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Techno-economic assessment of post combustion CO₂ capture in an air blown IGCC plant

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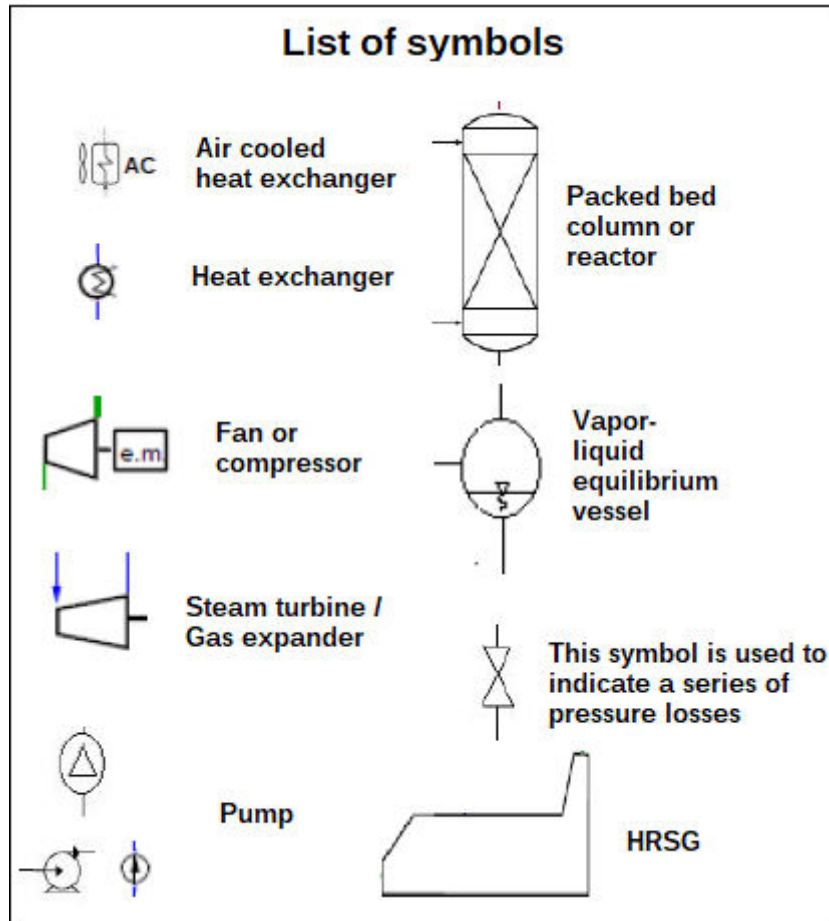
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ABSTRACT

In this thesis it is offered a techno-economic evaluation between different technologies for CO₂ post combustion capture by chemical absorption.

The power plant employed as a baseline is an air-blown IGCC plant operating with a sulfur-rich coal quality.

The investigated technologies are:

An amine-based solvent (Shell-Cansolv) which is meant to represent the state-of-the-art of post combustion CO₂ capture. Two innovative systems, known as *chilled ammonia* and *cooled ammonia* process, operating with aqueous ammonia.

Moreover, it is studied the opportunity of creating a close integration between the CO₂ capture unit and the AGR section of the IGCC plant. Two different processes for the recovery of sulfur are therefore investigated, creating a total of 6 different power plant configurations under scrutiny. The first process is a Claus unit, while the second one is a sulfuric acid plant working with a catalytic vanadium pentoxide converter.

Calculations were performed in a Matlab or Excel environment.

The thermodynamic model has been constructed by combining together in an original way a series of results drawn from existing sources in the open literature and modifying them, if needed, by means of approximate heuristics.

The economic evaluation has been developed according to a set of guidelines provided by the US Department of Energy. Capital costs are estimated with a scaling approach. The results have been challenged with a robustness analysis and a critical review.

The results indicate that the technology featuring the lowest LCOE is the cooled ammonia configuration integrated with a sulfuric acid plant.

SOMMARIO

In questa tesi viene effettuata una analisi tecno economica tra diverse tecnologie per la cattura della CO₂.

L'impianto di riferimento è un IGCC ad aria operante con una qualità di carbone ricca di zolfo. Le tecnologie di cattura studiate sono:

Un solvente amminico (Shell-Cansolv) che vuole rappresentare lo stato dell'arte della cattura della CO₂ post combustione. Due sistemi innovativi basati sull'ammoniaca acquosa, noti come ammoniaca in condizioni rispettivamente cooled and chilled.

Inoltre viene studiata l'opportunità di creare una stretta integrazione tra l'unità di cattura della CO₂ e l'unità AGR dell'impianto IGCC. Due differenti processi per la cattura dello zolfo vengono dunque considerati, creando un totale di 6 diverse configurazioni d'impianto. Il primo processo è una unità Claus, mentre il secondo si tratta di un impianto per la produzione di acido solforico mediante convertitore catalitico ad ossido di vanadio.

I calcoli sono svolti in un ambiente Matlab o Excel. Il modello termodinamico è stato costruito combinando in maniera originale una serie di risultati ricavati dalla letteratura pre-esistente, e modificandoli, se necessario, attraverso l'utilizzo di euristiche semplificate.

La valutazione economica è stata svolta secondo un insieme di linee guida fornite dal US Department of Energy. I costi del capitale sono stimati con un approccio di scaling. I risultati sono stati oggetto di un test di robustezza, nonché analisi critica.

Si è determinato che la tecnologia che offre il minore costo dell'elettricità è la configurazione con ammoniaca in condizioni cooled integrata con un impianto per la produzione di acido solforico.

EXTENDED ABSTRACT

1. Introduction

The increasingly high concentrations of carbon dioxide in the atmosphere are considered a grave reason of concern by the scientific[1] and international[6] community. In order to reduce the greenhouse gas emissions generated from the electricity generation sector, new sources of renewable energy such as solar PV and wind turbines have been introduced in the power supply. However, despite the achievement of important targets of cost reduction and market share [7], the intrinsic unpredictability of the natural element is expected to require growing costs for the integration of larger shares of renewables in the electricity network. Moreover, there are serious questions left open regarding the grid stability and the likelihood of power outages in a power system with an high penetration of the renewables in the event of adverse meteorological conditions[15].

Numerous energy storage techniques have been proposed to facilitate the integration of the renewable sources. Yet, there are no commercialized solutions for the storage of massive quantities of energy over time-periods in the order of several months[10], and there are only a few large scale projects involving key technologies for energy management applications[10] like Thermal Energy Storage (TES) and underground Compressed Air Energy Storage (CAES)[11].

For this reason, and given the large number of fossil-fuelled power plants of recent construction[16], it appears likely that conventional power plants based on fossil fuels and nuclear energy will continue to be the backbone of the power supply sector for the foreseeable future. On this regard, particular concerns are associated to the use of coal, which is scheduled by the IEA to reach a new record by 2025, driven by the growing electricity demand in East Asia and a rebound in coal consumption in Europe[16].

It appears therefore of fundamental importance the development of systems and methods capable of reducing the environmental impacts associated to the use of coal. The following fields of study appear especially relevant on this regard: i) methods aimed at minimizing the energy requirements associated to the process of coal mining[123], ii) techniques for the reduction of the coalbed methane emissions from existing coal mines[124], iii) new separation methods (beneficiation techniques) aimed at cleaning the coal from the gangue at the mine mouth while at the same time improving productivity and reducing waste by-products[23], iv) land rejuvenation activities aimed at restoring the original soil properties and local ecosystems in a degraded surface coal mine site after the end of the extraction operations[125], v) innovative chemical processes, like carbon capture and storage (CCS), aimed at reducing the emissions associated to the use of coal in industrial and energy-related applications.

On this last point, it is possible to draw the attention in particular on the following non-exhaustive list of topics of research interest: i) advanced methods for coal gasification, ii) techniques for ash removal and desulfuration of syngas streams at high temperature [30][31], iii) development of hydrogen-fired gas turbines [126], iv) development and testing of power plants incorporating oxycombustion[46] or chemical looping concepts[43], v) development of new chemical solvents for post combustion capture, vi) development of new high-performance carbon dioxide turbo-compressors[107], vii) methods of long term carbon sequestration[39].

1.1 Main methods of post combustion CO₂ capture

At present, the most mature and flexible CO₂ capture technology to reduce the greenhouse gas emissions from the existing coal-fired power plants is post-combustion capture by chemical absorption with amine mixtures[49]. This method can be considered long known since the first processes for the separation of acid gases by amine mixtures were patented in the first half of the 20th century[29], and they have been largely employed in the petrochemical industry ever since. The state-of-the-art for post combustion capture applications is considered to be a proprietary amine

mixture developed by Cansolv Inc (controlled by Shell from 2008). In particular, the Cansolv solvent has been successfully tested in two large scale projects for a maximum capacity of about 240 MWel (PetraNova project), and the net LHV efficiency of a modern USC coal-fired plant implementing such system is estimated around 32,7% according to a 2019 report from the US Energy Department [73].

An alternative solvent that has been studied in recent years by AlstomPower and Delta Electricity is aqueous ammonia. Alstom conducted from 2006 to 2013 a series of pilot projects (Pleasant Prairie, Mountaineer coal plant, Mongstad refinery) capturing the emissions from different typologies of thermal plants for a maximum capacity of 54 MW [97]. The specific design tested (and patented) by Alstom is the so called Chilled-Ammonia-Process (CAP), which according to the literature features an inlet temperature of the solvent in the absorber within the range 0-10°C [98].

The pilot project managed by Delta Electricity at the Munmorah coal power station in Australia was instead operating at near ambient conditions (15-20°C) [99].

As a matter of fact, each of the abovementioned systems features some characteristic issues that have to be taken into careful consideration. In particular, chemical absorption systems that employ amines are known to be generally very aggressive against carbon steel, meaning that more expensive stainless steel grades have to be used for the construction of the equipment[55]. Moreover, amine-based systems tend to discharge into the surrounding environment a non-negligible amount of solvent over time, the entire consequences of which are still unknown due to a lack of environmental data regarding the spatial and temporal patterns, as well as the effects, of the amine concentrations in the surface waters[60][65].

On the other hand, CO₂ capture systems based on ammonia are affected by problems of solid precipitation[102], which may cause plugging, and evaporation of the solvent in the absorption and stripping columns (also known as *ammonia slip*[100]), which causes the solvent to leave the system.

In general, solid precipitation can be either avoided through a proper choice of the operating parameters, or managed through proper equipment selection, although at the price of a greater operational complexity. Ammonia slip can be instead managed by introducing a further series of treatments (*water wash* or *acid wash*) before discharging the flue gases in the atmosphere.

1.2 State-of-the-art of CCS costing methods

Given the potential importance of the CCS technologies for greenhouse gas reduction, there are several different types of public, each of them with a different background and objectives, that may be interested in reading a study in which the cost of a CCS system is estimated. Various institutions (NETL, EBTF, IPCC, IEA), as well as groups of independent researchers, have published a series of lengthy reports over the years on this topic. However, as documented by E.Rubin et al. in a series of articles[71][66][70][72], this has led to a proliferation of different accounting standards and methodologies in the field of CCS costing, with the effect of creating the potential for confusion or misunderstandings in unexperienced readers.

In particular it is highlighted the fact that important cost objects are often treated differently in different studies, some example of systematic bias are documented, and it is pointed out how the estimate of the equipment cost is an activity which is outsourced in many important studies to third party engineering contractors, which creates a problem of lack of transparency from the point of view of the final reader.

In conclusion, the authors call for *a concerted effort to improve the understanding and communication of CCS cost estimates within the technical and policy communities (2012)*.

It is here worth to report a useful categorization that the author uses to classify the majority of the published studies on CCS costing[72]. The first type of analysis is called technological assessment, and it can be broadly described as a study where, given a certain set of baseline assumptions,

several technological systems are compared each other in order to determine the best configuration. These analysis are often interesting for researchers and R&D planners, but, due to the arbitrariness of some of the assumptions sometimes adopted, they may not be good predictors of the cost of a certain system in absolute terms.

The second category of analysis is called policy assessment, and it can be described as a study in which the objective is instead to produce precisely the most realistic estimate possible for the cost of a specific project.

1.3 Objective of the work

In the present work, it is offered a techno-economic analysis investigating the most viable solution to reduce the GHG emissions from an air-blown IGCC plant operating with a sulfur-rich coal quality. Three different technologies for carbon capture are considered: i) the Cansolv solvent, ii) the chilled ammonia process, iii) the so-called cooled ammonia process. All power plants are considered to be green field facilities. The carbon capture ratio is 90% in all cases.

Moreover, it is studied the opportunity of creating a close integration between the CO₂ capture unit and the acid gas removal section of the IGCC plant. For this reason, two different processes for the recovery of sulfur are also investigated, creating a total of 6 different power plant configurations under scrutiny. The first sulfur recovery process is a Claus unit, while the second one is a sulfuric acid plant working with a catalytic vanadium pentoxide converter[87][89].

Importantly, the analysis is structured as a technological assessment relying on strong baseline hypothesis. Therefore the results should not be taken as a proxy for the real cost of the projects investigated, but rather as part of a preliminary screening process in which the most economically attractive alternatives are identified before further analysis.

2. Methodology

The development of the thermodynamic model of the power plants investigated has been performed with a modular approach by independently constructing a series of performance models for each of the functional areas of the system, which have been then aggregated. The performance models are generally consisting in a block-diagram associated with a series of mass/energy flows, together with a rough estimate for the main utilities (electricity, steam, cooling water) required, often sourced from the previously existing literature.

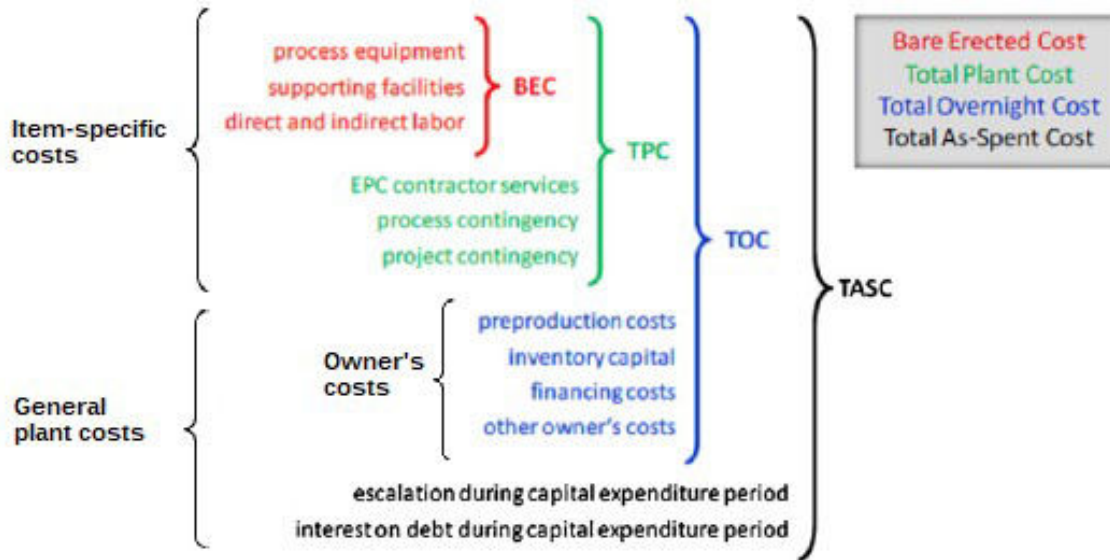
More in detail, the gasification train and the CCS plants operating with ammonia have been modelled according to the information provided in the article [75] (from Bonalumi, Giuffrida). The CCS plant operating with the proprietary Cansolv solvent has been modelled according to the data provided in open source reports[73][91]. The CO₂ compression unit, which is imagined to be consisting of a multi stage integral gear intercooled compressor[106], has been modelled according to the data provided in Case 12B of the 2019 Cost and Performance Baseline Report[73] published by the NETL (National Energy Tech. Laboratories). The bottoming steam cycle has been built from scratch according to a series of parameters that have been assumed to be model the behavior of the steam turbine (the parameters are listed in Table 7 in Chapter 3). New simple models have been built in a Matlab environment to simulate also other subsystems such as the water treatment unit and the sulfur recovery unit. Water was modelled with the RK equation of state.

Non rigorous heuristics have been employed in some instances to avoid overly complicated calculations. For example a similar approach has been adopted to quantify the ammonia slip phenomenon in the cases with cooled NH₃, as well as the amount of cooling water required from the various CCS plants.

The economic analysis has been performed following the approach presented in the NETL guidelines[74].

In detail, an itemized list of the equipments constituting the power plant has been compiled, and a cost estimate has been developed for each item through the capital cost scaling method described in the NETL guidelines[84]. The general structure of the method is schematically depicted in Figure I.

Figure I: breakdown structure of the capital costs considered in the economic analysis (from[74])



The capital cost quotations supplied in Case 1A of the 2019 NETL report have been used as initial data points, while the scaling exponents have been sourced from the document “Capital Cost Scaling Methodology: Revision 4 Report”[84].

The estimates for the capital cost of the CCS plants operating with ammonia have been instead developed using the equipment cost values from the article [104] (from Bonalumi, Valenti) as data points for the scaling method.

The total capital cost budget necessary for each equipment has been then determined by adding engineering fees, project contingency and process contingency costs, which were expressed as percentages of the equipment cost. Project contingencies have been set to the values suggested by the NETL, ranging from 17.5% (Detailed Design) to 35% (Preliminary Design) for the majority of the items. Process contingencies have been set to the same value of 17% (full sized modules have been operated) for the Cansolv process and the cooled NH₃ process. An higher value of 50% (new concept with limited data) has been set for the chilled NH₃ process due to the uncertainties related to the handling of solid precipitates.

To further validate the results, a sensitivity test has been performed at the end of the economic analysis. In particular, the robustness of the main conclusions has been challenged by testing the behavior of the model in presence of a series of less favourable hypothesis.

All the calculations have been performed in an Excel or Matlab environment.

3 – IGCC and CCS plant layout

Each of the power plants considered in this work consists in two gasification islands and two combustion turbines, with two heat recovery steam generators sharing the same steam turbine. At the same time, each of the heat recovery steam generators is coupled with a dedicated CCS unit. In the following paragraphs it is provided a general description of the IGCC and CCS plants.

3.1 The IGCC plant

In Figure II and Table I it is provided a schematic representation of the gasification island (in the reference case without CO₂ capture) and the main subsystems connected to it.

Input of the process is a stream of sulfur-rich coal Illinois n6, that is fed into a gasification reactor. The gasifier (MHI gasifier) is a double stage entrained flow reactor with water walls[76].

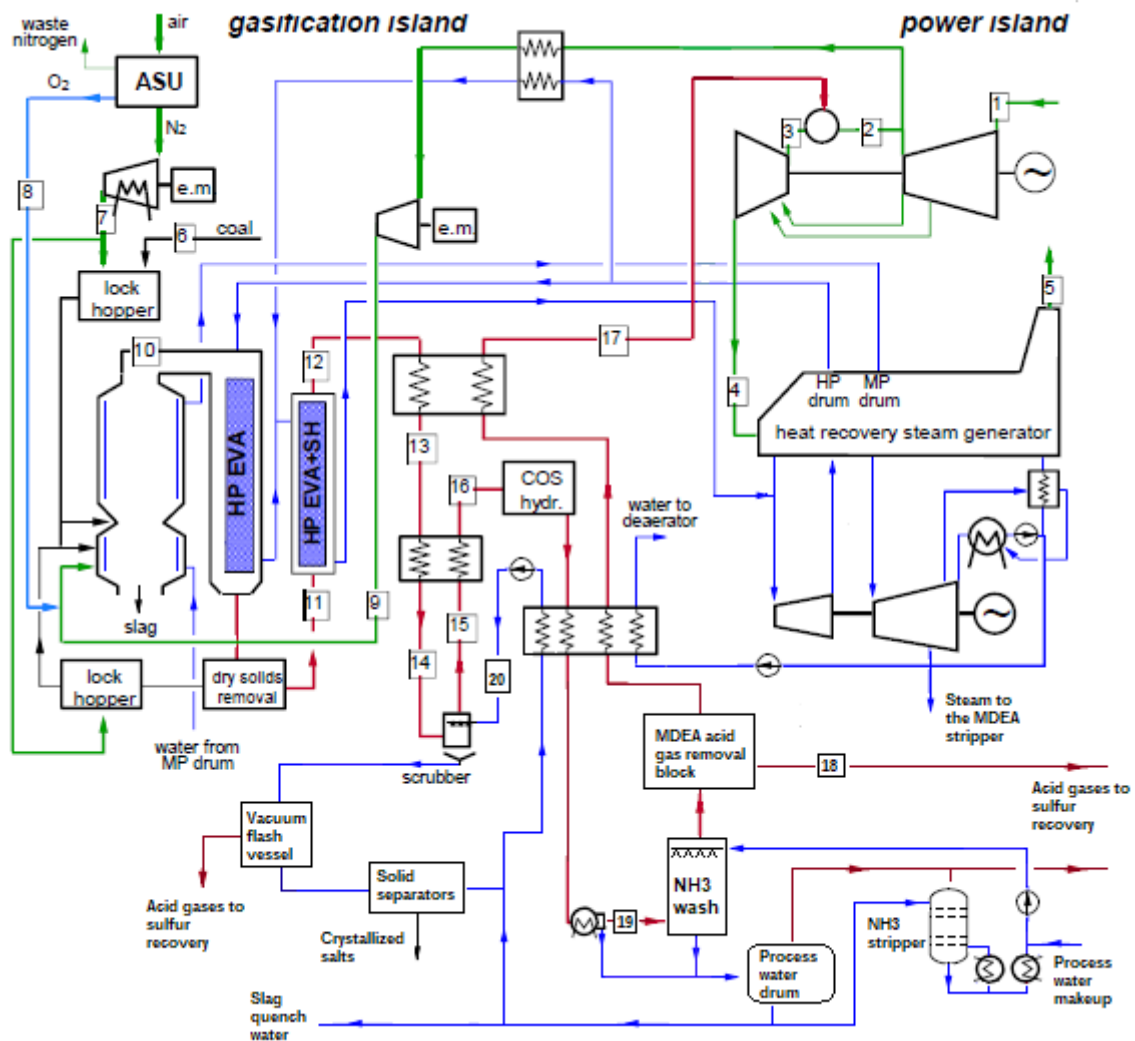
The technology that is utilized to feed solid coal in the pressurized reactor is a dry lock hopper system, which pneumatically conveys batches of coal by means of an high pressure flow of inerts. Nitrogen is the inert gas used in the reference plant, and it is delivered by a small air separation unit (ASU) specifically for this purpose.

