantages of revolving machines are considerable, which means that the $O_2/CO_2$ concept is most suitable for large power plants, preferably ones with large carbon dioxide emissions.

## 2.2 Emissions

$O_2/CO_2$ combustion can be designed to be a near zero emission concept. The carbon dioxide and sulphur dioxide can, except for some leak flows, be almost eliminated. Another advantage is the reduction of nitrous gases directly from the combustion. Okawa *et al.* (1997) and Okazaki (1997) have showed that $NO_x$ formed during $O_2/CO_2$ combustion is lower than from conventional combustion and they explain the reduction by the fact that 50% of the recycled nitrogen oxide is reduced into nitrogen. Croiset *et al.* (2000) notice that the flame temperature is lower for $O_2/CO_2$ combustion than in a corresponding air mixture because carbon dioxide has a higher specific heat capacity than nitrogen. In addition, they show that the measured carbon dioxide values during $O_2/CO_2$ combustion correspond well with the theoretical values calculated for complete combustion. Similar to Okawa *et al.* and Okazaki, Croiset *et al.* showed a significant reduction in nitrous emissions from $O_2/CO_2$ combustion compared with conventional combustion in air, Figure 2.2.

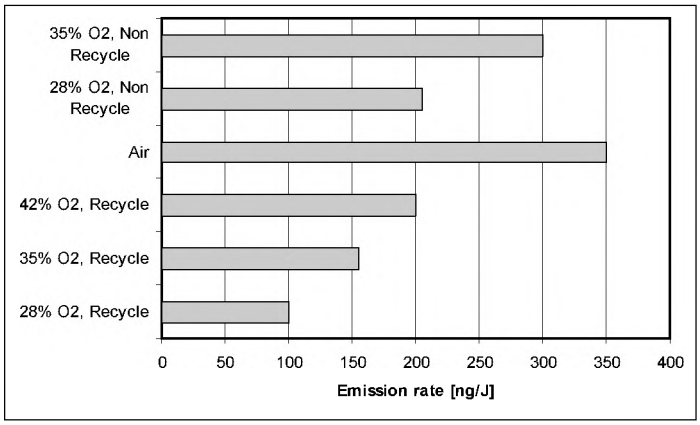


Figure 2.2: *$NO_x$ formed two meters from burner (Croiset et al. 2000)*
*The experiments were carried out in a 300 kW furnace, burning bituminous coal.*

The amount of sulphur emissions is not affected by oxygen concentration but instead by the sulphur content of the fuel. Almost all flue gases that contain sulphur dioxide ($SO_2$) also hold a small, but significant amount of sulphur trioxide ($SO_3$). This is important since sulphur trioxide is highly reactive and thereby has a significant influence on the boiler chemistry (corrosion, dew point etc). It also adds to the pollution problem if emitted to the atmosphere. The amount of sulphur trioxide is a critical factor in the calculation of flue gas dew point, which is further explained in Appendix A1.

However, there are a number of methods to decrease the levels of sulphur oxides in the flue gas. The most common method of reduction in a conventional plant is to add a wet desulphurisation unit after the precipitators. This unit stands for a large part of the internal power consumption as well as investment costs. It is not unusual that the desulphurising unit takes up 10 to 15% of the investment cost in a new power plant, but even with a highly modern and effective desulphurising unit sulphur is emitted to the atmosphere, normally a desulphurising efficiency of 95% is achieved.

In the $O_2/CO_2$ concept the sulphur dioxide can be left in the flue gas and the desulphurisation unit becomes unnecessary because of the sequestration. To dispose of the sulphur dioxide instead of separating it in a wet-scrubber therefore means less sulphur emissions to the atmosphere.

To sum up, there are problems with the $O_2/CO_2$ concept that are still to be solved, especially the power consumption, but the advantages are several as shown below.

**Advantages**

- High concentration of carbon dioxide in the flue gas – easy to capture
- Lower $NO_x$ emissions than normal air combustion
- Less investment cost since the desulphurising unit can be removed
- Reduced $SO_2$ and dust emissions
- Potential for increased combustion efficiency

**Disadvantages**

- Decreased electrical efficiency of the power plant
- Expensive and complex technology

# 3. Aim and Method

This work will apply an $O_2/CO_2$ concept to data from the 2x933 MW lignite-fired Lippendorf power plant. Actual values of mass flows, excess air and emissions from the plant have been obtained from the plant owner VEAG (mostly during on an on-site visit). The aim of the present work is to show what practical problems as well as which opportunities can occur with respect to the flue gas treatment when the $O_2/CO_2$ concept is applied to a plant of the Lippendorf size. In principle, the studied conditions could be seen as corresponding to repowering of the reference plant onto $O_2/CO_2$-firing. Differences between the original power plant and the corresponding $O_2/CO_2$ plant are identified and possible consequences for power consumption and investment costs are evaluated.

Parallel to this work a study of possible process integration of the interaction and air separation unit and the combustion process is performed (Andersson and Maksinen, 2002).

## *3.1 Reference Power Plant*

The Lippendorf power plant, near Leipzig in central Germany, has been chosen as a reference plant in this study. There is a long tradition of electricity production at Lippendorf. The first unit was commissioned in 1926 and had an efficiency of 17%. The new plant is a 2x933 MW lignite-fired power plant (2x865 MW net electrical power output) that was commissioned in 1999. Lippendorf also delivers 230 MW of district heating to the city of Leipzig. The coal, which is mined in the close vicinity of the plant is combusted in sixteen tangentially fed pulverized burners on two levels. The boilers are of the once-through type and the steam output for each boiler is 2,420 t/h, (supercritical) 267.5 bar and 554°C. Two cooling towers provide the cooling, which means that the lowest temperature available in the plant is 16°C on a yearly average basis. The electrical efficiency is 42.6%, which at present is the world's highest efficiency for a lignite-fired plant.

Much of the design efforts of the Lippendorf plant have been concentrated on reducing environmental impact. For example the desulphurising units bear as much as 10 to 15% of the total plant investment cost. In the desulphurising units the flue gas is exposed to lime slurry in a scrubber and the sulphur dioxide forms calcium sulphate – gypsum. Approximately half the amount (400 million kg/year) of the gypsum produced in

Lippendorf is sold to the building industry to be used in gypsum boards and the rest is disposed of together with some ash in the open pit to be used as a stabilizing agent. The Lippendorf's environmental performance under normal conditions is given in Table 3.1 for an oxygen excess of 6% on dry basis, which is the standard state to use for comparisons, Table 3.2. Actual emissions [kg/h] are calculated in Appendix A2. Total emissions from Lippendorf and the $O_2/CO_2$ scheme are compared in chapter 4.9.

Table 3.1: *Lippendorf actual emissions [mg/m$^3_n$] during the year 2000, Block S (6% $O_2$ dry) (App. A2)*

| | |
|---|---|
| $SO_x$ | < 355 |
| NOx | < 145 |
| HCl | < 12 |
| HF | < 2 |
| Dust | < 2 |

Table 3.2: *The environmental regulations [mg/m$^3_n$] in Germany (6% $O_2$ dry) (VEAG)*

| | |
|---|---|
| $SO_x$ | < 400 |
| NOx | < 200 |
| HCl | < 15 |
| HF | < 2.5 |
| Dust | < 20 |

The nitrous gases from the Lippendorf plant are reduced by effective boiler design and combustion control, there is no addition of ammonia, nitrogen catalysts or other common denitrification units.

In this study, values from one Lippendorf unit are used for comparisons with the $O_2/CO_2$ concept with respect to investment costs and internal power consumption. When it comes to transport issues and emissions both units are considered.

## 3.2 Method

All equipment and flows in the $O_2/CO_2$ concept studied in this report are considered identical with the reference plant, if not otherwise stated. To minimize the need of redesign of burners, convection surfaces etc, an air like mixture of 20 vol% oxygen and 80 vol% recycled flue gas has been chosen. This makes it easy to compare equipment and flows with those of the original plant. Figure 3.1 shows the principle combustion process and mass flows for the 933 MW $O_2/CO_2$ power plant. The flue gas values will be further discussed in chapter 3.3.1.

