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Coal-fired Power Plants based on Oxy-combustion with Carbon Capture: Combustion Conditions and Water Consumption

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MASTER THESIS

for

student Karen Almåas

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Coal-fired power plants based on oxy-combustion with carbon capture: combustion conditions and water consumption

*Kolfyrte kraftverk basert på oksy-forbrenning med karbonfangst:
forbrenningsvilkår og vassforbruk.*

Background and objective

CO_2 capture processes have not yet been commercially applied for capture from power plants for CO_2 capture and storage. However, new technologies are currently being developed and will soon be demonstrated in large-scale CCS demonstration projects. One concept is oxy-fuel combustion or oxy-combustion, which is combustion with pure O_2.

A critical aspect in such developments is to demonstrate technological feasibility of the proposed concept. To predict and demonstrate feasibility, process modeling and simulation is often performed as part of qualifying the technology. Det Norske Veritas (DNV) has recently released a recommended practice for the qualification of CO_2 capture technologies. Currently, DNV is expanding upon this by developing simulation platforms and guidelines for the verification and validation of process simulations.

The main focus of this work is twofold: the modeling of the combustion conditions within oxy-fuel combustors and the water consumption for the oxy-combustion process with carbon capture. The latter is motivated by the fact that post-combustion carbon capture processes often use substantial amounts of water, which can be a limiting factor in many areas. The former issue is a continuation of the specialization project conducted autumn 2011.

The following tasks are to be considered:

- A brief literature study on carbon capture from coal combustion in general, and oxy-coal combustion specifically; including technology and modelling.
- Continue investigations of combustor/reactor models (from the specialization project), their uncertainties, strengths and limitations.
- Use Hysys to model a power-plant process with oxy-coal combustion and carbon capture, to quantify the water consumption. If time allows, also model a process with coal combustion and post-combustion carbon capture for comparison.
- Investigate and discuss the water consumption of oxy-coal combustion with carbon capture.

Within 14 days of receiving the written text on the master thesis, the candidate shall submit a research plan for his project to the department.

When the thesis is evaluated, emphasis is put on processing of the results, and that they are presented in tabular and/or graphic form in a clear manner, and that they are analyzed carefully.

The thesis should be formulated as a research report with summary both in English and Norwegian, conclusion, literature references, table of contents etc. During the preparation of the text, the candidate should make an effort to produce a well-structured and easily readable report. In order to ease the evaluation of the thesis, it is important that the cross-references are correct. In the making of the report, strong emphasis should be placed on both a thorough discussion of the results and an orderly presentation.
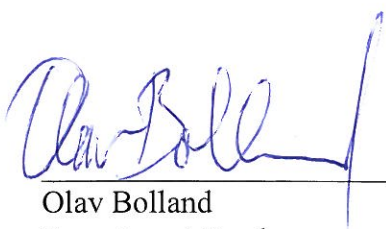
The candidate is requested to initiate and keep close contact with his/her academic supervisor(s) throughout the working period. The candidate must follow the rules and regulations of NTNU as well as passive directions given by the Department of Energy and Process Engineering.

Risk assessment of the candidate's work shall be carried out according to the department's procedures. The risk assessment must be documented and included as part of the final report. Events related to the candidate's work adversely affecting the health, safety or security, must be documented and included as part of the final report.

Pursuant to "Regulations concerning the supplementary provisions to the technology study program/Master of Science" at NTNU §20, the Department reserves the permission to utilize all the results and data for teaching and research purposes as well as in future publications.

The final report is to be submitted digitally in DAIM. An executive summary of the thesis including title, student's name, supervisor's name, year, department name, and NTNU's logo and name, shall be submitted to the department as a separate pdf file. Based on an agreement with the supervisor, the final report and other material and documents may be given to the supervisor in digital format.

Department of Energy and Process Engineering, 22. January 2012

Olav Bolland
Department Head

Ivar S. Ertesvåg
Academic Supervisor
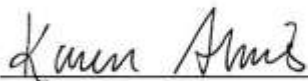
Research Advisor:
Tore Myhrvold, DNV

Preface

This paper, Coal-fired power plants based on oxy-combustion with carbon capture: combustion and water consumption is written as a master thesis at the Norwegian University of Science and Technology, in collaboration with Det Norske Veritas (DNV). This thesis comprises 30 out of 30 credits in the 10th semester for the 5-year Master of Science Degree in the field of Energy and Process Engineering.

During this period, my academic supervisor, Ivar Ståle Ertesvåg, has been available and helpful at all times. He is a great source of knowledge, and deserves my greatest thanks. I would also like to express my gratitude to my research advisor Tore Myhrvold from DNV, for showing interest and enthusiasm in this master thesis.

Further, I would like to thank PhD candidate, Fu Chao, for his guidance, and for allowing me to use, his simulation model of the oxy-combustion coal-fired power plant in Aspen Plus, in this work. Research assistant, Kjetil Vinjerui Ekre, also deserves great thanks for his help, with the MEA absorption model in Aspen HYSYS.

Any questions regarding the content of the report can be sent to my email address: karenalmaas@gmail.com.

Trondheim, August 26th 2012

Karen Almås

Remarks

In agreement with the academic supervisor, Ivar Ståle Ertesvåg, some modifications where made to the original tasks considered in the thesis description. The task based on the specialization project, continuation of reactor investigation, was eliminated. Instead, the main topic, water consumption was extended. In addition to simulation of an oxy-combustion power plant, an amine absorption CO_2 capture process was to be simulated. In that way, it was more convenient to compare the water consumption of the two CO_2 capture technologies, which is part of the aim in this thesis.

Abstract

Electricity production from nearly all conventional steam plants rely on water as a heat sink. Besides from process cooling, there are several other subsystems that require water for operation. Additionally, integration of CO_2 capture tends to increase the water usage and consumption of a power plant. In some parts of the world, water is a limited resource and must be used with concern. Concurrently as the world population is growing, the focus on CO_2 emissions is also sharpened. The promising CO_2 capture technology based on oxy-combustion appear to rely less on water compared to post-combustion capture using amine absorption.

Relating CO_2 capture and water consumption, this work presents a case study of two similar power plants based on different CO_2 capture methods. A 561 MW oxy-combustion power plant and a 550 MW post-combustion capture plant are considered in order to quantify and compare the water consumption of the plants.

The results show that the cooling tower is by far the largest water consumer in both power plants. Evaporative losses are dominant, but cooling tower blowdown also accounts for a significant portion of the total water consumption, assuming a mid-range water quality. However, due to larger cooling water requirements in the post combustion capture plant, the cooling tower water losses are 17.8% higher compared to the oxy-combustion case.

The second largest water consumer is the flue gas desulfurization (FGD) process. In air-fired plants, the main cause of water loss in FDG systems is evaporation of water. Other water losses are related to the production of gypsum and to the purge system. A high fraction of water vapor in the flue gas from oxy-combustion eliminates evaporative losses, if the flue gas dew point is below the operating temperature of the FGD system. In this study, the FGD process in the post-combustion capture plant consumed eight times more water than that of the oxy-combustion plant.

A significant amount of condensate is available during air separation and CO_2 recovery as water also enters the system via humid air intake, moisture content in fuel, and hydrogen bound in the fuel. Water recovery could contribute to strongly reduce the water consumption in both power plants

A scenario where the power plants are located near the sea utilizing a once-through cooling system is also considered. Calculations show that the oxy-combustion plant only consumes 17 % of the water consumed in the post-combustion capture plant. This trend also applies in a smaller extent, when evaporative cooling towers are employed.

Sammendrag

Nesten all konvensjonell dampkraftproduksjon er avhengig av vann for kjøling. Tilgang til vann er også nødvendig i drift av flere andre delsystemer i et slikt kraftverk. I tillegg medfører integrasjon av CO_2 fangst ofte at vannforbruk i et kraftverk går opp. Deler av verden har stor vannmangel og vann bør derfor brukes med omhu. Samtidig med en stadig økende verdensbefolkning, skjerpes fokuset på CO_2 utslipp. Det forekommer at den lovende CO_2 fangstteknologien basert på oksy-forbrenning, er mindre avhengig av vann sammenlignet med CO_2 fangst basert på etterrensing av eksosgassen via kjemisk absorpsjon.

To liknende kraftverk med ulik fangstprosess er studert i denne oppgaven. Vannforbruket i et 561 MW oksy-forbrenningskraftverk og et 550 MW kraftverk med etterrensing av eksosgassen, er henholdsvis blitt beregnet. De to kraftverkene sammenlignes deretter opp mot hverandre med hensyn på vannforbruk.

Resultatene viser at kjøletårnet utgjør det desidert største vannforbruket i begge kraftverk. Vanntap grunnet fordamping er dominant, men nedblåsing av urenheter er også en betydelig årsak til vanntap når en middelmådig vannkvalitet brukes. CO_2 fangst basert på etterrensing av eksosgass har et betraktelig større kjølevannsbehov, noe som resulterer i 17,8% høyere vanntap i kjøletårnet sammenlignet med oksy-forbrenningskraftverket.

Det nest største vanntapet skjer ved fjerning av SO_x fra eksosgassen (FGD). I luft-fyrte kraftverk er fordamping av vann hovedårsaken til vanntap i FDG-systemer. Andre vanntap er knyttet til produksjon av gips og til nedblåsing av urenheter. En høy vanndampandel i røykgassen fra oksy-forbrenning eliminerer fordampingstap, fordi røykgassens duggpunkt ofte ligger under driftstemperatur til FGD systemet. I denne studien, utgjorde vanntapet i FGD prosessen i etterrensing kraftverket åtte ganger mer enn i oksy-forbrenningskraftverket.

Et scenario hvor kraftverkene er lokalisert nært havet og sjøvann brukes som kjøling er også studert. Beregningene viser at oksy-forbrenningsanlegget bare bruker 17% av det vannet som forbrukes i etterrensingskraftverket. Denne trenden er også synlig i noe mindre grad, i tilfellet hvor kjøletårn er brukt.

En betydelig mengde kondensat er tilgjengelig i oksygenproduksjonen og CO_2 utvinningsprosessen, fordi vann tilføres systemet via inntak av fuktig luft, fukt i brenselet og via hydrogen bundet i brenselet. Gjenvinning av vann kan bidra til betraktelig reduksjon det totale vannforbruket i begge kraftverkene.

Nomenclature

Acronyms and Abbreviations

ACC	Air Cooled Condenser
ASU	Air Separation Unit
°C	Degrees centigrade
CCS	CO_2 Capture and Storage
CDR	Carbon Dioxide Recovery
CPU	Compression and Purification Unit
C.W.	Cooling Water
DCA	Direct Contact after cooler
DCC	Direct Contact Cooler
ESP	Electrostatic Precipitator
FGD	Flue Gas Desulfurization
FWH	Feedwater Heater
HHV	Higher Heating Value
HP	Higher Pressure
IGCC	Integrated Gasification Combined Cycle
IP	Intermediate Pressure
kW	Kilo Watt
LHV	Lower Heating Value
LP	Low Pressure
Mol%	Molar percent
MW	Megawatt
MEA	Monoethanolamine
NETL/DOE	National Energy Technology Laboratory/ Department of Energy
NGCC	Natural Gas Combined Cycle
ppm	Parts per million
ppmv	Part per million per volume
PC	Pulverized Coal
SRC	Selective Catalyst Reduction
Wt%	Weight percent
WGS	Water-gas-shift

Chemical Symbols

Ar	Argon
C	Carbon
$CaCO_3$	Calcium carbonate
$CASO_3$	Calcium sulfite
$CASO_3 \cdot 2H_2O$	Gympsum
CO	Carbon monoxide
CO_2	Carbon dioxide
H_2O	Water
H_2	Hydrogen
N_2	Nitrogen
NO	Nitrogen monoxide
NO_2	Nitrogen dioxide
NO_x	Nitrogen oxide
O_2	Oxygen
SO_2	Sulfur dioxide
SO_3	Sulfur trioxide
SO_x	Sulfur oxides

Greek Letters

η	Efficiency	[%]

Roman Letters

P	Pressure	[bar]
P_i	Partial pressure	[bar]
y_i	Mole fraction	[-]
W	Work	[MW]
Q	Heat	[MW]
$\dot{m}$	Mass Flow	[kg/s]
h	Specific enthalpy	[kJ/kg]
CC	Cycle of Concentration	[-]
x	Humidity ratio	$[kg_{H2O}/kg_{dry\ air}]$

Contents

Chapter 1

Introduction

1.1   Background

With more stringent environmental regulations concerning emissions of greenhouse gases, and other environmental pollutants, CO_2 capture and sufficient flue gas cleaning, plays an important role in modern coal fired power plants. Implementation of CO_2 capture requires a great portion of additional power, for plant operation. The overall power output per unit of fuel is thereby reduced, consequently, lowering the overall power plant efficiency. To deliver the same net power as that from a power plant without CO_2 capture, fuel consumption is increased. This leads to a chain reaction of increasing factors such as flue gas mass flow, emissions, power plant dimensions, need of process cooling, etc.

Electricity production from nearly all conventional steam plants rely on water as a heat sink. Besides from process cooling, there are several other subsystems that are dependent on water for operation. The flue gas desulfurization (FGD) unit, and the steam power cycle, are examples of such subsystems. Nevertheless, use of water tends to introduce minor or more significant water losses in the power plant.

In addition to increasing concerns regarding greenhouse gas emissions, the world population is in constant growth. The demand of power generation and water will keep on increasing in the future. Water is primarily needed for essential purposes such as drinking, cooking, and cleaning, secondarily for power generation. In areas with limited water resources, water needs to be used with concern. Power plant water usage and consumption has got more and more attention over the recent years, and the National Energy Technology Laboratory (NETL), has previously performed some studies, [1], [2], and [3], of water usage and consumption in steam power plants from coal generation.

1.2 Motivation

By introducing CO_2 capture to a power plant, the water usage and consumption, tend to increase in the power plant, as opposed to a power plant without CO_2 capture. For instance, post-combustion technology based on amine/ammonia absorption relies on significant quantities of circulating water within the capture process, at the same time, external process cooling is needed. Evaporative water losses are also present. The alternative and more novel CO_2 capture technology, based on oxy-combustion, appear to rely less on water, compared to the post-combustion technology.

Performance and optimization of CO_2 capture technologies is widely studied in literature. However, few studies related directly to water consumption exist for power plants with integrated CO_2 capture, as most scientific papers have a techno-economical perspective. Water is a limiting factor in many areas; therefore, a comparison of the water consumption by the two CO_2 capture technologies is a very relevant case study, in order to quantify this difference in water consumption.

1.3 Methodology

After building knowledge about coal based power generation and the technology of CO_2 capture by amine absorption and oxy-combustion, processes of water usage and consumption were identified. Further, a pulverized coal oxy-combustion power plant was established through simulations in Aspen Plus. Data for a similar CO_2 capture plant, based on MEA absorption, was extracted and used to simulate an amine absorption plant in Aspen HYSYS. From these simulations, a water material balance was generated to quantify the water consumption of the power plants. The CO_2 capture technologies were then compared in terms of water consumption and power output.

1.4 Report outline

The report is organized in 9 chapters including this introduction. Chapter 2 gives an introduction to the current CO_2 capture technologies. In chapter 3, the water usages and consumption in conventional steam power plants, and plants with CO_2 capture, are presented, and the principle of water losses is explained. Chapter 4 introduces power generation from coal-firing, including the steam cycle, cooling system and flue gas desulfurization process. In Chapter 5, the technologies behind the oxy-combustion, and post-combustion CO_2 capture, are explained in detail. Water dependent and consuming process are also pointed out. Chapter 6 provides the design basis for the simulations. In Chapter 7, a description of both the pulverized coal oxy-combustion plant, and the post-combustion capture plant, are given. Chapter 8 provides a

presentation and discussion of the water consumption in both power plants. A comparison of the water consumption, by the two different CO_2 capture technologies, is also included. Chapter 9 is the conclusion of the report, including suggestions for further work.

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Chapter 2

Introduction to CO$_2$ Capture

2.1   Greenhouse Gases and Global Warming

The natural presence of several chemical compounds in the atmosphere, mainly water vapor H_2O, carbon dioxide CO_2, methane CH_4 and ozone O_3, forms what is called the "Greenhouse effect". Radiation from the sun is trapped in the atmosphere, because these greenhouse gasses absorb some of the thermal radiation, emitted by the Earth's surface, and reradiate them back. This keeps the atmosphere, and the Earth's surface, warm.
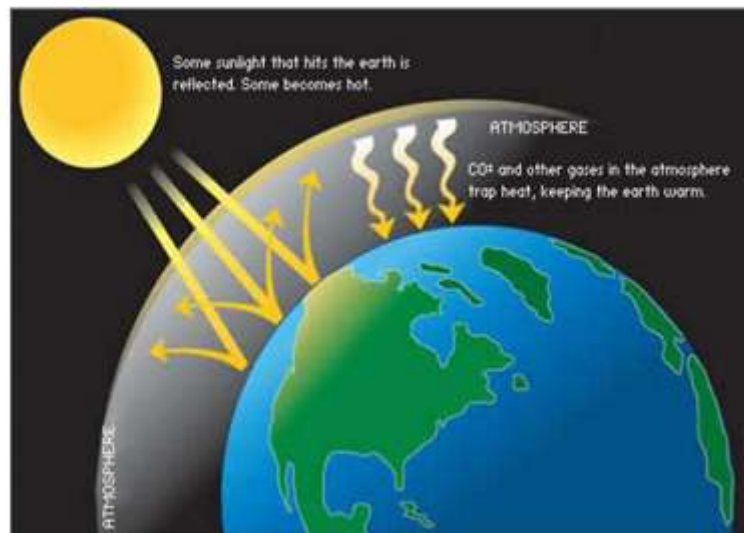
Figure 2.1: The greenhouse effect [4]

Ever since the late industrial period began in the late 1800s, the concentration of greenhouse gasses in the atmosphere has increased dramatically, with CO_2 as the dominant contributor. Research shows that the atmospheric concentration of CO_2 has increased by more than 25 % over the past 100 years [5]. This change reflects on combustion of fossil fuels such as coal, oil, and gas, as energy sources in developed

countries. Parallel with increasingCO_2 concentration in the atmosphere, measurements show that the earths' surface temperature has increased by approximately 0.74 ° C from 1906 to 2005 [5]. A mutual compliance among the world's climate researchers state that the recent temperature rise most likely is due to the observed changes in the atmospheric concentration of greenhouse gases, primarily CO_2.

If this trend, which nowadays is recognized as global warming, continues, the long term consequences are potentially melting poles, and rising sea level, which can lead to serious destruction of coastal environments [5]. Public awareness and legislation have resulted in a strict policy of reducing greenhouse gases, whereof the "Kyoto protocol", and "the Intergovernmental Panel on Climate change", are examples on such action. An important part of the solution to global warming is to rely more and more on alternative energy sources such as nuclear power, and especially renewable energy sources. The problem related to nuclear power concerns waste management and safety. Until renewable sources are mature enough to deliver, a significant portion of the energy demand, the use of fossil fuels as an energy source, will continue leaving us no choice, but to reduce emissions of greenhouse gases associated with combustion of fossil fuels. This can mainly be achieved in two ways [6] :

- ➢ Improving efficiency of power plants. In that way the fuel consumption is reduced, hence the CO_2 emission per energy output is reduced.
- ➢ By capturingCO_2 and storing it in ground or possibly in the sea.

The concept of reducing CO_2 emissions by introducing taxes is a political initiative that was implemented in the 1990s. The purpose is to motivate major oil/coal companies to reduce their own emissions by developing new low-emission technologies [7].

2.2 CO_2 Capture and Storage

Ever since the importance of reducing CO_2 emissions reached a level of action, technologies on CCS have been under development. Currently there are three main processes to capture the CO_2: 1) post-combustion CO_2 capture, 2) pre-combustion CO_2 capture and 3) oxy-combustion for CO_2 Capture. Figure 2.2 illustrates the most important steps in these processes.
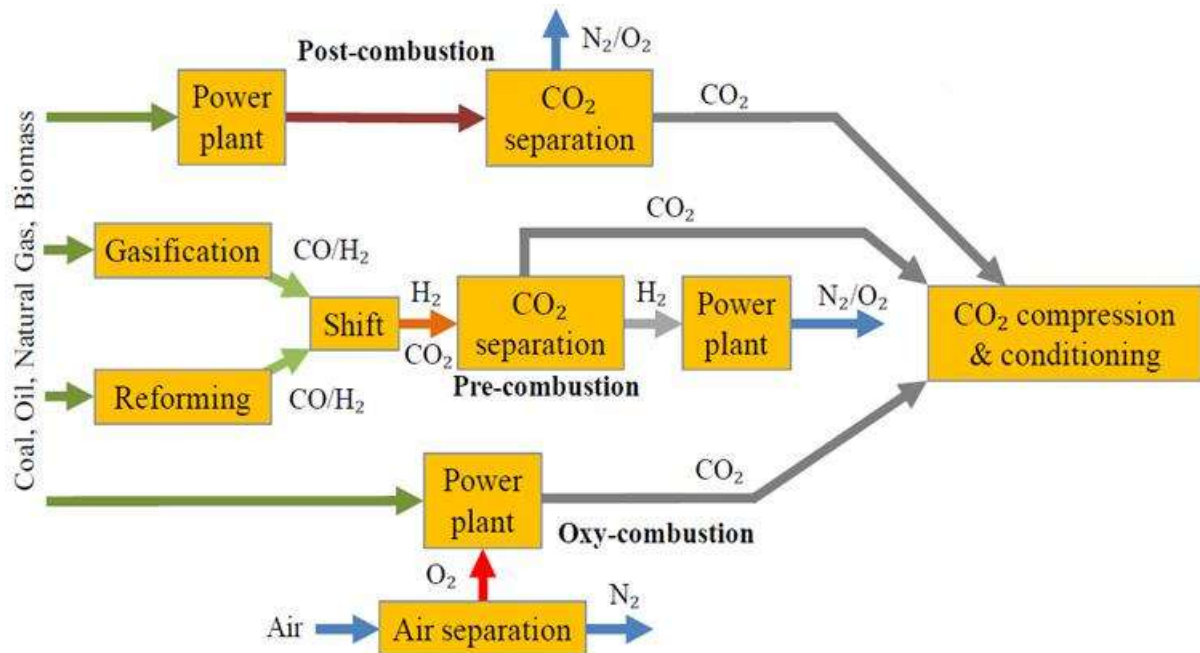
Figure 2.2: Main technologies for CCS [8]

- ➢ Post-combustion capture: a chemical solvent such as amine or ammonia is used to scrub CO_2 out of the combustion flue gas.

- ➢ Pre-combustion capture: A solid fuel is gasified with oxygen to produce syngas, a gaseous fuel consisting mainly of CO and H_2. Water-gas-shift (WGS)reaction is employed to convert CO and water to H_2 and CO_2 and a physical sorbent is then used to capture CO_2.

- ➢ Oxy-combustion capture: Pure oxygen is used for fuel combustion, thereby producing a CO2 enriched flue gas ready for sequestration once water and other impurities has been removed from the flue gas.

2.3 CO$_2$ Storage

To actually have a positive effect on global warming, the captured greenhouse gas CO_2 has to be kept away from the atmosphere, for a sufficient long period by underground storage. This comprises three possibilities; storing CO_2 in depleted oil and gas reservoirs, in unused coal seams, or in aquifers. It is considered likely that 99% or more of the injected CO_2 in such structures will be retained for 1000 years. By storing CO_2 in depleted oil and gas reservoirs, enhanced oil and gas production can be achieved. Figure 2.3 shows the concept of deep underground CO_2 storage [8].
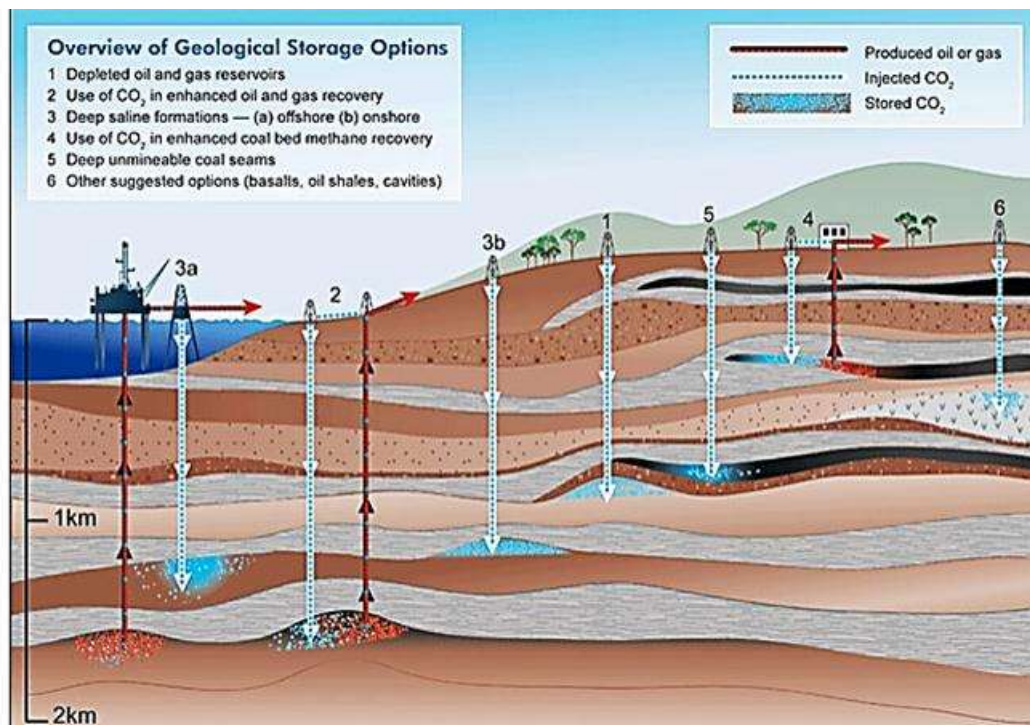
Figure 2.3: Options for storing CO2 in deep underground geological formations [8]

An aquifer is an underground layer of water-bearing permeable rock or unconsolidated materials such as gravel, sand, slit or clay. Aquifers suited for CO_2 storage are located deep in the ground, and contain highly mineralized brines (saline), in which the CO_2 dissolves. The process takes several hundred years. Among the storage options, aquifers are considered to have the greatest potential for CO_2 storage [8].