As a by-product, the ASU also produces a stream of oxygen that is used to enrich the air used as an input for the gasification.

In the first stage of the gasifier (combustor) compressed air and coal produce an extremely hot (1900°C) flow of combustion products.

In the reductor stage the remaining coal is added, without further oxygen. Due to the high temperatures, solid coal particles react endothermically with the surrounding atmosphere, cooling down the system of about 700°C[76].

Figure II: schematic of the IGCC plant (reference case - without co₂ capture)



The flow exiting the top of the reactor is a pressurized mixture of carbon monoxide and hydrogen importantly diluted with nitrogen and other inerts. Such mixture is called syngas. It also contains ashes and unconverted carbon, as well as various gaseous pollutants such as H₂S and COS. The syngas is cooled from 1200°C to 900°C and subsequently passed into a cyclone and a candle filter where solid particles are collected and circulated back to the reaction chamber. The recirculation of solids is needed to obtain a complete conversion of solid carbon, which is not granted through a single passage due to the relatively poor kinetic conditions in the reductor stage. An important feature to notice in the design of the MHI gasifier is that the system is supposed to be able to handle with great flexibility coals with different characteristics, and in particular with relatively low ash melting point[77][78]. As a consequence, the syngas at the gasifier outlet (~1200°C) can be cooled in a normal convective heat exchanger without the need of exotic solutions like radiant coolers or quench baths.

After filtration, the syngas is further cooled down to 137°C in a series of heat exchangers, and it is purified in a ejector venturi-type scrubber from the particulate matter. The syngas is then reheated up to 180°C, and it is passed into a catalytic reactor aimed at converting COS into H₂S. Downstream to the COS hydrogenation reactor, the syngas is cooled down to 35°C in order to facilitate the removal of H₂S. This cooling section is called low temperature heat recovery (LTHR). The H₂S removal process is performed by chemical absorption in a packed column. The solvent adopted is mono-diethanol-ammine (MDEA).

Since the AGR solvent is sensible to the presence of NH₃ in the syngas[79], it has been decided to consider as part of the process also a water wash column preceding the AGR section. Aim of the water wash is to reduce to traces the ammonia in the syngas by spraying it with fresh water. After the sulfur removal block, the syngas flow is reheated in a regenerative heat exchanger and burned in the combustor of a combustion turbine. The compressor of the CT unit provides also the stream of compressed air that is used in the gasifier.

The combustion turbine inlet temperature is considered to be 1360°C. Being the syngas already significantly diluted with nitrogen, it is not necessary to mix the syngas with steam or nitrogen to obtain a stoichiometric flame temperature compatible with acceptable NO_x emissions at the turbine outlet.

Table I: Temperature, pressure, flow rate and composition of the main streams of the air-blown IGCC shown in Figure II

	Temperature °C	Pressure Bar	Mass flow (kg/sec)	Ar	CO	CO ₂	H ₂	H ₂ O	H ₂ S	N ₂	O ₂	CH ₄
1	15	1,01	611	0,92	0	0,03	0	1,03	0	77,28	20,73	0
2	417,6	18,16	378,8	0,92	0	0,03	0	1,03	0	77,28	20,73	0
3	1438,8	17,61	540	0,89	0	10,62	0	5,65	0	75,24	7,6	0
4	611,1	1,04	665	0,89	0	8,59	0	4,77	0	75,63	10,12	0
5	115	1,01	665	0,89	0	8,59	0	4,77	0	75,63	10,12	0
6	15		39,1	Coal as received (%wt: 61.27 C, 4.69H, 8.83 O, 1.1 N, 3.41 S, 12 moisture, 8.7 ash)								
7	80	56,13	19,9	0	0	0	0	0	0	100	0	0
8	15	35	7,3	3,09	0	0	0	0	0	1,91	95	0
9	477,4	33,02	107,1	0,92	0	0,03	0	1,03	0	77,28	20,73	0
10	1200	28,06	172,4	0,61	25,28	3,62	11,91	4,22	0,62	53,19	0	0,56
11	900	28,06	170	0,61	25,28	3,62	11,91	4,22	0,62	53,19	0	0,56
12	350	27,5	170	0,61	25,28	3,62	11,91	4,22	0,62	53,19	0	0,56
13	206,6	26,95	170	0,61	25,28	3,62	11,91	4,22	0,62	53,19	0	0,56
14	137	26,41	170	0,61	25,28	3,62	11,91	4,22	0,62	53,19	0	0,56
15	112	25,89	172,2	0,6	24,83	3,56	11,7	5,92	0,6	52,24	0	0,55
16	180	25,37	172,2	0,6	24,83	3,56	11,7	5,92	0,6	52,24	0	0,55
17	250	23,16	161,2	0,64	26,74	2,93	12,6	0,24	0	56,26	0	0,59
18*	45	1,01	3,93	0	0	58,36	0	0	41,64	0	0	0
19*	35	~24,5	165,13	0,63	26,33	3,77	12,4	0,23	0,63	55,4	0	0,58
20*	137	25,89	65,45	0	0	0	0	100	0	0	0	0

The layout of the gasification island in the cases with CO₂ capture is very similar to that depicted in Figure I. The most important exceptions are the ones following:

- the lock hoppers are fed with a flow of high pressure carbon dioxide extracted from the CO₂ capture unit. As a consequence, the ASU is not necessary
- since the gasifier is not fed with oxygen from the ASU, a greater quantity of pressurized air is channelled to the gasifier through the CT compressor and the air booster, with the end result of increasing the mass flow of syngas of about 10% with respect to the configuration without CCS

3.2 The CCS plant in chilled mode

As depicted in Figure III, the stream of flue gases exiting the HRSG is cooled in a series of three contact coolers where it releases moisture. The first contact cooler uses ambient air (assumed at 15°C) as a refrigerant, while the others use chilled water. The gas is forced to circulate in the various columns by means of a fan. Then, at a temperature of 7°C, the gas enters the bottom of the CO₂ absorption column.

The CCS section is assumed to be identical to that described in the reference article [75].

In particular, the system is formed by an absorption and a regeneration columns, which are connected with a regenerative heat exchanger. Other key components are: i) an high pressure pump which has to circulate the rich solution to the regenerator, ii) an heat exchanger (and a chilling plant) which has to cool down the lean solvent to about 7°C before entering the absorption column, iii) an hydrocyclone, which has to separate the solid precipitates from the liquid at the bottom of the absorber. The solids are sent to the regenerator with a small fraction of liquid, while the bulk of the liquid is chilled in an heat exchanger and recycled back to the absorber.

The ammonia slip effect from the absorber is contained by washing the flue gases with sulfuric acid in a packed column (WT1) before releasing them from the stack.

The CO₂ stream exiting the regenerator is instead passed into a condenser and a water wash tower aimed at reducing the ammonia slip. The stream is then compressed from 5 bar to supercritical conditions in a seven-stage integral-gear centrifugal intercooled compressor. The stream of carbon dioxide is also dehydrated through a series of water knockouts and a scrubbing operation with a Triethylene-glycol (TEG) solution[113]. Finally, after recycling a certain amount of CO₂ for coal loading, the supercritical CO₂ stream is pumped to storage.

3.3 The CCS plant in cooled mode

The layout in Figure IV for the CCS plant in cooled mode is relatively similar to that of the plant in chilled mode. The main differences to be highlighted are the following ones, as evidenced also in the reference article [75] from Bonalumi:

- Only two contact coolers are sufficient at the outlet of the HRSG unit, and the process of cooling is performed only down to the temperature of 20°C, considering an air temperature of 15°C and a difference of 5°C in the exchangers.
- The temperature at the top of the absorber is 20°C, while it is 25°C at the bottom.
- Ammonia slip is more significant in cooled mode, therefore an absorption-desorption cycle by means of fresh water is necessary before the acid wash column in order to obtain a level of ammonia slip reducible with the amounts of sulfur available from the coal entering the plant. This step is called water wash. In particular, a fraction of the gaseous ammonia exiting from the top of the absorber is reduced in a wash tower (WT3), producing an ammonia-rich solution which can be pumped to a stripper, where the ammonia is finally recovered and delivered back to the absorber.

The main operational parameters defining the CCS units with cooled and chilled ammonia are listed in Table II.

Figure III: schematic representation of the chilled-ammonia CO₂ capture unit (from [75])

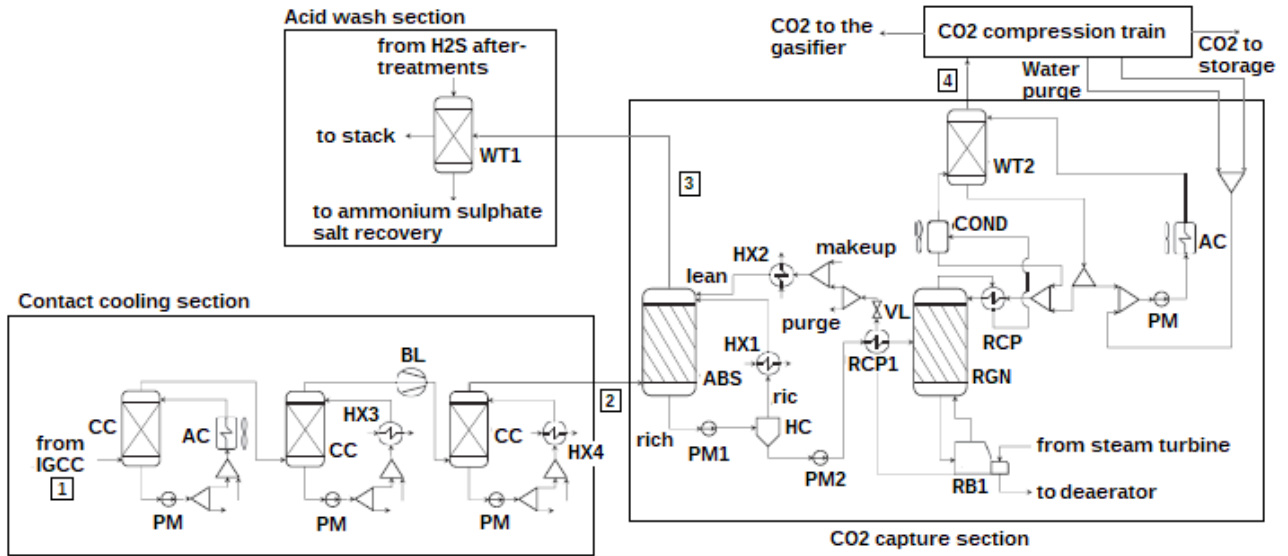


Figure IV: schematic representation of the cooled-ammonia CO₂ capture unit (from [75])

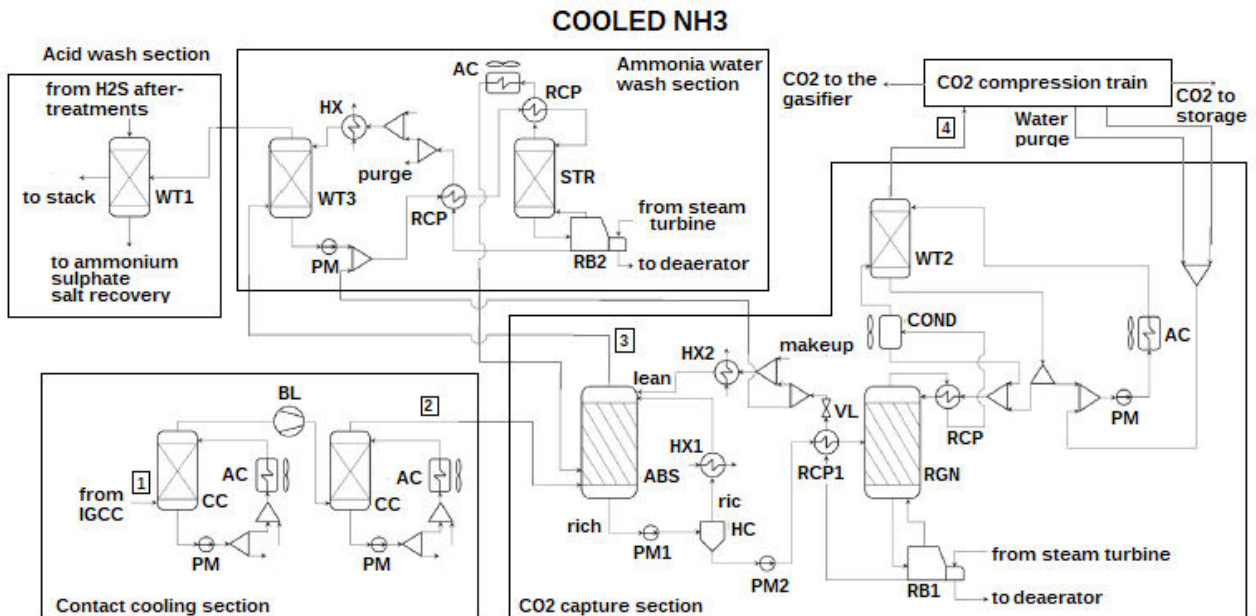


Table II: operating parameters for the ammonia CO2 capture process (from [75]):

	Unit of measure	Chilled	Cooled
Temperature absorber inlet streams	°C	7	20
CO2 capture ratio	-	90,00%	90,00%
Ammonia initial wt concentration (upon entering the absorber)	-	20,00%	10,00%
NH3/CO2 ratio	-	5	4,75
Regeneration pressure	bar	5	5
Solvent recycle ratio (around the absorber)	-	80,00%	10,00%

Where:

$$\text{Ammonia concentration} \stackrel{\text{def}}{=} \frac{m_{NH_3}}{m_{NH_3} + m_{H_2O}} \Big|_{\text{lean}} \quad \text{NH}_3 \text{ to CO}_2 \text{ ratio} \stackrel{\text{def}}{=} \frac{n_{NH_3} \Big|_{\text{lean}}}{n_{CO_2} \Big|_{\text{gas-inlet}}}$$

3.4 The Cansolv-type CCS plant

The layout of the CCS plant using the Cansolv solvent is represented in Figure V. The design is relatively similar to the one of the previous cases. It is possible to recognize the absorption and regeneration column, the circulation pumps, the regenerative heat exchanger, the lean heat exchanger and the condenser. The flue gases exiting the HRSG are quenched in a contact cooler and fed in the absorption column by means of a fan. The absorption column consists in three packed bed intercooled stages. In the upper part of the column the flue gases are washed with water in order to minimize the loss of solvent.

The regeneration of the solvent is realized in the cansolv system not only through a thermal input from the reboiler, but also with the help of a mechanical compressor.

The system is finally equipped with a tank for the solvent, as well as a thermal reclaimer, which is a component aimed at eliminating from the system eventual impurities (heat stable salts HSS) caused by the progressive degradation of the solvent over time.

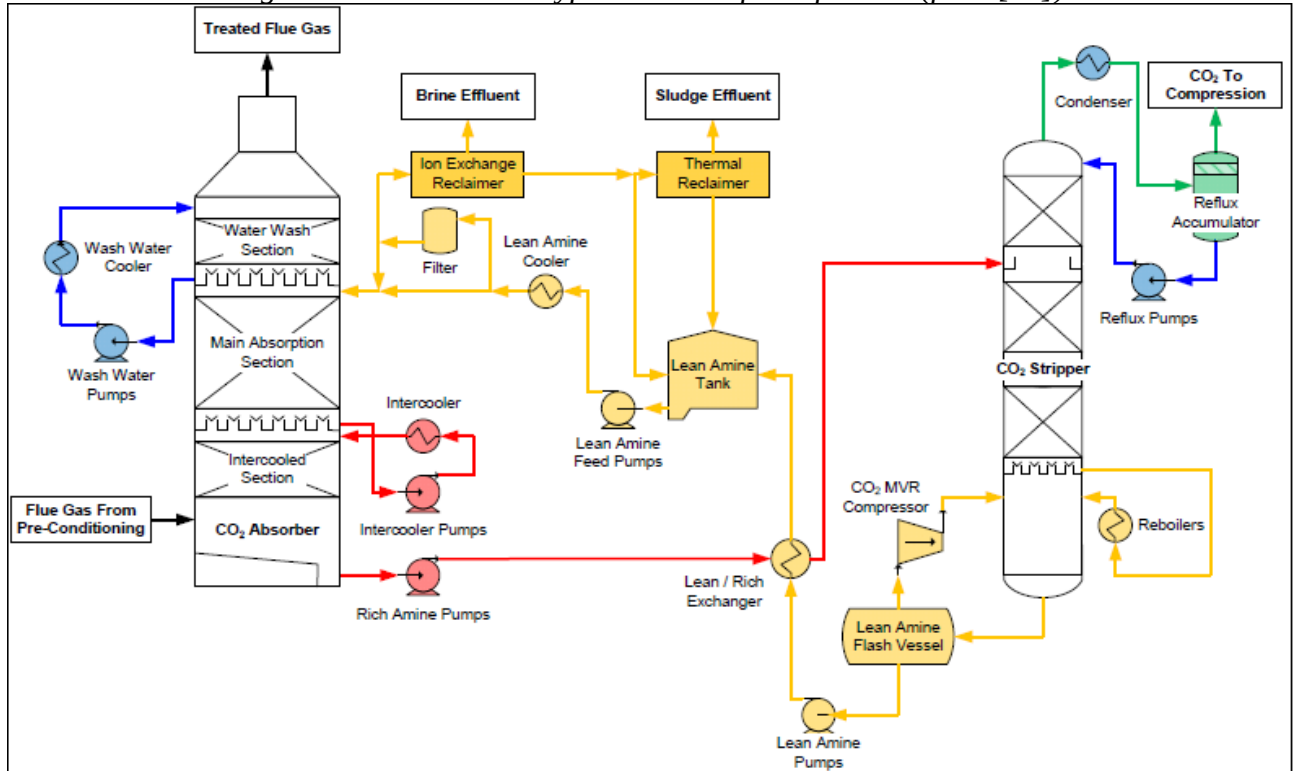
Another element to keep into consideration is the regeneration pressure, which is only 2 bar, whereas the value in the previous cases was 5 bar. As a consequence, an additional stage is needed in the compressor.

The main parameters used for the modellization of the Cansolv CCS unit are listed in Table III.

Table III: main parameters and utilities associated to the Cansolv CO2 capture system

	unit of Measure	value	source of the data
Reboiler Heat Duty	MJ(th)/kg co2 captured	2,33	Direct data from [91]
Auxiliary electric * Consumption	kWh/kg co2 captured	0,0523	Modified data from Case 12B in [73]
CO2 discharge* Pressure	bar	2	Data directly employed
Steam pressure * In the reboiler	bar	5,1	From case 12B in [73]

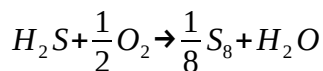
Figure V: Shell Cansolv typical CO₂ capture process (from [91])



3.5 The sulfur recovery plant

The sulfur recovery plant is the unit which is tasked with the objective of treating the flow of acid gases, mainly composed by hydrogen sulfide and carbon dioxide, separated from the syngas during the desulfurization treatment in the AGR section.