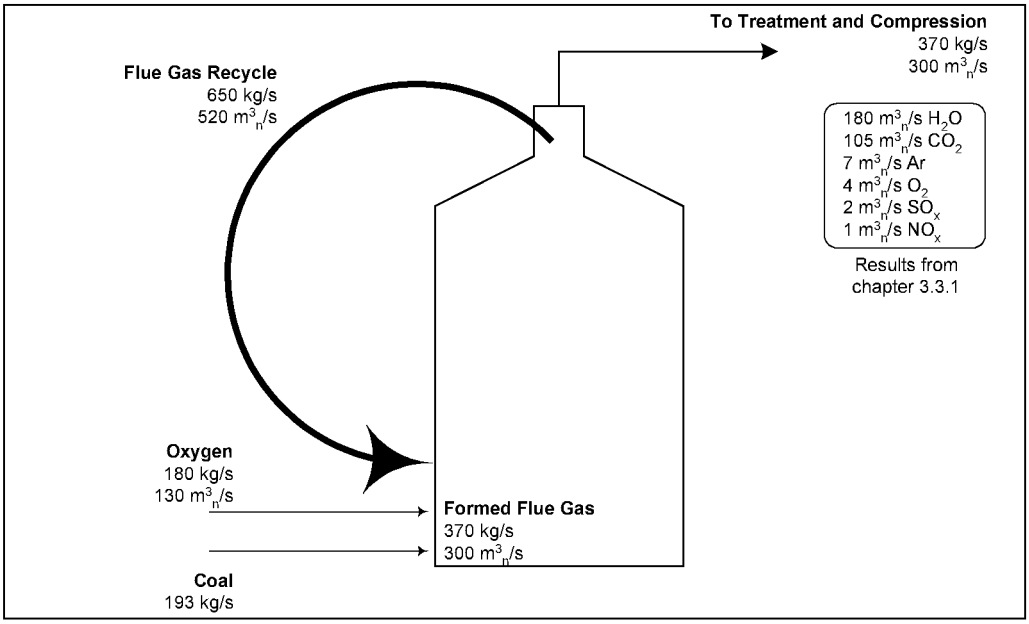


Figure 3.1: *Flue gas flows in the lignite-fired $O_2/CO_2$ power plant (Values from chapter 3.3.1)*

This work shows that several units have to be added to the flue gas treatment passage, compared with the original plant. As seen in Figure 3.1, the flue gas outlet from the boiler consists, besides of carbon dioxide, of a high concentration of water, some sulphur and some non-condensable gases (argon, oxygen, nitrous oxides and nitrogen). To be able to compress and further transport the carbon dioxide, those components must be separated and removed. The components needed for the separation are identified and given an approximate sizing and cost estimate in chapter 4.

Two main boundary conditions (inlet conditions and outlet conditions) had to be determined to identify the required treatment equipment. At the inlet: flue gas composition and flow (chapter 3.3.1) and at the outlet: transport conditions (chapter 3.3.2). Further on, certain process equipment sets sub-boundary conditions in the treatment line. For example: the carbon dioxide must be in liquid state when entering the gas separator and practically all the water must be removed before the flue gas enters the compressors. The results of the identification are closely described in chapter 4.

Chemical process simulators have been used to simulate the complex course of events in the flue gas condenser. ChemCad (v5.0), which in this case is the most accurate, uses electrolyte reactions together with thermodynamical models (Peng-Robinson) to solve the problem. Normally the accuracy is good, better than 5% (Vamling, 2001). As a comparison to ChemCad and to determine the condensation energy another program, Hysys (v4.2), is used. Hysys does not consider the electrolyte reactions, hence, for dissolved compounds the accuracy is lower than the ChemCad program. It should be noted, however, that not all substance interaction parameters were known. Some had to be estimated, which increases the uncertainty, especially for equilibrium calculations.

Proprieties of gas mixtures have been obtained from the NIST standard reference database with the program RefProp (Reference Fluid, Thermodynamic and Transport Properties). The adiabatic compressor work is drawn into a pressure-enthalpy diagram, using RefCalc (Refrigerator Calculator). The RefCalc program can also output numeric values for selected compressor cycles.

**Assumptions**
The air leakage into a modern power boiler is small. Disregarding the leak flow in the air preheater, the leakage is measured in parts per thousand (Kerff, 2001). Since the air preheater is removed in an $O_2/CO_2$ scheme, the leak flows can be neglected.

As found in chapter 2.2 there is a reduction of nitrous gases in an $O_2/CO_2$ plant (Croiset, 2000). Figure 2.1 shows that compared with the air combustion case only a third of the amount of nitrogen oxide is formed. Consequently, a third of the nitrous oxides values generated in Lippendorf have been used in this study. The nitrous gases are considered to be composed mainly of nitrogen oxide. None of the condition (long time or slow cooling) needed to let the nitrogen oxide transform into nitrogen dioxide is found in the boiler.

The excess of oxygen in an $O_2/CO_2$ combustion unit is 5% (Croiset, 2001), but this is for a small, non-optimised, laboratory sized unit. The air excess in Lippendorf is 1.15 before the air preheater, which corresponds to an oxygen excess level of 3 vol% (dry basis). This is a design value and the actual excess is lower. It is likely that the air excess could be further decreased on a full scale $O_2/CO_2$ plant, due to the flue gas recycle. Hence, a value of 1.5% $O_2$ excess is used in this study (Strömberg, 2002).

## *3.3 Technical Boundary Conditions*

This chapter describes calculations of the flue gas flows and why the particular state of transportation proprieties has been set.

### *3.3.1 Composition and Flow of Flue Gas*

The fuel flow and coal composition at the Lippendorf plant are given in the document Babcock: Kraftwerk Lippendorf (1995). In this report seven examples of coal are examined of which three cases are of interest for this work: the guarantee coal, the max water coal and the max ash coal, Table 3.3. The guarantee coal is used for all calculations regarding efficiency and flue gas flows. The plant must nevertheless be able to handle other coal as well. Hence the max water and max ash coals are used for dimensioning the flue gas condenser and the electrostatic precipitator respectively.

Table 3.3: *Proximate analysis [kg/kg] of coal (Babcock, 1995)*

| | **Guarantee** | **Max Water** | **Max Ash** |
|---|---|---|---|
| $H_i$ [MJ/kg] | 10.5 | 9.7 | 9.7 |
| C | 0.29110 | 0.29170 | 0.27178 |
| H | 0.02470 | 0.02475 | 0.02306 |
| O | 0.08190 | 0.08207 | 0.07646 |
| N | 0.00300 | 0.00301 | 0.00280 |
| S | 0.01430 | 0.01433 | 0.01335 |
| Cl | 0.00010 | 0.00010 | 0.00010 |
| F | 0.00005 | 0.00005 | 0.00005 |
| Ash | 0.06500 | 0.05414 | 0.08495 |
| Moisture | 0.52000 | 0.53000 | 0.52760 |

With the Lippendorf values, $W_{el,out}$=865 MW, $\eta_{el}$=0.426 and $H_i$=10.5 MJ/kg the fuel flow, $\dot{m}_{fuel}$, becomes:

$$\dot{m}_{fuel} = \frac{\left(W_{el,out} / \eta_{el}\right)}{H_i} \qquad [3.1]$$

which gives a fuel flow of 192.9 kg/s. This result for the Guarantee coal flow corresponds well with the value in the Babcock report, given in Table 3.4, which will be used for the subsequent calculations.

Table 3.4: *Fuel flows [kg/s] (Babcock, 1995)*

| Guarantee | Max Water | Max Ash |
|---|---|---|
| 192.6 | 223.7 | 223.7 |

For the Max Water and Max Ash coals the values stated in the Table have been compensated for soiling of the boiler and are therefore the maximum values. (The corresponding value for the new, clean boiler is 215.4 kg/s with the low heat value coals.)

The gas flows are calculated based on the coal analysis given in Table 3.3. The total fuel flow is known and the oxygen demand is obtained by means of stoichiometry. The oxygen demand is then adjusted to the oxygen excess (1.5 vol%) and an argon flow is calculated. A corresponding argon flow for 99.7% oxygen purity is also calculated for comparison. The calculations for the guarantee coal are found in Appendix B1 and Max Water and Max Ash in Appendix B2, B3 respectively. The results of flue gas composition for the guarantee coal are given in Table 3.5. The high volume percentage of water is a result of the moisture and the hydrogen content in the fuel.

Table 3.5: *Design composition of flue gas for guarantee coal during $O_2/CO_2$ combustion (Appendix B1)*

| Component | kg/s | wt% | $m^3_n/s$ | vol% |
|---|---|---|---|---|
| $H_2O$ | 142.7 | 38.4 | 179.8 | 60.4 |
| $CO_2$ | 205.4 | 55.3 | 105.3 | 35.4 |
| $SO_2$ | 5.4 | 1.5 | 1.9 | 0.6 |
| $SO_3$ | 0.1 | 0.0 | 0.0 | 0.0 |
| $O_2$ | 5.2 | 1.4 | 3.7 | 1.2 |
| $N_2$ | 0.5 | 0.2 | 0.4 | 0.2 |
| NO | 0.0 | 0.0 | 0.0 | 0.0 |
| Ar | 11.8 | 3.2 | 6.7 | 2.2 |
| **Total** | **371.1** | **100** | **297.8** | **100** |

### *3.3.2 Transport Condition*

Storage alternatives have been mentioned earlier in this report. However, the storage alternative alone does not fix the boundary outlet from the compression process. The density of the carbon dioxide is a function of temperature and pressure. It is important that the carbon dioxide is not allowed to transform into a gaseous state after it has been injected into the aquifer. If that happened, there would be a risk for gas pockets or other safety hazards. The pressure and temperature must be kept over the critical value, which means that the carbon dioxide must be stored 800 to 850 meters below the surface (Holloway, 1996). When injected into an aquifer the gas should be in super critical state, at least 73 bar, at a temperature exceeding 31°C, Figure 3.2. The vertical pressure gradient inside the aquifer is 105 bar/km and the temperature gradient is 25 to 35°C/km (Holloway, 1996). To manage loss of pressure and increase in temperature during injection, the carbon dioxide must have a pressure of at least 80 bar, when reaching the 800 meters level, to remain supercritical.