Chapter 3

Water Usage in Power Plants

3.1 Conventional Steam Power Plants

In electricity production from conventional steam power plants, supply of water is essential. Nearly all steam electric plant cooling relies on water as the heat sink, thus, the plant cooling system is very dependent on the availability of water [9]. As already mentioned, the flue gas desulfurization (FGD) and the steam power cycle, are other subsystems that require water for operation. Water is often circulating in closed loop; nevertheless, use of water tends to introduce minor or more significant water losses in the power plant. The water usage of a power plant is defined as the water needed for plant operation provided by a raw water source such as a river, lake, ocean, or groundwater. Distinguishing between water usage and water consumption, the water consumed, is referred to as the amount water, which is used and not returned to its source. Processes water losses to the environment can occur in two ways:

➢ as physical losses such as process blowdown streams, water entrained in solids, or gas streams vented to the atmosphere (evaporation)
➢ through chemical reactions such as in gasification process, or in FGD

On the other hand, water is attributed to the balance via humid air intake, moisture content in fuel, and hydrogen bound in the fuel. These are all sources of water addition to the system. In conventional coal firing without CO_2 capture, all this water leaves the system with the cleaned flue gas to stack.
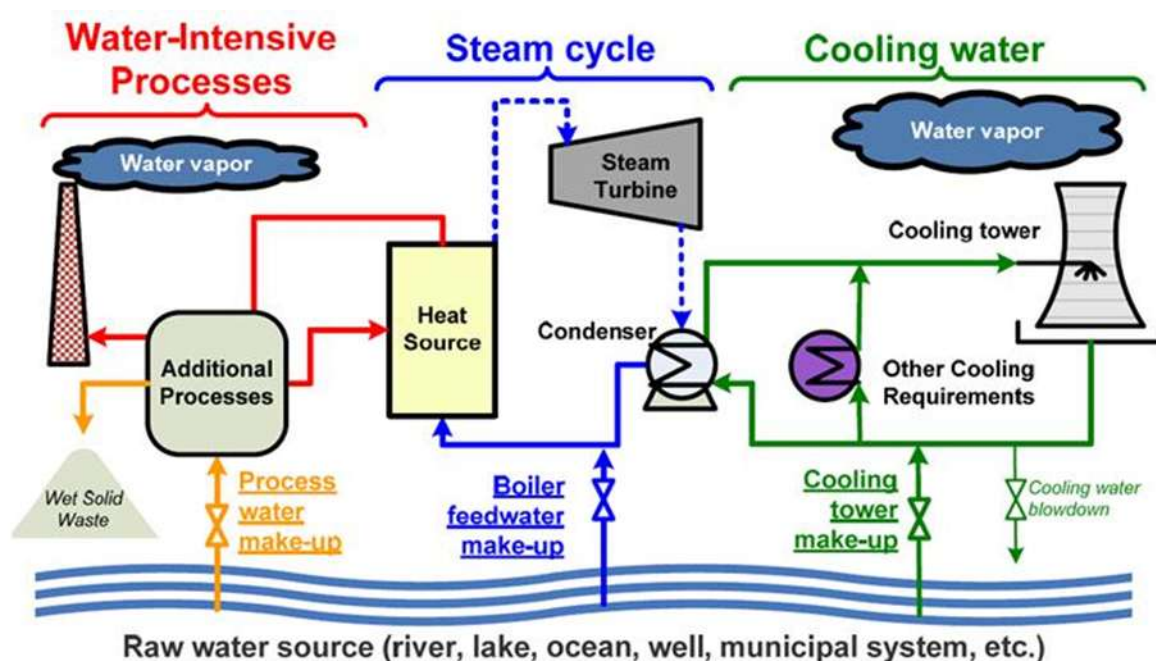
Figure 3.1: Typical water flow schematic for power plants [2]

Figure 3.1 shows a conventional steam power plant with an evaporation cooling tower. Such a power plant include several points of water losses, like cooling tower evaporation and blowdown, flue gas desulfurization, and other process blowdowns. To maintain the amount of water required by plant operation, makeup water is constantly supplied to compensate for water losses.

3.2 CO_2 Capture and Water Usage

Integration of CO_2 capture increases the water requirement per net power generation of a plant, due both to a reduction in the power plant efficiency and to the cooling water and process water requirements associated with CO_2 capture and compression.

A study performed by NETL/DOE [2], Water Requirements for Existing and Emerging Thermoelectric Plant Technologies, presents a comparison of different thermoelectric power plants in terms of water consumption and performance. Power plants considered are Pulverized Coal (PC), natural gas combined cycle (NGCC), and integrated gasification combined cycle (IGCC), including a case with carbon dioxide recovery (CDR) for each plant. The CO_2 recovery method for PC and NGCC plants is based on monoethanolamine (MEA) absorption. For IGCC plants, a high level of CO_2 recovery is achieved by a water-gas shift reactor and a physical-absorption based scrubber. Figure 3.2 show the result of the study[1].

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[1]The steam cycle blowdown is not considered in the study
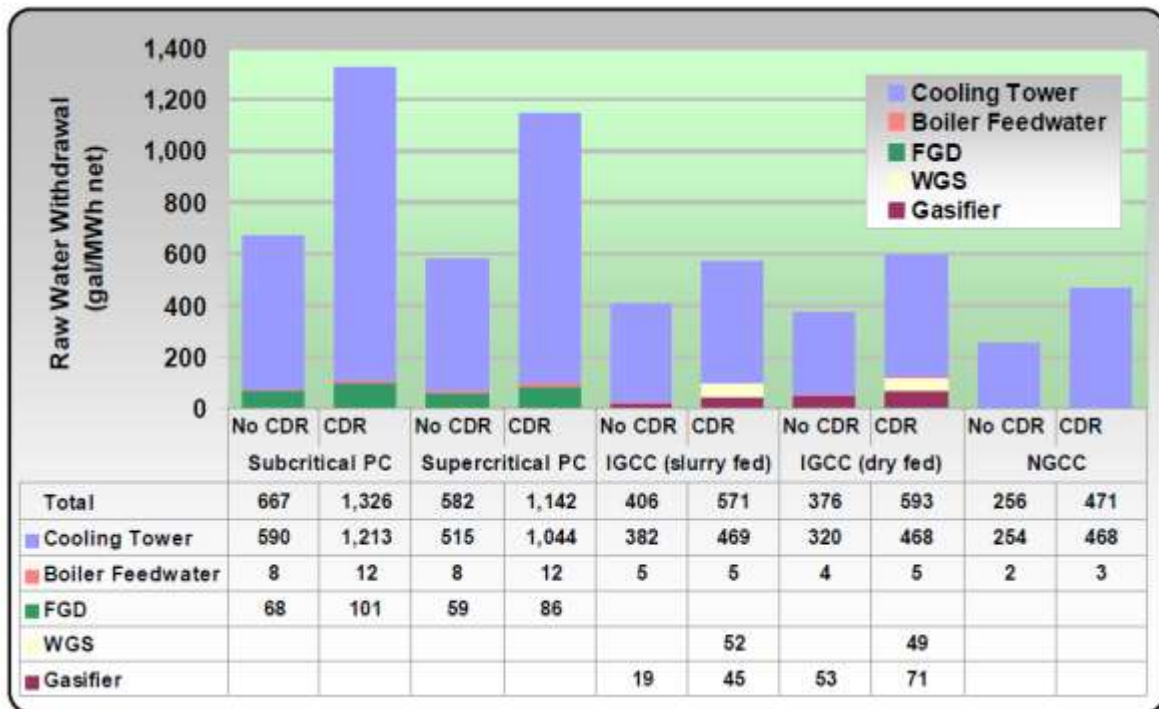
Figure 3.2: Comparison of raw water consuming factors with and without CO_2 capture for greenfield plants using wet recirculating cooling towers – net power basis [2]

	No CDR	CDR	No CDR	CDR	No CDR	CDR	No CDR	CDR	No CDR	CDR
	Subcritical PC		Supercritical PC		IGCC (slurry fed)		IGCC (dry fed)		NGCC	
Total	667	1,326	582	1,142	406	571	376	593	256	471
Cooling Tower	590	1,213	515	1,044	382	469	320	468	254	468
Boiler Feedwater	8	12	8	12	5	5	4	5	2	3
FGD	68	101	59	86						
WGS						52		49		
Gasifier					19	45	53	71		

From Figure 3.2, it can be seen clearly that implementation of CDR leads to increased specific water consumption in each plant configuration. Additional process cooling is the main cause of higher water consumption in power plants with CDR. Focusing on the coal-fired power plants, water losses due to FGD also increases when integrating CO_2 capture.

3.3 Water Availability

The water availability varies from site to site and with season. Factors affecting water sources can be listed as precipitation, ambient temperatures and stream flow. Precipitation is maybe the most important factor for water availability, as it helps refill both surface and groundwater sources. Higher temperatures are usually associated with reductions in water supply, due to increased evaporation, and uptake by heat-stressed vegetation, but also because of sublimation from glaciers. Considering a river, channel or stream, the stream flow, which constantly changes, is affected primarily by precipitation runoff in the watershed, but also by other factors such as evaporation, groundwater discharge among others. If the power plant is located anywhere near an ocean, the advantage of utilizing sea water as the coolant, instead of limited fresh water, is an opportunity.

3.4 Physical Water Losses

Physical water losses are maybe the most common cause of water loss in power plants, and often related to evaporation of water to vented streams or process blowdown.

3.4.1 Evaporative Losses

Evaporative losses typically occur in operation of evaporating cooling towers, water wash towers, FGD, or any other process where water is in direct contact with a gaseous mixture. Focusing on processes with a relatively small pressure drop; the temperature is the governing factor for the quantity of water that evaporates.

Considering a gaseous mixture at around atmospheric pressure and a temperature range of moderate temperatures, the mixture can be approximated to behave as an ideal gas. The partial pressure of any component i in this mixture, is given by the equation

$$P_i = y_i \, P_{total}$$

<div align="right">3.1</div>

where y is the mole fraction of the component i. The partial pressure of water vapor in a gaseous mixture at a given temperature, which is in contact with sufficient amount of water, can be defined as the saturation pressure of water. At close to atmospheric pressures, the mole fraction roughly equals the saturation pressure of water. Figure 3.3 shows how the saturation pressure of water vapor increases steeply with temperature. As evaporation of water requires heat, it can be concluded that at any direct contact between liquid water and a warmer gas, unsaturated with water, will cause evaporation of water.
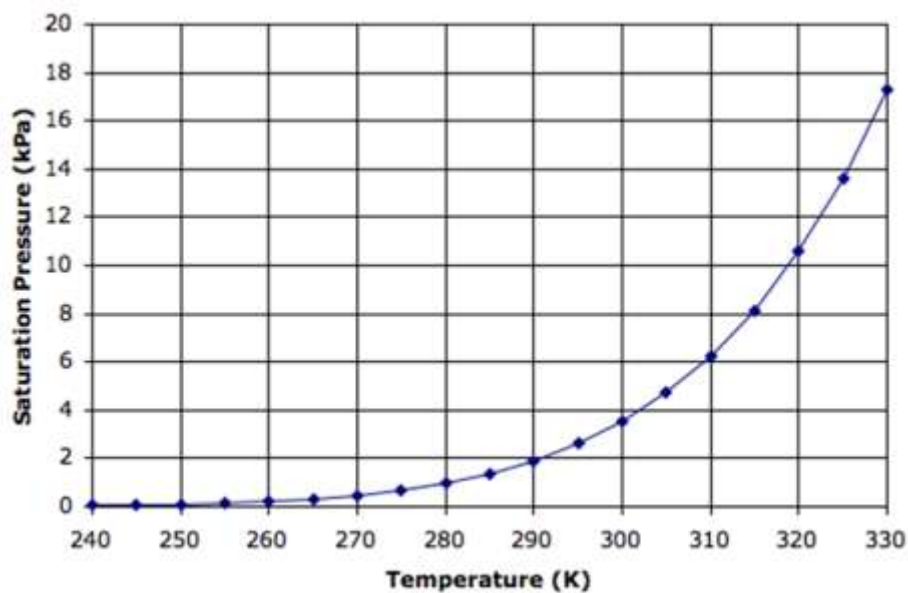
Figure 3.3: Plot of saturation pressure of water vapor at different temperatures [4]

3.4.2 Blowdown

Blowdown is a periodic discharge of recirculating water to remove buildup of solids and other undesirable components in water/steam handling equipment. Regular addition of makeup water is needed to compensate for blowdown losses. The blowdown rate varies with water quality of the circulating water. For instance, the blowdown range in the boiler/steam cycle can be from less than 1% when extremely high-quality feedwater is available, to greater than 20% in a critical system with poor-quality feedwater [10]. In evaporative cooling towers, the concentration of salt and other impurities increase as pure water is evaporated and exits with discharged air. Consequently, the blowdown rate in evaporative systems is dependent on the amount of evaporated water as well as the water purity. The water/steam used in a Rankine cycle is restricted by very high feedwater purity in order to avoid corrosion and overheating on heat transfer surfaces caused by accumulation of solids [11]. Steam cycle blowdown is commonly routed to the cooling system/cooling tower to back up makeup water [1]. Process blowdown, including cooling tower blowdown, may be returned to the original source or sent to a water treatment facility [12]. However, in this report, all process blowdown will be considered as contribution to the total water consumption.

Chapter 4

Power Generation from Coal Power Plants

The intention of the following chapters is to give an understanding of how conventional power generation from coal combustion is carried out, including the main components. Water using and consuming processes are also to be identified.

4.1 Pulverized Coal Combustion

The most common process for coal-fired power generation is pulverized coal combustion, and takes place close to atmospheric pressure. Coal is being pulverized to a fine powder in a mill, before it is fed to the boiler. A drying process might be introduced prior combustion, depending on the coal moisture content. Pulverized coal is blown together with preheated air into the boiler, where combustion takes place at around 1300-1700 °C, depending on the heating value of the coal [13]. At various heights of the boiler, secondary and tertiary air, may also be introduced. The combustion heat is rejected to a steam generator inside the boiler, producing high pressure steam of the incoming feed water. In the steam cycle, also referred to as Rankine cycle, heat is converted into work by expanding steam through a turbine. The steam exhaust is condensed in a condenser, and pumped to a higher pressure, before the feedwater is returned to the boiler. The efficiency of the coal power plant, based on the lower heating value of the fuel, can be expressed as follows

$$\eta_{power\ plant} = \frac{W_{net}}{\dot{m}_{fuel} \cdot LHV_{fuel}} \qquad 4.1$$

The flue gas exiting the boiler passes through a selective catalytic reduction (SCR) process for NOx removal, where NO is reduced to N_2 and H_2O by reaction with ammonia in a catalytic environment. The flue gas is then used to preheat combustion air. Fly ash is collated in an electrostatic precipitator, or in a fabric filter, and the flue gas enters a wet scrubber for SO_2 removal by limestone absorption. Cleaned flue gas is then sent to stack.

4.2 Ideal Rankine Cycle
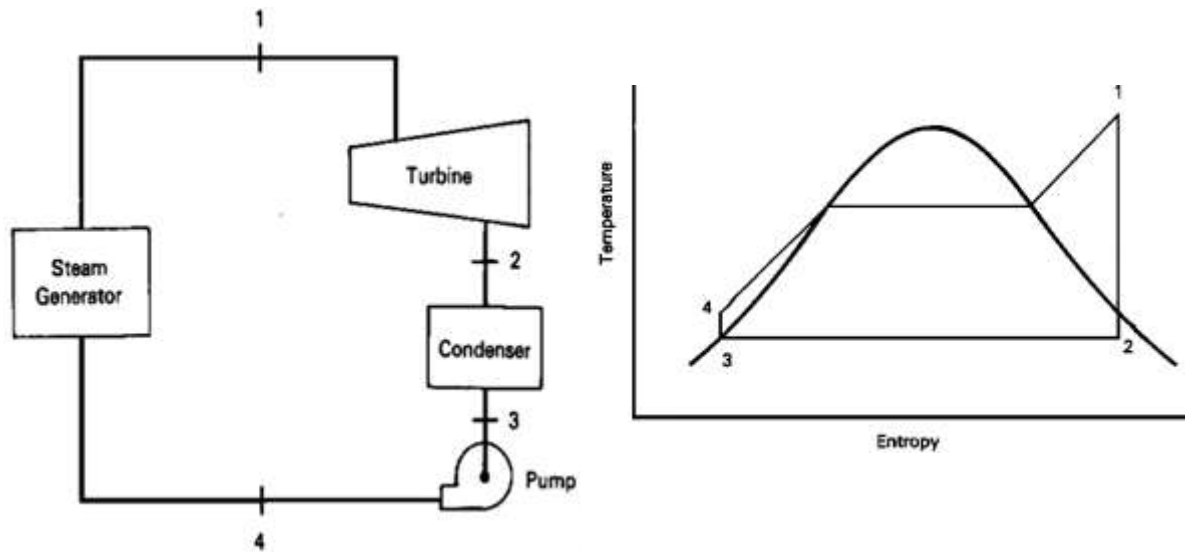
Figure 4.1: Ideal Rankine cycle. Flow diagram and T-S diagram [14]

Figure 4.1 shows a steam power plant flow diagram of a reversible Rankine cycle. Superheated vapor from the boiler enters the turbine at state 1, where it expands isentropically to the exit pressure at state 2. The condenser pressure in state 2 is limited by the steam quality. Presence of liquid droplets can cause erosion on the turbine blades that can result in decreased turbine efficiency. Common practice is to keep the steam quality at least around 90% [15]. Steam is then condensed at constant pressure, and temperature, to saturated liquid at state 3. The condenser is part of a cooling system that will be introduced in Section 4.5. The saturated liquid, often referred to as the feedwater, is then pumped to the boiler pressure in state 4. The feedwater enters the boiler again, where it is heated along a constant pressure line to sate 1. Assuming that the steam turbine and feedwater pump operate adiabatically, the thermal efficiency can be expressed by the following equation

$$\eta_{thermal} = \frac{W_{cycle}}{Q_{in}} = \frac{(h_1 - h_2) - (h_4 - h_3)}{h_1 - h_4} \qquad 4.2$$

In general, thermal efficiency tends to increases with the average temperature of heat addition, or by lowering the temperature of heat rejection [15]. It is from the thermodynamic point of view, desirable to have as high boiler pressure and as low condenser pressure, as possible.

Due to irreversibilities in the processes, mainly in the turbine expansion (state 1-2), the temperature-entropy diagram will not have constant entropy lines. Figure 4.2 shows a T-S diagram of an irreversible Rankine cycle.
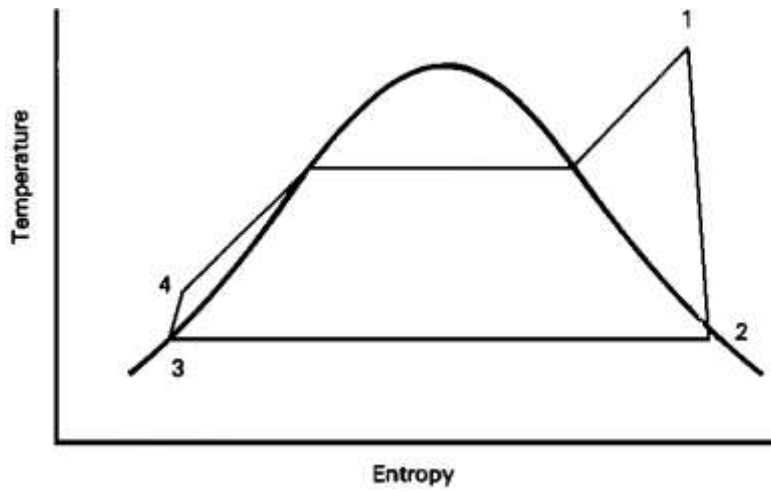
Figure 4.2: T-S diagram of a irreversible steam power plant [14]

4.3 Improved Performance of a Rankine Cycle

To improve the performance of the standard Rankine cycle, some common modifications including reheat and feedwater preheating, are comprised to increase the average temperature of heat addition. Figure 4.3 shows a flowsheet and T-S diagram of such an improved Rankine cycle.
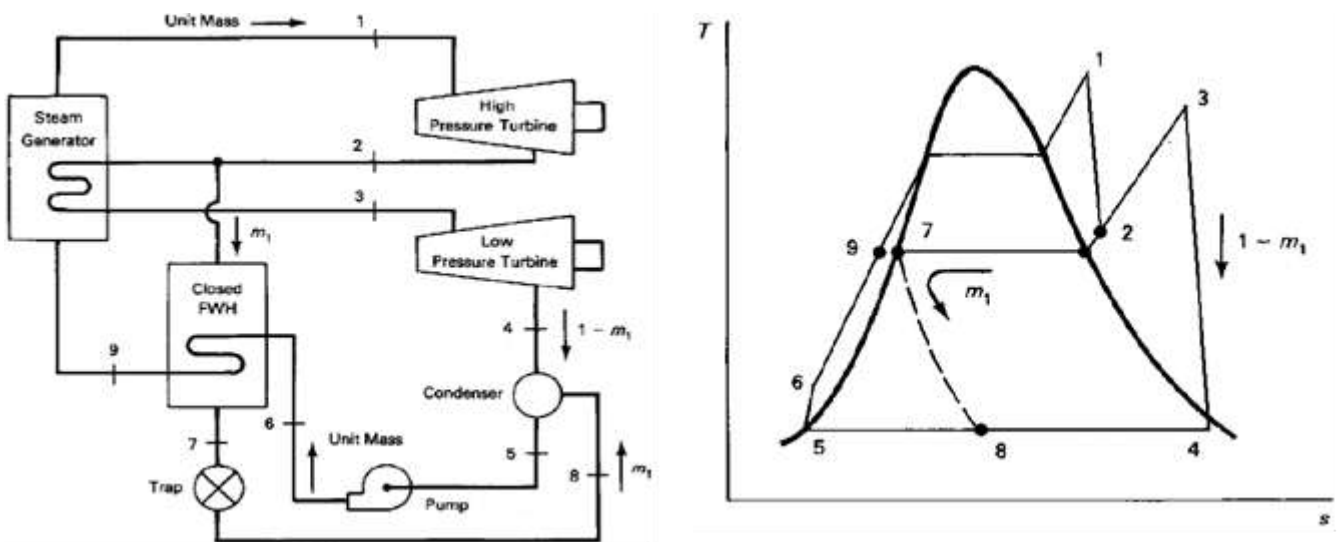
Figure 4.3: Regenerative Rankine cycle with reheat and one closed feedwater heater. Flow cycle and T-S diagram [14].

Reheat: Employing reheat, the steam at state 1 is not expanded to the condenser pressure in a single stage, but reheated from state 2 to state 3, as shown in Figure 4.3,

and then expanded to the condenser pressure in state 4. The advantage of this is to operate with a higher boiler pressure, and yet avoiding low steam quality at the exit of turbine. Reheat will also contribute to increase the average temperature of heat addition. Modern gas turbines are designed with both one and two reheat sections.

Feedwater heating: Another common configuration that contributes to increasing the average temperature of heat addition is boiler feedwater preheating. The heat is supplied by steam extractions from the turbine, as illustrated in Figure 4.3. Heat is exchanged either, directly or indirectly, in an open or a closed preheater. In power plants with several preheaters, it is common to have at least one open preheater operating above ambient pressure to vent oxygen and other dissolved gases from the cycle. Such deaeration is a consequence of Henry's law; the solubility of a gas in a liquid decreases with its partial pressure. This unit is often referred to as a deaerator, and is needed to maintain the purity of the working fluid in order to avoid corrosion of equipment [15]. Steam extractions reduce the mass flow of steam through the turbine, thus, the net turbine power output is reduced. Consequently, feedwater heating decreases the power plant efficiency, but tends to increase the thermal efficiency. Additionally, feedwater preheating is employed to keep the flue gas outlet temperature above its dew point. The presence of acidic gases such as NO_x and SO_x from acid in contact with free water, which is very corrosive.

Supercritical and ultra-supercritical cycles: New material technologies allow the steam pressures at state 1 to operate above the critical pressure of 221 bar, thereby increasing the average temperature of heat addition. Due to limitations of material tolerance, the steam pressure can only be increased until a certain point. The positive effect of increasing the pressure will diminish at high pressures due to increased feedwater pump work. However, the pressure at which the increase in efficiency stops is considerably higher, than the pressures allowed by current material technology [16].