In this work two different sulfur recovery plants are taken into consideration. The first one is a conventional three stage Claus process, identical to that described in [73], for the production elemental sulfur. The system (not represented here) consists in an adiabatic furnace in which one third of the gas is oxydized with air in order to produce sulfur dioxide. Then the gas is passed into a series of intercooled catalytic beds in which sulfur dioxide and hydrogen sulfide react to form elemental sulfur and water. The overall reaction is:

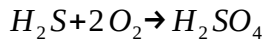


The conversion rate for a three stage system is around 98%. According to the calculations performed, given the amount of hydrogen sulfide available, it is possible to recover around 8 MW of heat in each train of the Claus process. Such heat has been supposed to be of sufficient quality to be able to address the demand for steam of various internal processes like waste water treatment and ammonia stripping.

The second plant (represented in Fig VI) is a system for the production of sulfuric acid. The flow of acid gases is completely oxidized with air to form sulfur dioxide in an adiabatic furnace followed by an heat recovery steel tube boiler. In particular, given the amount of sulfur available in the coal, it should be possible to produce about 16,5 MW of steam at 36 bar for each train. Then the gas is dried and passed into a three bed vanadium pentoxide catalytic converter, where sulfur dioxide is further oxidized to sulfur trioxide. Finally, the gas is passed into a bubble tank reactor

where sulfur trioxide and water react to form liquid sulfuric acid. The remaining gases are discharged from a secondary stack.

The overall reaction is:



The flow of sulfuric acid can be then used for the operation of acid wash, where it reacts with ammonia to form ammonium sulphate, which is a valuable product that can be sold to the market as a fertilizer.

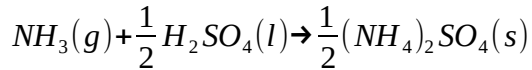
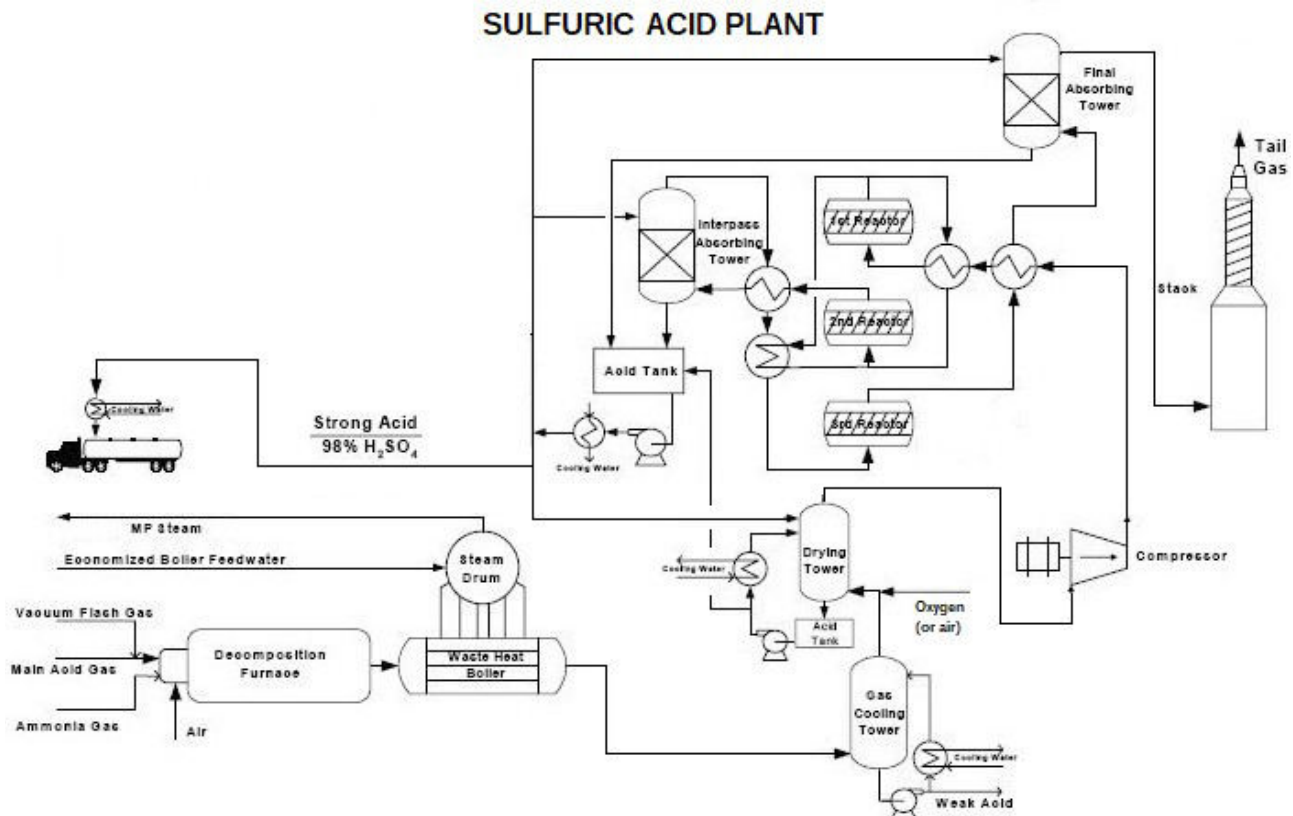


Figure VI: schematic representation of a sulfuric acid plant integrated with an IGCC plant (from [87])



4. Results and discussion

In this section the main results of the techno-economic analysis are presented and discussed. In Table IV the most significant indexes characterizing the thermodynamic performance of the various power plant configurations are listed. The following points are in particular worth of attention:

- The power production from the steam turbine decreases in the carbon-capture cases with respect to the reference situation due to the need of extracting steam for solvent regeneration. The reduction is relatively limited in the cases operating with chilled ammonia, while it is much more severe in the other cases.
- The power plant configurations using cooled-ammonia show the greatest overall thermodynamic efficiency, with Case 4 displaying the highest value.

The plant configurations integrated with the Cansolv-type CCS unit show instead the poorest performance, due to a very significant reduction in the power produced from the steam turbine.

The chilled-ammonia configurations are in an intermediate position: while on the one hand the

power production from the steam turbine is elevated, on the other hand the electric consumptions of the CCS auxiliaries are also very significant.

- Thanks to the greater amount of heat recoverable, the power plant configurations that are integrated with a sulfuric acid plant show a power production from the steam turbine which is around 8-9 MW higher than the value achieved with a Claus-type sulfur recovery unit.
- The IGCC-related auxiliary electric consumptions of the carbon-capture cases are significantly lower than the reference case. This is mainly a consequence of the elimination of the Air separation unit and the lock hoppers nitrogen compressors. The increase of work of the air boosters does not compensate for the elimination of such components.

Table IV: summary of the results in terms of thermodynamic performance

	Reference Case (without Capture)	Case 1 claus Cooled	Case 2 claus Chilled	Case 3 claus Cansolv	Case 4 sulfur Cooled	Case 5 sulfur Chilled	Case 6 cansolv Sulfur
Steam turbine gross power	485,89	424,98	473,48	401,17	432,86	481,43	409,46
Gross electric power output, MW	1039,09	956,18	1004,68	932,37	964,06	1012,63	940,66
IGCC plant Auxiliaries	107,202754	78,71	78,81	78,25	79,2	79,65	78,75
CCS plant Auxiliaries	0	69,36	145,14	81,99	69,36	145,14	81,99
Net power output	931,88	808,12	780,73	772,14	815,5	787,85	779,93
Net electric LHV Efficiency %	48,00	41,21	39,81	39,37	41,59	40,17	39,77
Specific emissions Kg co2 / Mwh	660,22	87,64	90,71	91,72	86,85	89,89	90,81
SPECCA MJ/kg co2	-	2,15	2,7	2,89	2,02	2,55	2,72

In Table V the results of the economic analysis are instead reported. It should be noted in particular that:

- The configurations with ammonia feature much lower variable costs when they are integrated with a sulfuric acid plant. This is explained from the obvious fact that sulfuric acid for the acid wash operation do not have to be purchased from the market anymore.
- The Claus - type sulfur recovery unit has been found to be about twice more expensive than the sulfuric acid plant. This result reinforces the conclusion suggested by the previous point, however, since it cannot be easily explained, it should be taken with care.
- The capital cost associated to the cooled ammonia CCS plant has been found to be lower than the capital cost associated to the Cansolv unit. This result can be explained from the higher corrosivity of the Cansolv solvent, which requires the use of stainless steel for the column internals. Moreover, the pressure of regeneration is higher in the configurations with ammonia, which allows for a more compact design of the regeneration column.
- The case number 4 (cooled nh3 with sulfuric acid production) features the lowest LCOE among the cases analyzed. In particular it is more competitive than the Cansolv CCS plant of about 10\$/MWh. The key factors of success appears to be the lower capital cost of the cooled NH3 CCS plant and the greater thermodynamic efficiency of case 4. It has been verified that these two factors have a nearly equal weight in numerical terms.
- If it is imagined that elemental sulfur, sulfuric acid and ammonium sulphate can be sold to the market at 150\$/ton, then Case 4 is even more competitive than all other configurations due to its large production of ammonium sulphate.
- Case number 5 (chilled ammonia with sulfuric acid production) is not particularly attractive due to its large capital cost. The situation changes drastically if it is made the assumption

that the electricity consumptions required for the refrigeration of the solvent are strongly reduced, which might be verified in cold geographical locations.

Table V: summary of the results of the economic analysis

	Case 1 claus Cooled	Case 2 claus Chilled	Case 3 claus Cansolv	Case 4 sulfur Cooled	Case 5 sulfur Chilled	Case 6 cansolv Sulfur
Capital \$/Mwh	79,88	85,05	87,88	76,72	81,82	84,67
Fixed	28,15	29,96	30,99	27,09	28,86	29,85
Variable	32,81	30,7	23,39	23,77	23,72	22,79
Fuel	17,77	18,39	18,6	17,61	18,23	18,41
Total LCOE	158,62	164,12	160,87	145,21	152,65	155,74
LCOE after all secondary sources of income are accounted	156,85	162,29	159,02	138,02	145,72	150,16

The sensitivity analysis has indicated that Case 4 would still be more economically convenient than Case 6 even if the following hypothesis were true: i) the cost of ammonia was 600\$/ton, ii) the steam for the CO₂ solvent regeneration in Case 4 was extracted at 2 bar from the crossover, iii) the isentropic efficiency of the steam turbine was 5% points lower than previously assumed, iv) the labor cost associated to Case 4 was to be doubled, v) the capital cost necessary to build the CCS plant was equal for Case 4 and Case 6.

As a matter of fact, the sensitivity analysis has shown that, even if all the abovementioned hypothesis were simultaneously verified, Case 6 would still not be significantly more competitive than Case 4. In other words, the sensitivity analysis shows that the conclusions of the economic analysis (Case 4 has a lower LCOE than Case 6) are verified in a large interval of values of the parameters.

5 Conclusions

This work indicates that a CCS configuration with cooled ammonia is clearly the best way to reduce the GHG emissions from an air blown IGCC plant, provided that sufficient sulfuric acid can be produced within the system.

It has been found that an important factor behind this result is constituted by the lower capital cost necessary for the cooled NH₃ CCS plant with respect to the Cansolv CCS plant. This is partially a consequence of the assumption of carbon steel usage for the pipings of the cooled NH₃ system (from [104]).

However it has not been possible to find an open-source study assessing the corrosivity of aqueous ammonia in a CCS environment, therefore the safety of carbon steel in similar applications is a topic that needs to be further investigation.

Another potential driver for the competitiveness of cooled ammonia has been found to be the possibility of producing ammonium sulphate crystals as a by-product of the CO₂ capture process. As part of this work, a tentative design has been proposed regarding the section of the plant dedicated to the operation of acid wash and recovery of the salt crystals.

The proposed process has not been characterized in a quantitative manner due to the lack of data regarding the properties of the mother liquor. A more detailed analysis regarding the costs associated to the salt recovery process is therefore necessary in the future.

1 INTRODUCTION

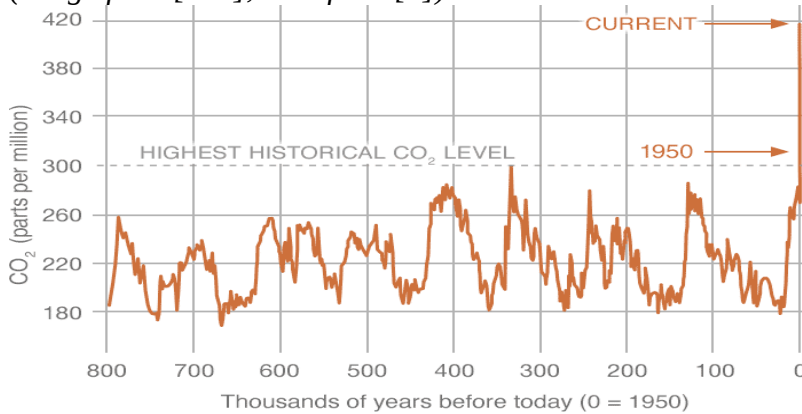
1.1 Recent trends in the power sector: renewables, energy storage and ongoing reliance on fossil fuels

Gaseous emissions from human activities are changing the composition of the atmosphere in what has been called “a vast geophysical experiment”[1].

According to air samples obtained from drillings in the Antarctic ice sheets, the current average level of carbon dioxide in the atmosphere is about one third higher than any point in the last 800 thousands of years [2]. This is widely believed to be a leading factor in the observed increase in the surface temperatures during the last century.

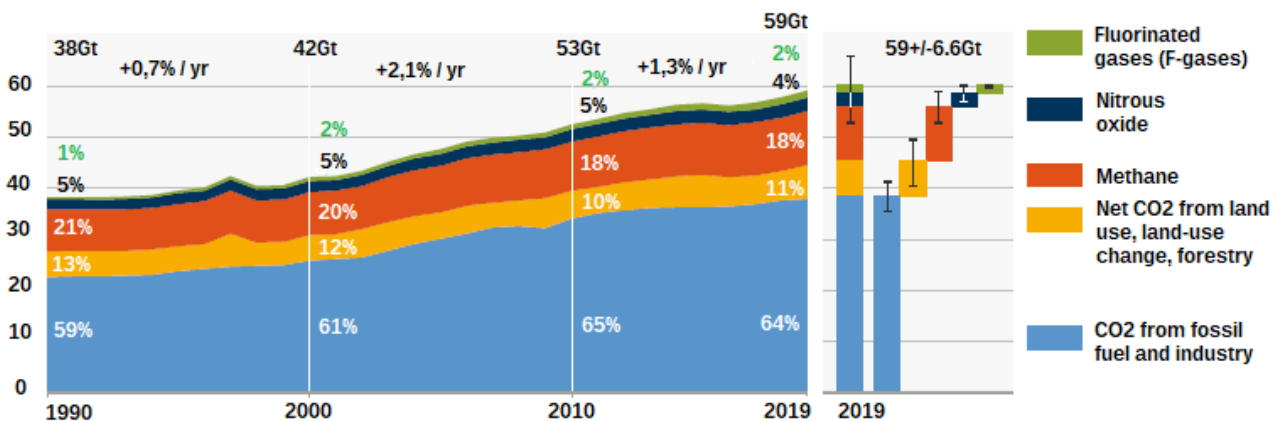
Moreover, it is acknowledged that abnormal levels of carbon dioxide in the atmosphere are partially being absorbed into the oceans, with potentially harmful effects on marine ecosystems [3]. This problem is usually referred to as *ocean acidification*.

Figure 1: CO₂ concentrations over the last 800,000 years as measured from the samples from the Antarctic ice sheets (image from [127], data from [2])



In 2019 global net anthropogenic emissions have been estimated[4] at 59 GtonCO₂eq, about two thirds (64%) of which are fossil fuels driven. Fossil fuel exploitation is also partially responsible for some of the methane emissions, which constitute the second most significant category of emissions in terms of global warming potential.

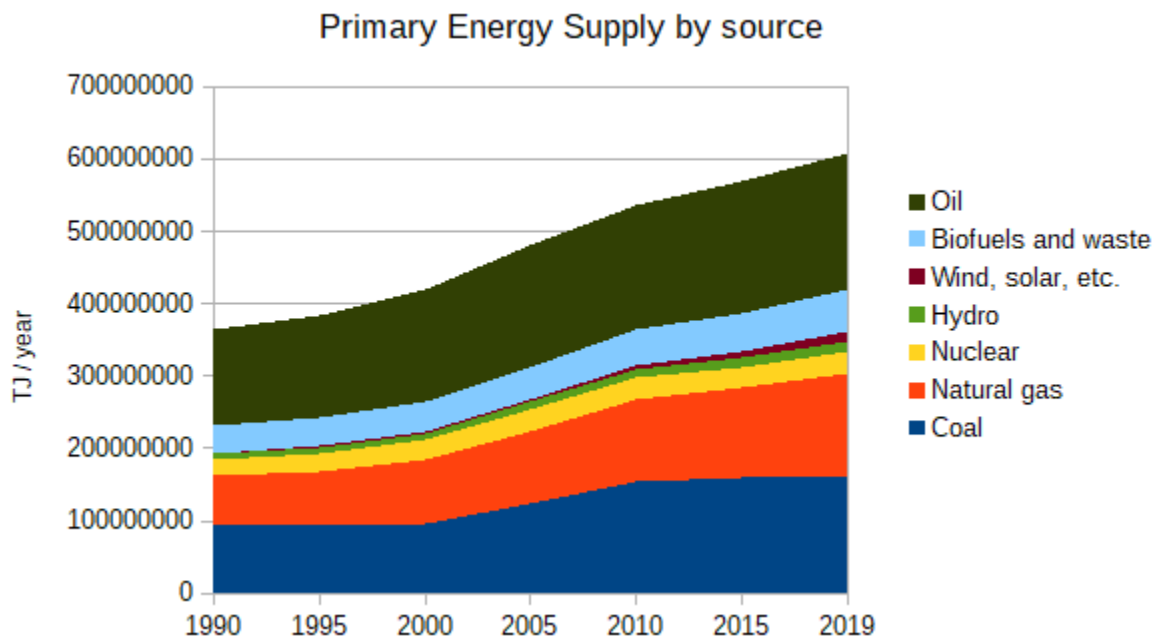
Figure 2: Global net anthropogenic GHG emissions 1990-2019 expressed in GtCO₂equivalent (from the IPCC sixth report)



Fossil fuels are currently employed primarily as a source of energy. In particular in 2019 the primary energy directly supplied by hydrocarbons was 80% of the total supply, which has been estimated at 600 millions of TJ / year worldwide [5].

The share of energy provided by Oil, Natural gas and Coal was roughly equal in terms of overall significance, although there are strong differences in the way the various fossil fuels are used. Oil in particular is employed mainly in the transportation sector, while coal and natural gas are mostly used for power generation or heating purposes.

Figure 3: World Primary Energy Supply by source 1990-2019 (IEA data[5])



In response to growing concern from the public, the governments of many countries have committed to reduce, in the medium-long run, the quantities of greenhouse gas emissions generated each year from the human activities.

The public commitments have resulted in the signature of the 2015 Paris Climate Agreement where 194 nations officially indicated decarbonization as a broad direction of public policy in the decades ahead (the official goal set in the agreement is to hold “the increase in the global average temperature to well below 2°C above pre-industrial levels” and pursue efforts “to limit the temperature increase to 1.5°C above pre-industrial levels”)[6].

If the attention is focused in particular on the electricity generation sector, which was responsible in 2021 of an estimated 14.6 Gigatonnes of CO₂eq [5], a solution that has been proposed and financially incentivized over the years to curb the reliance from fossil fuels is the deployment of technologies aimed at generating electricity from sources of renewable primary energy such as the wind or the solar radiation.