In liquid or super critical state the volume of carbon dioxide is approximately 500 times smaller compared to when in gaseous state and liquid phase is therefore preferable for

transportation. Below, two ways of transportation are described, both of which are commonly used commercially today.

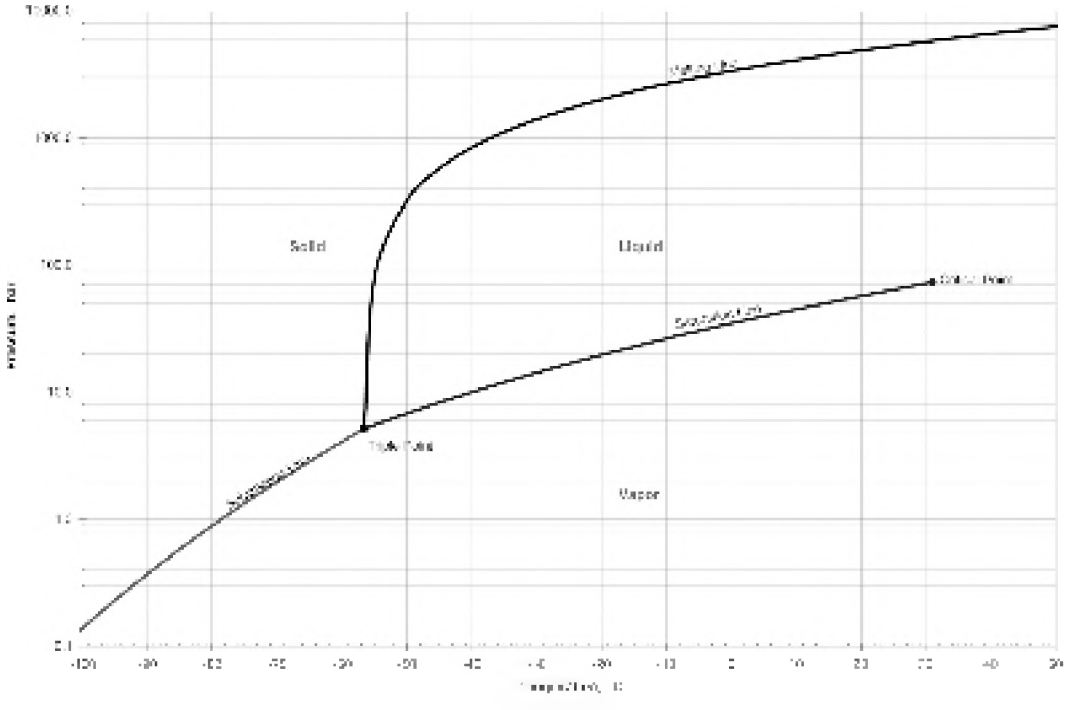


Figure 3.2: *Phase diagram for carbon dioxide (RefTab)*

**Tank Transport**

In tanks, carbon dioxide is transported at a pressure of approximately 20 bar and a temperature of approximately –30°C (Pettersson, 2001). The gas companies use this set of properties for transporting kinds of gases in a liquid state. These properties are based on economical and thermodynamical optimisation; 20 bar is the normal production outlet pressure, and few customers need higher pressure in their processes. Those who have a higher pressure demand have high pressure pumps on their sites and the transportation is still conducted at a pressure of 20 bar.

The wall thickness of the tank is directly proportional to the pressure; a doubling in pressure requires walls twice as thick and hence heavier and more expensive tanks. To avoid expansion of the gas and thereby pressure increase during transportation, the temperature must be kept at a constant level. The tanks are well insulated (by vacuum) and can hold the temperature several days without any external cooling or bleeding of the gas (Pettersson, 2001).

The gas manufacturer AGA AB has a ship for carbon dioxide transportation, which loads 1,250 tonnes and can manage five days without temperature problems, thus providing a radius of action of 2,600 km (Pettersson, 2001). A truck with a trailer can load 30 tonnes of carbon dioxide, while a train with 14 wagons can load 770 tonnes (Vattenfall Utveckling, 1997). Larger tanks could be designed and if the plant is located close to the sea or a major canal, super tankers may be suitable for transportation. The cost and area needed for storage tanks must be taken into consideration in such cases.

As the Lippendorf plant emits 1,500 tonnes of carbon dioxide an hour, it would mean in the tank transportation case that two full trains, 50 trucks or one ship would be needed every hour.

**Pipeline**
When transported in a pipeline the pressure of the carbon dioxide is 80 to 150 bar and the temperature approximately 20°C. The investment cost for a pipeline is high; however the annual cost is lower than in the tank transportation case. Pipeline transportation is cheaper in cases of large volumes, and it is therefore good if more than one plant can be connected to the same pipeline. It is the most practical and economical means of continuously transporting large volumes of gas. The inlet pressure in the pipeline must have a pressure margin to handle pressure loss and temperature during transportation.

In summary, tank transport is good for small volumes and long distances and pipelines for the opposite. In the case of a large plant like Lippendorf, pipeline transportation is the most realistic alternative and it is the one chosen in this report as the outlet boundary condition of the flue gas treatment and, thus, set to 100 bar and 25°C.

# 4. Evaluation of Flue Gas Treatment

Flue gas treatment basically involves the removal of water and non-condensable gases. This is carried out in several steps and Figure 4.1 shows the process parts required to prepare the flue gas for transportation with the numbers referring to the subchapter number where the step is described in detail.

1. Electrostatic precipitator (Chapter 4.1)
2. Flue gas condenser with water treatment (Chapter 4.2)
3. Compression steps (Chapter 4.3)
4. Active dehydration with TEG (TriEthyleneGlycol) (Chapter 4.4)
5. Heat exchanging (Chapter 4.5)
6. Removal of Non-condensable gases as $N_2$, $O_2$ and Ar (Chapter 4.6)
7. Pressure increase with booster pump (Chapter 4.7)

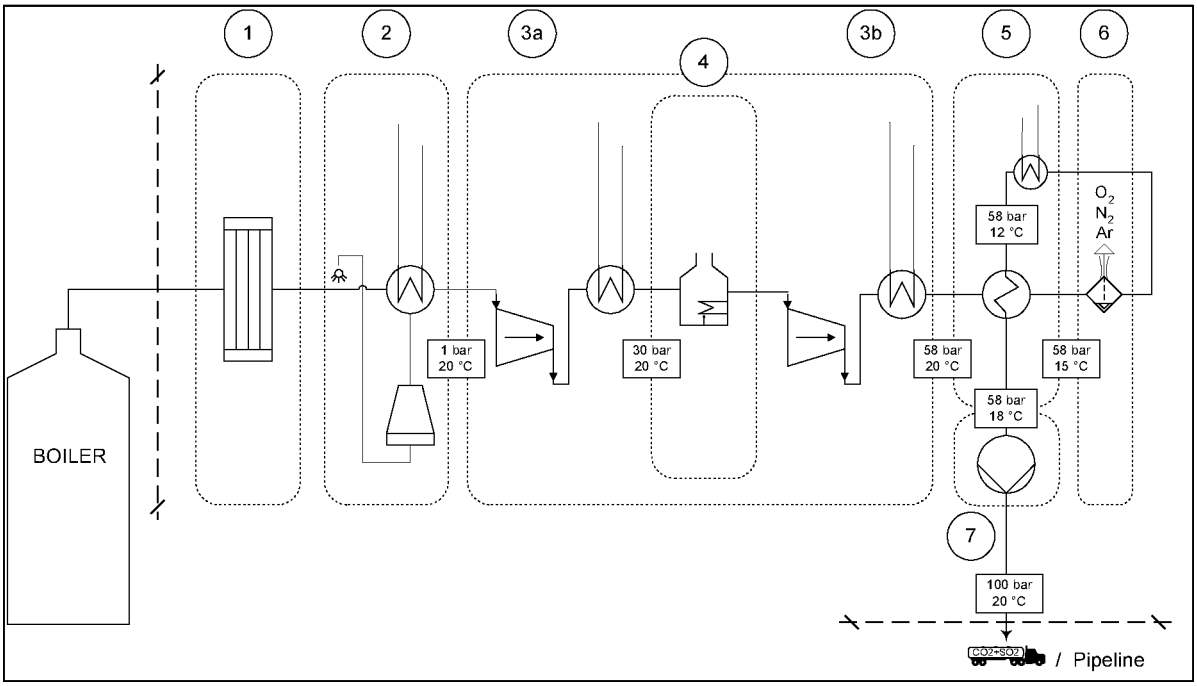


Figure 4.1: *Flue gas treatment steps*

A complete dehydration of the flue gas is important since it will:

- Reduce the mass flow
- Inhibit corrosion
- Inhibit the hydrate precipitation

If the flue gas is dehydrated to a dew point five degrees below that for transport conditions, the sulphur dioxide will behave almost as carbon dioxide and the two gases will not cause any corrosion problems. The gas must be dehydrated before reaching the high-pressure steps in the compression, to make the compression of the gas mixture possible (Lindeberg, 2001).