4.4 Cooling Water Requirements

As previously discussed, low condenser pressures are associated with higher turbine power output. At the same time, a low condenser pressure causes larger dimension of the condenser and steam turbine [6]. The condenser can be considered as a heat exchanger and is the largest cooling demanding process in a power plant. Figure 4.4 shows a T-Q diagram of a condenser, where latent heat is rejected indirectly to a coolant.
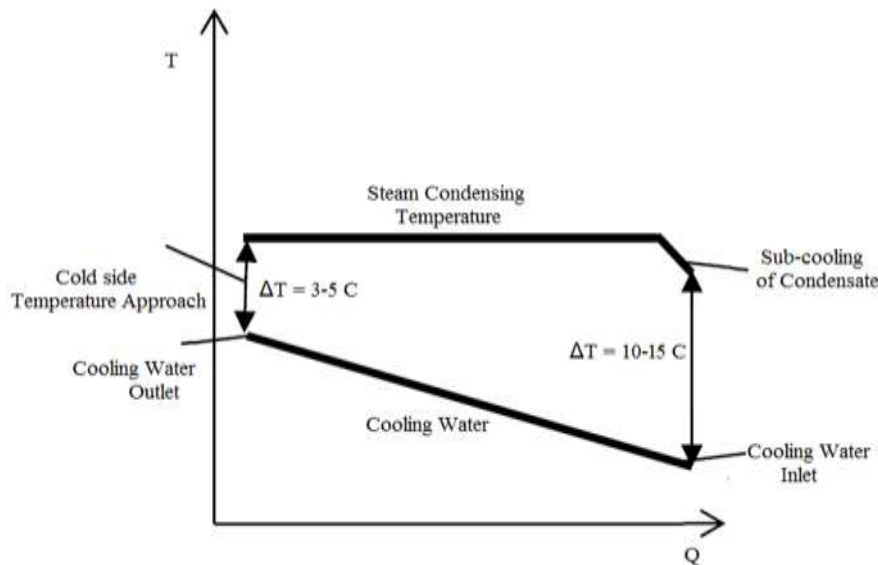
Figure 4.4: T-Q diagram of a condenser

As already pointed out, the water usage and consumption of a power plant is closely related to the plant cooling demand. Increased cooling requirements, also lead to larger dimensions and investment costs of the cooling system. Additionally, mechanical work is required in order to drive pumps in the cooling water circuit. Two main factors are important, when determining the amount of cooling water required.

➢ Cooling water temperature range
➢ Cold side temperature approach

The cooling water temperature range, often referred to as the cooling range, depends on the type cooling system used, and its site conditions. If a lake, river, or ocean, is used as heat sink in a once though cooling system, see Section 4.5, the cooling range is not only limited by the temperature of the cooling source, but also by the fact that the ecosystem in the water should not be affected. In an evaporative cooling tower, the cooling range is dependent of the wet bulb temperature of the air. Evaporated water will be returned to the environment as rain, through the lifecycle of water. The cold side temperature approach [2] is a parameter determined by the heat transfer

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[2] The minimum temperature difference between the cold and hot outlet of a heat exchanger

characteristics of the heat exchanger, and the flowing mediums. Metallurgy, size, pressure, temperature, composition, and phase are factors that affect the heat transfer. Purposed heat exchange reference values are given in Table 4.1

Table 4.1: Reference values for cold side temperature approach in heat exchangers [17]

Phase	Cold side temperature approach, °C
Gas/Gas	25
Gas/Liquid	10
Condensing/Liquid	3

When all stream temperatures are known, the cooling load Q can be used to calculate the mass flow of required cooling water using the following expression

$$m_{CW} = \frac{Q}{(h_{cold,in} - h_{cold,out})} \qquad 4.3$$

4.5   Cooling System

After process cooling, the heated cooling water is returned to its source, or passed to a heat sink, where it is cooled and recirculated. This arrangement is referred to as the power plant cooling system. The three most common types of cooling systems are [16]

- ➢ Once through water cooling using river water or see water
- ➢ Direct/indirect air cooling in an air cooled condenser
- ➢ Evaporating cooling in a wet cooling tower

In a once-through cooling system, chilled water is taken directly from a water source like the ocean, a river, or a cooling pond, used for process cooling, and then returned to its source. This is the most economical type of cooling system, and allows the lowest condenser pressure [13]. If water is very limited near the coal mine, it may be more convenient to locate the power plant close to a large water source, and transport the coal instead.

In an air cooled condenser (ACC), no cooling water is required. However, the overall power plant efficiency will be somewhat reduced due to the significant amount fan work required in an ACC. This cooling system is used in regions, where water resources are very limited, or simply just to minimize interruption with the environment in the power plant surroundings [16]. Thermal duty is removed by flowing air. Fans are used to impel air to flow through the finned tube bundles, as shown in Figure 4.5. This type of dry cooling system is three to four times more expensive, than a wet recirculating system [2].
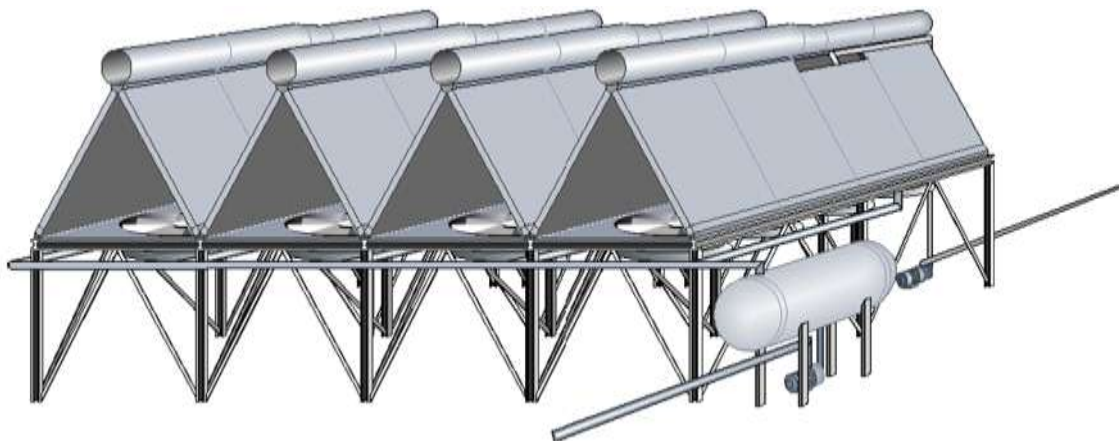
Figure 4.5: Direct Air-Cooled Condenser [18]

4.5.1   Evaporative Cooling Towers

Evaporative cooling towers are categorized as wet recirculating system, and widely used, when water resources are limited to some extent. The principal of the cooling system is the same as that for a once through cooling system, except that most of the cooling water is circulating in a loop. Heat rejection takes place in a wet cooling tower

by latent heat transfer to the ambient air. Figure 4.6 shows an illustration of a wet mechanical draft cooling tower. Temperature reduction is obtained by evaporation of water droplets in the unsaturated air. Hot cooling water is sprayed as fine droplets in a counter-current flow with the up-raising air stream. The heat transfer is taking place on the surface of the water droplets, at the adiabatic saturation temperature of the ambient air. This temperature is commonly referred to as the wet bulb temperature, and is theoretically, the lowest temperature that the water can reach [19]. Consequently, the wet bulb temperature sets a lower limit for the cooling water temperature, and the condenser pressure. Water is evaporated until the air is saturated with water vapor and exits at the top of the cooling tower. Cooled water droplets are collected in the bottom of the tower and return for process cooling. Makeup water is required to replace evaporation, blowdown and drift losses in the cooling tower.

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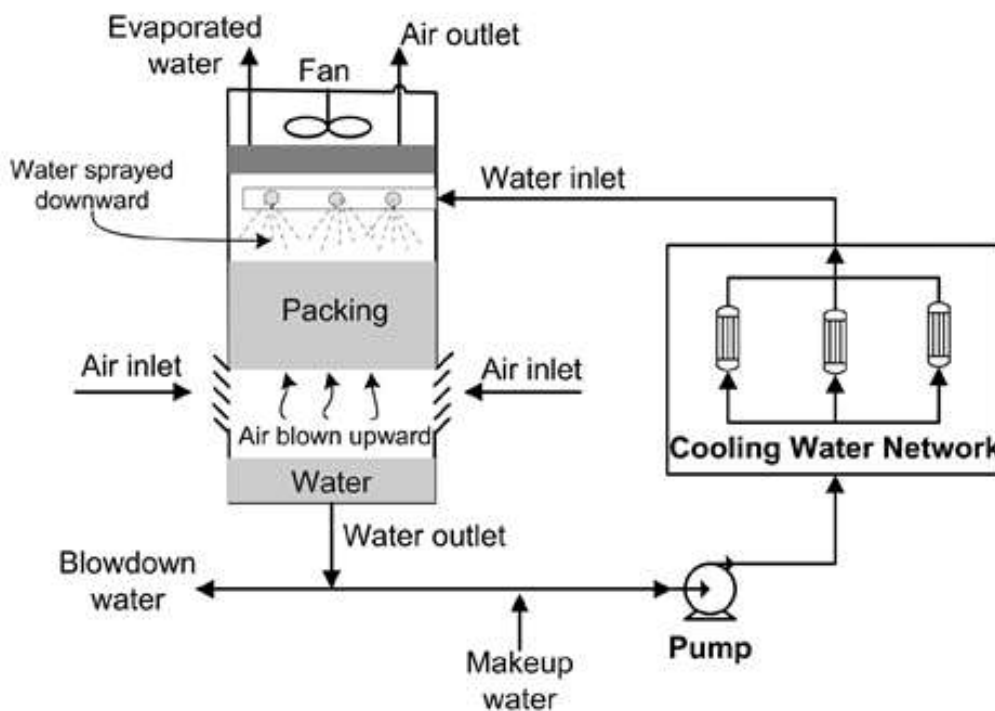
Figure 4.6: Schematic of mechanical draft, evaporative cooling tower [19]

4.5.2 Water Losses

Water losses in wet cooling towers are mainly due to evaporation of the cooling water itself that leaves with discharged air. The concentration of salts and other impurities will increase in the circulating water, as pure water is evaporated. To avoid buildup of solids and other undesirable components, within the cooling tower, water needs to be blow down and replaced with fresh makeup water. Drift losses due to entrainment of fine water droplets in the discharged air are also present. For modern cooling towers, the drift losses can be as low as 0,001% of the circulating water [2]. Higher cooling

loads, lead to more latent heat transfer and thus, increased evaporative losses. Consequently, the evaporation rate is dependent on the temperature range, and mass flow, of the cooling water. Fresh water makeup is added at the same rate as water is lost. The total amount of makeup water can be expressed by

$$m_{makeup} = m_{evap} + m_{drift} + m_{blowdown} \qquad 4.4$$

The quantity of blowdown water is estimated using the following equation

$$m_{blowdown} = \frac{m_{evap}}{CC - 1} - m_{drift} \qquad 4.5$$

Cycle of concentration (CC) is a term related to the water quality in the cooling tower. It is a measure of how many times the water is recirculated before blowdown. The operation at a higher cycle of concentration reduces cooling tower blowdown losses.

The fresh water needed to compensate for evaporative losses are estimated based on a material, and energy balance of the entering and exiting air and water, respectively.

$$m_{FW} = m_{W,evap} = m_{W,in} - m_{W,out} = m_a(x_1 - x_2) \qquad 4.6$$

$$\frac{Q}{m_a} = (h_{in} - h_{out}) - h_{FW}(x_1 - x_2) \qquad 4.7$$

Q represents the cooling load, m_a and m_w, the mass flow of dry air, and water, respectively. h is the enthalpy, and x, the humidity ratio of the moisture air.

4.6 Wet Flue Gas Desulfurization

The flue gas desulfurization process is of great importance, when considering the water consumption in coal fired power plant. The principle of most common desulfurization method used in industry will be explained in the forthcoming chapter. Description of water using process, and sources of water losses, will also be included.

Dissolution of SO_x in water leads to the formation of sulfuric acids, which are responsible for acid rain. SO_x emissions from coal fired power plants have to be controlled in order to avoid such consequence. Stringent environmental regulations against SO_x emissions have been implemented in many countries due to the harmful effect on the environment. The sulfur content in the coal varies widely with coal seam, geological age and location and is mostly found as pyrite or in organic forms. The oxidation of sulfur can be described by the simplified reaction scheme

$$S + O_2 \rightarrow SO_2$$

$$SO_2 + O + M \leftrightarrow SO_3 + M$$
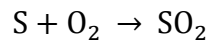
where M represents an inert.

However, under high temperatures such as during combustion, the equilibrium conversion of SO_2 to SO_3 is low, and as the gas cools downstream through the boiler, the reaction rate slows down. Consequently the amount of SO_3 is no more than 2-5% of the converted SO_2 [20]. SO_2 removal from the flue gas is there the main concern. The worldwide dominant process for SO_2 control from coal fired boilers is the wet flue gas desulfurization (FGD) process. SO_2 is absorbed from the flue gas using a solvent of an alkaline agent like limestone ($CaCO_3$) in a scrubber/absorber. The process takes place in an aqueous environment, where slurry of limestone is sprayed counter currently into the rising flue gas, scrubbing SO_2. The main reaction product, calcium sulfite ($CaSO_3$) has no commercial value and is used as landfill or ends up as waste [21]. By introducing oxidation air or pure oxygen to the system, gypsum ($CaSO_4 \cdot 2H_2O$) can be produced through forced oxidation of sulfite. Gypsum is commercially used in a wide variety of applications such as manufacturing of wall boards, in fertilizer, cement industry etc. Limestone forced oxidation system has become a very common SO_2 removal process and removal efficiencies up to 99% have been proved by this technology [1].The chemical process of a WFGD with forced oxidation is very complex, involving both dissolution of reactants and stripping of CO_2 gas. A very simplified reaction scheme is shown below.

$$SO_2(aq) + H_2O \leftrightarrow HSO_3^- + H^+$$

$$CaCO_3(s) + HSO3^- + H^+ + 1/2\ O_2 + H_2O \rightarrow CaSO_4 \cdot 2H_2O\ (s) + CO_2(g)$$
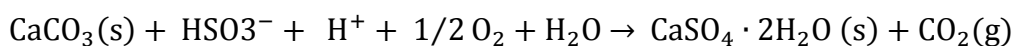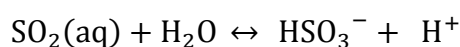
4.6.1   Process Description

A very basic block diagram of a wet FGD process is shown in Figure 4.7.
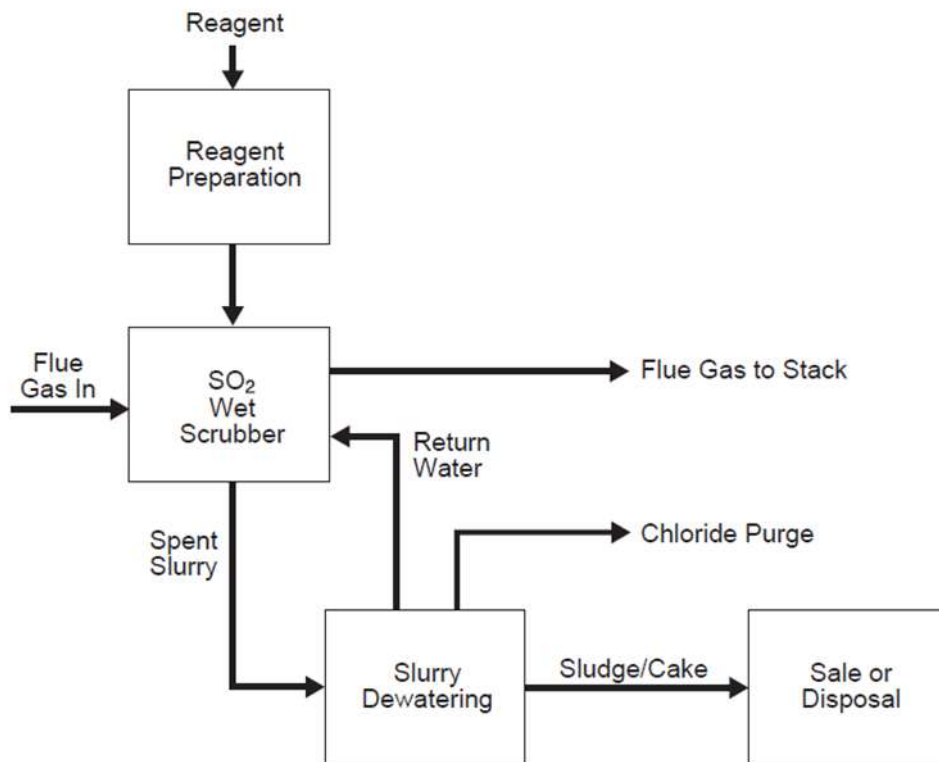
Figure 4.7: Block diagram of wet FGD process [7]

In the reagent preparation system, limestone is milled to fine particles, which are then mixed with water in the feed tank to limestone slurry containing of 25-35% of suspended limestone solids [21]. The flue gas enters the absorber in the temperature range of 121 to 177 °C and flows upward the absorption tower, as shown in Figure 4.8
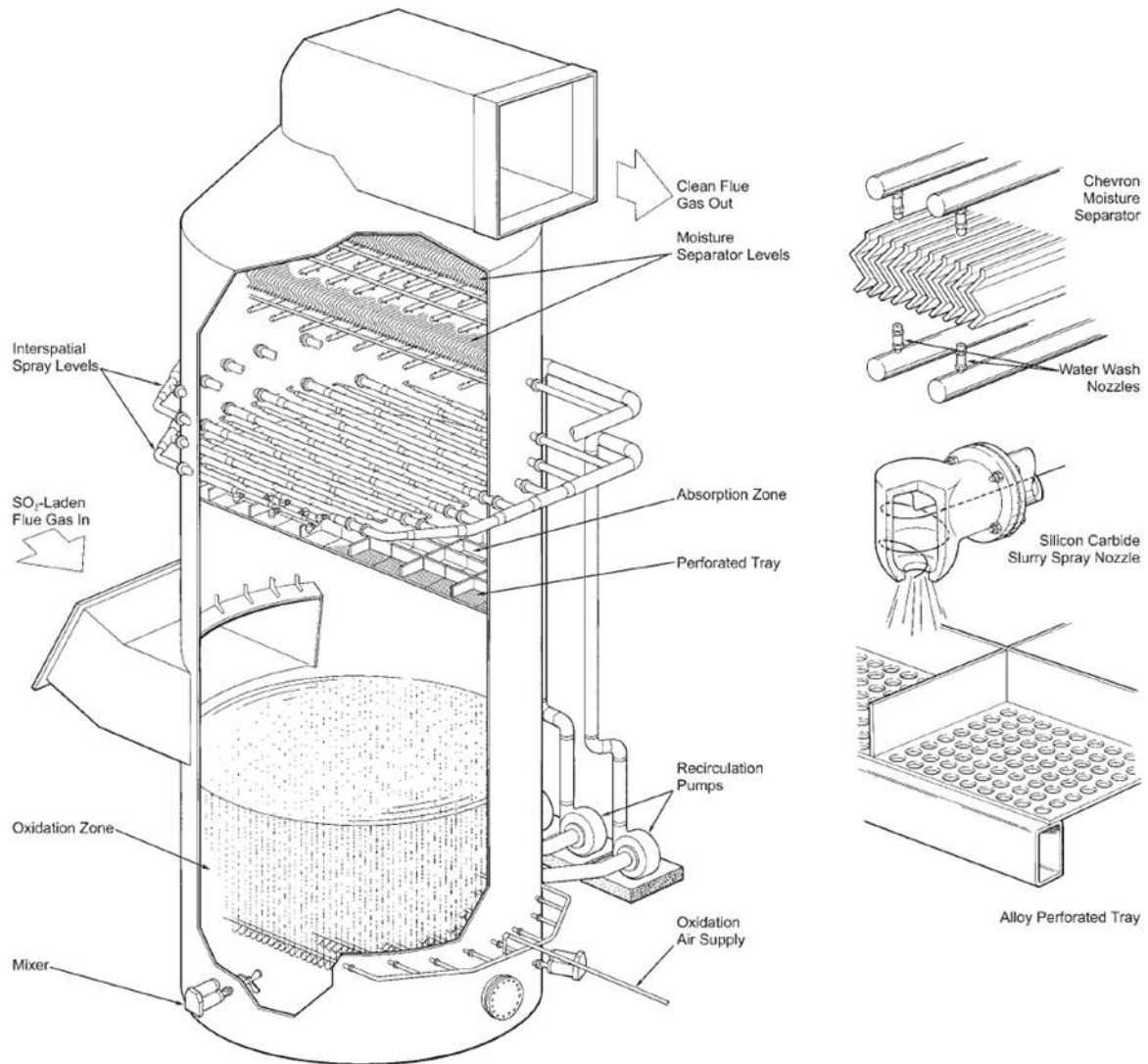
Figure 4.8: Absorber arrangment, wet FGD [22]

Slurry is pumped to a higher pressure, and sprayed as fine droplets into the absorption zone, where SO_2 is scrubbed from the flue gas by reaction with dissolved limestone. The flue gas passes through a moisture separator to remove slurry droplet entrained in the gas. Wash water is used to drain the droplets back to the absorber. Clean gas exits at its adiabatic saturation temperature, cooled by evaporating water. At the bottom, sulfite is oxidized to gypsum in the oxidation zone. Gypsum slurry composted of 20% suspended solids, is continuously extracted and sent to dewatering. Remaining limestone slurry is recycled back to the spray section in the absorber by a pumping system. In the dewatering system, a set of hydrocyclones are used to separate solids from liquid. A portion of the hydrocyclone overflow is sent to the purge system in order to remove fine particles and chloride. Waste water from the purge requires treatment before it is released. The sludge/cake is washed with freshwater in a vacuum filer to remove total dissolved salts and further dewatered to contain between 10% and 20% free moist, depending on application [22].

4.6.2 Oxy-combustion FGD

A major part of the flue gas in oxy-combustion is recycled for temperature control in the combustion chamber. In theory the sulfur dioxide containing flue gas could be recycled, and then the SO_2 would be removed by an FGD in the CO_2 compression and purification unit (CPU). If this was the case, the sulfur dioxide concentration in the boiler would increase significantly. The factor of increase would generally be as large as 3.4 to 3.5. To avoid excessive corrosion, the practical design limit in conventional coal firing is set to 3.5 wt% sulfur in the coal. With flue gas recycling, the limit would be a coal containing 1 wt% sulfur [23].

4.6.3 FGD for Air-fired Combustion with Post-Combustion Capture

MEA is the most common amine used as solvent in post-combustion CO_2 capture. The presence of SO_2 in the flue gas leads to degradation of this amine. A chemical reaction between SO_2 and MEA forms a non-reclaimable corrosive salt, which is very harmful for plant operation. Solvent losses are often more expensive than installing a FGD system. A typical limit of SO_2 in the flue gas, when applying chemical absorption by MEA is less than 10-20 ppmv SO_2 [13].

4.6.4 Water Usage and Consumption

The major water loss considering air-fired configurations in the FGD system is due to evaporating water to the hot flue gas. However, the flue gas from oxy-combustion has a much higher partial pressure of water vapor, thus evaporative losses are strongly reduced if present at all. There are also some water losses to the gypsum cake, which consists of 18.4 wt% water and additionally contains between 10-20% free moisture. Other water losses occur in the purge system. The amount is dependent upon the need to purge dissolved solids or the need to remove impurities that arise from the coal, sorbent and makeup water. The purge rate is often controlled by the chloride concentration, with a typically limit of 20 000 ppm or less [24].

Water is needed for several operations within the FGD system. Reclaim water is collected and reused in slurry preparation and in maintenance of the water level in the absorber reaction tank. The water quality regulations are not very stringent in a wet FGD process. Cake wash and filter pumps requires fresh water from river or lake, otherwise all other water demanding processes can use blowdown water or seawater [24].

Chapter 5

CO_2 Capture Technologies

In the forthcoming chapters, the technology behind CO_2 capture based on both post-combustion and oxy-combustion CO_2 capture will be introduced. Main components will be presented, and sources of water losses and cooling water relying processes will be identified.