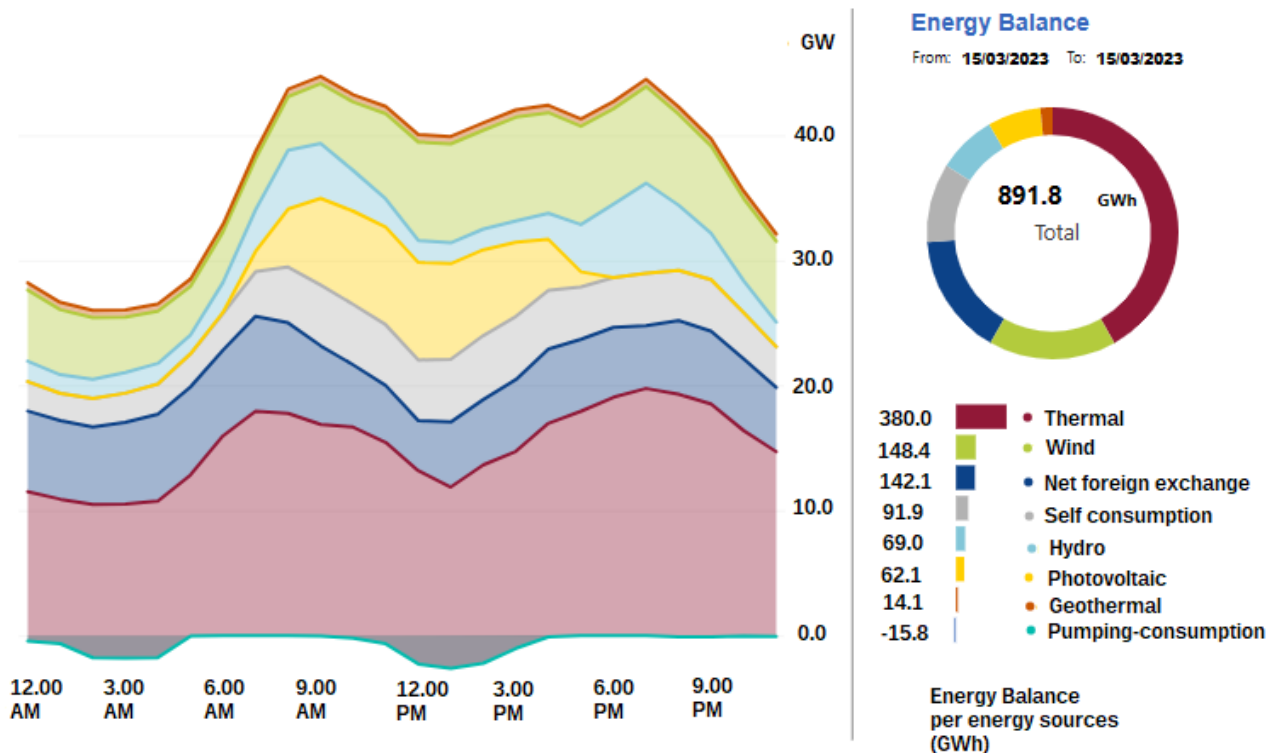
These new types of power plants have attracted increasing investments in recent years, to the point where in many countries they contribute in a percentage between 10% and 30% to the total annual electricity supply. At the same time the costs associated to the manufacturing of the renewable technologies have decreased dramatically with respect to a couple of decades ago[7].

While this strategy has proven successful in reducing the amount of fuel consumed for power generation in many regions of the world, the intrinsically undispachable nature of the renewable

energy sources poses a major roadblock in view of the creation of a fully decarbonized electricity grid.

To better understand the issues related to the integration of the renewables in the power grid it is useful to examine the profile (represented in Figure 4) of the electricity demand and the power generation from the various energy sources on a working Wednesday in a country like Italy, which has a total installed capacity of respectively 24 GW solar PV and 11 GW eolic.

Figure 4: Load curve, Italy, 15/3/2023 (data from Terna[8])



The electricity demand has a very characteristic shape, with a first peak at 9-10AM, a stable phase of operation for several hours, a second peak at 7PM, and finally a minimum during the night. It can be seen that a large fraction of the load is always covered from the thermal plants. In particular it is possible to distinguish between the base load plants and the peak load plants.

The first category includes systems, such as nuclear and coal plants, involving important phenomena of thermal inertia and long transitories of start-up. As a consequence, base load plants are operated always at a nearly constant load, made exception for periodical shut downs for planned maintenance.

Base load plants, which are also called *low-merit* plants, tends to have high investment costs and low marginal costs, largely dominated by the cost of the fuel.

Peak load plants, or *high merit*, on the contrary have low investment and high marginal costs. They are capable of performing a rapid start-up, and they are operated on a variable basis chasing the demand profile during the working week. According to the contracts for power delivery traded on the Forward Electricity Market, peak load plants are not supposed to operate during weekends, where the electricity demand tends to be significantly lower than the rest of the week.

To this category typically belong gas turbines and hydroelectric plants.

Natural gas combined cycles are in a somewhat intermediate position (*mid merit*) since they are able to fluctuate their output of a considerable amount if compared with base load plants; at the

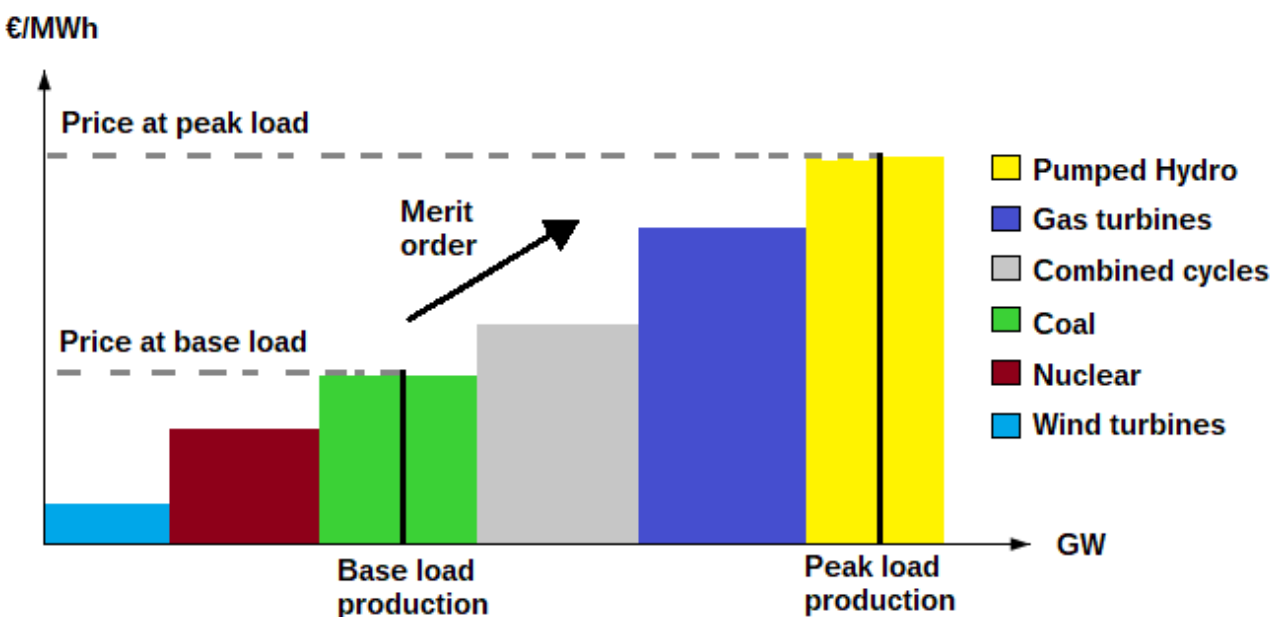
same time they are supposed to run over relatively stable periods of time without being frequently shut down. Moreover, they tend to suffer on average of lower levels of efficiency and greater deterioration, particularly due to fatigue, if they are operated on a daily basis in a cyclical trajectory at partial load[9].

In Figure 5 it is represented a possible distribution of the merit order for a fleet of generators. It can be noticed how the renewables have the lowest marginal cost, virtually close to zero. In other words, once the renewable plants are built and the fixed costs are paid, the renewable sources are able to outbid all the other producers.

This is generally not a problem during the hours of peak production, where the high merit plants are able to control their production in order to account for the presence of the renewables. On the other hand, problems of excess of generation may arise at the times of minimum production, especially during the night or during the weekend, because all the variable generators are already turned off and the base load plants are not able to adjust. When this occurs, the price of electricity tends to fall, and the hydroelectric plants are incentivized to buy cheap electricity in order to pump back the water and refill their uphill basin, as it can be seen from Figure 4. Eventually the price may even go to zero, meaning that some energy is being wasted.

The base-load power plants, in particular if they are numerous and not very efficient, are the ones most suffering these kinds of events, which may eventually lead to the decision of closing some of the oldest base-load units.

Figure 5: Example of a merit order distribution for a fleet of generators



The production from the photovoltaic panels, in Figure 4, is concentrated in the late morning and early afternoon, in coincidence with the period of maximum insolation.

The production from the wind energy is instead relatively constant during the day, however it features a moment of low production during the morning from 6AM to 9AM. This phenomenon is due to a deliberate curtailment performed by the grid operators. The early hours of the morning are indeed a phase of operation where the power suppliers have to steadily ramp up their production. Given that wind gusts tend to have an erratic and unpredictable behavior, and given that the other power suppliers would be called to suddenly increase their production in the event of a wind reduction, wind generators are curtailed in the morning to ensure that the load gradient remains always within the dynamic capabilities of the other plants.

Both the load and the renewable sources are subject to weekly or seasonal patterns. For example the peak load tends to be significantly (about 20%) lower during the weekends. On the contrary during the summer the load tends to be higher due to the electricity demand for air conditioning, reaching a peak load around 15-16PM.

At the same time the generation from wind turbines, at least in Italy during the last few years (2018-2022), tends to be lower during the summer due to the stable atmospheric conditions.

If it is taken into account that the production must always match the demand in order for the grid to be stable, it can be concluded from the abovementioned considerations that the combination of solar and wind alone cannot be safely relied upon to satisfy the entire demand over all the possible circumstances, which creates a limit of the maximum amount of renewables that can be integrated in a certain area of the grid network.

In particular it is possible to identify a set of distinct problems related to the management of a power system with an high percentage of renewables. Some of the most notable ones are listed below:

1. The constantly fluctuating nature of the wind may cause power quality issues such as short voltage dips or frequency variations.
2. It is necessary for the system operators to always have a certain amount of spare generation capacity that can be rapidly activated, if needed, in order to follow random variations in the load. This capacity is known as *primary reserve*, and it is necessary for all power systems. However the required reserve needs to be greater when the system includes a relevant share of renewables since the load variability is compounded with the variability of the renewables.
3. The production from the renewable sources (solar PV in particular) tends to be mismatched with respect to the moments of high consumption. As a consequence there is likely to be an excess of production in certain periods of the day or the year, while in others there is likely to be a shortfall.
4. Wind farms are often situated in remote locations with favourable wind patterns, therefore it is not always guaranteed the availability of an adequate transmission line capable of efficiently distributing the energy in all conditions. It may instead happen that there are bottlenecks to the maximum transmission capacity during some parts of day.

A strategy that has been adopted in the current energy systems is to minimize the variance of the renewables by spreading them over a wide geographical area. Power quality problems have been instead solved in many instances by the use of power electronics devices, capable for example of decoupling the rotating speed of the rotor from the grid actual frequency.

Despite the power of these relatively simple solutions, it is widely recognized that, in order to create a power grid with an high share of renewables, it is fundamental to consider also some other technologies, often called *dispatchable* generation, capable of adjusting their electricity production according to a planned schedule.

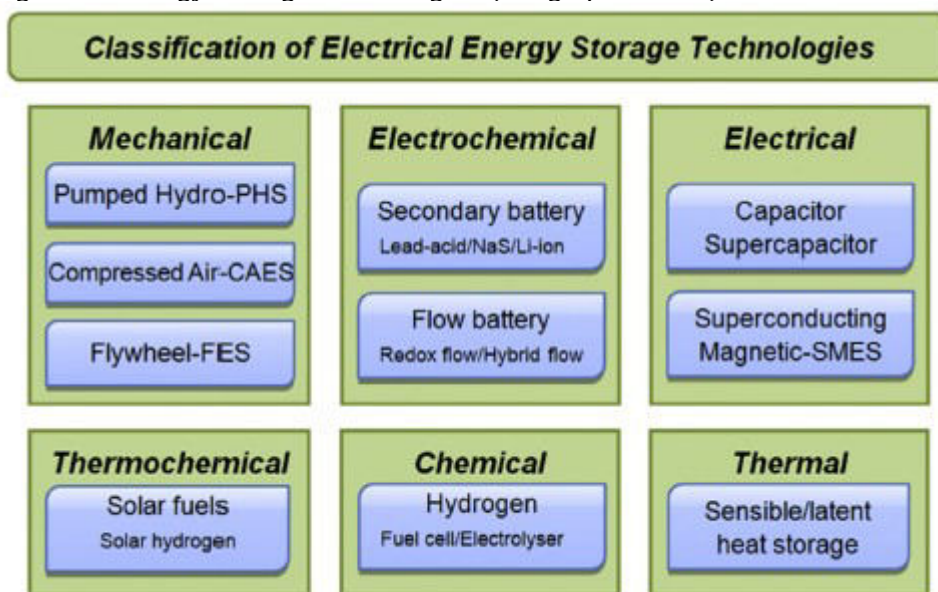
In particular a major field of R&D in the last two decades has revolved around the introduction of new systems capable of storing over time the energy produced from the renewable sources. In Figure 6 it is provided a classification of the main energy storage systems under investigation, according to a 2015 review from Luo et al [10]. It is possible to distinguish between:

- Mechanical energy storage. It involves the application of a force on a proper medium to deliver acceleration, compression or displacement (especially against gravity). The process can be reversed to recover the potential or kinetic energy. Pumped hydroelectricity storage

(PHS), compressed air energy storage (CAES) and flywheel energy storage (FES) belong to this category.

- **Electrochemical energy storage.** It involves the conversion of electrical energy into chemical energy and vice-versa through a series of electrochemical cells. Flow batteries and Lithium-ion batteries belong to this category.
- **Chemical energy storage.** It involves the dissociation of water into oxygen and hydrogen in an electrochemical cell called electrolyzer, which requires an electrical input. Hydrogen is then stored and converted back to electricity when needed. The inverse process may involve either an electrochemical transformation realized in a fuel cell, or a thermo-mechanical transformation realized typically through a gas turbine combined cycle.
- **Thermal energy storage (TES).** It involves the temporary storage, through a proper mechanism of thermal insulation, of high or low temperature thermal energy for later use. The stored thermal energy may be converted into electric energy when needed. An example of application of thermal energy storage can be encountered in some solar thermodynamic plants, where a fraction of the thermal energy produced during the daylight hours is stored into a tank filled with molten salts.
- **Electrical energy storage.** It involves the storage of energy in an electro-magnetic field generated through a proper device. Common examples of this concept are capacitors and inductors, as well as more advanced technologies such as supercapacitors and superconducting magnetic energy storage systems (SMES).
- **Thermochemical energy storage.** It involves the production of synthetic fuels from the energy obtained from a renewable source. It is a relatively new concept and there are currently several different approaches discussed in the literature. A possible approach consists in capturing the sunlight with a process of natural or artificially catalyzed photosynthesis in order to produce respectively carbohydrates or hydrogen. An alternative possibility is to concentrate the sunlight with a heliostat in order to generate a high temperature in a close environment, which may be used to drive an endothermic reactor like a steam reformer.

Figure 6: Energy storage technologies (image from [10])



In general, the various energy storage technologies feature different properties which make them more or less suitable for a specific type of application. The most important properties are:

- Power density, energy density, specific power and specific energy. These properties refer to the amount of space and weight that is needed to create a system with given parameters of discharge power and energy storage capacity.
- Cycle efficiency
- Self discharge. This property refers to the velocity in which the energy stored in a certain system tends to be dissipated
- Response time and discharge time at nominal conditions
- Lifetime expressed in terms of number of years and/or cycles of charge and discharge
- Technological maturity and economic costs

At present, according to the review paper from Luo et al[10], there is sufficient experience in the use of batteries, flywheels and electrical energy storage systems for solving the power quality issues often associated with the renewable energies.

Moreover, batteries and flow batteries are considered a proven technology to smoothen the transitories and fluctuations associated to the renewables in all the applications requiring a response time measured in seconds and a discharge duration measured in minutes. Flywheels and supercapacitors are also considered a promising alternative for similar applications.

The investment costs associated to the electrochemical systems, which are formed by the sum of a great number of elementary modules, tend however to become prohibitive for large scale energy management applications, like for example the distribution over the entire day of the solar energy produced during the daylight hours in a solar farm with a nominal capacity measured in the hundreds of MW.

Pumped hydroelectric storage, compressed air storage and thermal energy storage are therefore the only technologies considered mature for similar applications, although in practical terms there are only a few large scale TES and CAES facilities around the world.

In particular it is worth to mention the 290 MW CAES plant built in 1978 in Hurtorf (Germany) and the 110 MW CAES plant built in 1991 in McIntosh (Alabama, USA), both of which had the objective of selling during the peak hours of the day the energy produced during the night from the base-load plants [11].

The most notable projects involving a thermal energy storage facility are the Ouarzazate Solar Power Station (Morocco), the Cerro Dominador Solar Thermal Plant (Chile) and the Solana power plant (Arizona, USA), which consist in large (more than 100 MW) CSP plants of recent construction with thermal storage tanks for several hours (6-17.5 hours) of production after the sunset [12][13][14].

The situation is instead different for what concerns pumped hydroelectricity energy storage, which is a very well known method with high capacities already available. For example the hydroelectric capacity in Italy is 18.94 GW, and 1.5 TWh of energy were stored in 2022 [8].

The main issue associated with a future development of pumped hydroelectric storage is the fact that it is a space-intensive technology that relies heavily on the morphology of the territory, meaning that it cannot be always implemented or expanded, especially in urbanized areas.

In a scenario with an high share of renewables, an additional large-scale energy management application that the energy storage systems would have to handle is the seasonal energy storage, which consists for example in storing large quantities of solar energy during the summer in order to utilize it during the winter.

According to Lou et al[10] at the present moment there are no commercialized systems for this application, and storing fossil fuels is still a practical solution. On the other hand it is recognized that PHS, TES, CAES, electrolyzers and solar fuels have potential to serve this application.

In conclusion, it is possible to draw a picture of the electricity market where solar energy and wind turbines have become a relevant source of generation in the last two decades. However the cost of integration of increasingly large shares of renewable generators in the network is going to become more and more relevant due to the need of introducing new storage systems. Moreover, the problem of seasonal energy storage remains unresolved, which casts some doubts on the effective stability of the network in the event of a prolonged period of simultaneously low generation from the solar and wind generators [15].

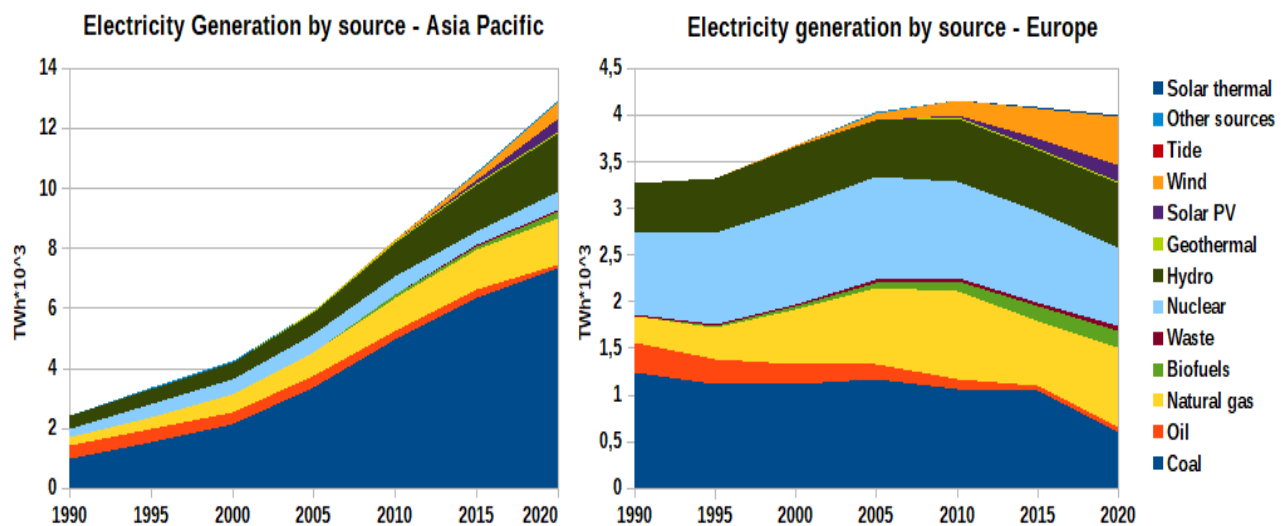
Therefore it can be stated that conventional power plants based on nuclear energy, hydroelectricity and fossil fuels remain still an indispensable element of the power supply system due to their properties of dispatchability, with no fundamental changes in sight.