Carbon dioxide alone can be corrosive in the presence of water (sweet corrosion), but also in this case dehydration to a dew point five degrees below the transport temperature is sufficient (Fayed, 1983). A general rule for pipeline transportation in presence of water, is that serious corrosion can be expected if the partial pressure of carbon dioxide exceeds 2 bar (Berry, 1983). Various mechanisms have been discussed for the carbon dioxide corrosion process and all involve either carbonic acid or bicarbonate ion formed when carbon dioxide is dissolved in water, Equation 4.1. The reactions with the carbon steel are given by Equations 4.2 and 4.3. If the gas is dehydrated sweet corrosion is not a problem, since dry carbon dioxide is not corrosive at temperatures below 400°C (Kermani and Smith, 1997).

$$CO_2 + H_2O \Leftrightarrow H_2CO_3 \Leftrightarrow H^+ + HCO_3^- \quad [4.1]$$

$$Fe \rightarrow Fe^{2+} + 2e^- \quad [4.2]$$

$$CO_2 + H_2O + Fe \rightarrow FeCO_3 + H_2 \quad [4.3]$$

Corrosion is not the only possible problem if there is water present in the transport gas. Water vapour in the gas can form solid ice-like crystals called gas hydrates, which can block pipelines, valves or other equipment in the transmission and result in an emergency shutdown of the pipeline. To avoid this, the maximum water content should not exceed 60 to 100 $mg/m^3_n$ (Sloan, p. 545, 1998). The hydrates are formed when water "encages" gas molecules smaller than 1.0 nm (which is the case for both carbon dioxide and sulphur dioxide) at low temperatures and elevated pressures (temperatures below 25°C and pressures greater than 15 bar).

The gas will be dehydrated in two steps. The first is a traditional flue gas condenser where most of the water is removed, together with remaining particles, sulphur trioxide etc. The second dehydration step is the Tri Ethylene Glycol (TEG) unit, which will remove the remaining water down to a value of 60 ppm, or a dew point of -5°C at 100 bar in the transmission gas. It is not possible to perform the whole dehydration in the flue gas condenser alone. The cooling needed to dehydrate the remaining water in a condenser is considerably larger than the energy spent in the TEG absorber.

Since the TEG normally requires a pressure of 30 bar to be efficient (chapter 4.4), a compressor step with intercooling is installed before the TEG. Some water is also separated in the cooling steps in the compressor.

Figure 4.2 shows the mass flow of the flue gas components throughout the treatment steps. A more precise composition analysis can be found in Appendix C1.

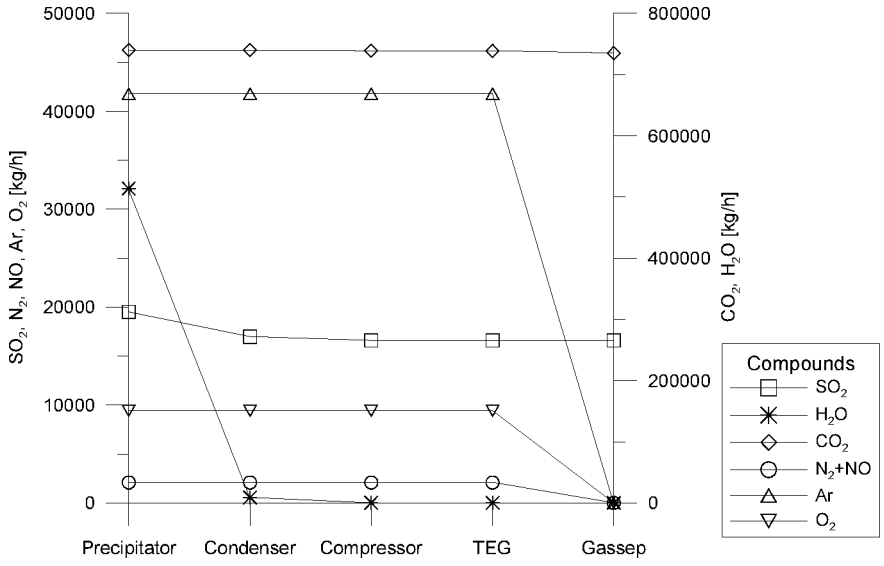


Figure 4.2: *Flue gas mass flow [kg/h] through the treatment steps*

## 4.1 Electrostatic Precipitator

When burning lignite only 20% of the ash is left at the bottom of the boiler (bottom ash) the rest follows the flue gas stream out of the boiler (fly ash) and must be removed before entering the condenser. In the Lippendorf case this means that 82,000 kg of ash has to be captured every hour (VEAG, 2001). (The dust is normally further reduced in the desulphurising unit.) In a modern electrostatic precipitator around 99.9% of the dust is removed. In the Lippendorf plant, less than 175 kg/h is left in the flue gas stream in the outlet. A New European Union directive states that maximum dust emission to the atmosphere for future power plants will be 30 $mg/m^3_n$ (Påle, 2001); *i.e.* this limit is already fulfilled by the Lippendorf plant (see Table 3.1).

The precipitator in the $O_2/CO_2$ scheme is the same or of the same type as in the original plant. A procedure for evacuating the flue gas stream must be arranged in case of failure of the precipitator. The condenser and compressor downstream will only manage a few minutes before they are totally filled with ash if the precipitator fails.

To avoid low temperature corrosion, the inlet temperature to the precipitator must be chosen with care. It depends of the amount of water and sulphur trioxide that is present in the flue gas. If the temperature decrease too much sulphuric acid will start to condensate on the walls. The dew point temperature for the $O_2/CO_2$ plant is found to be slightly higher than in an ordinary power boiler, mainly because of the high partial pressure of water. In the original Lippendorf plant the inlet temperature to the precipitator is 180°C and in the $O_2/CO_2$ scheme it has to be 190°C (Appendix A1).

To avoid accident filling of the boiler with ash, due to the flue gas recycle, the major part of the ash in the flue gas recycle stream must be removed. It is not effective to let the recycle stream through the precipitator, since the temperature of the flue gas recycle must then be lowered. A cyclone is therefore installed for ash separation in the recycle flow. The cyclone is outside the treatment line studied in this report.

The flue gas flow at the inlet of the precipitator is lower than at Lippendorf, because of the lack of nitrogen in the flue gas and lack of air leakage in the air preheater. Together with

the cyclone this probably means that the precipitator could be downscaled compared with the original Lippendorf precipitator.

Table 4.1: *Main differences in precipitator Lippendorf vs. $O_2/CO_2$*

| | **Lippendorf** | **$O_2/CO_2$** |
|---|---|---|
| Inlet temperature | 180°C | 190°C |
| Flue gas flow | 2x1,700,000 $m^3_n$/h | 1x1,100,000 $m^3_n$ /h |

The ash produced by burning lignite is relatively easy to trap in an electrostatic precipitator. However, it is important that the ash is to some extent conductive; otherwise it will not be affected when entering the electrostatic field in the precipitator and will then not be separated. If the ash has too low a conductivity, sulphur trioxide could be added, or the inlet temperature slightly decreased nearer the flue gas dew point.

**Approximate cost and sizing:**
An estimate of the size of the precipitator shows that the external measurements will be approximately 15 meters high and 20 meters wide (Påle, 2001). A surface contact area of 37,000 $m^2$ would probably be sufficient. The type of fuel and combustion environment is of great importance. As an example, a 500 MW power plant requires a surface area of 250,000 $m^2$ for the ash separation. In other words it is hard to make general conclusions on the precipitator without a simulation. A contact with a manufacturer gave the approximate cost and power need: 4 to 5 million EUR and 0.5 MW respectively.