5.1 Post-Combustion CO_2 Capture

Post-combustion capture is referred to as CO_2 capture from flue gases from air-firing of fossil fuels and biomass. The most common capture processes involve absorption, a chemical process, where molecules are taken up by a liquid (solvent). Two different types of solvents are used; chemical and physical solvents. Using a physical solvent, the absorbed molecules dissolve in the solvent, but remain chemically unchanged. In chemical absorption, the solvent reacts with the gas component to from a new chemical compound. Whether chemical of physical absorption is used, depends on the partial pressure of the gas component [8]. The principle of gas separation by absorption mainly involves a two-step process; after the CO_2 has been absorbed by the solvent, it has to be desorbed in a stripper to release pure CO_2 and to regenerate the solvent. Membrane technologies can also be applied to eliminate the CO_2 from the flue gas, but require further development to become competitive with the current post capture technology based on chemical absorption/desorption [8]. In air-fired configurations, the most efficient way of capturing CO_2 is through chemical absorption, using a chemical solvent such as monoethanolamine (MEA) or ammonia to bind the CO_2 in a new chemical from.
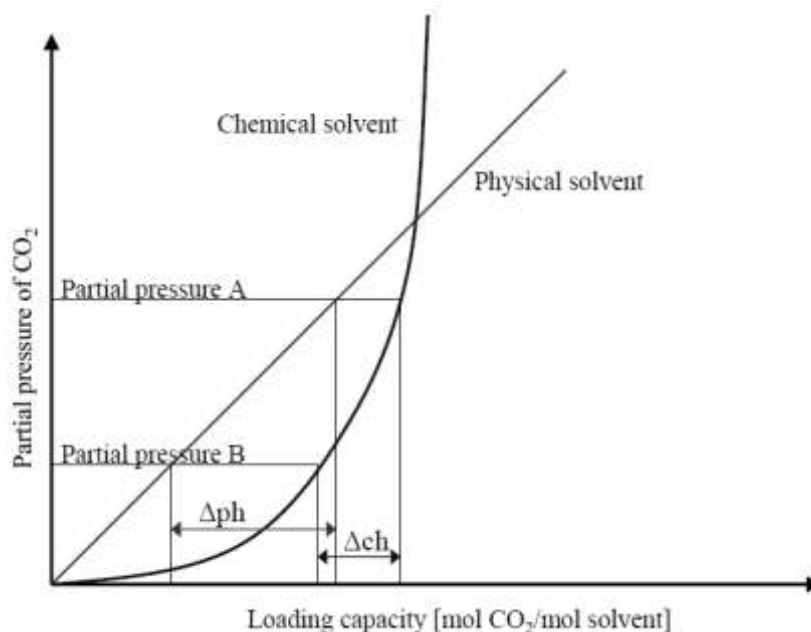
Figure 5.1: Illustration of the difference between a chemical and physical solvent for a given temperature [13]

As seen from Figure 5.1, a chemical solvent has a higher loading capacity at lower partial pressures of CO_2. The loading capacity of a solvent is defined as the number of moles CO_2 absorbed, per mole of solvent. A typical flue gas from coal contains up to 15 mol% CO_2, varying with the type of coal used [13]. At atmospheric conditions, the partial pressure of CO_2 in the flue gas is around 0.15 bar, consequently a chemical solvent is preferred. For higher partial pressures of CO_2, such as in gasification, a physical solvent serves much better.

5.1.1 CO₂ Capture by Chemical Absorption

Chemical solvents are already widely used in the industry both in natural gas processing and in CO_2 capture from petrochemical plants. Demonstrations of post-combustion capture from conventional coal firing are currently underway [20]. Because most of the flue gas from air-fired combustion is composed of nitrogen (~70%), the CO_2 fraction is relatively small (≤15%), resulting in large dimensions of the capture process. For retrofitting of existing coal fired power plants, CO_2 capture by absorption is an attractive option to reduce CO_2 emissions. Capable capture efficiencies have been proved to be over 90% of the CO_2 from the feed gas, and at very high purity levels.

5.1.2 Chemical Solvent

The most common amine used for flue gas CO_2 removal is monoethanolamine (MEA), with the chemical formula CH_2-CH_2-OH-NH_2. Due to the corrosive characteristics of MEA, the solvent exists in an aqueous solution of no more than 12-30 wt% dissolved MEA. The optimum process conditions for MEA absorption are within a temperature and pressure range of 40-55 °C and 1-5 bar, respectively, and

suited for gas streams with low CO_2 partial pressures [13]. Such operational conditions are very convenient for a typical flue gas coming from air-fired combustion of pulverized coal or natural gas, when as much heat as possible is recovered.
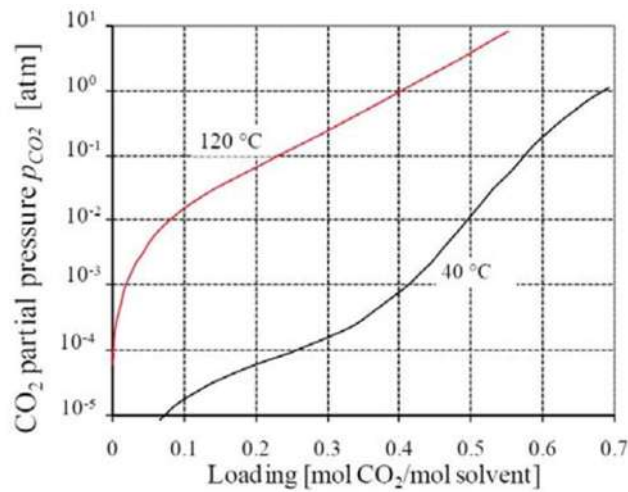
Figure 5.2: Loading curve for an alkanolamine as a function of partial pressure at two different temperatures [13]

Figure 5.2 shows a loading curve for different partial pressures of CO_2; at a low and an elevated operational temperature. The loading curve represents the equilibrium between the CO_2, and the solvent. At the elevated temperature, the loading capacity is significantly reduced, which is ideal for desorption. At the lower temperature, the loading capacity is at its highest, suited for the absorption process.

5.1.3 Process Description

Figure 5.3 shows a flowsheet of a typical post-combustion capture process based on temperature swing absorption. The pressure only varies slightly through the process; absorption takes place at atmospheric pressure and desorption at a pressure of 1-2.7 bars. This is a typical capture process for a feed gas in which the pressure is just above atmospheric, and MEA is used as the chemical solvent [13].
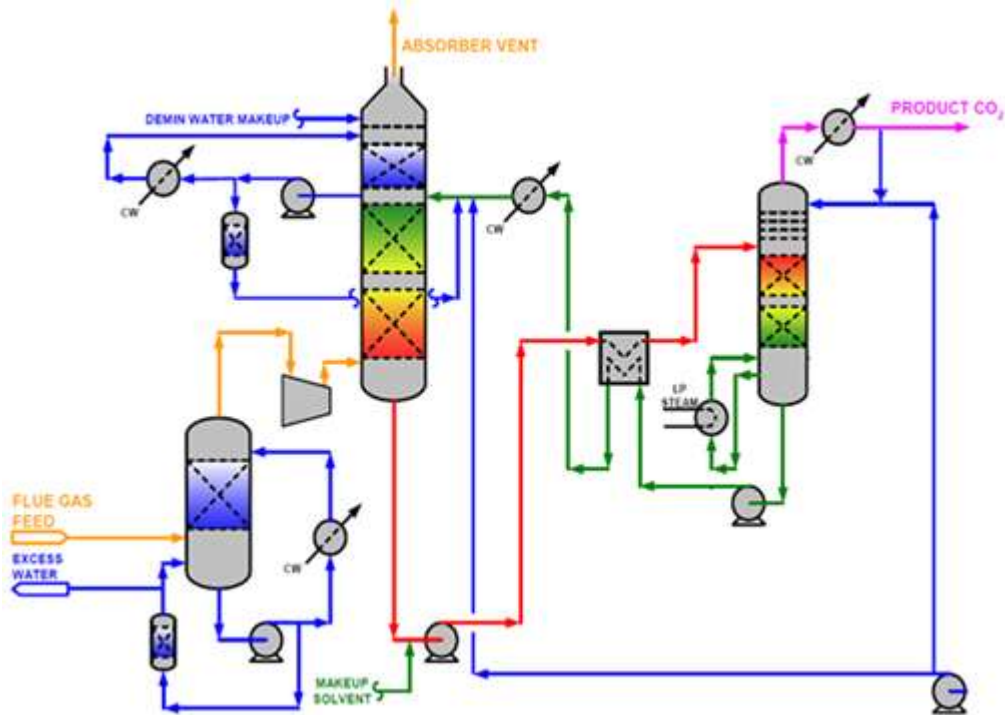
Figure 5.3: Typical flowsheet for MEA absorbtion process [25]

Referring to Figure 5.3, before the flue gas enters the absorber, it has to be cooled in a direct contact cooler (DDC). Depending on the sulfur content in the fuel, additional SO₂ removal may be necessary in order to meet the stringent limitations of SO₂ content in the absorption process. A polishing step might be combined in the DCC, or an existing FDG unit may be improved. In the DCC, the flue gas temperature is reduced to the low temperature bounded by the solvent. Due to the temperature drop, a major part of water vapor is condensed from the flue gas, and excess water is available from the DCC. This DCC excess water contains some dissolved carbon dioxide and other water soluble impurities. A portion of this water is used to maintain the water balance around the DDC. The rest can be sent to a water treatment facility to be used as plant makeup water. The cooled flue enters at the lower part absorber, induced by a fan in order to overcome the pressure drop in the abosber. The regenerated solvent, commonly referred to as the lean solvent, is equally distributed at the higher level of the absorber. CO₂ from the flue gas is absorbed by the solvent forming carbamate as the main product. This stream leaves the absorber at the bottom and is commonly referred to as the rich solvent.

Absorber chemistry

The most important equilibrium reactions associated with CO_2 absorption by a primary amine[3] such as MEA are listed in Table 5.1.

Table 5.1: Equilibrium reactions associated with CO2 absorption [26]

	Reaction	
Dissociation of water	$H_2O \leftrightarrow OH^- + H^+$	1
Dissociation of carbon dioxide	$CO_2 + H_2O \leftrightarrow HCO_3^- + H^+$	2
Dissociation of protonated amine	$RNH_2 + H^+ \leftrightarrow RNH_3^+$	3
Bicarbonate to carbamate	$RNH_2 + HCO_3^- \leftrightarrow RNHCOO^- + H_2O$	4

where R referrers to the group CH2-CH2-OH (MEA).

The carbamate, $RNHCOO^-$ formed in reaction 4, is the dominant product for a primary amine in a CO_2 absorption process. Additionally, carbamate binds up one RNH_3^+ to produce ammonium carbamate, $RNHCOO^-$ $^+RNH_3$. Consequently, it takes to moles of amine, to bind one mole of CO_2. Reaction 4 is a strongly reversible reaction; at low temperatures, the equilibrium shifts towards the right, while at high temperature, the equilibrium shifts towards the left. This is the basis for the temperature swing absorption/desorption process as previously indicated.

Absorber intercooling

The operating temperature of the absorber plays a significant role in the overall performance of a MEA absorption process. Heat of reaction is released through the chemical reaction between CO_2 and MEA. Evidently, higher flue gas CO_2 concentrations, lead to more heat release in the absorber and increased operational temperature. In general, higher temperatures lead to faster reaction kinetics. On the other hand, the solvent loading capacity is reduced since optimum loading is found between 40 and 55 °C. A flue gas from coal firing has a much higher CO_2 fraction, compared to that coming from natural gas. In CO_2 absorption, from such a flue gas, it is beneficial to remove some heat. This is commonly accomplished by absorber intercooling; a semi-rich fraction of solvent is extracted at the lower part of the absorber, cooled, and returned at approximately the same height.

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[3] N bound to two H atoms and one R-group
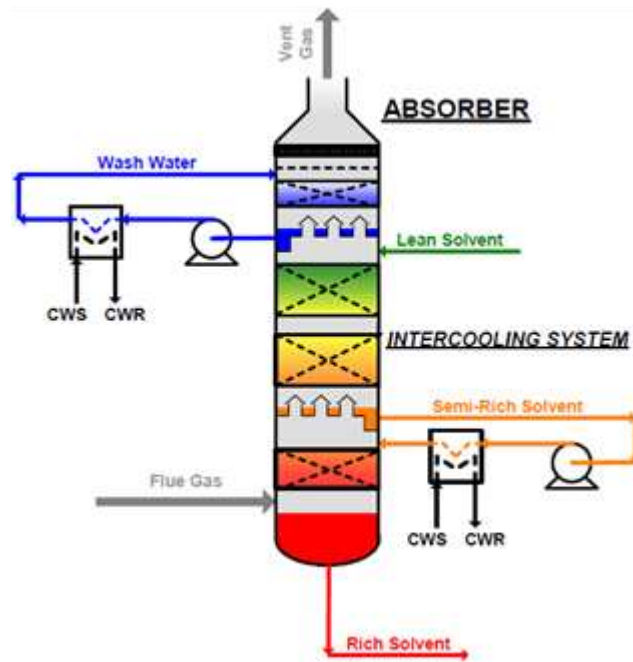
Figure 5.4: Absorber design with intercooling configuration [25]

Figure 5.4 illustrates the principle. Cooling water is provided from the plant cooling system. In this way, the total solvent circulation rate in the absorption/desorption process is reduced, thereby decreasing both the reboiler duty and the capital cost of the solvent circulation equipment [25].

The treated flue gas passes through a water wash section before it is vented to the atmosphere. The water wash reduces solvent losses by removing unreacted solvent entrained in the flue gas and helps to maintain a certain water level in the absorber. Additionally, the water wash section cools the treated gas by an intercooling circuit as shown in Figure 5.4. The lower the outlet temperature of the treated gas is, the lower are the evaporative water losses, but more cooling water is needed for intercooling.

Stripper

The rich solution leaving the absorber is sent to the disorber/stripper where solvent regeneration takes place, and pure CO_2 is released. In order to break the chemical bounds in carbemate, i.e. reverse reaction 4 in Table 5.1, elevated temperatures are required. The regeneration process takes place at a temperature and pressure range of 100-130 °C and 1-2.7 bars respectively. The stripper operates as a distillation column with a reboiler and condenser. In addition to vapour-liquid equilibrium, chemical reactions are also occurring in the stripper column. Indirect heat supply to the reboiler is provided by LP steam from the steam cycle. Nearly pure CO_2 gas leaves the top of the stripper with water vapor as the major contaminant. Water and solvent is condensed and leaves the stripper as the lean solvent. Before returning to the absorber, the lean solvent has to be cooled.

The major challenge for post capture technologies is the intensive energy demand in the reboiler of the stripper column. To minimize some of these energy losses, a cross flow heat exchanger between the cool rich solvent, exiting the absorber and the hot lean solvent exiting the stripper is used.

5.1.4 CO₂ compression

A dense liquid phase of the purified CO_2 is obtained through compression and cooling. The CO_2 stream from the stripper is compressed in a three stage compressor with intercooling and condensate removal. The stream is then pumped to a pressure above 110 bar, and further cooled to a supercritical condition suitable for transportation and storage.

5.1.5 Water usage and consumption

Large amounts of cooling water are required in an amine-based CO_2 capture system to support cooling operations. This includes flue gas cooling, water wash cooling, absorber intercooling, reflux condenser duty, the lean solvent cooler, and CO_2 compression interstage cooling. Cooling water is provided from the plant cooling system and returned to the plant cooling system. If a wet cooling tower is employed, evaporation and blowdown losses are increased, when integrating CO_2 capture in the power plant.

Furthermore, makeup water has to be continuously added in the water wash section above the absorber, compensating for evaporated water to the atmospheric vent stream and the purified CO_2 stream. As already pointed out, a higher vent stream outlet temperature will increase the evaporation rate of the wash water. Two reports, [27] and [23], have been studied to identify appropriate temperatures. The vent stream outlet temperature was found to be 36.9 °C and °31.5 C, respectively. Another report, [28] states that 0.8% of mass flow rate of flue gas entering the absorber is a realistic estimate of the required water wash makeup.

5.2 Oxy-combustion CO_2 Capture

A simplified block diagram given in Figure 5.5 shows the concept of the alternative CO_2 capture technology based on oxy-combustion. Any oxy-combustion power plant consists of four main sub-systems; the air separation unit (ASU) that provides high purity oxygen for combustion, the boiler section where heat exchange takes place, the steam cycle producing power, and the compression and purification unit (CPU).
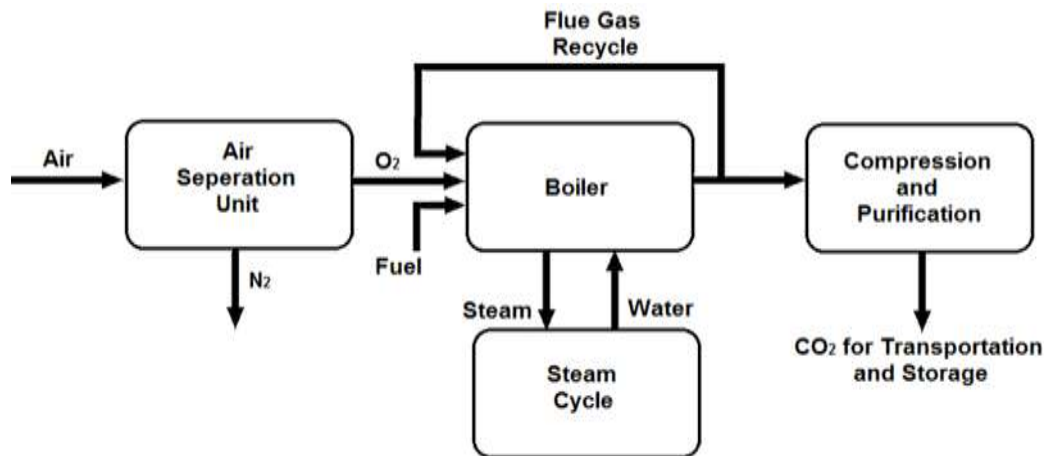
Figure 5.5: Simplified block diagram for oxy-combustion power plant

Conventional boiler technology is used in oxy-combustion to combust the fuel and for heat exchange with the working fluid. A large portion (~70%) of the flue gas is normally recycled back to the boiler to control the flame temperature and to reconstitute the flue gas volume, ensuring proper heat transfer [20]. The main combustion products are CO₂ and water vapor as nitrogen is mostly eliminated from the air. A very simplified reaction mechanism of oxy-combustion of coal is shown below.

$$C + O_2 \rightarrow CO_2$$

$$2H + \frac{1}{2}O_2 \rightarrow H_2O$$
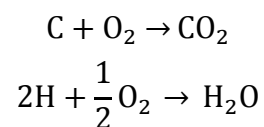
Due near complete absent of atmospheric nitrogen, NO_x formation is significantly reduced, and a selective catalytic reduction process for NO_x removal is not necessary as in conventional air-firing. Some NO_x can arise from fuel bound nitrogen, but such NO_x forms best in fuel lean environments, and at low temperatures. In oxy-combustion, near stoichiometric amounts of oxygen are used, because of the high production costs of pure oxygen. Still, it is very likely that some nitrogen will enter the boiler section as air leakage [20].

Conventional flue gas cleaning equipment like electrostatic precipitators (ESPs) or fabric filters and flue gas desulfurization (FGD) are also used in oxy-combustion of coal. Due to a higher moisture content and higher SO_x concentration, the operative

temperature range of the ESP/fabric filters has to be above the dew point temperature of the flue gas [20]. As discussed in Section 4.6.2, whether the FGD system is placed before or after the flue gas recycling, depends on the sulfur content in the coal.

Due to the high fraction of CO_2 in the flue gas, no chemical solvent or physical solvent is needed to separate the CO_2 from the flue gas. In the CO_2 compression and purify unit, the CO_2 goes through a multistage compression and cooling process, in which the flue gas is first dried and impurities such as the atmospheric gases oxygen, nitrogen and argon are removed.

5.2.1 Oxygen Ssupply

In oxy-combustion, high purity oxygen is a key parameter, normally with oxygen purities in the range of 95-97% [20]. The oxygen purity is closely related the reduction of NO_x emissions, but it is also a trade-off between the power consumption in air separation and flue gas compression and purification process. The technology that is mature enough at large scale oxygen production is cryogenic air separation. This is a very energy demanding process and leads to a loss in the total power plant efficiency, if compared to a conventional air-fired power plant without CO_2 capture. The energy consumption depends on the oxygen purity, as higher pressures in the high pressure column are associated with higher oxygen purity. Table 5.2 shows the performance of a typical ASU with increasing oxygen purity. The ASU penalty corresponds to the difference in net electrical efficiency of a base case power plant without CO_2 capture, compared to that with oxy-combustion power plant using cryogenic air separation.

Table 5.2: ASU performance with O₂ purity [29]

O₂ purity	mol%	85	90	95	97
O₂ stream flow	t/h	271.7	255.7	241.4	238.8
ASU power	MW	60.1	61.2	62.8	63.3
Specific consumption	kWh/t of O₂	221.3	239.1	260.2	268.7
ASU penalty	%-points	5.4	5.8	6.2	6.3

5.2.2 Cryogenic air separation technology

Cryogenic distillation is normally performed in an air separation unit (ASU) and is similar to any other distillation process, except that it is carried out at extremely low temperatures (in the area of 100 K) to achieve liquefaction of air. The process initializes taking in large volumes of air, which then are compressed, cooled and partially liquefied, followed by distillation. Depending on the required oxygen purity, the air is fully or partly separated into its major components; nitrogen, oxygen and argon [8].

5.2.3 Double Column Process

A wildly used principle is cryogenic air separation is the double column process. The arrangement of distillation columns is to combine the condenser in the high pressure (HP) column with the reboiler of the low pressure (LP) column, forming a heat exchanger unit. There are various configurations within the double column process that is used in well over 90% of oxygen producing plants [20]. The main steps in an air separation unit based on the double column process will be introduced in the following sub-sections.

Compression and cooling

The process initializes with compression of large quantities of filtered air in a multistage compressor with intercooling and condensate knock out. The discharge pressure is set by the pressure of the HP distillation column [30]. Further, the air is cooled in a water wash tower by direct contact of chilled water. The chilled water is usually obtained by water cooling in a heat exchanger against dry waste nitrogen. Water soluble impurities and fine particular matter are also removed in this process step.

Pre-purification of air

In the pre-purifier unit (PPU), moisture and CO_2 are removed to avoid freezing in the distillation process. Additionally, other impurities such as nitrous oxide and heavy hydrocarbons are removed to enhance process safety [20]. The PPU is based on temperature swing absorption (TSA), utilizing absorbent material such as molecular sieve. A double vessel process illustrated in Figure 5.6 is commonly used. Regeneration of one vessel takes place as the other vessel is used for absorption. The flue gas passes through the vessel until the vessel is saturated with impurities. Then, regeneration gas, typically dry waste nitrogen is heated by steam and fed to desorb the impurities. The impurities are vented to the atmosphere together with the regeneration gas. Cool regeneration gas is then used to cool the PPU to feed gas temperature again.
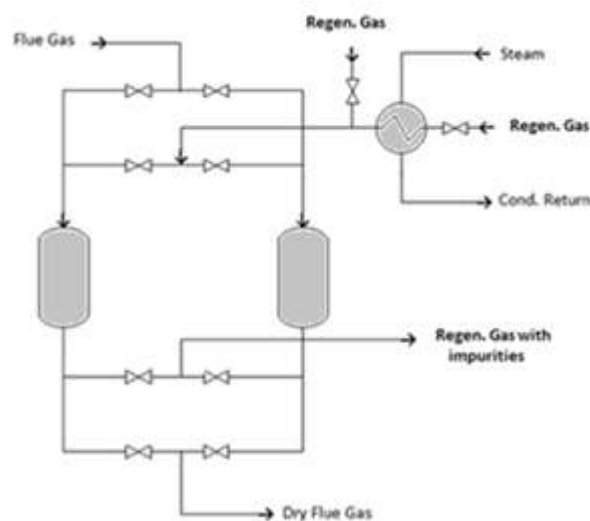
Figure 5.6: Double vessel process for temperature swing absorption

Heat exchange

After the purifying step, air is cooled in a large multi-stream heat exchanger against the refrigerate products from the distillation process. A heat exchanger technology based on brazed aluminum is used that enables very close approach temperatures [20]. The air exits slightly above its dew point.