1.2 Recent trends in the thermal coal sector: growing demand amidst environmental concerns and continuous pressure on the reservoirs

In Figure 7 on the right side it is represented the evolution of the power generation mix of the countries in Europe during the last three decades [5]. It can be noticed a recent tendency to close some of the old base-load coal plants and shift towards natural gas, which produces less greenhouse gas emissions and is more suited to work in a market with an high share of renewables, as previously discussed. The same trend can be encountered also in the United States, whose power generation mix is similar to that of Europe (the graph is not reported here).

On the other hand the transition away from coal has apparently come to an halt in 2022 due to the tensions between the European Union and the Russian Federation. In particular Germany has planned a “gas replacement reserve” of 11.6 GW of new capacity to address the interruption of the gas supplies from Russia. The new capacity includes 6.2 GW of coal plants which are allowed to return on the market by 2024. Moreover, the decommissioning of other 3.8 GW of coal capacity has been postponed[16]. A similar series of actions is expected also in a number of other EU countries.

Figure 7: Electricity generation by source 1990-2019 expressed in Thousands of TWh/year (IEA data)



As shown in Figure 7 on the left side, the situation is different in many of the so-called emerging markets, mostly in the Asia Pacific region, where coal is the predominant source of power supply, and it is projected[16] to further increase by 2025 driven by the growth in the electricity demand. India and China are the most important countries belonging to this group, although Indonesia and South Africa have also a power sector with similar characteristics.

There are several reasons behind the undying relevance of coal, the most important one being arguably the fact that coal is the fossil resource with the largest amount of proven reserves. In particular it is estimated[17] that the current reserves amount to 139 years at the current rate of exploitation. By comparison the ratio is equal to 48,8 years for natural gas and 53,5 years for oil. Similarly to the other fossil fuels, coal reserves are distributed unevenly around the world, with three-fourths of the known deposits being concentrated in the United States, Russia, Australia, China and India. Other countries with important coal reservoirs are also Germany, Ukraine, Indonesia, Poland, Kazakistan and Turkey.

Moreover, coal is easier to handle than natural gas. It can be transported by rail and ship without the need of special infrastructures. It is a key feedstock for the metallurgical industry and, to a lesser extent, other energy intensive application. The coal mining industry has been traditionally a major source of employment. Finally, it has to be remembered that, provided that the required infrastructure is made available, coal can be converted into other types of fuel if needed, as shall be discussed in the next pages.

In Figure 8 it is presented a detail of the main flows of thermal coal (to be distinguished from the coal used for metallurgical purposes) in terms of production, consumption and international trade. It can be immediately noticed how China is by far the largest consumer of coal in world, followed at great distance by India. China and India are also the two largest importing nations, although they are used to produce internally the largest part of their internal demand.

If one considers that India and China are countries not particularly rich in gas and oil, it appears evident how the choice of some nations of relying heavily on coal is driven also in part by the desire of not being dependent on imports for a strategic resource like energy.

It is worth on this regard to observe that, despite the large availability of coal within the chinese borders, the chinese reserves are expected to last only 37 years at the current pace of exploitation[17]. This particular topic is treated in an article from Zhang et al[18] in which the authors conclude that a key element for the success of the coal mining industry in China in the future will be the capacity of maximizing the recovery rate.

Figure 8: aggregate flows of coal in terms of production, consumption, import and export. All numbers are expressed in Megatons per year (IEA data)

	Thermal coal and Lignite Consumption		Thermal coal and Lignite Production		Imports of Thermal Coal and Lignite		Exports of Thermal Coal and Lignite
Asia Pacific	5351	Asia Pacific	5480	Europe	121	Australia	184
China	3542	China	3561	Japan	140	Colombia	52
India	1027	India	886	Korea	92	Indonesia	469
Japan	133	Australia	277	Chinese Taipei	56	Russia	157
Southeast Asia	350	Indonesia	615	China	240	South Africa	69
North America	479	North America	505	India	152	United States	35
United States	450	United States	480	Southeast Asia	124	Rest of the World	79
Central and South America	32	Central and South America	61	Rest of the World	111	World	1045
Europe	620	Europe	504	World	1035		
European Union	421	European Union	342				
Middle East	6	Middle East	0				
Eurasia	280	Eurasia	438				
Russia	173	Russia	308				
Africa	177	Africa	234				
World	6945	World	7221				

(all numbers are expressed in Megatons / year)

While the beneficial effects of coal on an industrial economy are clear, there is anecdotal evidence dating back to the english literature of the nineteenth century on the detrimental effects of the use of coal on the environmental and public health, with rigorous scientific investigations being performed in recent decades.

In detail, coal is known to be the fossil resource generating by far the greatest amount of greenhouse gas emissions per unit of thermal energy. In addition, the combustion products of coal, if not properly treated, typically include various pollutants such as sulfur dioxide and nitrogen oxides (Nox), which are linked to acid rains and respiratory problems [19]. Other pollutants emitted include small fly ash pieces called particulate, which is known to be a cause of cardiovascular and respiratory diseases[20], as well as carbon monoxide, mercury, lead, cadmium, arsenic and volatile organic compounds (VOCs).

The overall amount of air pollution generated by coal-fired plants is dependent on the specific characteristics of the coal employed as a feedstock. The properties of coal are indeed different from site to site. It is possible to distinguish between four broad categories of coals: Anthracite, Bituminous coal, Sub-Bituminous coal, and Lignite. Anthracite and Bituminous coal are the highest grades and the best ones for metallurgical applications. Sub-Bituminous coals and Lignite have instead a lower energy content and they are mostly used to generate electricity. Other important parameters to characterize a coal quality are its sulfur and ash content.

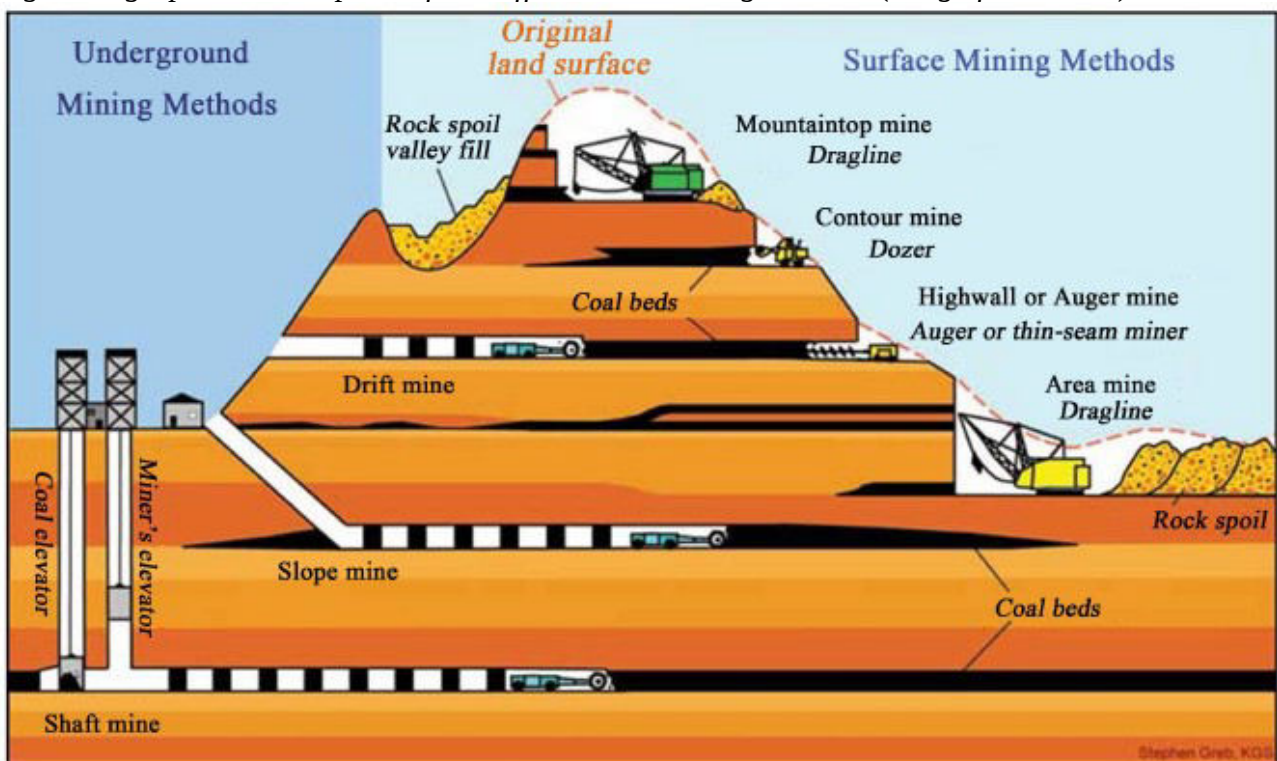
Besides the end-of-the-chain emissions, an additional series of concerns related to the use of coal is linked to the process of extraction. In Figure 9 it is graphically represented the set of practices usually adopted in the mining industry for the recovery of a coal bed. Two main categories can be singled out, namely *Underground Mining Methods* and *Surface Mining Methods*.

In surface mining, large machines remove the topsoil and layers of rock to expose a coal seam.

In underground mining, which is adopted when the coal bed is more than 60 metres below the surface, miners ride elevators and travel on small trains in long tunnels to reach the coal.

Both the approaches are equally employed in the contemporary mining industry, with a slight prevalence of surface mining[21]. The final choice is generally dependent on technical aspects and cost-benefit considerations for a specific site.

Figure 9: graphical description of the different coal mining methods (image from [128])



Immediately after being extracted, coal has to be separated from the other non valuable minerals, also called gangue, through a series of size reduction and screening steps, sometimes called *beneficiation*. In particular one of the most common separation techniques is froth flotation, which takes advantage of the natural (or artificially induced) hydrophobicity of the liberated minerals to separate them from one another. Some chemicals or reagents are often used to improve the efficiency and selectivity of the flotation process.

Due to the natural friability of coal, the mining and beneficiation processes have historically produced over time large quantities of fine and ultrafine coal particles that could not be effectively separated from the gangue with the technologies available, despite their potential economic value. These are known as coal tailings, and they can be either stocked into large piles near the mine, or dumped into pools of wet mud in order to prevent the erosion of the wind[22].

As documented in a review paper from Ramudzwagi et al[23], rising energy prices and environmental concerns have led to a renewed interest in recent years on the development of new separation processes capable of recovering these deposits.

On the whole, the fuel and electricity consumptions associated with surface and underground mining processes are a major source of greenhouse gas emission, although there may be variations in the magnitude of such impacts due to site-specific differences. In the mining stage, land clearing, drilling, blasting, crushing and hauling are the operations that require the greatest amount of energy. For underground mining, in comparison with surface mining, there are additional energy requirements due to ventilation and water pumping. Transportation of the coal to the processing and shipping facilities also contributes to the energy related emissions.

Besides land use and energy-related emissions, coal mines are responsible for the release of a substantial quantity of methane, which is normally adsorbed in the solid matrix of coal, and it slowly vented when the coal seam is exposed to the atmospheric pressure. Coalbed methane is a long known safety risk in the underground mining activities, and it is normally vented outside of the mine through proper ventilation. The emissions may continue well after the end of the commercial exploitation of a coal mine. An estimated 40.5 Mt of methane were emitted in 2022 due to coal mining activities, accounting for about 10% of total methane emissions[24].

However it is noteworthy to observe that, despite technical, economical and cultural challenges, there is tendency in the last decades, partially due to new regulatory standards, towards a stricter control of these emissions. The techniques adopted may involve recycle and utilization of the methane, or incineration by means of flare stacks.

Before concluding, it is important to add that, according to a 2015 review article from M.Hendryx[25], there is strong evidence associating some of the practices of surface mining to a poor health status in the communities and ecosystems surrounding the extraction sites, although the direct mechanistic links are still not well understood.

Underground coal mining has been known for a long time to increase the risk for workers of developing respiratory diseases due to the great presence of dust on the workplace.

A review of the environmental and epidemiological data, focused in particular on the effects of the techniques of mountaintop-removal (MTR) implemented in West Virginia (USA), suggest the communities surrounding the mines are also affect by poor air quality.

In addition, there is evidence associating the practices of mountain-top removal to water pollution in the surface waters downstream to the extraction sites.

The main source of this phenomenon is considered to be the contact of the rain and the atmospheric oxygen with the mine surface, which causes the soil and the rocks to leach long-sequestered minerals, metals and other chemicals. Improper waste water management from the coal treatment

facilities is also suspected to be a source of water pollution in the areas surrounding some coal mines.

The implementation of MTR activities is therefore strongly discouraged by the author on the basis of a public health risk assessment.

1.3 Introduction to the gasification and CCS technologies

In light of all the considerations described in the previous chapter, and given the fact that it is unlikely that many of the newly opened coal-fired plants will be phased out before many decades, it appears of fundamental importance the development of technologies aimed at valorizing the existing assets and coal reservoirs while at the same time reducing their environmental impacts within acceptable limits. Among these clean coal technologies, an important one, about which great technical know-how has been gained during the years, is the integrated gasification combined cycle (IGCC).

In order to introduce the IGCC power plants it is useful to first highlight the fact that great technical achievements have been reached in the field of gas turbines, and modern natural gas combined cycle (NGCC) reach efficiencies close to 60%, with a temperature of the gas entering the turbine as high as 1400°C [26].

However gas turbines are usually made of Nichel superalloys[27], which do not tolerate sulfur at high temperature. In addition, if a gas turbine was fired with a fuel rich in ashes there would be a serious risk of mechanical obstruction of the movement of the blades. For these reasons the usage of an heavy fuel like coal is problematic in similar systems.

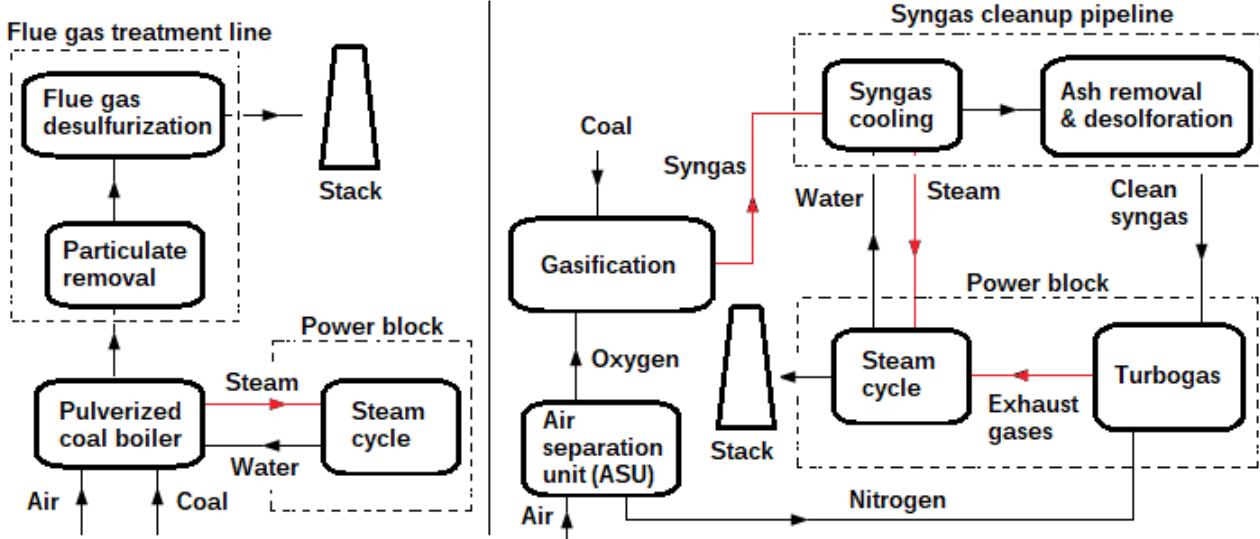
IGCC power plants bridge this gap by introducing a series of pre-treatments, the most important being the *gasification*, through which coal is converted in a clean synthetic gas (*syngas*) that can be safely fired into a gas turbine, allowing therefore to combine good thermodynamic efficiencies (46%- 48%)[26] with excellent performance of the plant in terms of air polluting emissions.

Gasification essentially consists in a process of incomplete oxidation of the feedstock, and it has been used at industrial scale for at least a century, with the first systems dating back even to the first half of the nineteenth century. It has been historically employed for the production of the so-called manufactured gas used for heating and illumination. Moreover, it has been employed for the production of synthetic fuels for transportation and other important chemicals like ammonia. In recent decades gasification has been re-discovered and it is nowadays adopted not only for electricity production from coal in a combined cycle (IGCC), but also, and to a greater extent, in the oil refinery industry for the processing of the heaviest fraction of oil (vacuum bottoms). Another notable use of gasification is in the field of biomass and urban waste disposal.

Currently, there exist about ten IGCC power plants of great size (several hundreds of MWs) operating around the world[26], some of which fuelled with coal, others with refinery heavy waste. In Italy in particular there are three large IGCC plants operating with refinery heavy materials. In general the gasification technologies offered by the various producers are relatively different from each other in many details, however a general feature shared by the majority of the reactors present on the market is to work with pressurized oxygen instead of air as oxidant.

A layout commonly encountered for an IGCC plant may be similar to the one sketched in a simplified manner in Figure 10 on the right, as compared to the layout of a coal fired power plant based on a more conventional Rankine steam cycle (on the left).

Figure 10: block diagram of a typical USC (left) and IGCC plant (right)



The following points should be noted in particular regarding the layout of a typical IGCC plant:

- In order to produce pressurized oxygen it is necessary to split air into oxygen and nitrogen, which is generally done in large facilities through a distillation process at cryogenic temperatures. This technique has been successfully employed for decades to supply oxygen for the gasification of a wide range of hydrocarbon feedstocks to generate synthesis gas for the production of fuels, chemicals and other value added products [28], therefore it can be considered mature and reliable.
The main drawback is that it requires a relatively important quantity of electricity (about 10% of the total gross power[26]), affecting negatively the efficiency of the system.
- Given the characteristics of the syngas with respect to natural gas, in order to obtain an acceptable temperature in the combustor of the gas turbine it is necessary to dilute the syngas with an inert, which is typically nitrogen from the air separation unit, possibly with the addition of steam.
- The reason for the superior environmental performance of the IGCC configuration with respect to the coal-fired Rankine cycle plants lies in the fact that all clean-up treatments (ash removal, desulfuration, mercury removal) are performed on a small flow of pressurized syngas before this is mixed with air and it is brought to atmospheric pressure. As a consequence, there is much greater thermodynamic driving force for the removal of the pollutants, and the equipment doesn't need to be of huge size.
- The hot syngas exiting the gasifier requires a clean up process. In the existing plants this operation is performed at near-ambient temperature after the syngas is cooled down in a series of recuperative exchangers. Desulfuration in particular is done in a wet scrubber with chemical solvents (usually amines) or with physical solvents like Rectisol or Selexol. Desulfuration techniques are in general very well known since they have been employed at massive scale for decades in the oil industry for the desulfuration (*sweetening*) of natural gas[29].

From a thermodynamic point of view it would be however preferable not to cool down the syngas and perform the clean up operation at high temperature with other methods (like Zinc oxide sorbents [30] and hot gas filtration[31]). While this route is interesting [32] and large demonstration trials[33][34] have been conducted for many years, it cannot still be considered a technology ready for commercial use. In particular two aspects are considered as critical. The first one relates to the difficulties of removing simultaneously multiple contaminants, some of which are present in the syngas only as trace elements, given the

intrinsically heterogeneous nature of coal. The second aspect relates to the durability of the equipment (particularly the hot filters [26]) and the number of regeneration cycles that the sorbent is able to accomplish before deactivation[32].

Notwithstanding that most of the research and experimentation activity on IGCC power generation revolves around oxygen-blown gasifiers, air-blown gasification is also an alternative studied and experimented. In particular this option allows to drastically reduce the costs associated to the ASU, both in terms of capital and auxiliary electric consumption, potentially reaching a greater plant efficiency.

The most significant activity concerning air blown gasification has been conducted in recent years by Mitsubishi Heavy Industries (MHI) in Japan, which launched in 2007 a 250 MW pilot plant near Fukushima[36]. Following many thousands of hours of successful demonstration (42-43% net LHV efficiency and extremely high reliability[35]) the plant has been run from 2013 to 2020 as a commercial utility. In 2020 it has been retired and substituted with a larger unit (525 MW)[38]. The producer claims in particular a very interesting performance of 53% gross power plant efficiency when the system is operated with a M501G gas turbine (1995 design)[37].

Besides gasification, a second group of technologies intended to limit the environmental impacts of coal, and specifically the GHG emissions associated to it, is carbon capture and storage (CCS). Carbon Capture and Storage is an expression that is used, in the broader sense, to refer to the possibility of *capturing* the carbon dioxide from a given source through a proper medium, and then *storing* it for a long time (most likely after some kind of *transportation*) in order to remove that amount of matter from the carbon cycle.