## 4.2. Flue Gas Condenser

The flue gas condenser is the first new component added into the flue gas treatment of the $O_2/CO_2$ power plant. It is however commonly found in biomass boilers, where the water content in the flue gas stream is quite high, and where it is used to recover heat used for district heating. As seen from Table 4.2, the flue gas in the study has a high concentration of water vapour, roughly 61 vol% that mainly originates from the moisture in the fuel. Quite a lot of energy can be regenerated in the condenser, but the main purpose here is to remove as much water as possible, at a cost as low as possible. A added bonus is that much of the remaining dust and electrolytes (sulphuric oxides, chlorides and fluorides) are absorbed in the condensing water. This means that the condensing water will be quite acidic (< pH 2) if not treated correctly and this will negatively affect the removal of sulphur trioxide. In order to solve the latter in water in an efficient way, the pH value should be 5 to 7 (Axby *et al.*, 2000).

Sodium hydroxide is added to the condensed water to increase the pH to approximately 6.5. A chemical simulation indicates that it would be necessary to add 1,600 kg/h of sodium hydroxide to the water treatment process (Appendix C2). Part of this water is recycled to the top of the condenser and sprayed on the tubes to clean them, remove particles in the flue gas and to improve the condensation. The amount of carbon dioxide and sulphur dioxide that are separated in the condenser is bound to the water and not emitted to the atmosphere. Chlorides and fluorides are easily dissolved in water and are almost completely eliminated from the flue gas stream. Of the remaining dust, which was not removed in the precipitator, 50% (Råbne, 2001) is washed out with water droplets that are collected in the last step in the condenser.

Table 4.2: *Properties at inlet and outlet of flue gas condenser*

| | **Inlet** | **Outlet** |
|---|---|---|
| Temperature | 190°C | 20°C |
| Pressure | 1 bar | 1 bar |
| Flow | 1,065,000 $m^3_n$ /h | 430,000 $m^3_n$ /h |
| Moisture | 60.8 vol% | 2.6 vol% |
| $CO_2$ | 35.6 vol% | 88.5 vol% |
| $SO_2$ | 0.6 vol% | 1.4 vol% |
| $N_2$+NO | 0.2 vol% | 0.4 vol% |
| $O_2$ | 0.6 vol% | 1.6 vol% |
| Ar | 2.2 vol% | 5.6 vol% |

The condensation will take place between the inlet temperature 190°C and down to 20°C, which is the available temperature in the plant cooling system. In the interval down to 30°C the heat is used to feedwater preheating (Andersson and Maksinen, 2001) of 412 MW. Plant cooling water is used to lower the temperature to 20°C (7 MW). To condensate more vapour is not considered economical because of the low water content per degree at this temperature. A temperature reduction from 20°C to 15°C will only reduce the water content in the flue gas by half a percent, but requires approximately 7 MW of cooling to be produced in a refrigerating machine. As can be seen in Figure 4.3 most water condensates in the interval between 90°C and 60°C. The water left in the flue gas after the condenser will be removed both in the cooling steps in the compression and in the TEG dehydration unit (Chapter 4.4).

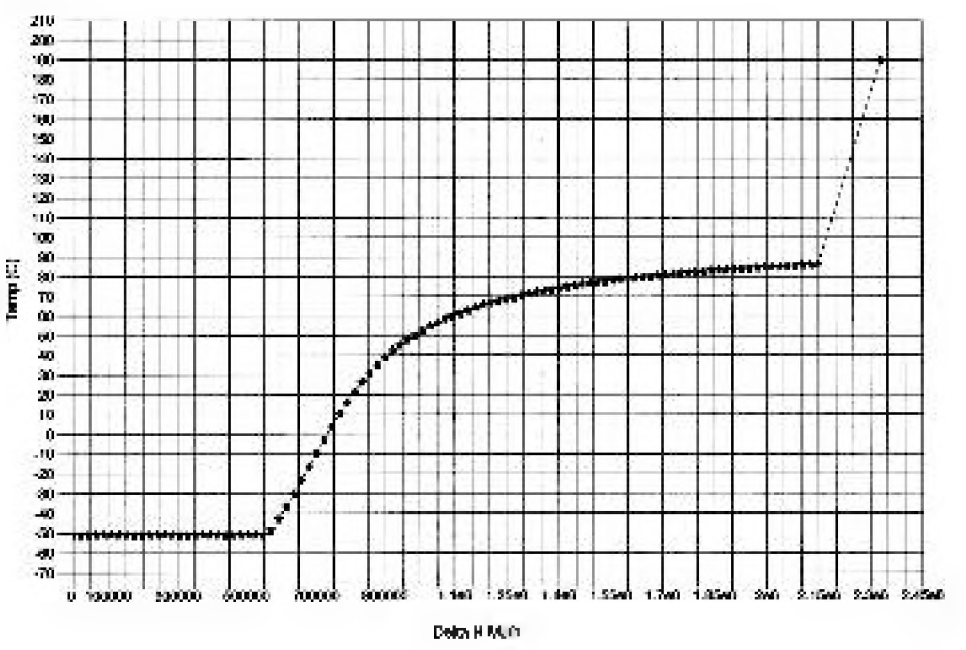

Figure 4.3: *T-Δh diagram for the flue gas (ChemCad)*

The cleaning of the condensed water requires some equipment. A sand filter is used to collect particles (forms mud) followed by a coal filter to collect organic compounds before the waste water is released. An example of a complete flue gas condenser with a water treatment plant is shown in Figure 4.4.

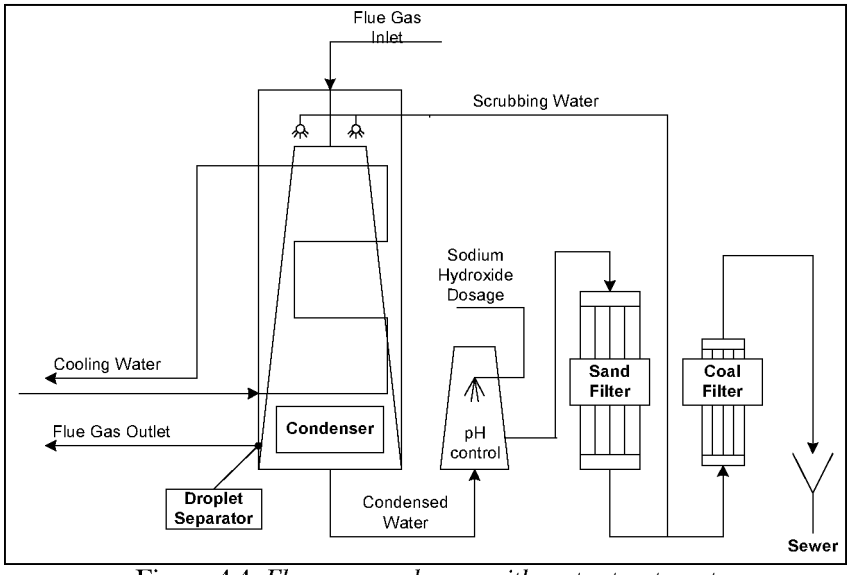


Figure 4.4: *Flue gas condenser with water treatment*

A simulation of the condensing reaction was performed with the chemical simulation software ChemCad and it was found that the water reduction is almost as expected from the stoichiometric calculations. The other chemical simulation program, Hysys, was run as a comparison. Both simulations gave similar results. The result from the ChemCad program is presented in Table 4.3. Dust, hydrochloric acid and hydrofluoric acid values in the Table were obtained from the condenser manufacturer. The complete output table from ChemCad is found in Appendix C2 and from Hysys in Appendix C3.

Table 4.3: *Amounts separated [kg/h] in the condenser unit (ChemCad and Råbne)*

| | |
|---|---|
| Water | 505,000 |
| $SO_2$ | 2,550 |
| $CO_2$ | 780 |
| $SO_3$ | 311 |
| Dust | 0.5 |
| HCl, HF | <0.9 |

**Material**
Mild carbon steel can be used in the flue gas stream before the condenser. Because of the corrosive environment when the water and other compounds condensates, the condenser and condenser housing requires stainless steel. A higher grade of stainless steel is required in the condenser tubes.

**Approximate costing and sizing:**
Contacts with a condenser manufacturer showed that approximately four condenser units with a diameter of 3.5 m and tube length of 10 m are needed. Each unit will have a weigh of 80 tonnes empty and 100 tonnes when filled with water (Råbne, 2001). The condensers together with buildings and water treatment plant will cost 5 to 10 million EUR.

## 4.3 Compression

The compression of the flue gas is the most power consuming process on the most complex to design and has the highest investment costs in the flue gas treatment scheme. To reduce the size and power consumption as much as possible, compression up to the transport pressure is not carried out in the compressor. The pressure should only be increased high enough to transfer the flue gas (mostly carbon dioxide) into a liquid state at a reasonable

cost. The final pressure increase is carried out in the last treatment step (chapter 4.7) by means of a high-pressure pump. The power needed to raise the pressure for a liquid (with a pump) is less than for a gas (with a compressor), Figure 4.5 $\Delta h_{pump} \leq \Delta h_{comp}$.