Double column distillation

Referring to Figure 5.7, the cooled process air enters the base of the HP column. The HP and the LP column make up a conventional double column, in which the reboiler and condenser is the same unit. In a three component mixture such as air, nitrogen is the lightest component, oxygen the most volatile, while argon is slightly lighter than oxygen. Accordingly, the vapor flowing upwards the HP column consists of nearly pure nitrogen. This stream is condensed in the condenser and part of it is provided as reflux for both distillation columns. The remaining nitrogen is vented to the atmosphere after recovering its cold exergy and pressure exergy. The liquid stream enriched with oxygen from the HP column is sub-cooled and fed to the LP column. Cooling is provided by the pure gaseous nitrogen from the LP column that holds an even lower temperature obtained by expansion of the incoming streams. High purity liquid oxygen is extracted from the bottom of the LP column. All streams are heated against incoming the air in an large multi-stream heat exchanger before exiting the air separation unit.
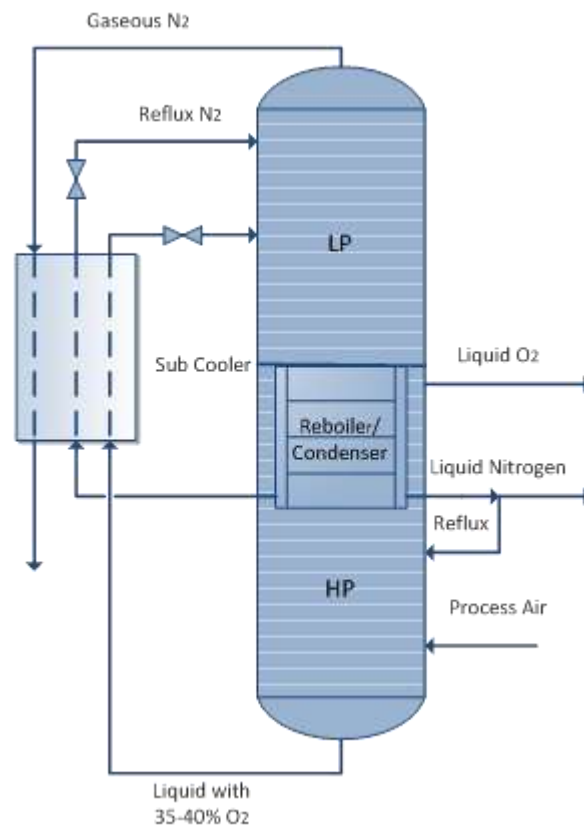
Figure 5.7: Conventional double column system

5.2.4 Oxygen Transport Membrane

A future alternative for large scale oxygen generation is currently under development. This new promising technology introduces oxygen separation from atmospheric air through mixed conducting membranes. The oxygen transport membrane integrates oxygen separation and combustion in one unit, keeping the air and the fuel streams separated. The concept is to transfer the oxygen in the air as ions and electrons through a membrane wall, using the difference in partial pressures of O₂ at both wall sides as the driving force. The membrane has a complicated structure to favor as high transportation flux of oxygen as possible [20]. An illustration is given in Figure 5.8.
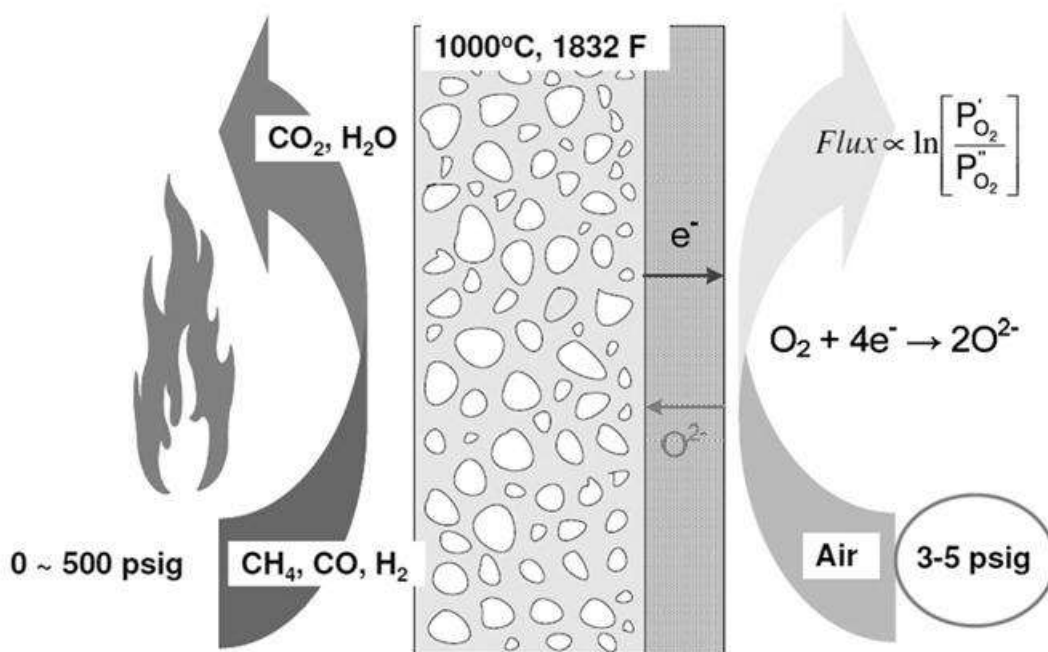
Figure 5.8: Oxy-combustion using oxygen transport membranes [4]

5.2.5   CO₂ Compression and Purification

The CO₂ enriched flue gas from oxy-combustion of coal contains a number of impurities. These impurities have to be removed in order to reach a CO₂ quality appropriate for transportation and storage. A large fraction of impurities come from atmospheric gases such as nitrogen, argon, and water vapor. The flue gas will also contain small fractions of CO, SO$_x$, NO$_x$, particulate matter, mercury and other trace impurities. Studies show that the presence of impurities increases the energy consumption of purification, because of their impacts on the thermodynamic properties of the CO₂-streams [31]. After purification, the CO₂ is compressed to a supercritical condition for transportation and storage. Figure 5.9 indicates the transportation and storage conditions for CO₂ in a phase diagram.
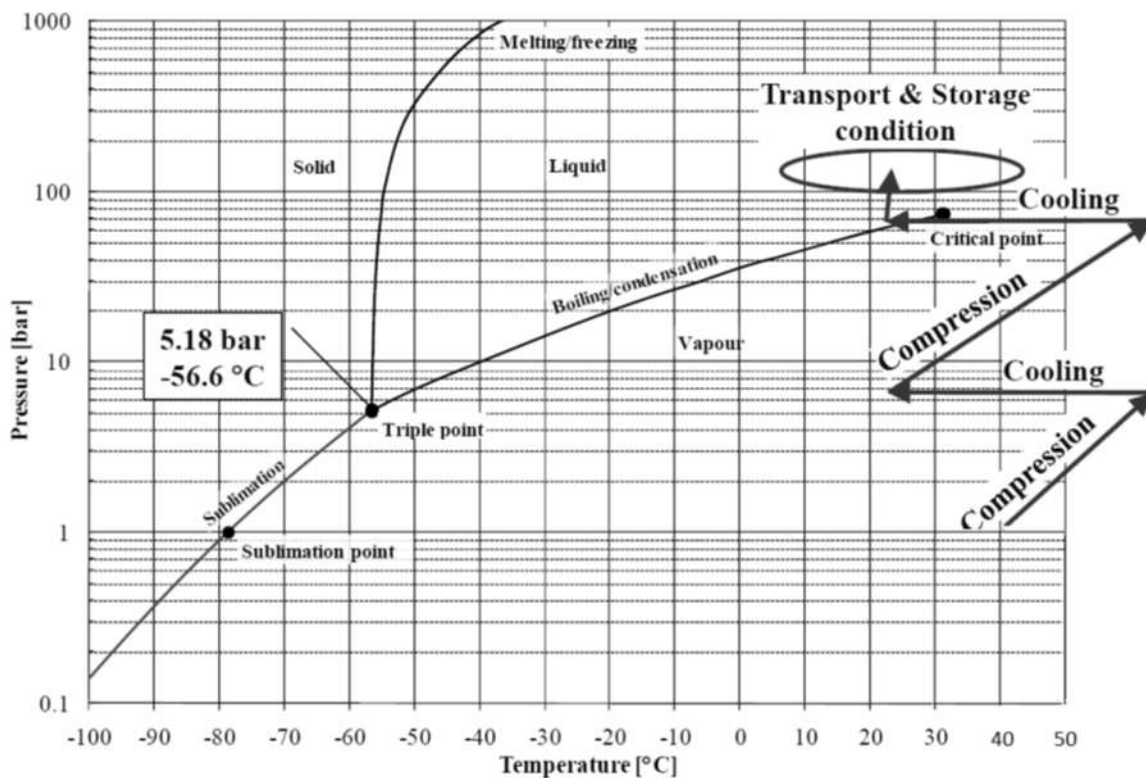
Figure 5.9: Phase diagram of CO₂ [13]

5.2.6   CO₂ Purity Requirements for Transportation and Storage

There are at present no common standard for the level of impurities for transportation and storage of CO₂. However, the limit of H₂O for pipeline transportation very stringent and commonly set to 10 ppmv. The presence of free water in the compressed CO₂ gas can cause corrosion in contact with acidic gases. Additionally, hydrate formation can also occur. For EOR, higher concentrations of O₂ can lead to reaction with bacteria, potentially causing overheating at the injection point. Furthermore, oxidation reactions inside the reservoir can result in a higher oil viscosity, which is

associated with higher extraction costs of the oil [13]. Three different specifications for CO$_2$ transport and storage are purposed in Table 5.3. It is suggested that the CO$_2$ gas should be purified to at least 95% pure CO$_2$.

Table 5.3: Proposed CO$_2$ purity specifications for sequestration [20]

Composition by volume	Medium purity	High purity	Very high purity
CO$_2$	> 95%	> 99.5%	> 99.5%
N$_2$+O$_2$+Ar	< 5%	< 10 ppm	< 10 ppm
H$_2$O	< 10 ppm	< 10 ppm	< 10 ppm
NO$_x$	< 1500 ppm	< 1500 ppm	< 1500 ppm
SO$_x$	< 1500 ppm	< 1500 ppm	< 1500 ppm
CO	< 100 ppm	< 10 ppm	< 10 ppm

5.2.7 The CO$_2$ Compression and Purification Unit

The CO$_2$ enriched flue gas that enters the compression and purification unit (CPU) is about 30% of the original flue gas. At first, the flue gas is cooled in a heat exchanger and flashed for moisture. This heat exchanger is often a spray tower where water and flue gas are in direct contact, providing removal of any water soluble impurities such as HCl, HF, NH$_3$, SO$_3$ and NO$_2$ [20]. The cool flue gas is now sent through a compression-train. Compression up to 25-35 bar takes place in a multistage compressor with intercooling and condensate removal. The discharge pressure is selected to minimize power consumption in the further purification process; higher CO$_2$ purity also requires higher pressures. Some trace impurities such as SO$_x$, NO$_x$ and Hg will drop out with condensate during the compression process. To avoid hydrate formation and pipeline corrosion, further water removal is necessary.

Water absorption
Water absorption is accomplished using the same technology as that used in the air pre-purifier unit introduced in Section 0; a double bed vessel is used such that the drying process takes place in one bed, while the other bed is being regenerated. Heating of regeneration gas is accomplished by LP steam supplied from the steam cycle.

Cold box purification
Further purification to remove non-condensable gases such as N$_2$, O$_2$, and Ar takes place in a flash drum or distillation column. The boiling point of CO$_2$ is much higher than that for nitrogen, oxygen and argon, so separation is fully possible at much lower temperatures than those used in air separation. The final purification of CO$_2$ occurs in what is referred to as "a cold box", where temperatures are at sub ambient levels. The cold box uses brazed aluminum heat exchangers (BAHX), see Section 5.2.3. The cold

box also includes phase separators and/or distillation columns. Two different cold cox configurations are purposed, which one depends on the CO_2 purity specification.

Flash separation

In a flash separation, also called one stage distillation, the liquid and vapor phase reaches equilibrium inside a flash drum, and the two phases are carried out separately. Figure 5.10 shows a cold box based on flash separation. This technology is used to obtain CO_2 purities greater than 95%.
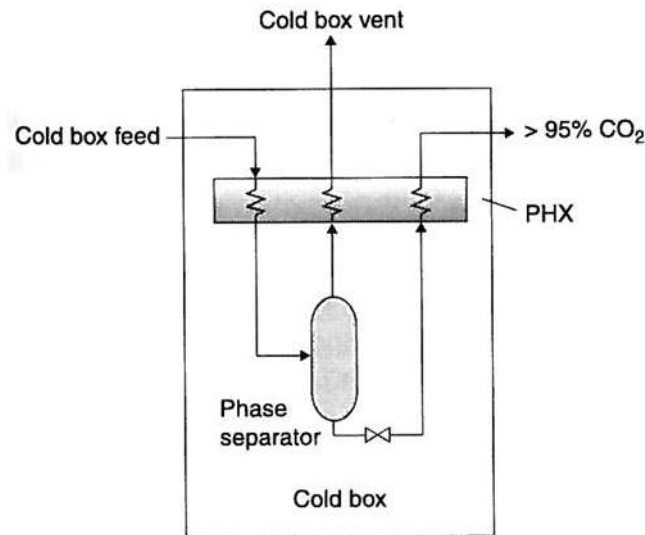
Figure 5.10: Cold box flash separation [20]

Referring to Figure 5.10, the compressed flue gas is cooled in a BAHX to temperatures between -45 to -50 °C, at which the majority of CO_2 is in liquid phase. A flash drum separates the liquid from the gas phase. The less volatile components O_2, N_2, Ar, and CO are recovered in the vapor stream. The liquid stream contains greater than 95% CO_2 by volume. Throttling of this stream provides refrigeration in the BAHX. To avoid phase change from gas to solid in the BAHX, throttling is controlled to ensure temperatures above the CO_2 triple point. The vent stream and the purified CO_2 stream are both warmed against the incoming flue gas stream. Various configurations within the flash separation have been studied. For instance, a to stage condensation with double flashing is used to improve CO_2 purity and reduce power consumption [20].

Distillation

If CO_2 purities greater than 99.5% are desirable, a cold box configuration based on distillation is used. Figure 5.11 shows the arrangement of cold box distillation process.
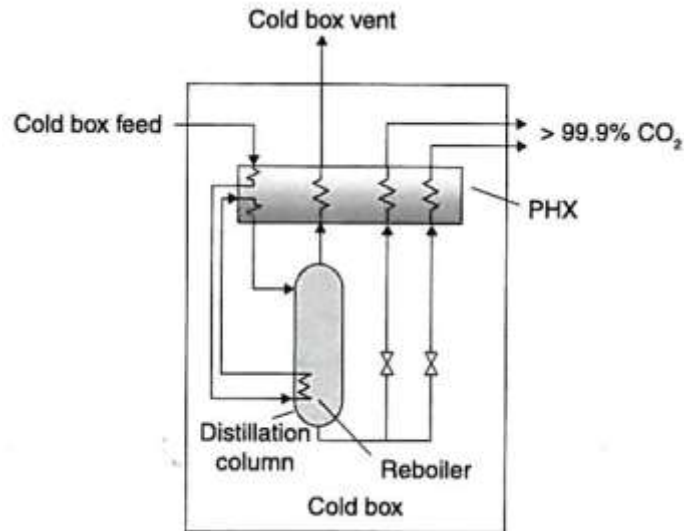
Figure 5.11: Cold box for distillation process [20]

As seen in Figure 5.11, the compressed flue gas feed is first partially cooled in a BAHX. This semi-cool gas is used to reboiler the liquid in the distillation column, before returning to the BAHX for final cooling. The cool stream enters partially liquefied at the top of column. Liquid and gas are separated through a number of equilibrium stages, before the two phases are carried out separately. Both liquid streams are vaporized in a valve and used to cool the incoming flue gas.

Final compression and cold box vent processing
The high purity CO_2 stream is prepared for transportation and storage in a final compression process. A multi sage compressor with intercooling is used to compress the CO_2 to a pressure of 110-150 bar. An after cooler is then used to cool the CO_2 stream to ambient temperatures.

The vent gases from the cold box are at elevated pressure and expanded to recover power. The expanded vent steam is used for cooling in the water absorption process before vented to the atmosphere.

5.2.8 Water Usage and Consumption

Water usage in the ASU and CPU unit is mainly related to cooling water for removal of compression heat. Thus, additional water losses in an oxy-combustion power plant are mainly concerned cooling tower evaporation and blowdown, if that is the cooling system used in the power plant.

Chapter 6

Design Basis

To quantify and compare the water consumption of the two different CO_2 capture technologies; oxy-combustion and post-combustion CO_2 capture, a study of two such power plants have been carried out. This chapter will give a basis for the simulation.

6.1   Pulverized Coal Oxy-combustion Plant

A supercritical oxy-combustion power plant with 561 MW net power output was simulated. 95 % purity oxygen is used for combustion, and the CO_2 recovery system is designed to deliver CO_2 purity of 96.3%. The purified CO_2 stream is compressed and cooled to the transport and storage condition of 150 bar and 25 °C.

6.2   Pulverized Coal Post-combustion Capture Plant

A supercritical air-fired power plant of 550 MW net power output with CO_2 capture by MEA absorption was studied. The power plat is based on a study performed by DOE/NETL [23] The CO_2 removal is approximately 90% from the feed gas, and a CO_2 purity of 99.7% is obtained. The pure CO_2 is compressed and cooled to the same transport and storage conditions, as the CO_2 product in the oxy-combustion plant.

Due to the time consuming process of modeling such a power plant, only the CO_2 absorption process was simulated, see Section 7.2.1 Necessary stream data to perform a reasonable comparison of the water consumption in the two power plants was extracted directly from [23].

6.3   Coal Characteristics

The bituminous coal with the characteristics presented in Table 6.1 is used for power generation in both the oxy-combustion plant and the post-combustion capture plant.

Table 6.1: Illinois No. 6 bituminous coal characteristics [23]

Proximate analysis	As- received	Dry
Volatile Matter	34.99%	39.37%
Moisture	11.12 %	0.00%
Ash	9.70%	10.91%
Fixed Carbon	44.19	49.72
Total	100%	100%
Ultimate analysis	As- received	Dry
Carbon	63.75%	71.73%
Hydrogen	4.50%	5.06%
Nitrogen	1.25%	1.41%
Sulfur	2.52%	2.82%
Chlorine	0.29%	0.33%
Ash	9.70%	10.91%
Oxygen	6.88%	7.74%
Moisture	11.12%	0.00%
Total	100%	100%
Reported heating value	As received	Dry
HHV, kJ/kg	27 135	30 5131
LHV, kJ/kg	26 171	29 447

6.4   Limestone Composition

Table 6.2 contains the solvent composition of the limestone used for wet flue gas desulfurization.

Table 6.2 : Solvent composition [23]

Component	Chemical formula	Mass fraction
Calcium Carbonate	$CaCO_3$	80.40
Magnesium Carbonate	$MgCO_3$	3.50
Silica	SiO_2	10.32
Alumina Oxide	Al_2O_2	3.16
Iron Oxide	Fe_2O_3	1.24
Sodium Oxide	Na_2O	0.23
Potassium Oxide	K_2O	0.72
Balance	-	0.43
Total	**-**	**100**

6.5 Cooling System

In this work, the plants are assumed to be equipped with evaporative mechanical draft cooling towers. Cooling water with the range, see Section 4.4, of 16/27 °C is available based on the site conditions and cooling tower design specifications given in Table 6.3 andTable 6.4. All circulating cooling water, including the condenser cooling water is assumed to be routed to the cooling tower. A minor part of the cooling processes, require cooling to temperatures out of range by the available cooling water provided by the cooling system. These are for simplicity assumed to utilize chilled water from another source with the range of 10/19 °C.

Table 6.3: Ambient conditions

Parameter	Value
Atmospheric pressure, bar	1.013
Design Ambient Temperature, Dry Bulb, °C	15.0
Design Ambient Temperature, Wet Bulb, °C	10.8
Design Ambient Relative Humidity, %	60

Table 6.4: Design specifications cooling tower

Parameter	Value
Cooling water range, C°	16/27
Cooling water approach to wet bulb temperature, C°	5.2
Drift loss	0.02% of recirculating water
Cycles of concentration	4 (mid-range)

The software Thermoflex was used to simulate the cooling tower. Water losses due to due to evaporation, blowdown, and drift where quantified. Thermoflex is a modular program that is part of the Thermoflow packaged. It is mostly used to calculate design and off-design conditions for steam/water systems.

Chapter 7

Process Descriptions and Simulations

7.1 Pulverized Coal Oxy-combustion Plant

The PC oxy-combustion power plant in this study is based on the model by Fu Chao in [32], exclusive of the FGD and cooling water system. The simulation software Aspen Plus, version v.7.3, is used for process simulation. An overall block-diagram of the power plant is given in Figure 7.1. Detailed flowsheets of the three sub-systems: the ASU, the CPU, and the steam/power cycle can be found in Figure 7.2-Figure 7.4.

7.1.1 Aspen Plus Software

Aspen Plus is a process modeling software provided and manufactured by Aspen Tech. The software is suitable for a wide range of steady state modeling applications within chemical/process engineering. The Aspen modeling environment is based on "blocks" or unit operations (compressors, turbines, rectors, distillation column, separators etc.). By interconnecting the blocks using material, work, and heat streams a complete process flowsheet can be constructed. Aspen Plus includes several databases containing physical, chemical, and thermodynamic data for a wide variety of chemical compounds, as well as a selection of thermodynamic models required for accurate simulation of any given chemical system. The simulation is performed by specifying: (1) flow rates, compositions, and operating inlet conditions of the streams; (2) operating conditions of the blocks used in the process, e.g. temperature, pressure, number of stages and (3) heat and/or work inputs into the process. Based on the input data, ASPEN calculates flow rates, compositions and state conditions of all outlet material streams, as well as the heat and work output of all outlet heat and work streams.

7.1.2 Thermodynamic Property Package

In this simulation, the Peng-Robinson property package was used in the ASU, Boiler, FGD, and CPU. The property package is expected to give reasonable results at all temperatures and pressures. It is recommended for gas-processing, refinery, and petrochemical applications. Sample applications include gas plants, crude towers, and ethylene plants [33]. For the steam cycle and cooling water calculations, the STEAMNBS steam tables where used for water/steam properties.

7.1.3 Design Specifications

The design specifications given in Table 7.1 were used to run the simulation.

Table 7.1: Design specifications for PC oxy-combustion power plant [32]

Parameter	Value
Turbo-machinery	
HP steam turbine isentropic efficiency	0.9
IP steam turbine isentropic efficiency	0.9
LP steam turbine isentropic efficiency	0.88
Steam turbine mechanical efficiency	0.996
Generator mechanical efficiency	0.985
Compressor isentropic efficiency	0.82
Compressor mechanical efficiency	0.97
Fan isentropic efficiency	0.88
Fan mechanical efficiency	0.98
Tail gas turbine isentropic efficiency	0.9
Tail gas turbine mechanical efficiency	0.999
Pump efficiency (including motor driver)	0.736
Compression intercooler temperature, °C	308.2
ASU and CPU	
Minimum temperature difference in sub-ambient heat exchangers, °C	2
Temperature difference of the condenser/reboiler, °C	1.5
Pressure drop in the pre-purification unit, bar	0.1
Pressure drop in sub-ambient heat exchangers, %	1-3
Pressure drop in HP column, bar	0.05
Inlet/outlet temperatures of cooling water, °C	298.2/308.2
Inlet/outlet temperatures of seawater, °C	288.2/298.2
Minimum temperature difference in cooling water heat exchangers, °C	8
Cooling water pressure, bar	2
Steam cycle	
Pressure loss in feed water heaters, bar	0.34
HP steam turbine inlet pressure, bar	242.3
IP steam turbine inlet pressure, bar	45.2
LP steam turbine inlet pressure, bar	9.5
Condenser pressure, bar	0,069

7.1.4   Process Description

Stream data for all material streams in Figure 7.1 can be found in in Appendix A.
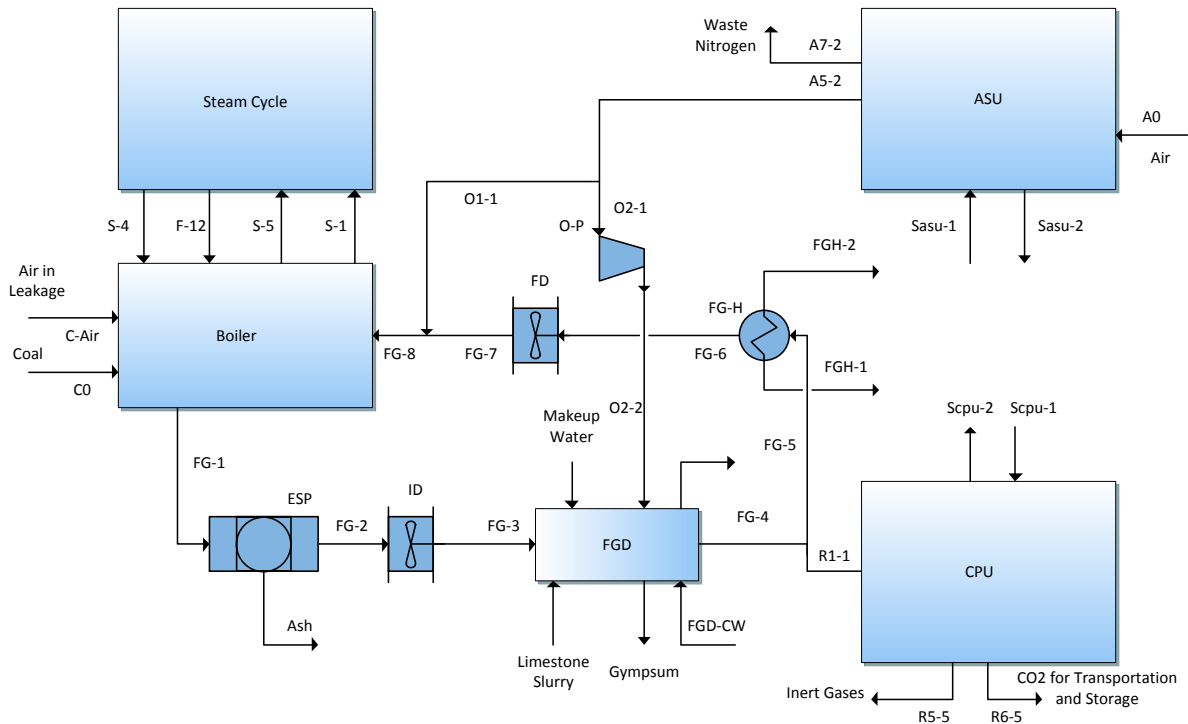
Figure 7.1: Block diagram of PC oxy-combustion power plant

Referring to the block diagram in Figure 7.1, pulverized coal (C0) with the characteristics given in Table 6.1 is combusted with oxygen (O1-1) provided by the ASU. The ASU will be described in detail in Section 7.1.6. To ensure complete combustion, excess oxygen in the combustor is modified to a value of 2.5 mol% O_2 in the flue gas. The combustion process takes place at 1.1 bar and a small amount air in leakage is assumed. The heat of combustion is converted to power by the steam cycle, which will be described in more detailed in Section 7.1.5. Particulate matter is removed from the flue gas in an electrostatic precipitator (ESP), assuming 100% removal efficiency. The ash-free flue gas (FG-2) containing 0.027 mol% SO_2 is induced by an ID fan to the wet FGD system. The FGD system will be introduced in the next subsection. To control the flame temperature in the combustor ~ 72% of the flue gas is recycled to the combustor after desulfurization, while the reminders (R1-1) are sent to compression and purification. The molar composition of that flue gas is: CO_2-69.1%, H_2O-15.0%, N_2-8.5%, O_2-2.5%, Ar-2.9%, CO-1.7%, H_2-0.1%, NO-0.1%. The recycled stream (FG-5) is at this point saturated with water vapor and must be reheated to prevent the entrained water droplets in the fan (C-P3). Heat is supplied from the warm feedwater in the steam cycle. The recycled flue gas (FG-7) is then mixed with the O_2 (O1-1) and preheated against flue gas before entering the combustor.