The option that is most often cited as suitable for long term storage is geological sequestration[39], which involves the injection of high pressure carbon dioxide into depleted oil/gas reservoirs, as well as other geological formations such as salt mines or deep aquifers.

This solution offers the advantage of exploiting technologies and methods well known in the oil&gas industry, and some large demonstration plants have been successfully tried [40][41][42], even though its practical viability remains undemonstrated on the scales and the timeframes theoretically required.

For what concerns the possible strategies of carbon capture several alternatives have been proposed. In the following pages are presented the principal methodologies that have been tested at large scale and are considered more or less ready for commercial deployment, provided that proper incentives are supplied. The more advanced technologies instead are not discussed since, although they are in some cases potentially very innovative (this is true in particular for Chemical Looping Combustion[43]), they nonetheless feature unresolved technical issues that make their deployment unlikely in the short-medium term.

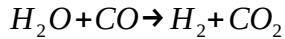
Pre combustion capture. It is a form of carbon removal that, as the name suggests, is performed before the combustion. The aim of this type of process is to produce a partially or completely decarbonized fuel, which may be then used as a feedstock for power production or other applications.

The most glaring example of pre-combustion separation of carbon dioxide can be encountered in the production of hydrogen from coal or natural gas. It is a process well known for decades, since hydrogen is commonly used in large volumes for oil refinement or in the production of ammonia for fertilizers. However it has the disadvantage of requiring new infrastructure and not being suitable to retrofit the existing coal plants.

The process, which is slightly different depending on whether the feedstock is natural gas or coal, can be summarized as follows:

1) A flow of hot and pressurized syngas is generated in a gasifier or a steam reforming reactor. From a chemical point of view the syngas is a mixture of hydrogen and carbon monoxide with also variable amounts of other gases.

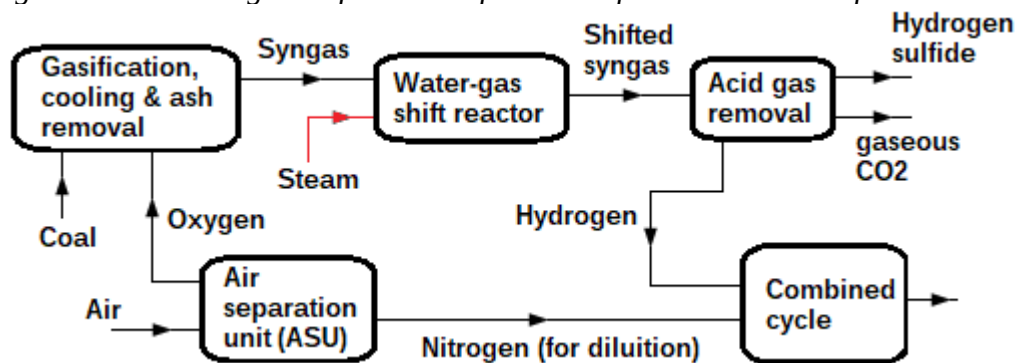
2) The syngas, after being saturated with steam, is fed into a series of water-gas-shift reactors where carbon monoxide is almost completely converted into carbon dioxide according to the reaction:



3) The flow of shifted syngas is finally cooled down at near ambient temperature and passed into a unit for the removal of the acid gases (AGR), meaning hydrogen sulfide and carbon dioxide. Given the fact that the partial pressure of carbon dioxide is elevated, this operation can be performed with a physical solvent (Dual-stage Selexol is mentioned by many authors), requiring only a pressure swing to be regenerated.

4) The flow exiting the AGR is constituted mostly by hydrogen, with the degree of purity being dependent on the specifics of the process employed. Some of the possible applications of hydrogen, one of them being for example the use as a fuel for transportation, may require a further purification, for example through pressure-swing-absorption beds[44]. However, focusing the attention on the process starting from coal, the scheme that is usually proposed (illustrated in Figure-11) is one in which a flow of impure hydrogen is directly fed into a gas turbine in order to fuel a combined cycle.

Figure 11: block diagram of an IGCC plant with pre-combustion capture



In this way it is possible to realize an IGCC plant with pre-combustion CO₂ capture. It is estimated [45] that such a solution might achieve a HHV efficiency of about 31%, with a loss of 8-9 percentage points if compared to a similar plant without capture.

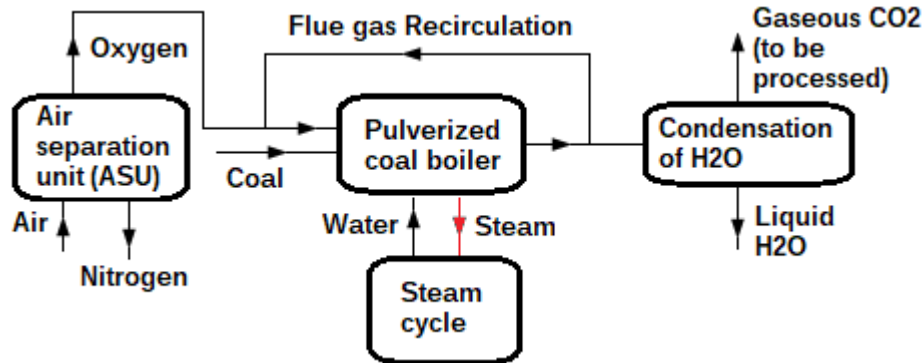
The maximum temperature allowable at the inlet of the gas turbine is in particular a limiting factor and a field of active study[126]. The gas turbines currently available, and more specifically the premixed combustors, are indeed less performing when they are operated with hydrogen, with the consequence that it is necessary to lower the flame temperature in order to limit the production of NO_x.

Oxycombustion. It is a strategy of carbon capture in which a certain reaction of combustion is realized in an atmosphere of pure oxygen not diluted with nitrogen[46]. In order to obtain a stream of concentrated carbon dioxide it is then sufficient to cool the exhaust gases and eliminate the water condensate. The carbon capture ratio that is achieved in this way approaches 100%.

This method has the advantage of being relatively simple if compared with the other strategies proposed, and existing USC coal plants could be retrofitted according to the oxycombustion scheme without substantial modifications of the components, made exception for the boiler, which would need instead to be completely redesigned. Oxycombustion is indeed very demanding for the boiler

since a combustion in oxygen features much higher flame temperatures. It is therefore necessary to dilute the oxygen by recirculating a fraction of the exhaust gases in the boiler feed (like represented in Figure 12).

Figure 12: block diagram of the oxycombustion process applied to a USC coal plant

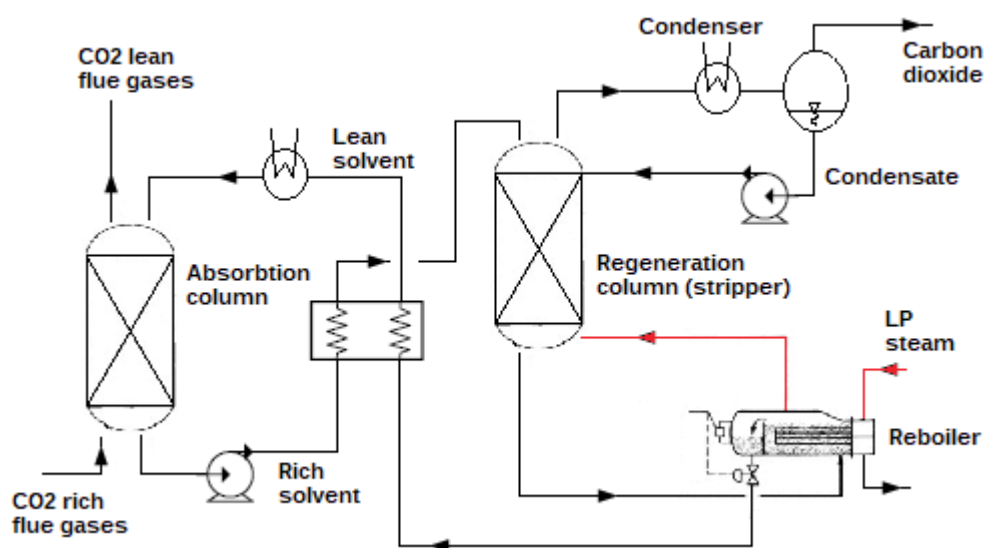


Another disadvantage of oxycombustion is that large volumes of oxygen have to be produced in a dedicated air separation unit with considerable expense of electricity. The expected HHV efficiency for a new coal fired USC plant implementing an oxycombustion scheme is estimated around 32% [45]. To date a couple of medium size (<100 MW) oxy-combustion demonstration projects have been successfully tested for thousands of hours of operations [47][48].

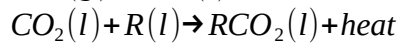
Post combustion capture: it is a process in which the separation of carbon dioxide is realized at atmospheric pressure as an after-treatment to a flow of exhaust gases that is subsequently released. It is the most flexible carbon capture option and, in principle, could be applied to retrofit not only existing power plants, but also other industrial sources of carbon emissions.

Although there have been theoretical studies and lab scale experiments on the use of selective membranes, the most viable methodology for post combustion capture is by far the absorption of carbon dioxide with a liquid chemical solvent[49], usually belonging to the family of amines. In Figure 13 it is schematically represented the typical process and equipment employed.

Figure 13: typical process used for CO2 chemical absorption



In the absorption column the molecules of carbon dioxide diffuse into the liquid phase, then they react homogeneously with the solvent creating a new species with a low vapour pressure. The absorption process tends to generate heat, with the consequence that in some cases it might be necessary to split the column into multiple intercooled beds in order to properly control the temperature.



The *rich* liquid solvent is extracted from the bottom of the column, containing a certain fraction of the carbon dioxide molecules “trapped” in the reacted form.

The rich solvent is regenerated endothermically through vapour stripping in a second column, where the inverse reactions take place. The *lean* solvent is finally recirculated back to the top of the absorber.

Carbon dioxide mixed with steam is obtained instead at the top of the regenerator, and can be easily separated by condensing the steam.

A major drawback of chemical absorption is represented by the fact that the regeneration of the solvent requires the introduction of an important quantity of heat in the reboiler at the bottom of the stripping column. The heat is obtained by extracting a certain amount of steam flowing in the LP turbine. As a consequence, the power output and the efficiency of the plant are reduced.

For example it is estimated that, if the solvent employed is 30% aqueous MEA (monoethanolamine), the reboiler duty is in the range 3-3.5 MJ/kgCO₂, with the temperature of evaporation of the liquid in the reboiler usually in the range 120-140°C. Given the limits of thermal stability of the solvent, the regeneration column usually works at a pressure around 2 bar, meaning that significant additional work needs to be spent to compress the carbon dioxide output stream. A realistic figure for the HHV efficiency of a new USC coal-fired power plant implementing a similar system is around 31-32%, while the efficiency of the same plant without capture is about 10 percentage points higher[45].

Historically, 30 wt% MEA has served as the standard for the evaluation of processes for post combustion capture. While there haven't been new breakthroughs or discoveries changing in a fundamental way the paradigm of post combustion capture, which continues to be regarded as essentially an energy intensive process, new solvents have been proposed and tested over the years improving the performance of MEA in terms of solvent stability, corrosivity and regeneration energy needed.

The composition of some of these second generation solvents, which may be constituted by a blend of various chemicals, is often a proprietary information not disclosed to the public. This is for example the case of the solvent KS-1 used in the Petra Nova project(2016-2020)[50], which has been so far the largest project (240MWel) with post combustion capture operating at commercial scale.

Among the solvents that are most frequently investigated in the open literature there are 2-amino-2methyl-1propane (AMP), Diethanolamine (DEA), Methyl-diethanolamine (MDEA), Morpholine (Mor) and Piperidine(PD). A solvent that seems particularly promising due to its stability and low regeneration energy is Piperazine (PZ)[51], although it has the disadvantage that it can be used only in blended form together with other amines due to its low solubility. Apart from amines, ammonia is also a solvent that has raised significant interest in the literature.

The process of chemical absorption used for post combustion capture may apparently resemble the operations of sweetening that are employed in the chemical industry (or an IGCC plant with pre

combustion capture) to separate the acid gases from methane or hydrogen. However the conditions are such that in reality that process is more difficult.

First, the flue gases entering the absorber are typically at near atmospheric pressure and have a concentration of carbon dioxide around 10-15%. Therefore the solvent has to be regenerated with a purely thermal process, instead of a simple depressurization like in the case of pre-combustion capture.

In addition, solvents for post combustion capture have to resist to a degradation process caused by the oxygen present in the flue gases (typically up to 5% vol).

As a matter of fact, solvent degradation over time, both of thermal and oxydative nature, is an important factor to consider for many of the solvents that have been proposed. It is estimated that the costs associated to solvent degradation account for about 10% of the total cost of a post combustion capture system based on 30% MEA[52].

The processes of degradation of the solvent consist in the occurrence of a series of undesired chemical reactions that are not reversed in the absorption/regeneration cycle and tend to diminish over time the CO₂ loading capacity of the solvent. The products of degradation belong to a wide category of species, and they are often referred to as “heat stable salts” HSS. In particular the chemistry of the degradation of amines is a wide topic of ongoing theoretical and experimental research.

The current understanding is that factors of influence for the solvent degradation (at least in the case of MEA) are not only the operating parameters such as CO₂ loading and the temperature in the stripper, but also the eventual presence of various dissolved contaminants (Fe³⁺, Cu⁺, other metals, SO₂⁻, NO₂⁻, NO₃⁻) coming from the flue gases or the corrosion of the equipment, although the precise causal relationships are quite complex.

For example it is reported that sulfite ions have a significant inhibition effect on oxydative degradation in a 30% MEA solution. At the same time they increase corrosion, which has a catalytic effect on oxydative degradation of the solvent due to the properties of Fe³⁺ ions[53]. Moreover, sulfite ions are believed to have a parallel influence on the mechanism of thermal degradation[54].

Generally amines are not intrinsically corrosive, since they associate high PH and low conductivity. However they can become corrosive when they absorb acid gases such as H₂S and CO₂, or when they become enriched in degradation products[55]. No consensus has yet been reached regarding the mechanisms of corrosion by amine solutions[55], however it is a well recognized fact in the oil and gas industry that corrosion is a major operational issue in sweetening plants, with a long history of documented failures[56].

MEA is reported to be the most corrosive amine, while secondary and tertiary amines are less aggressive. Other known factors of risk are 1) high temperature, 2) high CO₂ loading, 3) high solvent concentration, 4) turbulence of the flow, 5) presence of oxidative degradation products. The authors of a 2014 review[55] conclude that carbon steel cannot be used in safety in all operating conditions encountered in a MEA capture unit. Therefore they suggest the use of a more expensive austenitic steel grade 316L for new plants based on MEA, as well as for demonstration pilot plants operating with other amines.

The degradation of the solvent is problematic, above all, because of the costs of its replacement. However additional concerns are related to the eventual possibility of hazardous effects caused by volatile degradation products that might be released from the CO₂ capture unit into the atmosphere, especially when the solvent belongs to the family of amines.

It has been indeed demonstrated that, at least in theory, any solvent based on amines may degrade to produce stable nitrosamines[57], which are known to be carcinogenic. In particular secondary and tertiary amines can form stable nitrosamines, while primary amines (among which belongs

MEA) can form nitrosamines only through their degradation products[57]. While it is admitted that more study is necessary to fully understand all the pathways of nitrosamine formation, it is known that nitrite ions, which enter the absorber because of the Nox in the flue gases, are responsible of the degradation of various secondary amines to nitrosamines in a reaction catalyzed by formaldehyde[58], which is among the degradation products of the amines[59]. Other classes of chemicals that may result as degradation products are respectively the nitramines, which can be formed also from the primary amines, and the alkylamines.

The authors of a 2013 review [60] argue that, according to the limited data on emissions available from the existing plants, the emissions from a capture plant operating with MEA are expected to consist primarily of ammonia and MEA itself, with nitrosamines and nitramines (and other degradation products) present at ppb levels. For other solvents it is also expected a similar behavior, although the authors highlight a general lack of data on real plants.

The authors suggest particular attention for the plants operating with secondary and tertiary amines, considering that their potential for nitrosamines formation is higher. Moreover, they point out that the emissions of secondary and tertiary amines might be more problematic, since they have greater potential to form nitrosamines after being emitted.

According to experimental studies performed on the atmospheric degradation of amines, it has been indeed verified that, while nitrosamines are not formed through photo-oxidation from MEA, nitramines can be formed with a yield between 0,3% and 1,5% according to typical concentrations of Nox that can be encountered in the rural or urban environment[61].

Another study[62] investigating methylamine, dimethylamine and trimethylamine reports a yield in nitrosamine between 0,6% and 2,3% of the initial quantity of dimethylamine photo-oxidized. The same study reports a yield between 2,5% and 8% in nitramines for dimethylamine.

Given the limits of exposure proposed by the Norwegian Institute of Public Health[63], according to a modelization performed for the Technology Centre Mongstad [64] these levels are well below a reasonable threshold of safety.

Finally it needs to be noticed that according to the authors of the review another potentially serious issue might consist in the accumulation of the amines emitted from the CO₂ capture plant into the water reservoirs, with potentially detrimental effects on ecosystems and drinking water supplies. However the authors of a 2014 review on the present status of amine-related contamination into the surface waters [65] conclude that *“there are surprisingly few data on the concentrations of amines in surface waters and on the toxicity of these compounds. In particular, there are apparently only a few measurements of amines in unpolluted rivers and lakes. This is surprising, given that amines are widely used in various industrial, pharmaceutical and chemical applications and present in wastewaters”*.

This implies that in order to evaluate the potential impact of a CO₂ capture plant it should be first gathered more information on the natural levels of amines in surface waters, as well as the spatial and temporal patterns and the factors influencing the concentration. Moreover, it should be carefully investigated the toxicity of amine-related water pollution for the aquatic ecosystems. Finally, it should be better understood the degree in which carcinogenic compounds such as nitrosamines and nitramines are formed from amines in the natural environment. This in particular would help to clarify the risks for the health of a chronic exposure to low levels of amines in the water supply.

2 INTRODUCTION TO THE CCS COST ESTIMATES

2.1 Why a cost estimate?

Carbon dioxide capture and storage is acknowledged to be a technology of critical importance to reduce greenhouse gas emissions in various sectors, however its cost constitutes a major obstacle to a more widespread deployment in the industrial landscape.

Due to its potential significance, information on the costs of CCS is sought by a broad group of actors for purposes ranging from policy analysis, investment decisions, technology assessments, R&D activities, and policy-making.

From a practical point of view, cost estimates are usually developed to answer to a specific problem or question, which may be framed explicitly or implicitly. In particular, Rubin[66] identifies two families of problems/questions that are typically answered by cost estimates.

The first set is labelled “what if” problems, and, according to the author, is typical of the R&D community, which is interested in setting R&D targets and quantifying the expected cost reductions obtained by a certain system if those targets are met.

The question asked is an hypothetical construct: “what would be the cost of an advanced technology if it successfully achieves specified R&D goals?”.

The second family of problems are the so-called “what will” problems. The question that is generally asked here is: “what it is likely to be the cost of technology X in the moment of time Y?” Generally this family of problems is important to people interested in the cost and competitiveness of a given technology in the future, that includes final users, investors and policy analysts.

“What if” and “What will” problems are also called respectively *technological* and *policy* assessments, and for each category of cost estimate it is necessary to employ a different set of methodologies and/or assumptions in order to produce a result that addresses the posed problem in an effective way.

The methodology and the assumptions adopted in a given cost estimate must then be taken in consideration when the document is read, since it might have been structured in order to answer to a different question from that for which it is consulted.