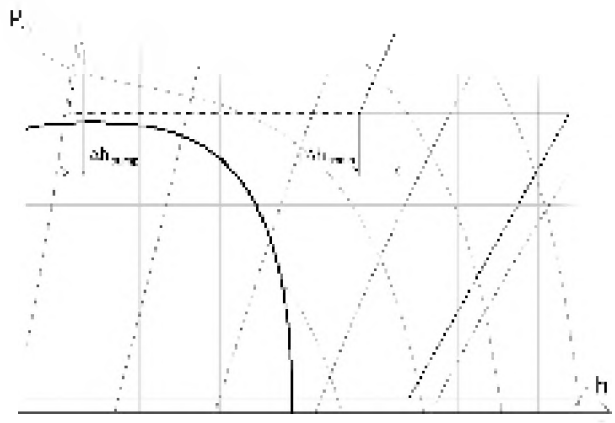


Figure 4.5: *Comparison of pump work vs. compressor work*

The compressor stream is divided into two units. The first unit raises the pressure from 1 bar to 30 bar, which is the inlet pressure of the TEG (chapter 4.4). After active dehydration the flue gas is compressed in the second unit from 30 bar to 58 bar. At a pressure of 58 bar, the carbon dioxide and sulphur dioxide will, if cooled to 20°C, liquefy and 20 °C is the temperature found in the main cooling system. The heat of evaporation is so large that it is not worth trying to condensate at a lower temperature, to save compressor work. The outlet temperature from the last step in the compressor is at least 70°C, which means that no condensation of gases will occur inside the compressor.

A simulation of the approximate gas mixture (97.5 wt% carbon dioxide and 2.5 wt% sulphur dioxide) gases indicates that the pressure needed for condensation at 20°C is slightly lower than 58 bar, Figure 4.5 (the simulation accuracy is 5%). However, carbon dioxide, which has the highest partial pressure in the mixture, is fully liquefied at 58 bar and 20°C, Appendix D1.

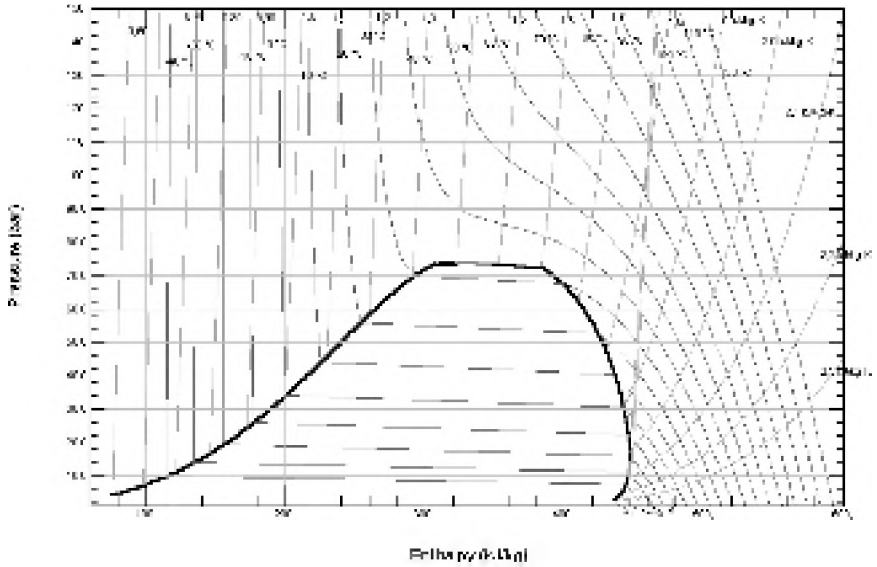

Figure 4.6: *p-h diagram for 97.5 wt% $CO_2$ and 2.5 wt% $SO_2$ (RefProp)*

There are three optional types of compressors: axial, radial or a combination of both axial and radial steps. The radial compressor is more robust than the axial compressor and can handle impurities in the gas. Axial compressors have higher efficiency and are smaller than the corresponding radial compressors but have a higher investment cost, which however, can be compensated for by the higher efficiency. The total pressure relation is lower for the axial compressor and it will therefore require more compression steps. An axial step has a maximum total pressure relation of 1.2 compared with a maximum of 5 for a radial. The compressor manufacturers are recommending radial steps for this application.

The higher efficiency of axial compressors compared to radial could motivate the higher investment costs and increased need of maintenance due to the large flows in the flue gas stream. The radial compressor is likely to be able to handle contamination of the flue gas. Brown (1996) argues that an axial compressor is the only type able to handle the large flows and at the same time maintain a high efficiency. To increase the efficiency, cooling is required between steps.

Adiabatic estimations of the work needed to compress the gases are found in Table 4.4 and the calculation is given in Appendix C4. The work is a function of pressure, flow, pureness of gas, efficiency and number of compression steps. Based on the above-mentioned facts, a radial compressor with three steps with a total pressure relation of 3.1 is chosen for the compression up to 30 bar.

Table 4.4: *Estimations of adiabatic shaft work for compression of flue gas [MW]* $\eta$*=0.85*

| | |
|---|---|
| Total -> 30 bar | 58.0 |
| Total -> 58 bar | 67.0 |
| Total -> 80 bar | 80.8 |
| Total -> 100 bar | 84.4 |
| Total -> 150 bar | 91.6 |

**Material**
In the first compressor unit water will condensate in the cooling parts, which means that they have to be made of stainless steel.

**Approximate costing and sizing:**
Because of the large volume flow two compressors are used in a parallel arrangement, which also provides better flexibility than with a single compressor. The approximate power consumption is slightly higher than the estimates given in Table 4.4. A rough estimate of the investment cost gives 5 million EUR per compressor (without driver), *i.e.* a total of 10 million EUR (Balzereit and Steuer, 2002).

## 4.4 Active Dehydration with TEG

As stated in the introduction to this chapter the maximum water content in the gas is set to 60 ppm to avoid corrosion and hydrate precipitation during transportation. The gas should be transported as a one-phase flow in supercritical or liquid state. To achieve such a low water content, active dehydration must be considered. Most common and widely used commercially in the oil and gas industry is the Tri Ethylene Glycol absorber (TEG), Figure 4.7.

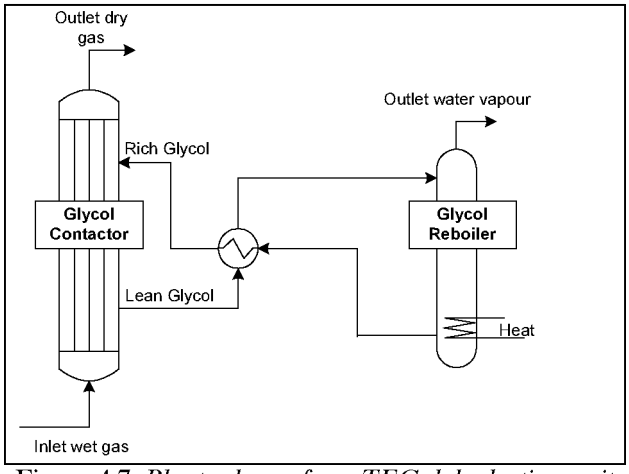


Figure 4.7: *Plant scheme for a TEG dehydration unit*

The flue gas is injected into the bottom of a contact column where the water is absorbed by the TEG solution. When the flue gas leaves the column, almost all the water is removed, Table 4.5. The grade of dehydration is determined by the quality of the TEG-solution. The TEG is regenerated in a separate boiler column where it is heated and the water leaves as steam.

Table 4.5: *Properties at inlet and outlet at TEG*

| | **Inlet** | **Outlet** |
|---|---|---|
| Temperature | 20°C | 20°C |
| Pressure | 30 bar | 30 bar |
| Flow | 415,000 $m^3_n$ /h | 415,000 $m^3_n$ /h |
| Moisture | 600 mg/$m^3_n$ (0.1 vol%) | 60 mg/$m^3_n$ |
| $CO_2$ | 90.9 vol% | 90.9 vol% |
| $SO_2$ | 1.4 vol% | 1.4 vol% |
| $N_2$+NO | 0.4 vol% | 0.4 vol% |
| $O_2$ | 1.6 vol% | 1.6 vol% |
| Ar | 5.7 vol% | 5.7 vol% |

Some carbon dioxide is dissolved in the TEG liquid and the amount is pressure and temperature dependent, Figure 4.8. This means that some of the TEG solution must be regenerated and that some carbon dioxide is lost. At 58 bar and a temperature of 20°C little is dissolved. A manufacturer simulation showed that 222 kg/h of carbon dioxide and a small amount of sulphur dioxide are absorbed by the TEG-solution in the $O_2/CO_2$ scheme. Those components are boiled off together with the water in the reboiler and therefore emitted to the atmosphere. If the water vapour is condensed to collect the heat, some of the carbon dioxide and sulphur dioxide will dissolve in the water and can be deposed of via the waste waster.