7.1.5 Flue Gas Desulfurization

Forced limestone oxidation with a SO_2 removal efficiency of 98% is assumed in this work. The data for the FGD process are taken directly from NETL/DOE [23]. The solvent composition is given in Table 6.2, containing 80.4% pure limestone, which is provided as slurry of 30 wt% solvent. The gypsum produced contains 90% suspended solids and 10% free moisture. The flue gas is directly cooled to the saturation temperature of 69 °C by evaporating water in the FGD absorber. A cooling water heat exchanger is used cool the gas further to 57 °C in order to reduce moisture content.

7.1.6 Air Separation Unit

The conventional cryogenic double column air separation process introduced in Section 5.2.3 is applied to produce O_2 with a mole fraction around 95%.
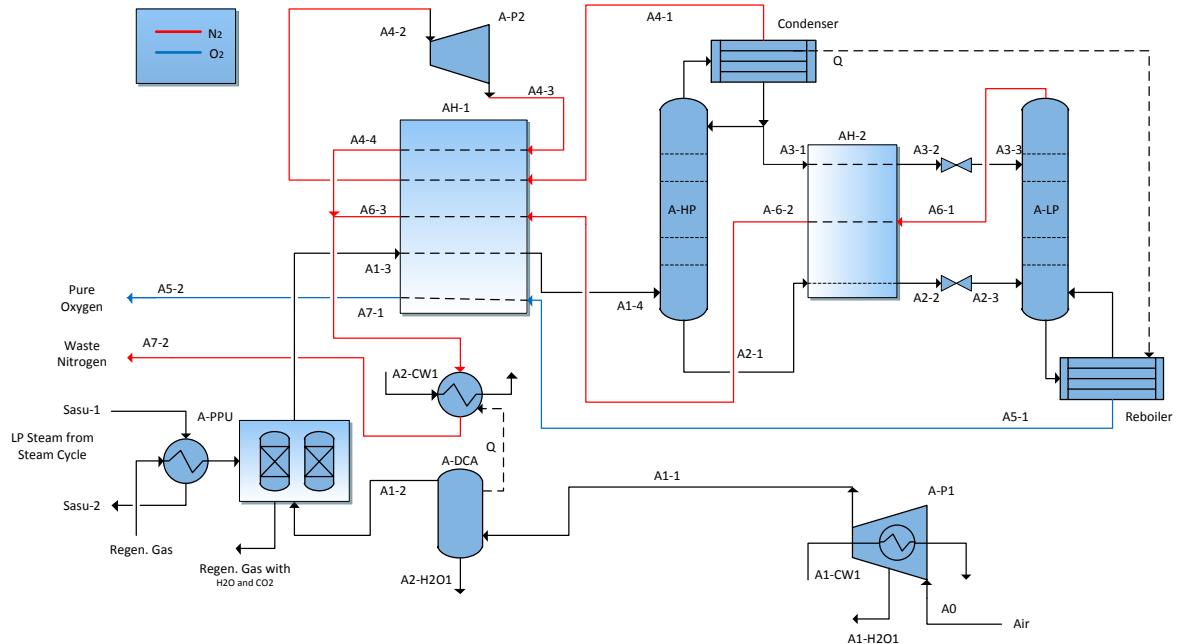
Figure 7.2: Flowsheet of air seperation unit

Referring to Figure 7.2, the incoming stream of ambient air (A0) is compressed to 5.6 bar in a two-stage compressor (A P1) with intercooling and condensate knock-out. Compression heat is removed by cooling water. The compressed air is cooled further in a direct contact aftercooler (A-DCA) and enters a front-end temperature swing adsorption pre-purification unit (A-PPU) to remove H_2O, CO_2 and other impurities. LP steam from the steam cycle is extracted for regeneration of the PPU. The dry compressed air (A1-3) is cooled to near its dew point temperature in the main heat exchanger (A-H1). The air is separated into O_2 (and Ar) and N_2 in the double distillation column, where the reboiler of the LP column (A-LP) serves as the condenser in the HP column (A-HP). The temperature difference of the condenser/reboiler is maintained at 1.5 °C. The N_2 stream (A4-1) from the top of the high pressure column mixes with another N_2 stream (A7-3), after heat recovery and expansion in a tail gas turbine (A-P2). The mixed N_2 stream provides cooling to the water used in the direct contact after cooler (A-DCA) before it is vented to the ambient (A-72). Normally, the waste nitrogen is used as regeneration gas in the PPU, however this was not included in the simulation. The O_2 product (A5-2) has a molar composition: O_2-95%, Ar-3.2%, N_2-1.8%.

7.1.7 Compression and Purification Unit

Stream data for all material streams in Figure 7.3 can be found in Table 10.3, in Appendix A.
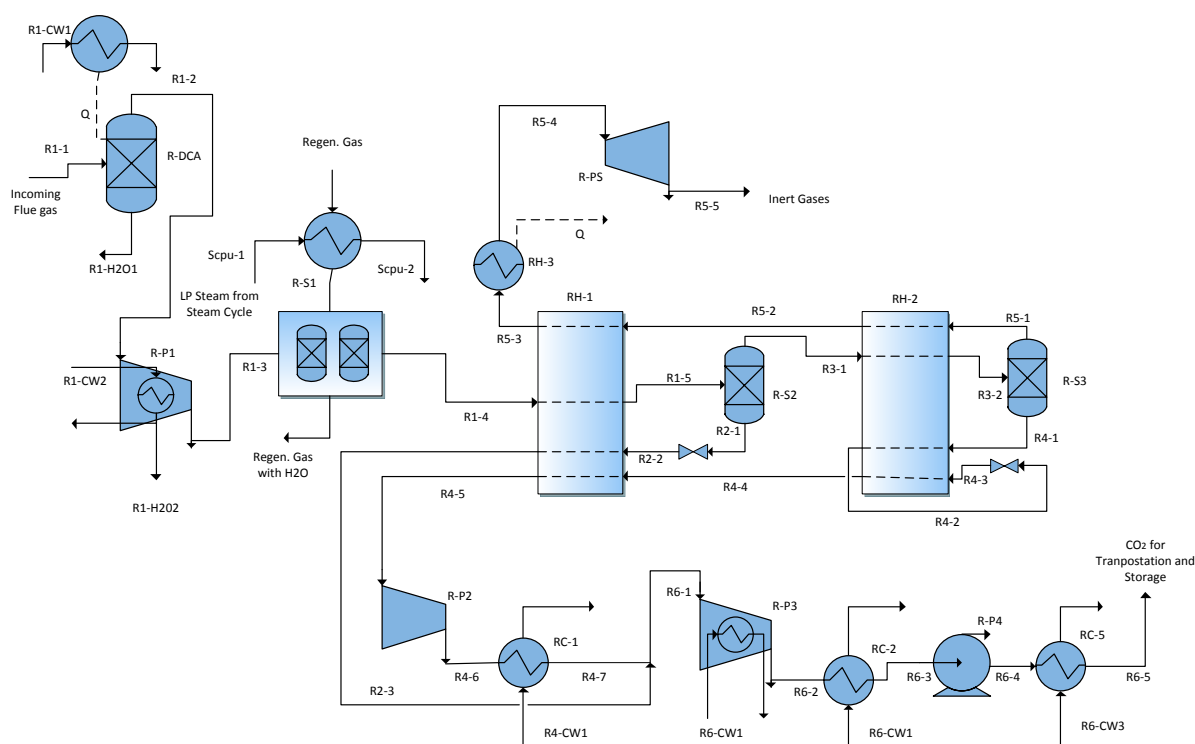
Figure 7.3: Flowsheet of compression and purification unit

The CO_2 rich flue gas (R1-1) in Figure 7.3, enters the CPU unit. The stream is cooled to 35 °C in a direct contact aftercooler (R-DCA) and the flue gas moisture content is reduced. A noteworthy amount of condensate is available as excess water (R1-H2O1). Further, the flue gas is compressed to 32 bar in a three-stage compressor (RP-1) with water intercooling (R-P1) and condensate knock-out. Complete water removal is achieved by water absorption using a molecular sieve twin bed drier (R-S1). Regeneration gas is heated by steam from the steam cycle. The dried flue gas is cooled to a sub ambient temperature of - 26 °C in a multi- stream heat exchanger (RH-1). The flue gas (R1-5) is now partially liquefied as it enters a flash drum (RS-2). Here it is separated into a liquid stream (R2-1) and a vapor stream (R3-1). The liquid stream composed of nearly pure CO_2 is expanded to 18 bar and -33.73 °C (R2-2) though a Joule-Thomson valve, and heated against the incoming flue gas in RH-1. The vapor stream (R3-1) is further cooled to - 54 °C in another multi-stream heat exchanger (RH-2) and separated in a second flash drum (R-S3). The vapor stream (R5-1), mainly containing inert gases, is heated in the two multi-stream heat exchangers (RH-1 and RH-2). These inert gases are further heated against the hot flue gas in the boiler area and expanded in a gas turbine (R-P5) to recover power. The liquid stream (R4-1) from

the second flash drum (R-S3) provides refrigeration to RH-2. Then it is expanded in a Joule-Thomson valve to -55.62 °C and 9 bar (R4-3) to provide more refrigeration in RH-2 and RH-1. The heated stream (R4-5) is compressed to the same pressure as stream R2-3 and cooled by cooling water from the circulating cooling system. The two streams are combined and compressed to 78 bar by a two-stage compressor with water intercooling to 35 °C (R-P3). The compressed CO_2 is further cooled to 25 °C by chilled cooling water. At this temperature and pressure, the CO_2 is in a dense phase and is pumped to 150 bar for transportation and storage. The purity of the captured CO_2 is 96.3 mol%.

7.1.8   Steam Cycle

The steam cycle of the power plant is a supercritical cycle based on configuration 5C in DOE/NETL [23]. The steam turbine is arranged in a high pressure (HP), intermediate pressure (IP), and low pressure (LP) section as shown in Figure 7.4. In total, there are six steam extractions that supply four low-pressure feedwater heaters, the deaerator, and three high-pressure feedwater heaters. Stream data for all material streams in Figure 7.4, can be found in Table 10.4, in Appendix A.

Referring to Figure 7.4, the HP steam (S-1), exits the boiler at the condition of 242 bar/598 °C, and enter the HP section of the turbine. The exhaust from the HP section (S-4), is extracted and reheated in the boiler, before entering the IP turbine section at 45 bar/894 °C (S-5). The major part of the stream is expanded all the way through the IP and LP sections to the condenser pressure of 0.069 bar. By means of cooling water supply, the exhaust steam exits the condenser as condensate (F-1). Makeup water is constantly supplied to compensate for blowdown streams in the cycle. The condensate is then pumped to 17.2 bar by the condensate pump, and preheated to 273 °C by four low pressure feedwater heaters. The feedwater (F-7) passes through the deaerator to remove non-condensable gases and is then pumped to 290 bar by a feed pump. Further feedwater preheating takes place in three high pressure feedwater heaters, before the feedwater (F-12) is introduced to the boiler again. The feedwater pumps are powered by a turbine driver. The steam to this turbine is supplied from IP turbine exhaust (TD-1). Steam extractions to the ASU and CPU are also provided by the IP exhaust. The hot water to the flue gas recycle heater, FG-H in Figure 7.1, is taken from the outlet of FWH 2 (stream FGH-1).
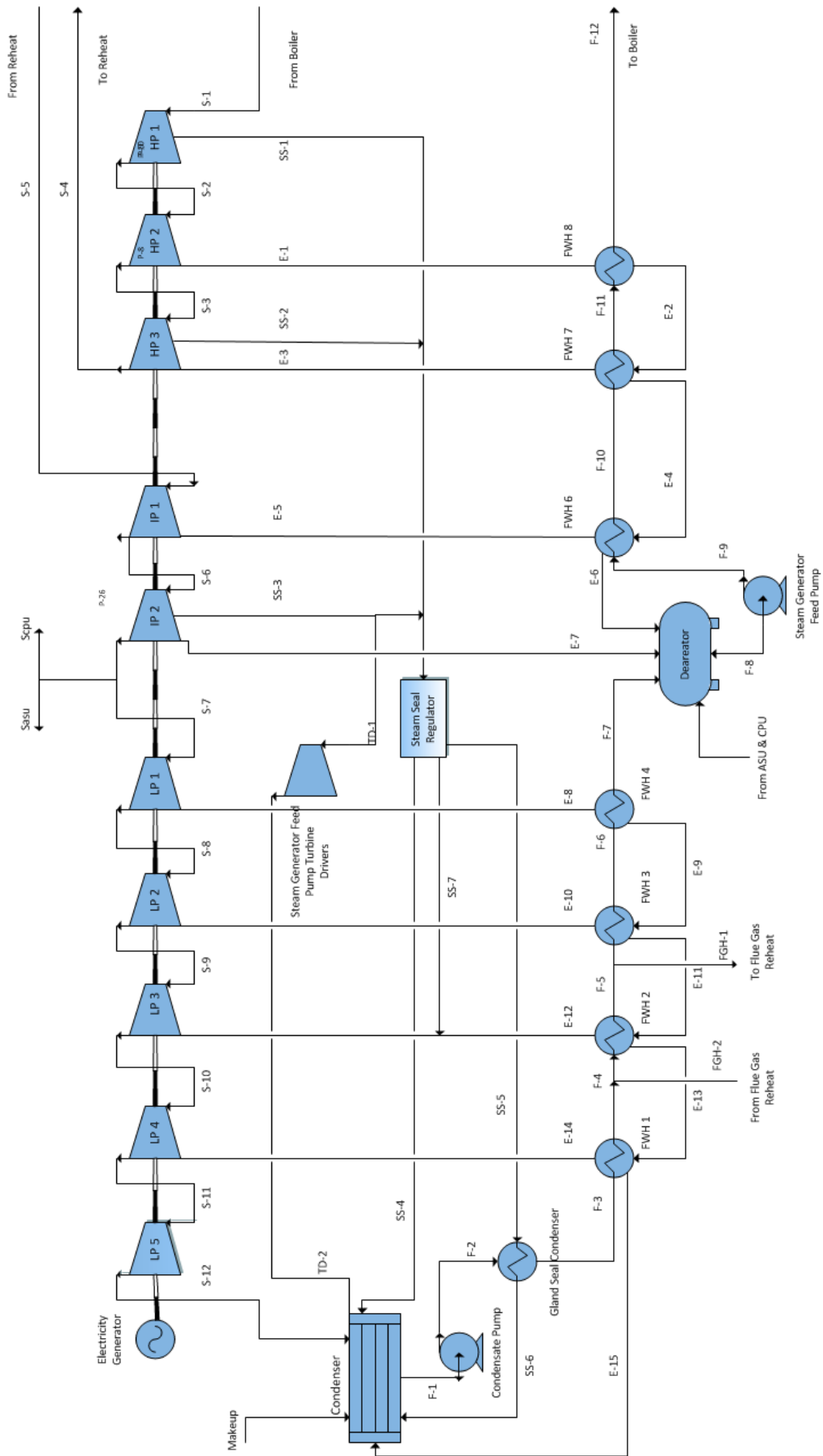
Figure 7.4: Flowsheet of steam cycle

7.2 Pulverized Coal Post-combustion Capture Plant

A simplified flowsheet of the post-combustion capture plant is presented in Figure 7.5. Stream data and operational data concerning the boiler, FGD, steam cycle, and condenser are directly taken from NETL/DOE [23]. The net power output is 550 MW, including power requirement by pumps, fans, compressors and the reboiler. As the objective of studying this power plant is to quantify the water consumption associated with plant operation, only relevant streams and units are considered in Figure 7.5. Stream data for the material streams 1-17 can be found in Table 10.5, in Appendix A. It should be noted thou, that Figure 7.5 is not in complete material balance.
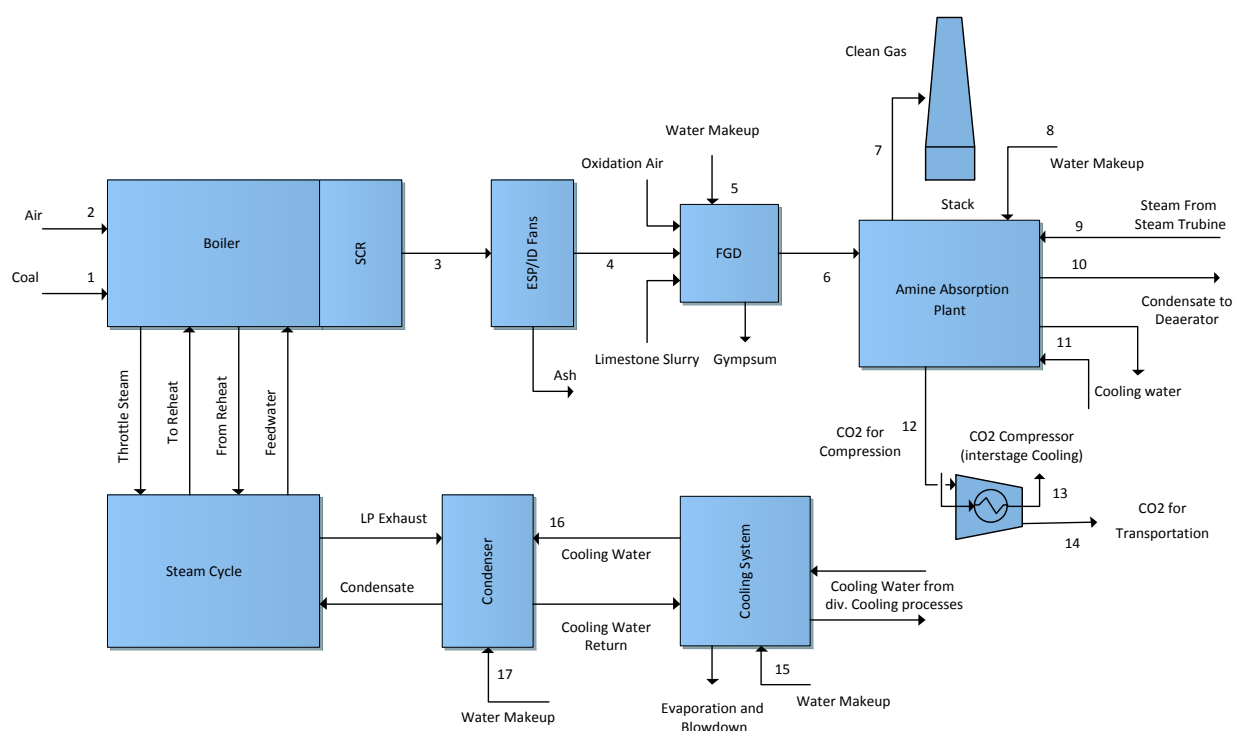
Figure 7.5: Block diagram of post-capture power plant

Air (2) and pulverized coal (1) is combusted in the boiler, generating steam from the circulating feedwater. The steam cycle configuration is the same as that in the pulverized coal oxy-combustion plant, based on a supercritical steam condition of 242 bar/598 °C with a single reheat to 45 bar/621 °C. Steam cycle blowdown is replaced by makeup water at the condenser inlet. Heat of condensation is removed by circulating cooling water, and rejected to the surroundings by an evaporative cooling tower. Evaporation, blowdown, and drift losses are replaced with fresh water makeup to maintain the water balance around the cooling tower. The cooling tower also serves as the heat sink for other cooling processes in the power plant. The entire cooling system is based on the assumptions and specifications in Section 6.5. The overall cooling

tower heat and mass balance was calculated using the simulation program Thermoflex 21.

The flue gas from the boiler passes through a SCR to control NO_x emissions. An ESP is used to control particulate matter and is assumed to provide complete ash removal. Ash-free flue gas (4) enters the FGD unit for limestone based SO_x control with 98% removal efficiency. The flue gas (6) is now sent to the amine absorption plant.

7.2.1 CO_2 Amine Adsorption

The CO_2 absorption process is a temperature swing process using the chemical solvent MEA. The process design is based in the conventional absorption/desorption arrangement introduced in Section 5.1.3. The entire process is simulated using Aspen HYSYS, version v.7.3, and is based on the flue gas stream (6) from Figure 7.5. The reboiler duty and the mechanical work required by the absorption/desorption process are not considered, since the values are included in the total power plant output of 550 MW calculated by NETL/DOE.

7.2.2   Aspen HYSYS Software

Aspen HYSYS is provided and manufactured by Aspen Tech. It is a very intuitive and easy to use process simulator, which can be used to model and simulate a wide range of industrial processes within the oil and gas and refining industry. Aspen HYSYS is similar to Aspen Plus software uses the same "block" modeling and input/output procedures. Further, Aspen HYSYS also includes a variety of databases containing physical, chemical, and thermodynamic data for a wide variety of chemical compounds as well as thermodynamic property methods.

7.2.3   Thermodynamic Property Package

The property package used in this simulation is Kent-Eisenberg. The property package is recommended for systems containing water, one of four ethanolamines (such as MEA), carbon dioxide, and other components typically present in gas-sweetening processes [34]. As this property package does not include the component argon, it is assumed in the simulation that the flue gas is free of argon. For the cooling water heat exchangers, the property package ASME steam was used. The CO_2 compression is simulated with the property package Peng-Robinson to provide accurate results.

7.2.4 Design Specifications

The design specifications given in Table 7.2 were used to run the simulation.

Table 7.2: Design specifications absorption plant

Parameter	
MEA, wt%	0.3
CO2 removal efficiency	90%
Absorber data	
Column pressure drop, 50 mbar	50
Lean solvent inlet temperature, °C	40
Number of stages	13
Diameter, m	10
Intercooling temperature, °C	35
Water Wash Section	
Wash water temperature, °C	35
Treated gas outlet temperature, °C	36.9
Stripper data	
Bottom pressure, bar	2
Pressure drop, mbar	20
Lean solvent outlet temperature, °C	121.5
Number of stages	25
Diameter, m	8
Overhead condenser temperature, °C	30
Various	
Cross-flow heat exchanger temperature approach, °C	5
Pressure drop over water coolers, mbar	10
Pump adiabatic efficiency, %	84

7.2.5 Process Description

Figure 7.6 shows the complete flow sheet for the amine absorption plant. Stream data for all material streams in Figure 7.6 can be found in Table 10.6, in Appendix A. The flue gas stream (6) from Figure 7.5 is slightly modified before entering the absorption plant. All sulfur oxides are assumed to be removed prior absorption, as it is common practice in order to avoid solvent degradation. Mole fractions are normalized to contain no argon, nor sulfur oxides.
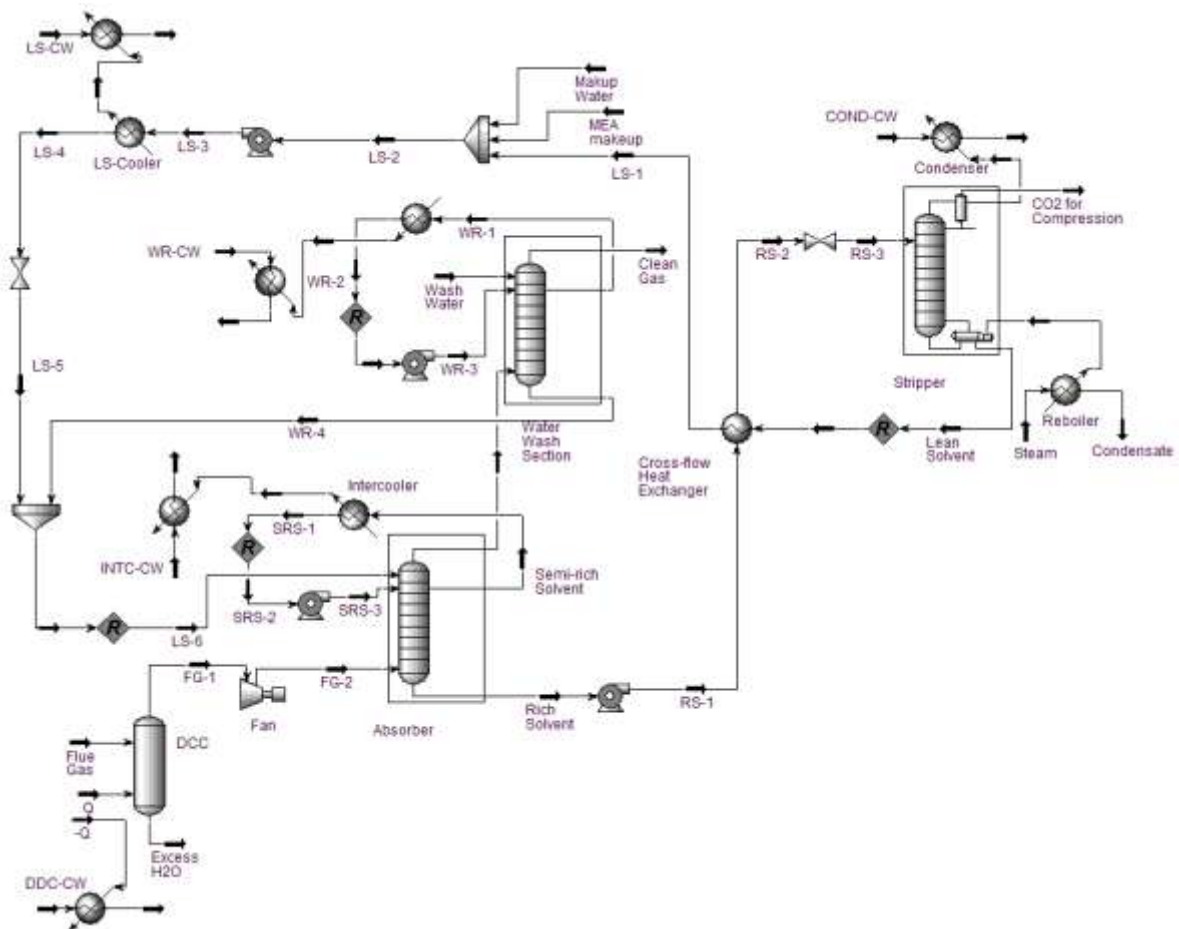
Figure 7.6: Flowsheet of amine absorption plant

The flue gas in Figure 7.6 enters a direct contact cooler (DDC), where it is cooled to 30 °C. A significant amount of condensate is available as excess water from the DDC. After passing through a fan, the flue gas (FG-2) enters the lower part of the absorber. Lean solvent (LS-6) enters the absorber at the top. This mass flow is adjusted in order to achieve 90% CO_2 removal. A semi-rich stream is extracted from stage 3 in the absorber and fed into a cooling water heat exchanger. The stream is cooled to 35 °C and pumped back into the absorber, at the same stage as the extraction point. The extraction stage is optimized with respect to reboiler duty. Treated gas flows through the water wash section to reduce solvent losses, and is then vented to the atmosphere (Clean gas). Circulation water (WR-1), from the water wash section, is cooled in a cooling water heat exchanger to 35 °C. The mass flow of the circulating water is decided by maintaining 36.9 °C of the vented gas, a temperature that seems realistic with respect to the available cooling water. Some evaporated water leaves with the clean gas, and wash water makeup is constantly added to maintain the water balance around the wash section. A mixture of water and washed MEA is reticulated to the absorber (WR-4). Rich solvent exits at the bottom of the absorber, passes tough the cross-flow heat exchanger, and is introduces to the stripper (RS-5). LP steam from the steam cycle is provided for reboiler duty, and clean CO_2 exits at the top of the

condenser at a temperature of 30 °C. A minor amount of water vapor leaves with the pure CO_2 that is sent to compression. The lean solvent exits the stripper and is cooled in the cross flow heat exchanger. Further cooling to 40 °C is provided by cooling water, before the lean solvent reenters the absorber (LS-6). MEA makeup, and water makeup, is added to maintain the balance in the system.