In particular, it is noteworthy to report a passage in which the author highlights some of the typical assumptions and limitations of technological assessments:

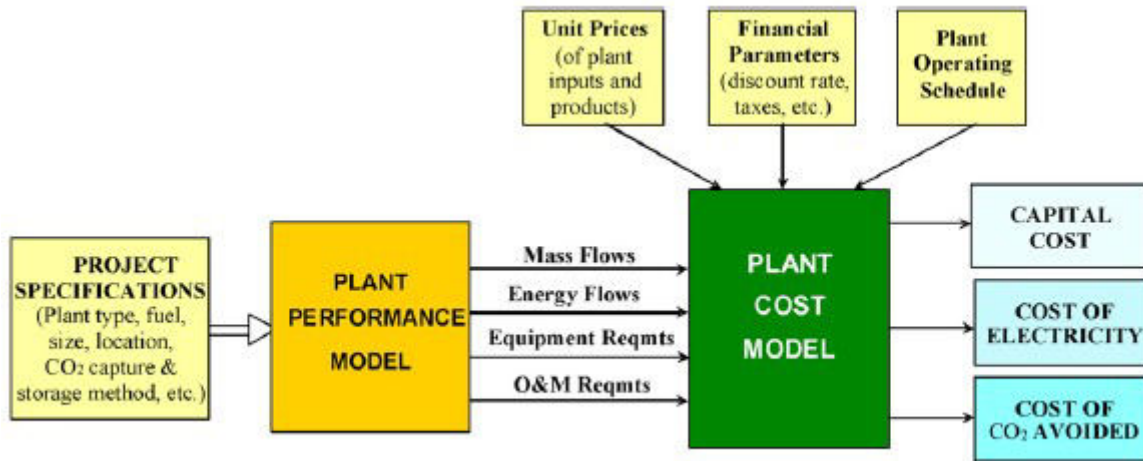
“technological assessments often seek to compare the expected cost of alternative CO₂ capture options for a particular application as a part of a feasibility or screening process... Technology-leveling assumptions are sometimes used to maintain uniformity of base plant assumptions that are not directly related to CCS subsystems (fuel type, operational availability, etc.), therefore these estimates are unlikely to be good predictors of the actual cost of projects because they don't account for the various in site and owner specifications that are included in real projects.

In contrast, cost estimates for specific projects aim to provide the owner with as accurate an estimate as possible of all the project costs that must be financed. In this case the technology already has been selected, and the focus is on the many site-specific elements that affect a project's cost.”[66].

2.2 Structure of a cost estimate

A possible general framework to understand the constituting elements of a CCS cost estimate is provided in Figure 14. (It will be always assumed hereafter that the main product of the plant investigated is electricity.)

Figure 14: logical blocks constituting a cost analysis (image from [71])



The starting point of the analysis is a set of specifications regarding the process realized in the plant. The specifications may be more or less detailed. In the simplest studies it is only requested a series of block flow diagrams for the critical sections of the plant, while the scope of the analysis tends to become larger in advanced studies, since a greater number of sub-systems and auxiliary processes need to be considered.

Besides having a larger project scope, finalized studies tend to have also a smaller unitary level of analysis. In other words block flow diagrams are gradually replaced with process flow diagrams of increasing complexity, including also aspects such as process control and piping isometrics, which usually are not considered at first.

Project specifications require the definition of a series of circumstantial factors like fuel characteristics, plant location and typical climate conditions.

Circumstantial factors may also be defined with increasing levels of detail and accuracy.

It is worth to notice, as evidenced by Rubin [67], that a precise and realistic modelling of circumstantial factors is often omitted from many studies on CCS. This is especially the case for what concerns the opportunities and challenges offered by the plant location.

However, the author continues, circumstantial factors are of paramount importance to correctly evaluate, in absolute terms, the cost of a real-world CCS project.

For example he observes that, if it is considered a CCS project involving the retrofit of an old unit, and there is not enough space to build the capture plant near to the existing power block, the realistic modelling of the project would need to include a particularly high level of pressure losses and capital costs associated to the pipings that convey the exhaust gases from the power block to the CO₂ capture unit. A study that does not consider the circumstantial factors related to the geography of the location might therefore end up underestimating the real costs of the project.

A second fundamental element constituting a cost estimate is the development of a performance model for the power plant. It is typically a thermodynamic model describing the transformations realized in the various systems considered within the scope of the project.

The thermodynamic calculations may be performed by pencil and paper, or with the aid of a process simulator.

Provided that the process is well understood, which is not necessarily the case for projects involving carbon capture, the modelling of the critical sections of the plant is generally developed from the early stages of the engineering process with the intention of being as adherent as possible to the physical reality. As a consequence, the results of the performance model for a certain power plant should not be radically different in a preliminary study and a finalized cost estimate developed by an engineering firm.

However there might be intrinsic uncertainty regarding the correct modelling of some processes. It is also frequent in early-stage studies developed by single students or researchers (or by small groups) to encounter approximations or omissions regarding the modelling of some cost items or parts of the process which are considered non-critical.

Therefore the way in which the performance model for a certain power plant is developed adds, in general, a degree of variability in the possible results of a cost analysis; and in particular there is a tendency for the expected performance to worsen when a feasibility study is converted into a more definitive engineering design.

An important result that can be obtained from the performance model is the list of mass and energy flows entering and exiting the fence line of the power plant. In particular, the exiting streams may be classified into:

- Researched products. The plant is built explicitly in order to satisfy a demand for these products, which are consequently sold generating a revenue. The electricity produced in a power plant is an example of explicitly researched product.
- By-products. This category includes the material (or energy) flows that remain as scraps of the transformations realized in the plant, being however not completely devoid of value. By-products may be sold generating a revenue if specific circumstantial conditions are verified. An example of by-product is sulfur in oil refineries.
- Waste streams. This category includes all the material and energy streams exiting the power plant which do not fall in the two categories just depicted. Waste streams are a liability for the owner of the plant, and they are associated to a cost of disposal, which might be more or less severe according to the hazard potential of the substance, as well as various other circumstantial factors like existing regulations or opportunities for recycle and landfill offered by the surrounding environment.

A typical example of waste stream is a flow of water contaminated with hazardous chemicals. Carbon dioxide, within a power plant implementing a carbon capture scheme, might be a waste stream or a by-product according to the circumstances.

In second place, a performance model can be used to identify the set of requirements that the equipment used to built the different parts of the system will need to satisfy.

In particular, based on the process flow diagrams determined for a particular section of the plant, it is possible to size a certain pump or heat exchanger, and develop a figure for the operation and maintenance costs associated to that specific piece of equipment.

Having obtained a coherent definition of the scope, the performance model and the financial framework of the project, it has to be developed a model for the cost of the plant.

There are two main strategies that can be followed for this purpose:

- Bottom-up method (or analytic estimate). The analytic estimate consists in a detailed estimate of the costs based on a careful analysis of the resources necessary for each activity (or the realization of each component). The greatest advantage of an analytic estimate lies in its accuracy, while the main disadvantage is the elevate quantity of time and effort needed to

formulate the estimates. Sometimes analytic estimates are not employed in the first stages of a project, where quicker and less accurate methods may be more appropriate[68].

- Top-down method (or analogic estimate). An analogic estimate is instead based on the experience gained from passed projects, rather than a careful analysis of all the materials and resources necessary for the construction of a certain process area[68].

The main advantage of analogic estimates is that relatively little time and effort is needed, however the results are highly approximated and have to be taken with care. Another problem is the fact that there might not be any passed experience relative to some processes or activities.

At the end it should be compiled a list of items including all process equipments required for the project, and each items should be associated with an estimate for the relative capital cost, together with the cost of all materials and labor needed to complete the installation. The sum of these terms is known in many published studies as Bare Erected Cost (BEC), and it constitutes the heart of the cost model.

The total capital cost is obtained as the sum of the Bare Erected Cost of all the items considered in the scope, often including also some fees for engineering services expressed as a percentage of the BEC[66].

The capital cost estimates can be classified in a relatively well-established set of categories according to the level of definition of the project and the effort spent to perform the analysis. In Table 1 it is reported the classification of cost estimates according to the AACE (American Association for Cost Estimates) best practices. In Table E-1 in Appendix E it is provided a more detailed description of the methodologies appropriate for the development of each type of class of analysis.

Table 1: Classification of cost estimates (from AACE [119])

Class of estimate	Level of project definition (as % of complete Definition)	Typical Purpose of estimate	Methodology (estimating method)	Expected accuracy range (+/- range relative to best index of 1)	Preparation Effort (relative to lowest cost index of 1)
Class 5	0% to 2%	Screening or Feasibility	Stochastic or Judgement	4 to 20	1**
Class 4	1% to 15%	Concept study or Feasibility	Primarily Stochastic	3 to 12	2 to 4
Class 3	10% to 40%	Budget, Authorization or Control	Mixed but Primarily Stochastic	2 to 6	3 to 10
Class 2	30% to 70%	Control or Bid / Tender	Primarily Deterministic	1 to 3	5 to 20
Class 1	50% to 100%	Check estimate or Bid / Tender	Deterministic	1*	10 to 100

* the accuracy of a Class 1 estimate is between -4% and +6%

** the cost to perform a Class 5 estimate is between 0,015% and 0,03% of the total installed cost of the plant

As previously observed, the scope of the project tends to grow when a feasibility study (Classes 4 and 5) is converted in a finalized study (Classes 1 or 2). Moreover, even if great effort is spent, a full definition of the activities necessary to realize the project cannot be realistically known until the construction of the plant is actually concluded.

It is therefore common in cost estimates to consider an additional capital cost item, usually called *project contingency cost*, accounting for the cost of equipment or other costs that would be identified in a more detailed design of a definitive project at a particular site[66]. According to the guidelines prescribed by the Electric Power Research Institute (EPRI), the project contingency is expressed as a fraction of the total capital cost of the power plant, and it is not applied to individual components. After project contingencies are applied, the results of the cost estimate are expected to lie in a specified confidence interval, which depends on the Class of the cost estimate[66]. In Table 2 it is provided the set of coefficients prescribed by the EPRI for each Class of analysis in order to determine the project contingency cost.

Table 2: guidelines for Project Contingency costs (from [120])

Cost classification	Design effort	Project contingency cost (%) ^a
Class I (similar to AACE Class 5/4)	Simplified	30–50
Class II (similar to AACE Class 3)	Preliminary	15–30
Class III (similar to AACE Class 3/2)	Detailed	10–20
Class IV (similar to AACE Class 1)	Finalized	5–10

^a Percentage of the total of process capital, engineering and home office fees, and process contingency.

Another capital cost of similar nature that should be included in the final account is the so-called *process contingency cost*. It is a cost that keeps into account the level of maturity of a particular process or component. It attempts to quantify the additional capital cost that will likely arise as the process matures into full scale commercial technology[66].

According to the EPRI guidelines (shown in Table 3), process contingencies should be determined as a percentage of the capital cost estimated for each item of the process equipment, with higher percentages applied to components or sub-units at earlier stages of development.

Table 3: guidelines for Process Contingency costs (from [120])

Technology status	Process contingency (% of associated process capital)
New concept with limited data	40+
Concept with bench-scale data	30–70
Small pilot plant data	20–35
Full-sized modules have been operated	5–20
Process is used commercially	0–10

Importantly, process contingencies should be based on the *current* level of development of a certain process, but, as evidenced by Rubin[67], the guidelines on process contingencies are not followed in many published studies. In particular it often happens that the cost of advanced systems for CO₂ capture, for whom little experience is available outside laboratory scale, is expressed like if they were mature processes commercially available, and with dubious or ill reported adoption of the process contingency costs.

As previously discussed, Rubin observes that in some types of context it might be actually interesting to investigate the future cost of a currently advanced technology in the hypothesis that certain R&D goals are encountered and the technology is brought to commercial maturity. However he notices that, if the objective is to estimate the future cost of an advanced technology in an hypothetical scenario, the use of process contingencies appears methodologically inappropriate since it is not possible to know in advance which is going to be the future design or the cost of a successful technology.

To address this problem some authors [70] have proposed an alternative hybrid method which is articulated in the following steps:

- It is calculated the current cost necessary to build the first plant employing a certain advanced technology never deployed at large scale. This cost is called First-of-a-kind (FOAK) cost and it need to be determined with a precise analytic estimate. The guidelines for process contingencies need to be followed.
- Made the assumption that technological maturity is achieved after N power plants have been built (the number N needs to be estimated), the cost necessary to build the N-th power plant (NOAK cost) is determined from the FOAK cost by adopting the same learning rate historically measured for a similar technology. In Figure 15 it is represented an example of learning trajectory from FOAK to mature plant.

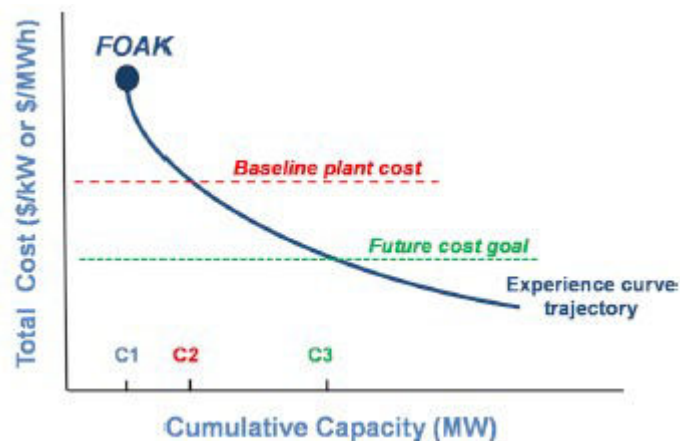


Figure 15: Illustrative cost trajectory of an advanced technology from FOAK plant to mature plant, showing the deployment of the technology needed to meet a given cost goal(image from [67]).

In addition to what has been discussed so far, it is necessary to consider, in the capital cost account, all those items that are not included in the BEC or the engineering fees. These costs, which are usually grouped under the label *owner's costs*, include a relatively wide range of terms which may be both general, this is for example the case of the fees that have to be paid to the financial intermediaries to raise the capital, or specific to a certain location. An example of location-specific cost is the sum required to upgrade the infrastructures (roads, railways) in the specific location in which the plant will be built.

Finally yet important, operation and maintenance costs need to be taken into account. These costs may be split in fixed costs and variable costs. Among fixed operative costs the main items are labor and fixed maintenance costs. Among variable operative costs, which are proportional to the operating schedule of the plant, it is instead possible to distinguish between disposal costs and the cost associated to all those inputs that are consumed in the normal operation of the power plant.

2.3 Typical metrics for CCS cost reporting

There is a great number of metrics used in the literature to report the cost of a CO₂ capture and storage system. In the following lines it is provided a brief summary of some of the most adopted ones:

- Levelized Cost of Electricity (LCOE). *The LCOE is a constant unit price (\$/Mwh) for comparing the costs of power plants that have different technologies, use different fuels,*

have different capital expenditure paths, differing annual costs (such as operating, maintenance, taxes, carbon prices), different net outputs, and different economic lives[66]. It is defined as the constant price of electricity that nullifies the Net Present Value of the project over its entire duration.

From a mathematical point of view the LCOE can be written as:

$$LCOE = \frac{\sum_t \left[\frac{\text{Capital Expenditures}_t}{(1+r)^t} + \frac{\text{OMcost}_t}{(1+r)^t} + \frac{\text{Fuel}_t}{(1+r)^t} \right]}{\sum_t \frac{\text{Electricity Sold}_t}{(1+r)^t}}$$

- Cost of CO2 avoided. It compares the CCS plant with a reference power plant without CCS. It is calculated as the additional cost necessary to avoid 1 ton of CO2 emissions without changing the power output. From a mathematic standpoint it is expressed as:

$$\text{Cost of CO2 Avoided } (\$/\text{tonCO}_2) = \frac{(LCOE)_{CCS} - (LCOE)_{REF}}{(\text{tonCO}_2/\text{MWh})_{REF} - (\text{tonCO}_2/\text{MWh})_{CCS}}$$

It must be noted that the reference plant needs to be always specified in order for the indicator to be usable.

- Specific Primary Energy Consumption per unit of CO2 Avoided (SPECCA). It is an indicator that compares the CCS plant with a reference power plant without CCS. The SPECCA is calculated as the additional fuel, expressed in terms of primary energy, which needs to be spent per each ton of CO2 emissions avoided in a power plant at constant production. Mathematically it is written as:

$$SPECCA (MJ/\text{tonCO}_2) = \frac{\left(\frac{1}{\eta_{CCS}}\right) - \left(\frac{1}{\eta_{REF}}\right)}{(\text{tonCO}_2/\text{MWh})_{REF} - (\text{tonCO}_2/\text{MWh})_{CCS}}$$

The SPECCA is frequently adopted to compare the performance of different systems for CO2 capture from a purely thermodynamic perspective. However it has the drawback of not including any information regarding the capital cost necessary to realize a certain process, therefore it needs to be complemented with other metrics in order to obtain a complete economic assessment of a certain system.

2.4 Frequent pitfalls of CCS cost estimates

A great number of studies has been published over the years investigating the cost of different carbon capture and storage systems.

Among these, it can be distinguished in particular between journal and conference papers, which are generally constrained by a limit of words, and technical reports of long format, often issued by organizations such as the National Energy Technology Laboratories (NETL), the European Benchmarking Task Force (EBTF), the International Energy Agency (EIA), the IPCC or others.

As documented in a 2012 review[71] from Rubin, an undesired consequence of such research effort has been the proliferation of different terminologies and methods adopted by the various organizations or research groups to obtain and communicate their results.

For example owner's costs and contingency costs, which constitute a significant fraction of the total capital cost of the plant, are calculated differently in the studies surveyed in the review, and are sometimes straightforwardly omitted.

Other areas where differences among the published studies can be frequently encountered are the definition of the scope of the project, the financial framework of the project and the estimation methodology of the process equipment cost, that is seldom reported in an accurate way.

While in some circumstances there may not necessarily be an accounting framework or a set of assumptions that is right or wrong in absolute terms, studies with different assumptions are not comparable and might generate confusion or misunderstandings for a reader that is not familiar with the details of CCS costing[71].

Moreover, Rubin points out that some assumptions commonly adopted in many published studies might be accused of being biased in the direction of a systematic underestimation of the costs of CCS, although there are no proofs that this is done with malice or intention.

An example of such kind is constituted by the value of 85% commonly attributed to the levelized capacity factor for coal plants. The levelized capacity factor (LCF) is a time-weighted discounted average of the availability of the plant over its entire lifetime.

If on the one hand there have been some baseload plants that have operated at 85% capacity for several years, the author observes that the average availability of the coal plant fleet in the US has been historically between 65% and 75%, and that the availability in the first years of operation, which are the most important ones in the calculation of the LCF, is always significantly lower than 85%.

Therefore he argues that the levelized capacity factor is systematically overestimated and that a more realistic value would be around 71%.

In conclusion Rubin points out that *a concerted effort to improve the understanding and communication of CCS cost estimates within the technical and policy communities is urgent* (2012), and in later publications with other authors he proposes a methodology[66] and a series of guidelines and best practices [67][72] for CCS cost reporting in order to improve clarity and consistency, highlighting in particular the fundamental importance of transparency (coherently with what is allowed in a given document format) in the reporting of the all the various assumptions employed to obtain a cost estimate.

3 DESCRIPTION OF THE REFERENCE CASE POWER PLANT

The chief objective of this report is to perform a techno-economic evaluation between different options for post combustion capture applied to a given reference plant.

In next pages it will be first offered a presentation of the methodology adopted to perform the economic evaluation of the power plant chosen as a reference.

It will be first provided a definition of the scope of the reference plant, which will be also described in its main processes and sub-units. Then, given the various circumstantial and financial factors, it will be developed a model for the performance and costs of the plant. Finally, the results will be commented.

The configurations with carbon capture will be instead described and analyzed in section 4 and 5.

3.1 Introductory notes on the methodology

The economic analysis performed regarding the reference case power plant is driven by the desire of establishing a common baseline in order to evaluate the various cases involving carbon capture. The main objective is therefore not to determine the most realistic estimate regarding the cost of the electricity produced, but rather to develop a coherent methodology to estimate the performance and the costs of the sections of the power plant that are present in all the cases investigated.

The following analysis can be in other words characterized as the starting point of a technological assessment answering a *what if* question.

The inquired question might be framed explicitly in the following way:

“Given the information currently available from the literature, which of the investigated technologies would be the most convenient way to reduce the emissions from the reference plant under the hypothesis that vast experience is acquired with all the subsystems and components included in the scope of the plant?”

The set of methodologies adopted does not fall in a clear way within any methodological framework of which it has been possible to have notice from the literature.

It has been made an effort to provide a wide definition of the scope of the project, meaning that a relatively large number of auxiliary systems has been considered. In particular it has been considered a project scope which is as similar as possible the case 1A analyzed in the 2019 edition of the Cost and Performance Baseline report published by the NETL[73].

At the same time components and sub-units have often been depicted as simple functional black-boxes univocally defined with a single parameter.

In other words, if on the one hand the terms *component* or *piece of equipment* have been frequently adopted, only occasionally physical objects have been described with precise and well defined specifics. More often components have been defined in terms of the series of functions that hypothetical groups of properly organized parts would be called to supply if the plant was to be built.

For example, no technical drawings will be provided in the next pages about the geometric specifics of the various heat exchangers. It will be instead defined the heat exchange area or the transfer of thermal power that the physical component should guarantee at nominal conditions if the plant was realized.

The performance model for the various units of the plant has been developed with an heterogeneous set of methodologies that will be described more thoroughly in the next pages, however, broadly speaking, it is possible to distinguish between two categories.

In the first category are those systems, like the steam cycle, for which it has been determined an ad-hoc thermodynamic model during the course of this work. The calculations have been performed in a Matlab environment or with the aid of an Excel spreadsheet.