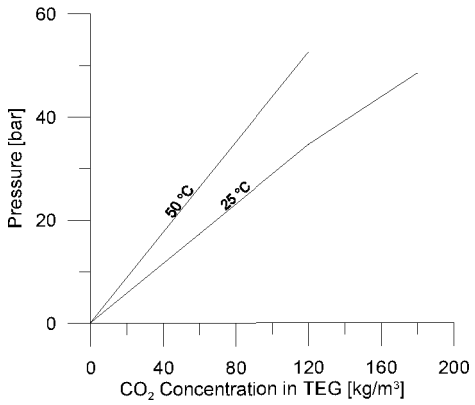


Figure 4.8: *Absorption of $CO_2$ in TEG-solution (Kohl et al., p. 974, 1997)*

Diaz and Miller (1987) have invented a method to substantially dry supercritical carbon dioxide with glycerol. With this method less carbon dioxide is dissolved in the absorber. This is accomplished by using glycerol instead of TEG as the active absorber fluid. The process is carried out at a temperature between 20 to 50°C and a pressure between 80 to 165 bar. No commercial units were found when performing literature studies, but the concept seems promising and should be further investigated.

**Approximate costing and sizing:**
Together with a TEG manufacturer a simulation of the TEG unit was made. A concentration of 99.6 wt% TEG is used in the simulation, which is a common value for similar commercial units. The unit requires approximately 125 kW of heat for the regeneration of the TEG solution and a small amount of electrical power for recirculation pumps etc (Bredahl and Karbo, 2002). The approximate investment cost is 2 million EUR.

## 4.5 Heat Exchanger

The carbon dioxide and sulphur dioxide is transferred into a liquid state after the final compressor step. At 58 bar, cooling water from the cooling towers can be used to condensate the gas in the condenser. The purpose of the heat exchanger is to supercool the flue gas stream and make sure that all carbon dioxide and sulphur dioxide are in a liquid state when entering the gas separator (Chapter 4.6). The cold is mainly recovered from the liquid steam after the gas separator before entering the booster pump. To compensate for eventual losses in the heat exchanger and in the gas separator, a refrigerating machine can be installed.

**Approximate costing and sizing:**
The investment cost for the condenser located after the compressor is estimated to 800,000 EUR and the heat exchanger to 75,000 EUR (Carlsson, 2001).

## 4.6 Separation of Non-condensable Gases

When reaching this unit the main part of the flue gas (carbon dioxide and sulphur dioxide) is in a liquid state; details are found in Table 4.6. However, the non-condensable gases (nitrogen, nitrogen oxide, oxygen and argon) are still in a gaseous state. A conventional gas-liquid separator, showed in Figure 4.9, is used to separate the liquid phase from the gas phase. The separator is quite simple in its construction, using gravitation to separate the flows, with an inlet for the mixed fluid on one side, an outlet at the bottom for the liquid and an outlet at the top for the gas. Several layers of demister nets cover the gas outlet and provide efficient droplet separation; approximately 99.5% of the liquid droplets are

removed (Karlsson, 2001). Consequently, 3,800 kg of the liquid phase is lost in the separator and emitted to the atmosphere.

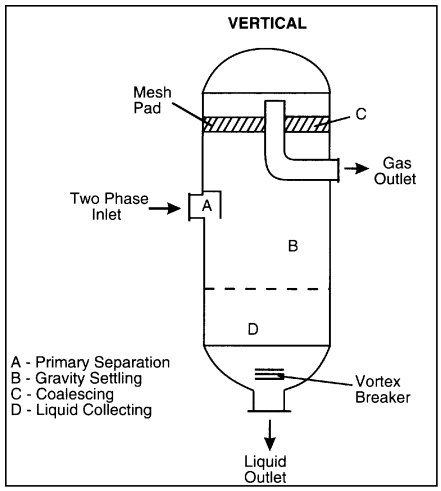


Figure 4.9: *Gas-liquid separator (GPSA, 1998)*

Table 4.6: *Properties at inlet and outlet of gas/fluid separator*

| | **Inlet** | **Outlet** |
|---|---|---|
| Temperature | 15°C | 15°C |
| Pressure | 58 bar | 58 bar |
| Flow | 415,000 $m^3_n$ /h | 385,000 $m^3_n$ /h |
| Moisture | 60 mg/$m^3_n$ | 60 mg/$m^3_n$ |
| $CO_2$ | 90.9 vol% | 98.5 vol% |
| $SO_2$ | 1.4 vol% | 1.5 vol% |
| $N_2$+NO | 0.4 vol% | - |
| $O_2$ | 1.6 vol% | - |
| Ar | 5.7 vol% | - |

The volume flows at 58 bar, can be calculated from the ideal gas law.

$$\frac{p_1 \cdot v_1}{n_1 \cdot R \cdot T_1} = \frac{p_2 \cdot v_2}{n_2 \cdot R \cdot T_2} \quad [4.4]$$

The pressure, volume and temperature are the only parameters that change between the normal state of the gas and the conditions in the separator. Hence, Equation 4.4 is simplified into Equation 4.5 and 4.6. The resulting volume flows at 58 bar are found in Table 4.7.

$$\frac{p_1 \cdot v_1}{T_1} = \frac{p_2 \cdot v_2}{T_2} \quad [4.5]$$

$$v_2 = \frac{p_1 \cdot v_1 \cdot T_2}{T_1 \cdot p_2} \quad [4.6]$$

Table 4.7: *Volume flows [$m^3/s$] at 58 bar*

| | **Inlet** |
|---|---|
| $CO_2$ | 1.90 |
| $SO_2$ | 0.03 |
| $O_2$ | 0.03 |
| $N_2$+NO | 0.01 |
| Ar | 0.12 |
| *Total liquid* | *1.93* |
| *Total gas* | *0.16* |

In the separator 42 000 kg of argon, 9 500 kg of oxygen and 2 000 kg of nitrogen are separated every hour. The concentrated stream makes it possible to further process the gas if there is a need or market for any of the gases. The oxygen flow in the combustion should be minimized, but be kept high enough to avoid unburned components in the flue gas.

**Approximate costing and sizing:**
When designing a gas-liquid separator a vertical gas velocity of 2 m/s is preferable, because the demister uses the kinetic energy in the liquid droplets to separate them from the gas. If the flow is too low there is a possibility that the liquid droplets will zigzag through the demister. However, too high a flow can choke the demister. With the present gas volume flows, a surface area of 0.11 $m^2$ and a diameter of approximately 0.4 m is sufficient to obtain a velocity of 2 m/s. Further on, a separator height of 10 m and a demister net thickness of 200 mm should be suitable for this process (Karlsson, 2002). A unit like this will cost approximately 21,000 EUR, of which the demister cost will be 6,000 EUR.

## 4.7 High Pressure Pump

The high-pressure pump is used to increase the pressure of the fluid to transport conditions. It has been located after the heat exchanger to avoid the heat generated in the pump from increasing the cooling power in the refrigerating machine. The heat exchanger must be designed so that all fluid is in a liquid state and negligible amounts in a gaseous state, otherwise there is a risk of pump cavitation and resulting pump damage. The fluid is a mixture of carbon dioxide (97.5 wt%) and sulphur dioxide (2.5 wt%). Viscosity, density and heat capacity for this mixture are found in Appendix C5.

To use a pump instead of a compressor to increase the pressure lower the investment cost, level of maintenance as well as power need. The pump required is specified in Table 4.8. It is a centrifugal pump with several steps. According to manufacturers there are probably not any standard pumps on the market of sufficient capacity, mostly because of the high inlet pressure.

Table 4.8: *Properties at inlet and outlet of high pressure pump*

| | **Inlet** | **Outlet** |
|---|---|---|
| Temperature | 18°C | 20°C |
| Pressure | 58 bar | 100 bar |
| Mass flow | 750,000 kg/h | |

**Approximate costing and sizing:**
Estimated shaft power (assuming a pump efficiency of 0.75) is 390 kW (Gerdin, 2001). The closest standard motor is 500 kW (efficiency 0.96). It should be noted that power consumption for increasing the pressure for a liquid is significantly lower than for the corresponding gas (Figure 4.4). The approximate cost for a pump and motor is 200,000 EUR for the pump and 100,000 EUR for the motor, *i.e.* in total 300,000 EUR.

# 5. Discussion

This chapter summarizes the costs and power consumption for the above described flue gas scheme. In addition, estimates of the final emissions are given. In Figure 5.1 all mass flows in the scheme are shown.

## 5.1 Overall Cost and Power Estimates

As always is the case with new technology, the cost is difficult to estimate. In Table 5.1 an attempt of comparison of the flue gas treatment between the original Lippendorf plant and the $O_2/CO_2$ process can be found. The equipment used in the compression of carbon dioxide is expensive, but savings are made in excluding the desulphurising unit and the stack. In a modern power plant like Lippendorf the stack is also removed and the flue gas is injected into the middle of the cooling towers.