7.2.6 CO2 Compression

Based on guidelines in DECARbit [17], the purified CO_2 stream is compressed and cooled to the same supercritical condition of 150 bar and 25 °C, as in the oxy-combustion case. A three stage compressor with intercooling and condensate knock-out, as shown in Figure 7.7, is used. Cooling water based on the specifications in Section 6.5 is used to remove compression heat. A final pressure of 150 bar will assure that the pressure drop in the transportation pipeline will not cause two-phase flow at any point.
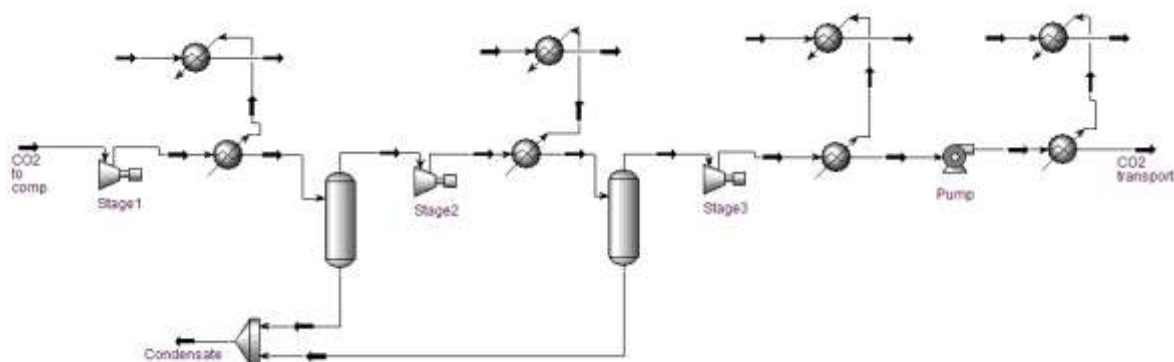
Figure 7.7: CO$_2$ compression

Chapter 8

Results and Discussion

The water consumption calculated in this study represents the total amount of fresh water to be supplied from water resources to provide the water needed by plant operation. An overall water balance, and a water balance for each major plant section, was considered when determining the water consumption of the two power plants. Condensed water from the incoming air, and flue gas, was also quantified to be considered for water recovery. A case where the power plants are located near the sea utilizing a once-through cooling system is also considered. The use of sea water is not counted as water consumption, which only concerns the use of fresh water.

8.1   Basis of Comparison

In order to compare the water consumption of two power plants, it is convenient to consider the water consumption per unit of power. The net power output of the oxy-combustion plant, was calculated to be 561 MW, with a power production of 800 MW at the steam turbine generator terminal. Table 8.1 shows the distribution of power consumption/production for the main components. Based on the LHV of the coal as it is received, the overall power plant efficiency is 30.9%. As already pointed out, the net power output of the post-combustion plant corresponds to 550 MW, with a power plant efficiency of 29.3% (LHV).

Table 8.1: Power consumption and power output of the oxy-combustion plant

Gross Power [kW]	800 179
Power Requirements [kW]	
ASU	
A-P1	133 720
A-P2	9 705
CPU	
R-PI	56 756
R-P2	2 202
R-P3	15 531
R-P4	2 602
R-P5	-9 705
Other	
Fan Work	12 028
Auxiliaries	26 421
Condensate pumps	880
Cooling tower	
Pump	3 006
Fan	5 366
Net Power [kW]	561 078

8.2 Water Consumption in Oxy-Combustion Plant

The total water consumption of the oxy-combustion power plant, was calculated to be 1 732 087.5 kg/h, or 3 0875 kg/h per MW of the net power output.
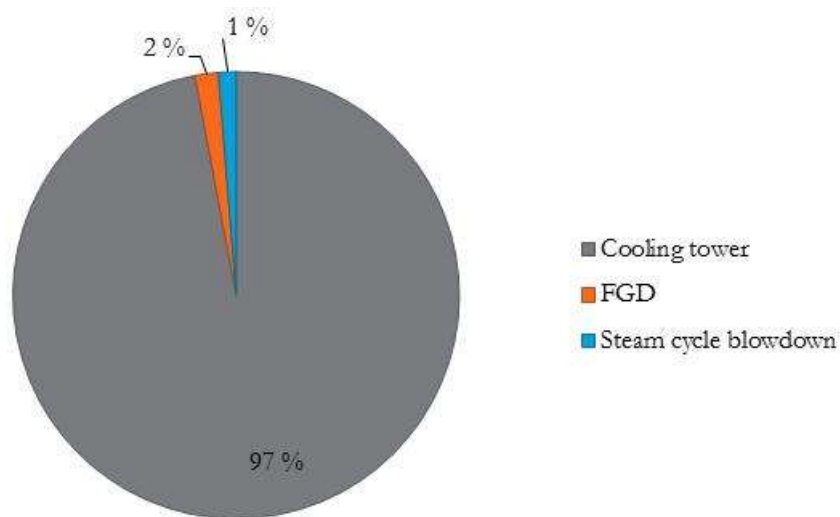
Figure 8.1: Distribution of water losses in the oxy-combustion power plant

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Figure 8.1 shows a pie-chart of the different water makeup requirements in the power plant. The cooling tower is by far the largest water consumer accounting for 97 % of

the water losses. The FGD process and steam cycle blowdown compromise for 1% and 2%, respectively, of the total power plant water loss.

8.2.1 Cooling Tower

The water consumption by the evaporating cooling tower corresponds to the difference between the circulating cooling water, and the cooling water required by process cooling. Table 10.7 in Appendix A shows the cooling water demand of the different cooling processes. The quantity of water lost in the evaporating cooling tower corresponds to 1 680 440 kg/hr.
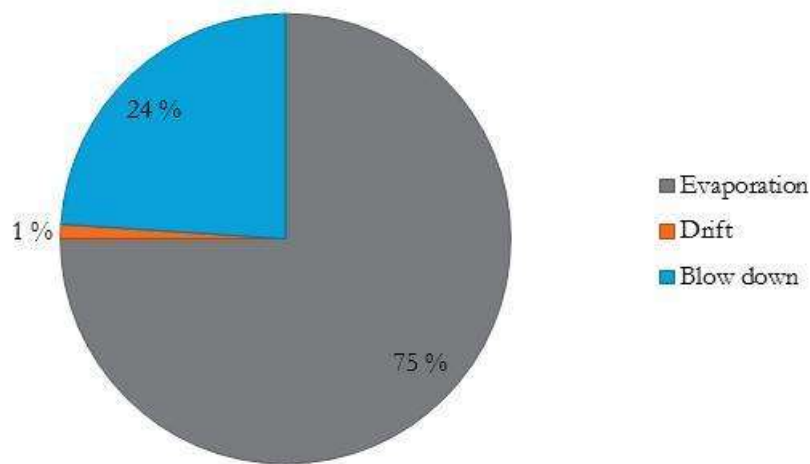
Figure 8.2: Distribution of water losses in the cooling tower

The water makeup supply to the cooling tower is distributed as shown in Figure 8.2. Losses due to evaporation are as expected, the dominant source of water losses and represents 75 % of the water consumption in the cooling tower. The blowdown rate is closely related to the evaporative losses. With the assumed makeup water quality of 4 cycle of concentration (CC), 24% of the makeup water supply is due to cooling tower blowdown. Drift losses represent 1% of the total cooling tower water loss, which corresponds to 2% of the circulating cooling water.

8.2.2 Flue Gas Desulfurization

The FGD system is an important source of water loss in any coal fired power plant. The total makeup water to the FGD was found to be 29 189 kg/h. What can appear to be an interesting aspect of the oxy-combustion FGD system, is that the incoming flue gas moisture content is reduced through the FGD process. This is because the flue gas dew point is below the operating temperature of the FGD system, a direct consequence of the high fraction of water in the flue gas from oxy-combustion. The FGD water losses are, therefore, mainly related to the production of gypsum and purge system. The water loss to gypsum production depends on the quantity of gypsum that

is produced, and on the residual moisture content of the gypsum. The quantity of gypsum formed, depends on the quantity of SO_2, which is influenced by the coal sulfur content, and the flue gas volume flow. The systems' SO_2 removal efficiency, which determines the conversion of SO_2 into gypsum, also has an impact on the water loss. The residual moisture of the gypsum is determined by the end use of the gypsum. The water loss to the purge system is dependent upon the need to purge impurities. In this FGD process, an 80.4% pure limestone solvent is used. This may have contributed to a larger purge stream, than if a more concentrated solvent was used.

8.2.3   Steam Cycle Blowdown

The water loss due to steam cycle blowdown was found to be 2 2068 kg/h. The blowdown rate depends on the makeup water quality, as discussed in Section 3.4.2, which again is influenced by the available water resources and water treatment facilities of the power plant.

8.3   Water Consumption in Post-Combustion Capture Plant

Total water consumption for the post-combustion power plant was calculated to be 2 296 368 kg/h, or 4 175 kg/h per MW of power produced.
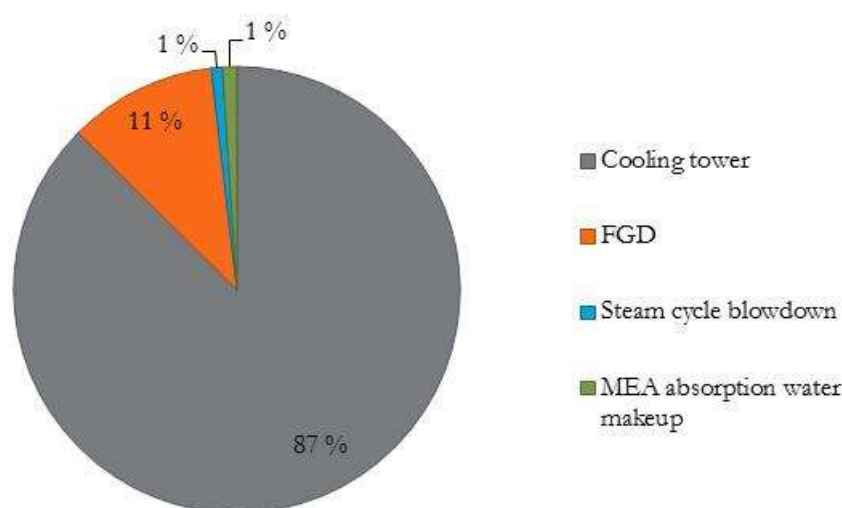
Figure 8.3: Distribution of water losses in the post-combustion capture plant

Figure 8.3 shows a pie chart of the makeup requirements in the post-combustion capture plant. As in the oxy-combustion plant, the cooling tower is the main consumer of water, accounting for 87% of the total water loss. The FGD process is the next largest source of water loss, responsible for 11% of the water consumption. Water

losses due to process blowdown in the steam cycle are 1%. Water losses in the amine absorption plant account for 1 % of the total makeup water supply.

8.3.1 Cooling Tower

The cooling tower water consumption was calculated on the same basis as described in Section 8.2.1. The total quantity of water lost to evaporation, process blowdown, and drift, corresponds to 2 007 432 kg/hr. Table 10.8 shows the cooling water demand of the different cooling processes in the post combustion capture plant.

8.3.2 Flue Gas Desulfurization

The quantity of water lost in the FGD process is 246 640 kg/h, corresponding to a number eight times higher than in the FGD system in the oxy-combustion case. The highest quantity of water is required for the saturation of the flue gas, consequently, the evaporative losses account for 56.8% of the water loss. This amount varies with the flue gas and depends on the volume flow of flue gas, the water content in the flue gas, and the absorber inlet temperature. The remaining water losses are related to the production of gypsum and to purge system. See chapter 8.2.2 for further discussion of the impact these factors have on the water consumption.

8.3.3 Steam Cycle Blowdown

The water loss due to steam cycle blowdown was found to be 18 432 kg/h. The blowdown rate depends on the makeup water quality, and will vary with the water resources available, and water treatment facilities of the power plant. What can be pointed out thou is that the blowdown rate in this power plant is somewhat lower than that in the oxy-combustion plant. This might be a consequence of the large amount of LP steam extracted from the steam cycle to provide for the reboiler duty. In that way, some sections of the steam cycle have reduced steam flow, hence the blowdown rate might be somewhat reduced

8.3.4 Amine Absorption Plant

The water losses in the amine plant were calculated to be 32 881 kg/h under the assumption that the treated gas is discharged at 36.9 °C. A material balance of the water entering and exiting the absorption plant is shown in Table 8.2. The stream names are according to Figure 7.6.

Table 8.2: Water balance of the absorption plant[4]

Mass flow H_2O, kg/h	
In	
FG-2	69 563
Required makeup water	23 882
Out	
Clean gas	88 414
CO2 for compression	5 031

Responsible for 84.3% of the water loss, the evaporated water to the treated gas is the dominant source of water loss. The cleaned CO_2 stream also contains some evaporated water, and is the other source of water loss in the amine absorption plant. The outlet temperature of the cleaned CO_2 stream is set by the condenser temperature in the stripper, which is operated at 30 °C. The outlet temperature of the treated gas is in theory only limited by the temperature of the circulating cooling water in the water wash section. However, more circulating water increases the dimensions of the water wash section, and the pump work, required for operation. No specific guidelines for this temperature were found in literature, except from the two temperatures of 31.5 and 36.9 °C mentioned in Section 5.1.5. An important aspect of the water losses in an amine absorption plant is this temperature, therefore a study of the effect that this parameter has on the overall water consumption was performed. Figure 8.4 shows a plot of the total makeup water as function of the treated gas outlet temperature.
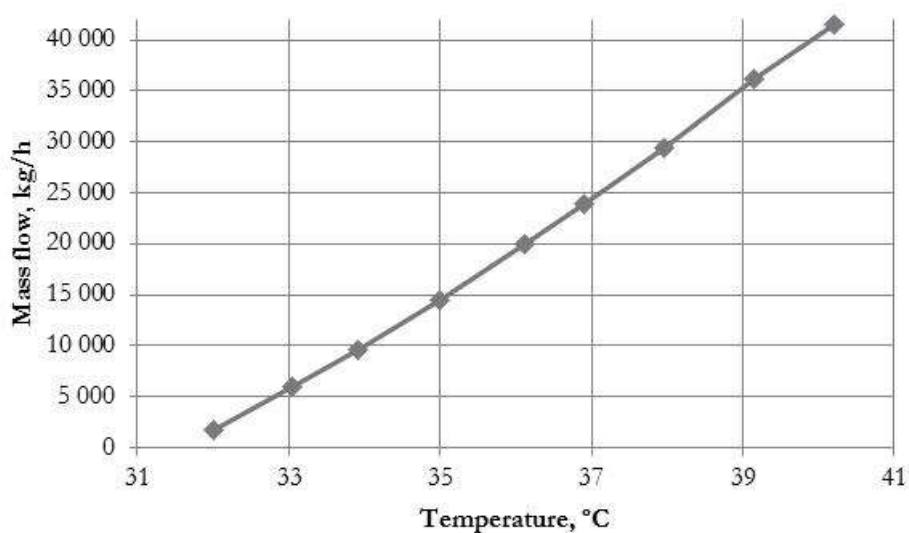
Figure 8.4: Total makeup water requirements as a function of treated gas outlet temperature

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[4] The material balance does not include the DCC.

From the plot it is evident that the makeup water requirements decrease steeply with the treated gas outlet temperature. As the temperature approaches 31 °C, the makeup water requirements are approaching zero. This indicates that the quantity of evaporated water to both the pure CO_2 stream, and the treated gas stream, is equal or less than that entering with the flue gas at the absorber inlet.

In this study of the amine absorption plant, it was assumed for simplicity that the flue gas was free of SO_2. In a real power plant; additional SO_2 removal would most likely be implemented, thus, introducing another source of the water loss to the plant.

As for any other water/steam circulating system; pure evaporated water departing the system, increases the concentration of solids in the circulating system. Process blowdown is necessary to maintain water purity and avoiding build-up of solids in equipment. The blowdown rate was not considered when calculating the water consumption in this amine plant. Consequently, the water losses calculated may be somewhat low.

8.4   Excess Water

Water enters the power plant via humid air intake, moisture content in fuel, and hydrogen bound in the fuel, as discussed in Section 3. In CO_2 capture power plants, it is of great importance to have high purity CO_2 for storage and transportation. The purifying process involves eliminating water. Excess water is thereby available as condensate during air separation and CO_2 recovery. This excess water also includes evaporated water in FGD process.

8.4.1   Oxy-combustion Plant

In oxy-combustion power generation, water is condensed from humid air in the ASU. This is the first source of excess water. The second source of excess water arises in CPU, where water is separated from the CO_2 enriched flue gas. Table 8.3 shows the amount of excess water available from the ASU and the CPU.

Table 8.3: Excess water in oxy-combustion power plant

Stream	Mass flow, kg/h	Pressure, bar
ASU		
A1-H2O1	2 903	2.4
A2-H2O1	16 203	5.6
CPU		
R1-H2O1	36 936	1.1
R1-H2O2	720	37.1

The excess water from the ASU contains traces of dissolved atmospheric gases; otherwise it is relatively pure and is available at two different pressures. Stream R1-H_2O is clearly the largest flow of excess water, coming from the direct contact after cooler (R-DCA) in Figure 7.3. The flue gas entering the R-DCA is saturated with water vapor as it is coming from the FGD unit. The water is originated from fuel moisture and hydrogen bound in the coal. This excess water is available at 1.1 bar, and is a bit sour, containing some dissolved CO_2 (11.2 ppm). Additionally, the water is contaminated with water soluble impurities such as HCl, HF, NH_3, SO_3 and NO_2 arising from coal combustion as discussed in Section 5.2.7.

8.4.2   Post-combustion Capture Plant

Before entering the absorption process, the hot flue gas cooled in a direct contact cooler. A large quantity of condensate is available as excess water. Some of this water is coming from the fuel and the combustion-air, but a significant amount is also evaporated water from the FGD process. Table 8.4 shows the quantities of excess water available.

Table 8.4: Excess water in post-combustion capture power plant

Stream	Mass flow, kg/h	Pressure, bar
Amine absorption plant		
Excess H_2O	222 0112	1
CO₂ compression		
Condensate	4 454	4.3

A significant amount of excess water is available from the direct contact cooler (DCC) in Figure 7.6. This excess water is available at 1 bar and is somewhat acidic, containing 199 ppm dissolved CO_2. Also, there is a small amount of water from the CO_2 compression condensate knock-out. This water is contaminated with 0.04 mol% dissolved CO_2. The condensate also includes other water soluble contaminants originated from the coal as mentioned in the previous sub-section.

Due to the corrosive characteristics of acidic water, it is suggested that CO_2 and other acidic gases are removed so that the excess water from both power plant can be used to back up makeup water. If the water is utilized in processes that are not restricted by high water purities, i.e. the cooling tower and the FGD, minimum pretreatment is needed.

8.5 Once-through Cooling System

Assuming that the two power plants are located near the sea, a once-through cooling system can be used instead of an evaporative cooling tower. The water consumption in both power plants is then dramatically reduced. The total quantity of water consumed by the oxy-combustion power plant corresponds to 51 249 kg/h. For the post-combustion power plant, the water consumption is reduced to 288 954 kg/h. A total reduction in water consumption of 97.0% and 87.4 %, respectively, is obtained. Figure 8.5 show how the water losses in the power plants are distributed when operating with a once through cooling systems.
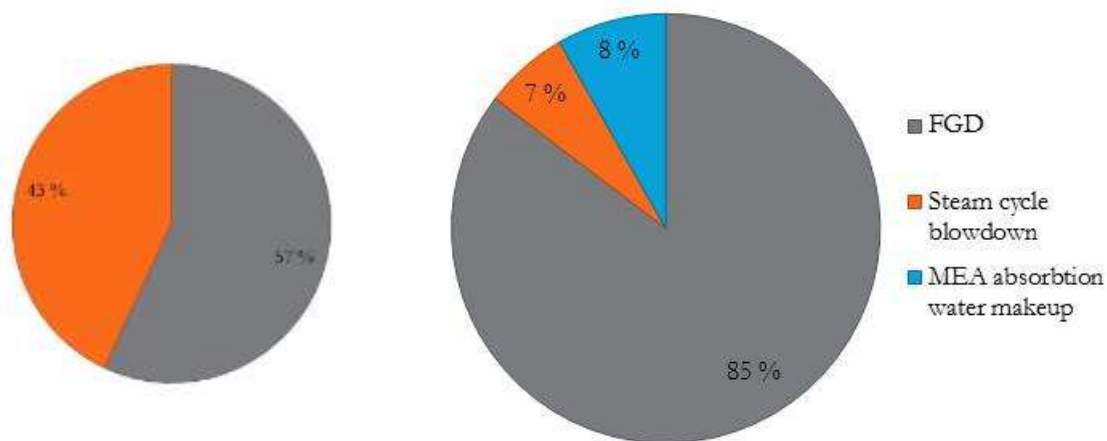
Figure 8.5 Pie-chart of the water consumption in :
a) PC oxy-combustion power plant with once through coolingsystem (total 51 249 kg/h)
b) RS: Post-combustion power plant with once through cooling system (total 288 954 kg/h)

The FGD system is now the major water consumer in both power plants. As previously pointed out, in the post-combustion capture plant, the FGD system is far more dominant than in the oxy-combustion plant. The MEA-absorption process is also consuming a significant amount of water in this context, accounting for 8% of the total water consumption of the power plant.