As can be seen from Table 5.1, it is estimated that the suggested flue gas treatment scheme for an $O_2/CO_2$ plant has an investment cost which is around 140 million EUR lower than of the original air-fired plant. Please note that only flue gas treatment is included in the investments cost above and that the compressor driving is demarcated. The air separation unit adds 100 million EUR to the $O_2/CO_2$ cost, has an approximate power consumption of 165 MW and is further discussed in Andersson and Maksinen (2002).

Table 5.1: *Comparison of approximate investment cost and power consumption for the flue gas treatment between the original Lippendorf plant and a corresponding $O_2/CO_2$-fired plant*

| | **Investment Cost** | | **Power Consumption** | |
|---|---|---|---|---|
| **Unit** | **Lippendorf** | **$O_2/CO_2$** | **Lippendorf** | **$O_2/CO_2$** |
| Precipitator | 2x4 M EUR | 4-5 M EUR | 2x1.2 MW | 2 MW |
| Flue Gas Condenser | - | 5-10 M EUR | - | -14 MW |
| Desulphurising unit | 160 M EUR | - | 15.8 MW | - |
| Compressors | - | 10 M EUR | - | 70 MW |
| TEG | - | 2 M EUR | - | [heat] |
| Heat exchangers | - | 875,000 EUR | - | [heat] |
| Gas/liquid separator | - | 21,000 EUR | - | 0 MW |
| Pump | - | 300,000 EUR | - | 0.5 MW |
| *TOTAL* | *168 M EUR* | *28 M EUR* | *18 MW* | *59 MW* |

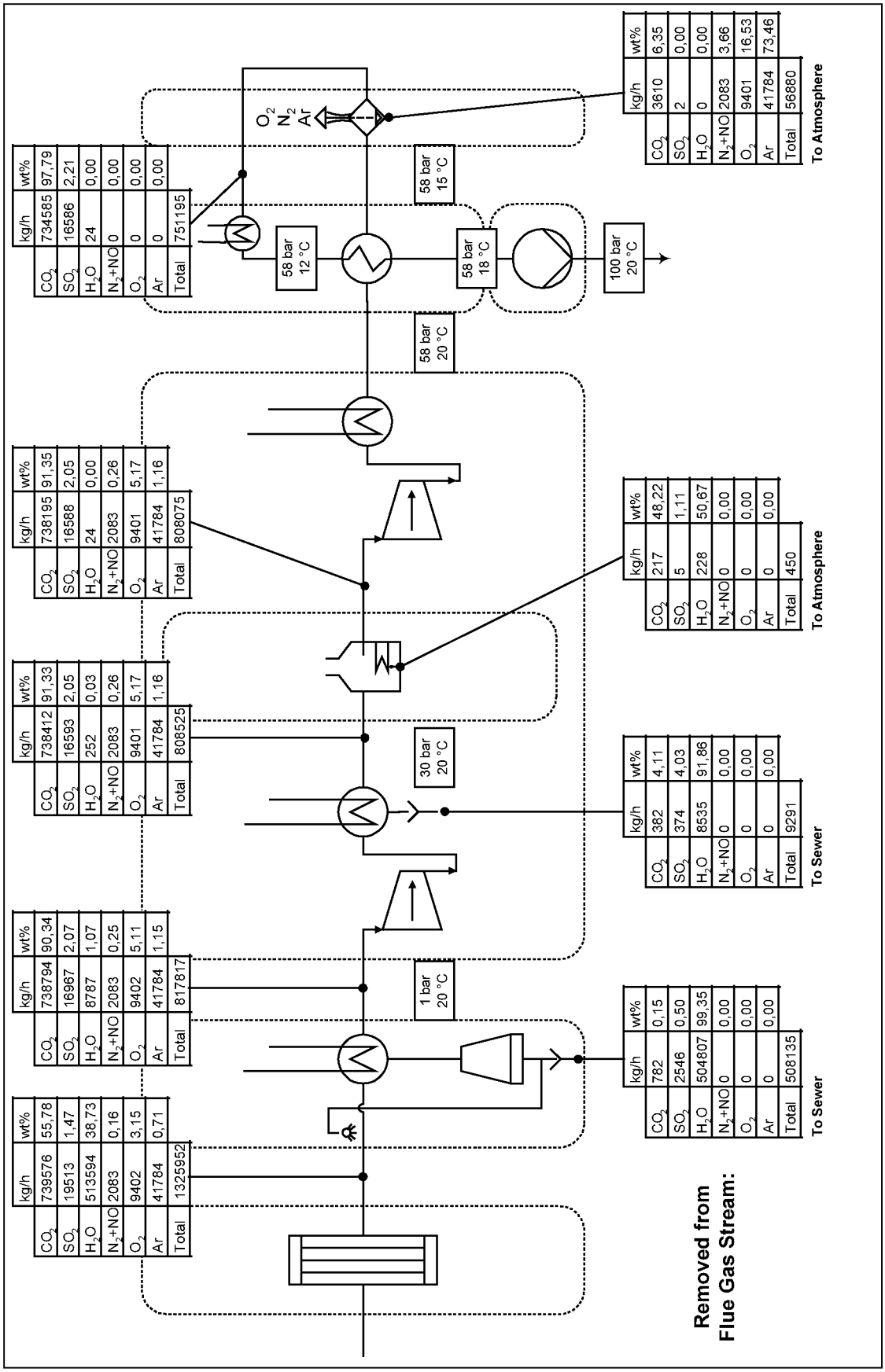


Figure 5.1: *Mass flows and separated streams in the scheme*

## 5.2 Overall Environmental Estimates

Table 5.2 shows a comparison of estimated emissions from Lippendorf and the corresponding $O_2/CO_2$-fired plant. The values are calculated for 6% oxygen excess on a dry basis. The $O_2/CO_2$ scheme has been converted to the corresponding flue gas flow for air combustion. Total emissions [kg/h] are calculated for 2x933 MW.

Table 5.2: *Comparison of estimated emissions between Lippendorf and $O_2/CO_2$ (Appendix A2)*

| **Emissions to air** | **Lippendorf** | | **$O_2/CO_2$** | |
|---|---|---|---|---|
| $SO_x$ | < 350 mg/m$^3$ | < 2,230 kg/h | < 6 mg/m$^3$ | < 20 kg/h |
| $NO_x$ | < 145 mg/m$^3$ | < 920 kg/h | < 141 mg/m$^3$ | < 310 kg/h |
| $CO_2$ | < 235 g/m$^3$ | < 1,480 tonnes/h | < 4 g/m$^3$ | < 8 tonnes/h |
| Dust | < 2 mg/m$^3$ | < 12 kg/h | < 1 mg/m$^3$ | < 1 kg/h |

# 6. Conclusions

This report presents a flue gas treatment scheme for an $O_2/CO_2$-fired plant. The data are given for a plant of 933 MW. In order to obtain relevant data, a state-of-the-art power station was used as reference. Also other commercial component data have been used as much as possible.

An almost complete dehydration is of great importance to avoid problems in the final flue gas treatment and in the transportation of the carbon dioxide.

The emissions from the $O_2/CO_2$ plant are that low that it can be considered as a zero emission concept.

The investment costs of the flue gas treatment are lower than in the conventional case, mainly because of the removal of the desulphurising unit. Even if the ASU increases the overall $O_2/CO_2$ plant cost, the total investment cost of the power plant is slightly lower than for the original plant. The internal power need is, however, increased and the electrical efficiency consequently decreased by 10 percent units. Improvements and optimisations can probably be made after a full-scale pilot plant has been built.

# Nomenclature and Conversion Factors

| | | |
|---|---|---|
| $H_i$ | Heat value | [MJ/kg] |
| $\dot{m}$ | Mass flow | [kg/s] |
| n | Substance amount | [mole] |
| $\eta$ | Efficiency | [%] |
| p | Pressure | [bar] |
| R | Molar gas constant | R=8.31451 J/mole K |
| T | Temperature | [°C] |
| W | Work | [MW] |

**Index**

| | |
|---|---|
| 1 | First state |
| 2 | Second state |
| el | Electricity |
| fuel | Fuel |
| i | Component number |
| n | Normal state (25°C, 1.013 bar) |
| out | Output |
| vol% | Percent of Volume |
| wt% | Percent of Weight |

**Concentration**

1 lb/MMscf = 16.02 $mg/m^3_n$

**Currency**

1 EUR = 1 US-Dollar = 10 SEK

**Mass**

1 ton =1 000 kg=2 205Lb

**Pressure**

1 bar=14.504 psi

1 atm = 14.696 psi

**Temperature**

1 degree C=(degrees F-32)*5/9=(degrees F-32)/1.8

1 degree F=(degrees C*9/5)+32=(degrees C*1.8)+32

**Volume**

1 MMscf=10^6 standard cubic feet=28317 $m^3_n$

1 Cuft = 28.317 liters = 0.02832 $m^3$

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