8.6 Summary

A comparison of the water consumption in the power plants with and without an evaporating cooling tower is presented in Figure 8.6. The values are given per MW power produced, also referred to as specific water consumption, in order to give a representative evaluation of the power plants.
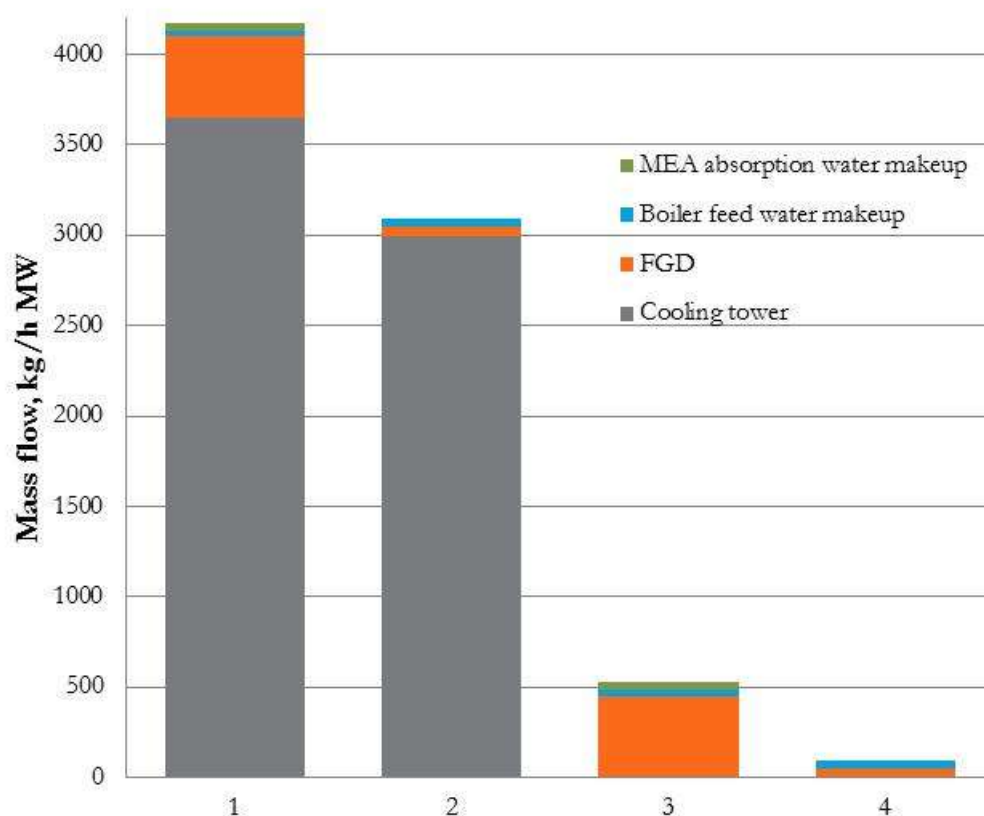
Figure 8.6: Comparison of the water consumption in
1) **Post-combustion capture plant with evaporative cooling tower**
2) **Oxy-combustion power with evaporative cooling tower**
3) **Post-combustion capture plant with once through cooling system**
4) **Oxy-combustion power with once through cooling system .**

The numerical values in Figure 8.6 are given in Table 8.5 in the same range that displayed in the figure.

Table 8.5: Data for Figure 8.6

	1	2	3	4
Cooling tower	3 650	2 996	0	0
FGD	448	52	448	52
Steam cycle blowdown	34	39	34	39
MEA absorption water makeup	43	0	43	0
Total	4 175	3 087	525	91

It can be seen from Figure 8.6 that the cooling tower is by far the largest water consumer in both plants. What can be pointed out is that the water losses due to process cooling in the post-combustion capture plant are 17.8% higher, than in the oxy-combustion power plant.

Considering the power plants with a once-through cooling system, it can be seen clearly that the post-combustion capture plant is the major water consumer of the two plants. Nevertheless, this trend also applies for the plants, when evaporative cooling towers are employed. The excessive use of water in the post-combustion capture plant, are partly subject to the FGD water losses. However, when the flue gas is sent to CO_2 recovery, the evaporative loss from the FGD condenses in a DCC and is available as excess water, in addition to water originated from the fuel and combustion-air. If this water is recovered, and used as makeup in the FGD process, significant water savings could be made. This also applies to the oxy-combustion power plant; if excess water is used to back up makeup water in FGD process, the water losses could be considerably reduced.

Considering the MEA absorption plant, the water losses could be strongly reduced by lowering the temperature of the treated vent gas. A lower cooling water temperature might be more appropriate when using sea water as the coolant. The water losses in the absorption process could be reduced to a minimum, if a treated gas outlet temperature close to 30-31 °C is obtainable.

Reuse of excess water and minimization of the evaporative losses in, the absorption process, could contribute to reduce the difference in water consumption between the two CO_2 capture technologies.

8.7 Consistency of Results

This study shows a representable comparison of the CO_2 capture technologies, as both power plants have a very similar base for evaluation. However, no additional SO_2 removal was included in the amine absorption plant, and amine plant process blowdown was not considered. The results might therefore be slightly in favor of the oxy-combustion power plant. The CO_2 capture technologies studied are both based on state of the art technology, which makes the study relevant in the context of today.

The water consumption of a given power plant varies with several factors such as type of cooling system, makeup water purity, and sulfur content in fuel. Water consumption due to process cooling will vary with site related factors such as wet bulb temperature, temperature of the river, ocean or lake, and the purity of the available water. Additionally, plant specific parameters such as efficiency, process design and technical solutions etc. will affect the water consumption.

The water consumption calculated, gives an indication of the water consumption of similar power plants. However, as already stated, there are many factors effecting the water consumption of a power plant, and the water consumption will vary from power plant to power plant, and with the location of the plant.

8.7.1 Comparison with Previous Studies

In order to get a perspective of the results obtained in this work, relevant information extracted from two previous studies of power plant water consumption, is included in this sub-section.

Study of post combustion capture plant with wet cooling tower
Referring to the study presented in Section 3.2, data regarding the water consumption of a PC supercritical power plant, with MEA absorption, are extracted and shown in Table 4.1. Condensate is assumed to be collected as the flue gas is cooled before CO_2 recovery.

Table 8.6: Water consumption of a supercritical PC post-combustion capture plant [2]

PC supercritical power plant with CDR		
Efficiency (HHV)	27.2 %	
Spesific water consumption		
	Gal (US) /MWh	kg/MWh
Cooling tower	1 044	3940
Steam cycle bowdown	12	45
FGD	86	325
Total	1 142	4310

Considering the results in Table 8.6, they show the same trend as those obtained in this work. This is not unexpected, as the power plants have nearly the same efficiencies, similar steam cycle configurations, equal cooling systems, and both are based on the same CO_2 capture technology. No water losses from the absorption process are present in the study, suggesting that collected condensate is used as makeup water. As the design basis of the plant is not identified, further conclusions cannot be drawn.

Oxy-combustion and post-combustion capture plants with once through cooling system
Another scientific paper [12], Evaluation and Analysis of Water Usage and Loss of Power in Plants with CO_2 Capture, presents the total water consumption of various CO_2 capture power plants. The paper briefly summarizes the results of a study carried out for the IEA Greenhouse Gas R&D Program. Results regarding the two most relevant plants in this paper is extracted and presented in Table 8.7. Expect from the

steam cycle configuration, which is based on an ultra-supercritical steam condition, the plants are quite similar to those in this work. A once through cooling system utilizing sea water is used by the two plants.

Table 8.7: Total water consumption of a post-combustion capture and an oxy-combustion plant [12]

	Efficiency (LHV) , %	kg/MWh
Post-combustion capture plant	34.8	410
Oxy-combustion plant	35.4	63

As the power plants are ultra-supercritical and has a higher efficiency, the specific water consumption is somewhat lower than that obtained in this work. The coal used also has a lower sulfur content (1.1 wt%) than the coal in this study, thus, leading to less water losses to FGD. Having these observations in mind, the results show good compliance with those obtained in this work, when considering a once through cooling system. Further, the recovery of the water from the flue gases by means of a direct contact cooler was also suggested in [12].

Chapter 9

Conclusion and Further Work

Integration of CO_2 capture increases the water requirement per net power generation of a plant, due both to a reduction in the power plant efficiency and to the cooling water and process water requirements associated with CO_2 capture and compression. However, the alternative and more novel CO_2 capture technology, based on oxy-combustion, appear to rely less on water, compared to the post-combustion capture technology.

Relating CO_2 capture and water consumption, a case study of a 561 MW oxy-combustion plant and a 550 MW post-combustion capture plant has been carried out in order to quantify and compare the specific water consumption of the two plants.

The results showed that the cooling tower is by far the largest water consumer in both power plants. Evaporative losses were dominant, but cooling tower blowdown also accounted for a significant portion of the total water consumption, assuming a mid-range water quality. Post-combustion capture have larger cooling water requirements, therefore, water losses in the cooling tower are 17.8% higher, than in the oxy-combustion case.

The second largest water consumer was the FGD process. The FGD process in the post-combustion capture plant consumed eight times more water than that of oxy-combustion plant. The high fraction of water vapor in the flue gas from oxy-combustion resulted in a dew point below the operating temperature of the FGD system, thereby, eliminating evaporative losses. In air-fired FDG systems, evaporation was the dominant source of water loss. Other water losses in the FGD system are related to the production of gypsum and to the purge system.

Shifting to a once-through cooling system, it can be seen clearly that the post-combustion capture plant is the major water consumer of the two plants. Calculation showed that the oxy-combustion plant only consumed 17 % of the water consumed in the post-combustion plant. This trend also applies in a smaller extent, when evaporative cooling towers are employed. The water consumption of the oxy-combustion plant accounted for 74% of that in the post-combustion power plant.

Considering the absorption process, the water losses were mainly related to evaporation of water in the treated vent gas. These losses could be reduced to a minimum, if a treated gas outlet temperature close to 30-31 °C is obtainable.

Water enters the power plant via humid air intake, moisture content in fuel, and hydrogen bound in the fuel. Consequently, a significant amount of condensate was available during air separation and CO_2 recovery. This water also included evaporated water in FGD process. Collection of condensate for water recovery could contribute to strongly reduce the water consumption in both power plants.

Suggestions for further work are to perform an extended study of the water consumption of CO_2 capture power plants, including an ultra-supercritical power plant. Additionally, investigation of water recovery and the impact a water treatment facility has on the overall power plant performance.

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Appendix A

Table 10.1: Stream data for ASU

		A0	A1-1	A1-H2O1	A1-2	A2-H2O1	A1-3	A-4	A2-1	A2-2	A2-3	A3-1	A3-2
Mass Flow	kg/s	627,29	626,48	0,81	621,98	4,50	619,27	618,25	335,96	335,96	335,96	6,03	6,03
Temperature	C	25,00	35,00	35,00	35,00	35,00	9,85	9,83	-173,75	-179,35	-188,97	-177,96	-181,46
Pressure	bar	1,01	5,60	2,38	5,60	5,60	5,55	5,50	5,45	5,40	1,45	5,40	5,35
Mole Fraction													
O2		0,205	0,206	0,000	0,208	0,000	0,210	0,210	0,393	0,393	0,393	0,006	0,006
N2		0,763	0,765	0,000	0,774	0,000	0,779	0,781	0,592	0,592	0,592	0,990	0,990
Ar		0,009	0,009	0,000	0,009	0,000	0,009	0,009	0,015	0,015	0,015	0,004	0,004
H2O		0,022	0,020	1,000	0,009	1,000	0,002	0,000	0,000	0,000	0,000	0,000	0,000
CO2		0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000
Total		1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000

		A3-3	A4-1	A4-2	A4-3	A4-4	A5-1	A5-2	A6-1	A6-2	A6-3	A7-1	A7-2
Mass Flow	kg/s	6,03	112,83	112,83	112,83	112,83	150,51	150,51	354,91	354,91	354,91	467,74	467,74
Temperature	C	-192,74	-177,96	5,35	1,20	1,17	-179,49	-6,21	-192,71	-178,38	-6,21	-6,25	25,00
Pressure	bar	1,40	5,40	1,00	1,00	1,00	1,50	1,47	1,40	1,38	1,35	1,17	1,14
Mole Fraction													
O2		0,006	0,003	0,003	0,003	0,003	0,950	0,950	0,002	0,002	0,002	0,002	0,002
N2		0,990	0,995	0,995	0,995	0,995	0,018	0,018	0,994	0,994	0,994	0,995	0,995
Ar		0,004	0,002	0,002	0,002	0,002	0,032	0,032	0,003	0,003	0,003	0,003	0,003
H2O		0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000
CO2		0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000
Total		1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000

Table 10.2: Stream data for boiler and FGD

		C-air	FG-1	FG-2	FG-3	O2-1	O2-2	FG-4	FG-5	FG-6	FG-7	O1-1	FG-8
Mass Flow	kg/s	12,60	737,81	737,81	737,81	2,10	2,10	714,28	514,28	514,29	514,29	148,41	662,70
Temperature	C	25,00	162,30	162,30	172,45	-6,21	71,27	57,22	57,22	65,67	73,20	-6,21	56,20
Pressure	bar	1,00	1,00	1,00	1,10	1,47	3,10	1,03	1,03	1,01	1,10	1,47	1,10
Mole Fraction													
O2		0,205	0,025	0,025	0,025	0,950	0,950	0,025	0,025	0,025	0,025	0,950	0,260
N2		0,763	0,080	0,080	0,080	0,018	0,018	0,085	0,085	0,085	0,085	0,018	0,068
AR		0,009	0,027	0,027	0,027	0,032	0,032	0,029	0,029	0,029	0,029	0,032	0,030
H2O		0,022	0,200	0,200	0,200	0,000	0,000	0,150	0,150	0,150	0,150	0,000	0,112
CO2		0,000	0,646	0,646	0,646	0,000	0,000	0,691	0,691	0,691	0,691	0,000	0,515
SO2		0,000	0,003	0,003	0,003	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000
H2		0,000	0,001	0,001	0,001	0,000	0,000	0,001	0,001	0,001	0,001	0,000	0,001
CO		0,000	0,016	0,016	0,016	0,000	0,000	0,017	0,017	0,017	0,017	0,000	0,013
NO		0,000	0,001	0,001	0,001	0,000	0,000	0,001	0,001	0,001	0,001	0,000	0,001
Total		1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000

		C0	Makeup water	Limestone slurry	Gympsum	FGD-CW
Mass Flow	kg/s	69,23	8,11	16,17	1,20	174,26
Temperature	C	20,00	15,00	15,00	57,00	16,00
Pressure	bar	1,00	1,00	1,10	1,00	2,00

Table 10.3: Stream data for CPU

		R1-1	R1-H2O1	R1-2	R1-H2O2	R1-3	R1-4	R1-5	R2-1	R2-2	R2-3	R3-1	R3-2
Mass Flow	kg/s	200,00	10,26	189,74	0,20	185,95	185,75	185,75	115,38	115,38	115,38	70,37	70,37
Temperature	C	57,22	35,00	35,00	35,11	35,00	35,11	-26,00	-26,00	-33,73	22,01	-26,00	-54,00
Pressure	bar	1,03	1,01	1,01	31,70	32,00	31,70	31,40	31,40	18,00	17,70	31,40	31,10
Mole Fraction													
O_2		0,025	0,000	0,028	0,000	0,029	0,030	0,030	0,005	0,005	0,005	0,065	0,065
N_2		0,085	0,000	0,095	0,000	0,100	0,100	0,100	0,015	0,015	0,015	0,223	0,223
AR		0,029	0,000	0,032	0,000	0,034	0,034	0,034	0,009	0,009	0,009	0,071	0,071
H_2O		0,150	1,000	0,047	1,000	0,002	0,000	0,000	0,000	0,000	0,000	0,000	0,000
CO_2		0,691	0,000	0,775	0,000	0,811	0,813	0,813	0,968	0,968	0,968	0,589	0,589
SO_2		0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000
H_2		0,001	0,000	0,001	0,000	0,001	0,001	0,001	0,000	0,000	0,000	0,003	0,003
CO		0,017	0,000	0,019	0,000	0,020	0,020	0,020	0,003	0,003	0,003	0,045	0,045
NO		0,001	0,000	0,002	0,000	0,002	0,002	0,002	0,000	0,000	0,000	0,003	0,003
Total		1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000
		R4-1	R4-2	R4-3	R4-4	R4-5	R4-6	R4-7	R5-1	R5-2	R5-3	R5-4	R5-5
Mass Flow	kg/s	39,10	39,10	39,10	39,10	39,10	39,10	39,10	31,27	31,27	31,27	31,27	39,10
Temperature	C	-54,00	-43,15	-55,62	-44,21	22,01	91,62	35,00	-54,00	-44,21	22,01	351,00	22,01
Pressure	bar	31,10	30,80	9,00	8,70	8,40	18,00	17,70	31,10	30,80	30,50	30,20	8,40
Mole Fraction													
O_2		0,009	0,009	0,009	0,009	0,009	0,009	0,009	0,119	0,119	0,119	0,119	0,009
N_2		0,023	0,023	0,023	0,023	0,023	0,023	0,023	0,418	0,418	0,418	0,418	0,023
AR		0,015	0,015	0,015	0,015	0,015	0,015	0,015	0,125	0,125	0,125	0,125	0,015
H_2O		0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000
CO_2		0,948	0,948	0,948	0,948	0,948	0,948	0,948	0,241	0,241	0,241	0,241	0,948
SO_2		0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000
H_2		0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,006	0,006	0,006	0,006	0,000
CO		0,005	0,005	0,005	0,005	0,005	0,005	0,005	0,084	0,084	0,084	0,084	0,005

NO	0,001	0,001	0,001	0,001	0,001	0,001	0,001	0,006	0,006	0,006	0,006	0,001
Total	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000

		R6-1	R6-2	R6-3	R6-4	R6-5
Mass Flow	kg/s	154,48	154,48	154,48	154,48	154,48
Temperature	C	25,29	35,00	25,00	44,95	25,00
Pressure	bar	17,70	78,00	78,00	150,00	150,00
Mole Fraction						
O2		0,006	0,006	0,006	0,006	0,006
N2		0,017	0,017	0,017	0,017	0,017
AR		0,010	0,010	0,010	0,010	0,010
H2O		0,000	0,000	0,000	0,000	0,000
CO2		0,963	0,963	0,963	0,963	0,963
SO2		0,000	0,000	0,000	0,000	0,000
H2		0,000	0,000	0,000	0,000	0,000
CO		0,003	0,003	0,003	0,003	0,003
NO		0,001	0,001	0,001	0,001	0,001
Total		1,000	1,000	1,000	1,000	1,000

Table 10.4: Stream data steam cycle

		S-1	S-2	S-3	S-4	S-5	S-6	S-7	S-8	S-9	S-10	S-11	S-12	E-1
Mass Flow	kg/s	612,78	612,22	564,65	503,09	503,09	478,21	423,73	390,45	373,97	358,33	342,85	342,85	47,57
Temperature	C	598,89	563,86	409,65	346,90	621,11	500,24	384,54	305,46	167,81	97,52	64,19	38,73	409,65
Pressure	bar	242,33	199,95	76,88	49,01	45,22	21,38	9,49	5,01	1,32	0,58	0,24	0,07	76,88
Vapor Frac		1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	0,974	0,929	1,000
		E-2	E-3	E-4	E-5	E-6	E-7	E-8	E-9	E-10	E-11	E-12	E-13	E-14
Mass Flow	kg/s	47,57	60,99	108,57	24,88	133,45	15,34	33,28	33,28	16,48	49,76	15,64	66,32	15,48
Temperature	C	290,39	346,90	260,84	500,24	214,26	384,54	305,46	108,87	167,81	86,81	97,52	66,62	64,19
Pressure	bar	74,81	49,01	47,54	21,38	20,74	9,49	5,01	1,38	1,32	0,62	0,58	0,27	0,24
Vapor Frac		0,000	1,000	0,000	1,000	0,000	1,000	1,000	0,000	1,000	0,000	1,000	0,000	0,974
		E-15	F-1	F-2	F-3	F-4	F-5	F-6	F-7	F-8	F-9	F-10	F-11	F-12
Mass Flow	kg/s	81,81	467,85	467,85	467,85	467,85	564,86	467,85	467,85	612,78	612,78	612,78	612,78	612,78
Temperature	C	45,13	38,39	38,54	39,06	60,85	81,43	103,49	147,26	176,38	181,97	214,78	259,71	290,08
Pressure	bar	0,10	0,07	17,24	16,89	16,55	15,86	15,51	15,17	9,21	289,58	289,24	288,89	288,55
Vapor Frac		0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000
		SS-1	SS-2	SS-3	SS-4	SS-5	SS-6	TD-1	TD-2	Sasu	Scpu	Makeup	Condenser-CW	
Mass Flow	kg/s	0,56	0,56	0,56	0,42	0,35	0,35	36,30	36,30	2,23	0,04	6,13	18 589,32	
Temperature	C	563,86	346,90	384,54	389,70	389,70	100,01	384,54	52,25	384,54	384,54	25,00	16,00	
Pressure	bar	199,95	49,01	9,49	9,49	9,49	1,01	9,49	0,14	9,49	9,49	1,01	2,00	
Vapor Frac		1,000	1,000	1,000	1,000	1,000	0,000	1,000	0,960	1,000	1,000	0,000	0,000	

Table 10.5: Stream data post-combustion plant

		1	2	3	4	5	6	7	8	9	10	11	12	13
Mass Flow	kg/s	71,65	2 830,28	787,04	781,48	17,07	826,08	36,90	6,63	222,05	222,05	12 545,78	29,98	13 026,50
Temperature	C	15,00	25,00	177,00	177,00	15,00	58,00	1,01	38,00	312,00	151,00	16,00	1,98	16,00
Pressure	bar	1,00	1,10	1,00	1,00	1,00	1,00	615,34	1,01	5,10	9,20	2,00	155,82	2,00
Mole Fraction														
O2		0,000	0,205	0.0247	0.0247	0,000	0,024	0,036	0,000	0,000	0,000	0,000	0,000	0,000
N2		0,000	0,763	0.7324	0.7324	0,000	0,678	0,895	0,000	0,000	0,000	0,000	0,000	0,000
AR		0,000	0,009	0.0087	0.0087	0,000	0,008		0,000	0,000	0,000	0,000	0,000	0,000
H2O		0,000	0,022	0.0870	0.0870	0,000	0,155	0,040	1,000	1,000	1,000	1,000	0,009	1,000
CO2		0,000	0,000	0.145	0.145	0,000	0,135	0,029	0,000	0,000	0,000	0,000	0,991	0,000
SO2		0,000	0,000	0.0021	0.0021	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000
H2		0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000
Total		0,000	1,000	0,000	0,000	0,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000

		14	15	16	17
Mass Flow	kg/s	25,00	525,30	12 311,00	5,12
Temperature	C	150,00	25,00	16,00	25,00
Pressure	bar	154,58	1,01	2,00	1,01
Mole Fraction					
O2		0,000	0,000	0,000	0,000
N2		0,000	0,000	0,000	0,000
AR		0,000	0,000	0,000	0,000
H2O		0,003	1,000	1,000	1,000
CO2		0,997	0,000	0,000	0,000
SO2		0,000	0,000	0,000	0,000
H2		0,000	0,000	0,000	0,000
Total		1,000	1,000	1,000	1,000

Table 10.6: Stream data amine absorption plant

		Flue Gas	FG-1	FG-2	Excess H2O	Semi - Rich	SRS-1	SRS-2	SRS-3	WR-1	WR-2	WR-3	WR-4	Clean Gas	Rich Solvent
Temperature	kg/s	58,00	30,00	40,18	30,00	51,34	35,00	35,00	35,00	38,25	35,00	35,00	38,25	36,90	44,52
Pressure	C	1,00	1,00	1,10	1,00	1,05	1,00	1,00	1,10	1,10	1,05	1,10	1,10	1,01	1,05
Mass Flow	bar	826,08	764,41	764,41	61,67	733,21	733,21	733,26	733,26	11 059,26	11 059,26	11 059,17	59,79	615,34	2 561,76
Mole Fraction															
N2		0,684	0,776	0,776	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,895	0,000
O2		0,024	0,027	0,027	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,036	0,000
H2O		0,156	0,042	0,042	1,000	0,829	0,829	0,829	0,829	0,995	0,995	0,995	0,995	0,040	0,825
CO2		0,136	0,154	0,154	0,000	0,056	0,056	0,056	0,056	0,000	0,000	0,000	0,000	0,029	0,059
MEA		0,000	0,000	0,000	0,000	0,115	0,115	0,115	0,115	0,005	0,005	0,005	0,005	0,000	0,117
Total		1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000
		RS-1	RS-2	RS-3	CO2	Lean Solvent	LS-1	LS-2	LS-3	LS-4	LS-5	LS-6	Conde-nsate	CO2 for transportation	
Temperature	kg/s	44,61	111,18	105,33	29,98	121,49	49,61	49,60	49,69	40,00	40,00	39,95	27,98	25,00	
Pressure	C	4,00	3,80	1,98	1,98	2,00	1,80	1,01	4,00	3,80	1,01	1,01	4,30	150,00	
Mass Flow	bar	2 561,76	2 561,76	2 561,76	155,82	2 405,95	2 406,12	2 408,76	2 408,76	2 408,76	2 408,76	2 468,38	1,24	154,58	
Mole Fraction															
N2		0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	
O2		0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	0,000	
H2O		0,825	0,825	0,825	0,009	0,853	0,853	0,853	0,853	0,853	0,853	0,652	0,996	0,003	
CO2		0,059	0,059	0,059	0,991	0,026	0,026	0,026	0,026	0,026	0,026	0,047	0,004	0,997	
MEA		0,117	0,117	0,117	0,000	0,121	0,121	0,121	0,121	0,121	0,121	0,302	0,000	0,000	
Total		1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	1,000	

Table 10.7: Cooling water requirement in the oxy-combustion plant

Stream	Mass Flow (kg/s)
ASU	
A1-CW1	2 988,11
FGD	
FGD-CW	174,26
CPU	
R1-CW1	653,41
R1-CW2	1 500,08
R4-CW1	46,13
Steam cycle	
Condenser	18 589,32
Total	23 951,31

Chilled water	
A2-CW1	198
R6-CW2	424,74
R6-CW3	237,61
Total	860,35

Table 10.8: Cooling water requirement in the post-combustion capture plant

Stream	Mass Flow (kg/s)
Amine Absorption	
DDC-CW	3 796,05
INTC-CW	914,73
WR-CW	3 319,30
COND-CW	4 515,69
LS-CW	1 863,57
CO2 Compression	
13	1 880,87
Steam Cycle	
16	12 311,00
Total	28 601,22

Chilled water	
Compression	210,73
Total	210,73