The second category includes the remaining systems, which are the majority, and for which a series of process flow diagrams and other quantitative data have been repurposed from existing sources, often with the introduction of a correction coefficient to adjust for the different scale or other factors.

The cost model has been constructed with a bottom-up methodology. However the capital costs for the various unitary items or functional units have not been determined through an analytic procedure, rather they have been extrapolated from existing sources with the use of correction parameters and experience-based curves. In particular most of the capital cost estimates have been determined from raw data contained in the NETL report.

Given that the data on capital costs provided by the NETL are supplied by external contractors (Black&Veatch in particular in the case of the 2019 cost and performance baseline report)[74] which do not disclose the procedures through which the capital cost estimates are obtained, a possible critique that might be moved against the present work is that there is dubious coherency between the performance model and the cost model developed.

In light of the fact that the quality of raw data is always a *conditio sine qua non* for the good outcome of a given study, this argument seems to be very poignant. On the other hand it has to be also kept into consideration that a non-neglibigle amount of project contingency costs has been included in the total capital cost of the plant to account, among other things, for eventual small inconsistencies in the definition of the scope.

In particular it has been followed the same convention adopted by the NETL report in order to determine the project contingencies, meaning that it has been considered an extra-cost around 17-23% for each capital cost item.

Given such level of project contingencies, it might be possible to frame this analysis as a preliminary study (Class 3) according to the AACE classification, although some of the methodologies adopted are suitable rather to a Class 4/Class 5 study (see table E-1 in appendix E). In order to be compatible with a Class 3 estimate the results of this study should yield an accuracy between -24%/+36% with respect to the real cost necessary to build the power plant.

This point will be raised again during the discussion of the results.

3.2 General characteristics of the reference plant

The reference case of this study is an air blown IGCC power plant. The project is supposed to be a greenfield installation in a generic location in the midwestern US at sea level. According to the suggestion of Rubin and colleagues[66], Table 4 has been filled with a list of important circumstantial factors needed to correctly frame the scope of the system investigated. Some of the recommended slots have been left empty due to the lacking of specific data on the subject.

Table 4: list of parameters needed to specify the scope and battery limits of a grassroots “reference case” power plant without CCS (from [66])

Plant size (net power output, MW)	931,89MW
Plant location (country, region of country, or state)	The plant is located at a generic plant site in the midwestern United States
Site characteristics	ISO conditions
– Plant elevation/ Atmospheric pressure	0(m) / 1,01 Kpa
– Design ambient dry/ Wet bulb temperatures	15°C / 10,8 °C
– Minimum/maximum design temperatures	
– Design ambient relative humidity	60,00%
– Site topography (e.g., assumed to be clear and level?)	Level site, greenfield installation, 300 acres
Generation technology	IGCC
– Specific technology features	
– Gasifier type	air blown MHI gasifier
– Steam conditions	144 bar maximum pressure
– Condenser pressure	0,068 bar
Fuel characteristics	Illinois n6 coal
– Coal ultimate analysis (including HHV and LHV)	Coal as received (%wt.: 61.27 C, 4.69 H, 8.83 O, 1.1 N, 3.41 S, 12 moisture, 8.7 ash) LHV 24,826 Mjth/kg
– Coal ash analysis (including ash fusion temperatures)	
– Coal delivery method (rail, barge, truck, conveyor, etc.)	Rail
– Other start-up fuel source (i.e., distillate, etc.)	Natural gas is the startup fuel
Air emission limits	
SO2	
NOx	
PM (Filterable)	
Hg	
Indoor or outdoor construction?	The steam turbine is located in a specific building
Makeup water source and typical quality	50% of the water is groundwater 50% comes from a publicly owned Water treatment facility (see the NETL report for a precise description of water solutes)
Cooling water system	mechanical draft cooling tower 4 cycles of concentration
Waste water disposal method	Zero Liquid Discharge
Electrical system	
– Grid frequency	60 hz
– Transmission system Interconnect voltage	345kV
– Switchyard included?	no
– Transmission line included? If so, how long?	no
Material storage assumptions	
– Indoor vs. outdoor storage	outdoor
– Coal pile (days of storage)	30 days
– Ash solids (days of on-site storage)	30 days
Any special noise limitations?	

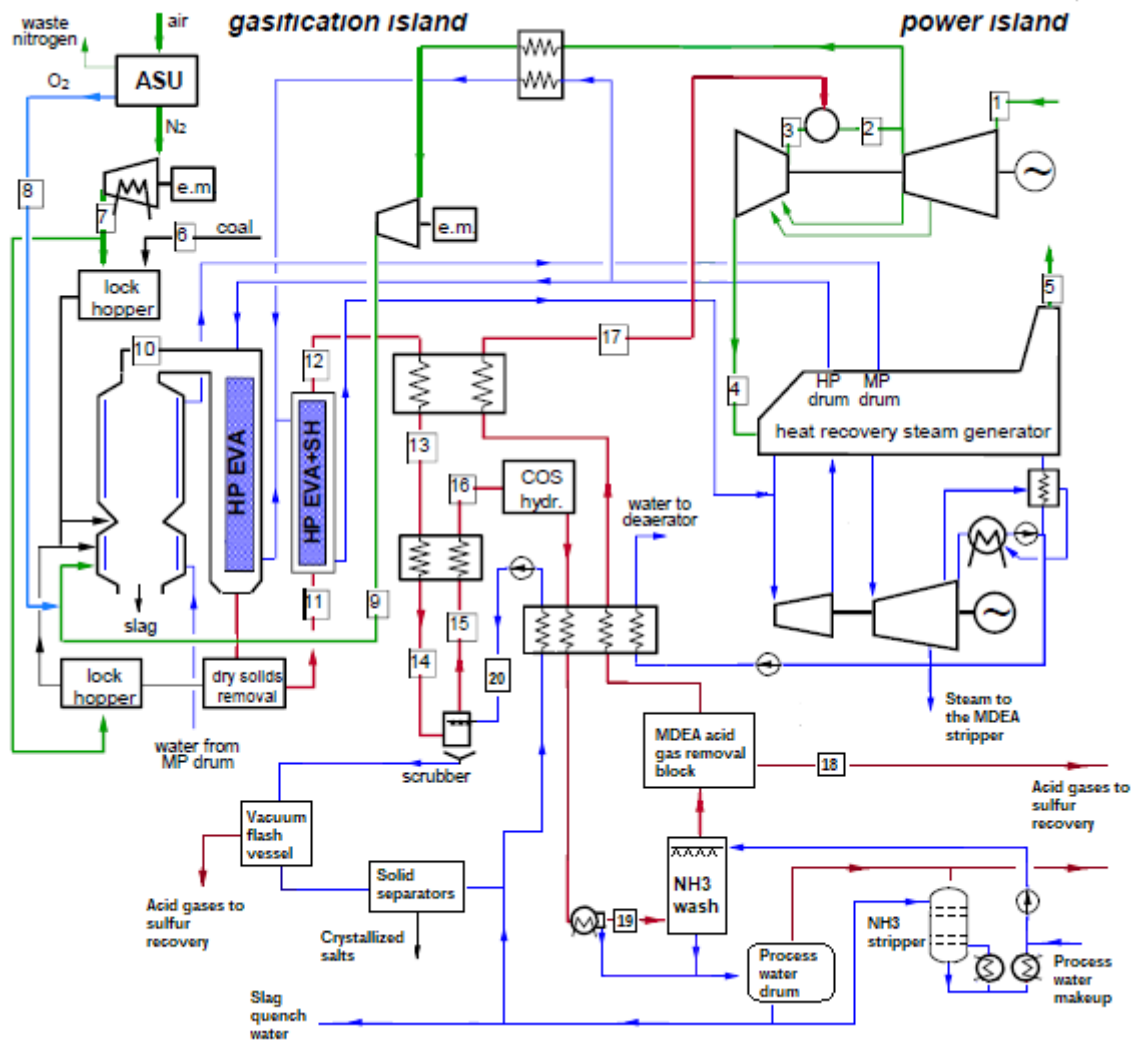
3.3 Description of the process

3.3.1 Gasification train

The main section of the power plant is constituted by two gasification trains like the one represented in the process flow diagram in Figure 16. Both of the gasification trains are linked to a steam cycle, which is unique for the entire plant.

For detailed information regarding the numbered streams in the PDF the reader is referred to Table 5. The process flow diagram in its general lines has been sourced from an article from Bonalumi et. Al[75]. However some additional streams have been included in the graphic representation (marked with an asterisk in Table 5) due to a different framework of assumptions that has been adopted in this analysis to model some systems.

Figure 16: schematic of the IGCC plant (reference case - without co2 capture)



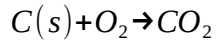
The most important input of the process is a stream of coal Illinois n6, that is fed into a gasification reactor. The specific gasifier chosen for this analysis (MHI gasifier) is a double stage entrained flow reactor with water walls[76].

The technology that is utilized to feed solid coal in the pressurized reactor is called dry lock hopper system, which pneumatically conveys batches of coal by means of an high pressure flow of inerts.

Nitrogen is the inert gas used in the reference plant, and it is delivered by a small air separation unit (ASU) specifically for this purpose.

As a by-product, the ASU also produces a stream of oxygen that is used to enrich the air used as an input for the gasification.

In the first stage of the gasifier (called combustor) compressed air and coal are reacted exothermically in order to produce an extremely hot (1900°C) flow of exhausts (mainly N₂, CO₂, H₂O).

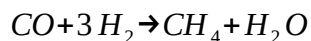
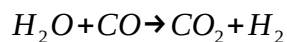
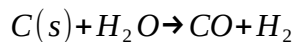


The complete oxydation of a significant amount of coal in the first stage of the reactor is necessary because the gasification reactions happening in the second stage (called reductor) are endothermic and need to be sustained by an exogenous source of energy.

Table 5: Temperature, pressure, flow rate and composition of the main streams of the air-blown IGCC shown in Figure 16

	Temperature °C	Pressure Bar	Mass flow (kg/sec)	Ar	CO	CO ₂	H ₂	H ₂ O	H ₂ S	N ₂	O ₂	CH ₄
1	15	1,01	611	0,92	0	0,03	0	1,03	0	77,28	20,73	0
2	417,6	18,16	378,8	0,92	0	0,03	0	1,03	0	77,28	20,73	0
3	1438,8	17,61	540	0,89	0	10,62	0	5,65	0	75,24	7,6	0
4	611,1	1,04	665	0,89	0	8,59	0	4,77	0	75,63	10,12	0
5	115	1,01	665	0,89	0	8,59	0	4,77	0	75,63	10,12	0
6	15		39,1	Coal as received (%wt: 61.27 C, 4.69H, 8.83 O, 1.1 N, 3.41 S, 12 moisture, 8.7 ash)								
7	80	56,13	19,9	0	0	0	0	0	0	100	0	0
8	15	35	7,3	3,09	0	0	0	0	0	1,91	95	0
9	477,4	33,02	107,1	0,92	0	0,03	0	1,03	0	77,28	20,73	0
10	1200	28,06	172,4	0,61	25,28	3,62	11,91	4,22	0,62	53,19	0	0,56
11	900	28,06	170	0,61	25,28	3,62	11,91	4,22	0,62	53,19	0	0,56
12	350	27,5	170	0,61	25,28	3,62	11,91	4,22	0,62	53,19	0	0,56
13	206,6	26,95	170	0,61	25,28	3,62	11,91	4,22	0,62	53,19	0	0,56
14	137	26,41	170	0,61	25,28	3,62	11,91	4,22	0,62	53,19	0	0,56
15	112	25,89	172,2	0,6	24,83	3,56	11,7	5,92	0,6	52,24	0	0,55
16	180	25,37	172,2	0,6	24,83	3,56	11,7	5,92	0,6	52,24	0	0,55
17	250	23,16	161,2	0,64	26,74	2,93	12,6	0,24	0	56,26	0	0,59
18*	45	1,01	3,93	0	0	58,36	0	0	41,64	0	0	0
19*	35	~24,5	165,13	0,63	26,33	3,77	12,4	0,23	0,63	55,4	0	0,58
20*	137	25,89	65,45	0	0	0	0	100	0	0	0	0

In the reductor stage the remaining coal is added, without further oxygen. Due to the high temperatures, solid coal particles react endothermically with the surrounding atmosphere, cooling down the system of about 700°C[76].



The flow finally exiting the top of the reactor is a pressurized mixture of carbon monoxide and hydrogen importantly diluted with nitrogen and other inerts. Such mixture is called syngas. It also contains ashes and unconverted carbon, as well as various gaseous pollutants such as H₂S and COS. Syngas is cooled from 1200°C to 900°C and subsequently passed into a cyclone and a candle filter where solid particles are collected and circulated back to the reaction chamber. The recirculation of

solids is needed to obtain a complete conversion of solid carbon, which is not granted through a single passage due to the relatively poor kinetic conditions in the reductor stage.

An important feature to notice in the design of the MHI gasifier is that the system is supposed to be able to handle with great flexibility coals with different characteristics, and in particular with relatively low ash melting point[77][78]. As a consequence, the syngas at the gasifier outlet (~1200°C) can be cooled in a normal convective heat exchanger without the need of exotic solutions like radiant coolers or quench baths.

After filtration, the syngas is further cooled down to 137°C in a series of heat exchangers, and it is purified in a scrubber from the particulate matter, which would be detrimental to the operation of the downstream equipment. In this analysis it is imagined that an ejector venturi type scrubber is employed.

The venturi scrubber is a device that separates the particulate matter from the syngas by exploiting the difference in inertia. In particular the syngas flow rich of particulate is passed into a narrow throat full of water. In this way, in order to satisfy the continuity equation, the gas needs to suddenly accelerate. On the other hand the solid particles remain entrained in the turbulence of the liquid and are carried away with it.

After the scrubbing operation the syngas is reheated up to 180°C, and it is passed into a catalytic reactor aimed at converting COS into H₂S.

Then the syngas is cooled down to 35°C in order to facilitate the removal of H₂S. This cooling section is called low temperature heat recovery (LTHR).

The H₂S removal process is performed by chemical absorption in a packed column. The solvent adopted is mono-diethanol-ammine (MDEA).

Since the AGR solvent is sensible to the presence of NH₃ in the syngas[79], it has been decided to consider as part of the process also a water wash column preceding the AGR section. Aim of the water wash is to reduce to traces the ammonia in the syngas by spraying it with fresh water.

The presence of an absorption column before the AGR absorber was not included in the aforementioned article from Bonalumi et. al[75], from which the process data on the gasification train were sourced. For this reason in the current analysis the impact of the column has only been accounted from the point of view of its capital cost, without considering, in a quantitative manner, the mass and heat transfer processes occurring in it.

After the sulfur removal block, the syngas flow is reheated in a regenerative heat exchanger and burned in the combustor of a combustion turbine. The compressor of the CT unit provides also the stream of compressed air that is used in the gasifier.

The combustion turbine inlet temperature is set at 1360°C. Being the syngas already significantly diluted with nitrogen, it is not necessary to mix the syngas with steam or nitrogen to obtain a stoichiometric flame temperature compatible with acceptable NO_x emissions at the turbine outlet. The operating parameters of the CT unit have been directly sourced, without any modification, from Bonalumi et. al[75]. In particular, it is considered an advanced CT unit.

The capital cost of the CT unit has been estimated through the following steps:

- It has been considered the combustion turbine technology employed in the 2019 NETL baseline report for the IGCC cases, which is a modified version of a state-of-the-art 2008 F class turbine originally intended to work with natural gas. The modifications include a redesign of the combustor and an adaptation of the nozzle areas of the turbine to permit the passage of a greater volumetric flow of gas. The generator is a standard H₂-cooled machine with static exciter. The net power output of the unit is 232 MW in ISO conditions for each train.

- Since the net power output of the combustion turbine included in this work is 20% higher than the one considered in the NETL report, it has been decided to consider, as a first degree approximation, an increase in the capital cost equal to 13,2% (in comparison with the NETL figure) according to the two-thirds rule[69].

3.3.2 Steam cycle

After exiting the combustion turbine, residual heat is recovered from the exhaust gases in a two pressure levels heat recovery steam generator (HRSG).

The process flow diagram considered in this analysis for the steam cycle is represented in Figure 17 and Table 6, being constructed assuming the operating parameters listed in Table 7. The parameters have been set to the values described in the article [75].

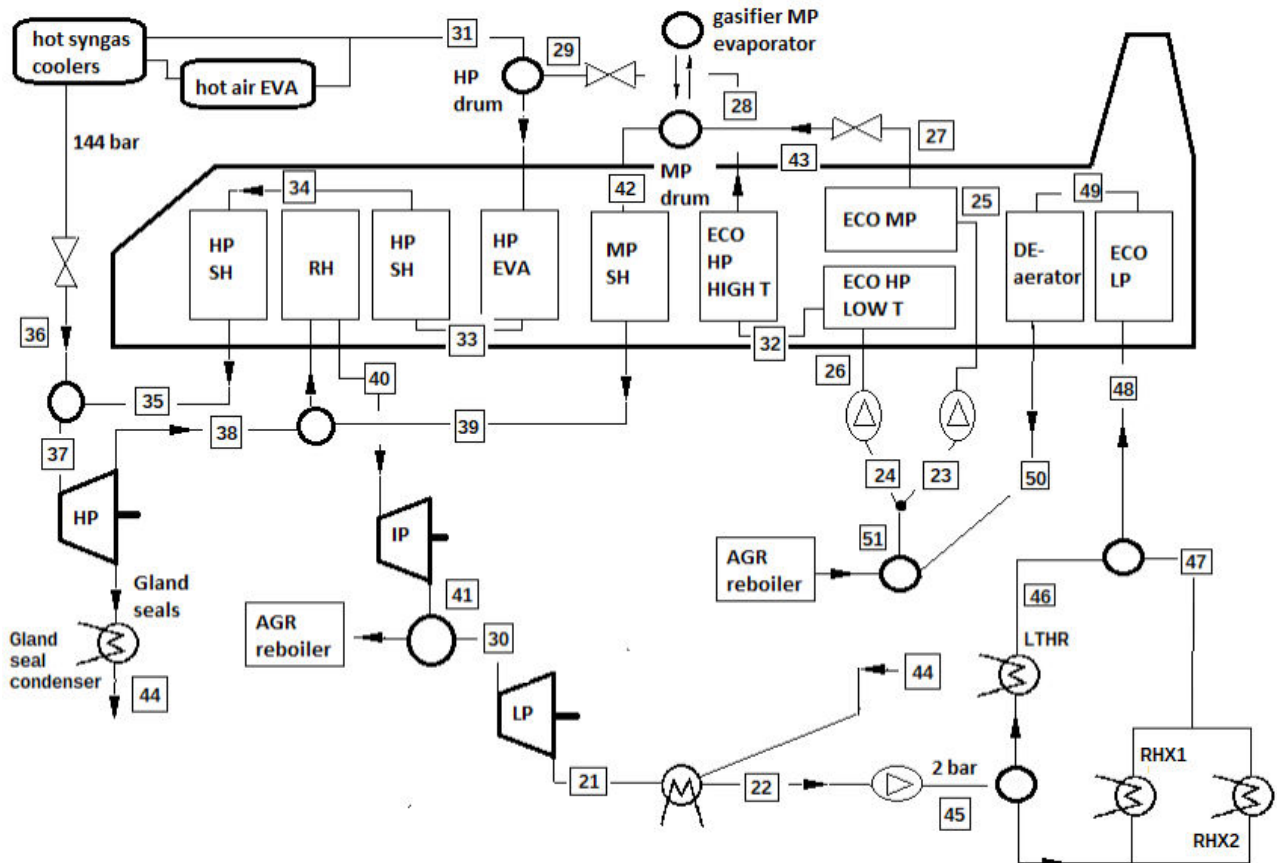
Two notable exceptions concern the condenser pressure and the LP turbine isentropic efficiency, which have been set to the values suggested by the NETL in its technical guidelines[80].

Steam has been modelled with the RK equation of state. The quantity of steam available has been determined modelling the flows of syngas or flue gases with the ideal gas law.

Table 8 shows the auxiliary electric consumptions that have been associated to the steam cycle.

In the following paragraphs the system is briefly described, and in particular, given that the capital cost of the equipment used in the steam cycle is determined from the data provided by the NETL 2019 report, it is provided a description of the scope of the unit as it is depicted in the NETL report.

Figure 17: schematic representation of the bottoming steam cycle (reference case – without CCS)



High temperature heat is introduced in the cycle mostly within two syngas cooling trains and two heat recovery steam generators (only one is represented in the diagram).

Main steam from the syngas coolers and the high pressure HRSG circuit is collected in a header and expanded in the high pressure turbine. The exiting flow is then mixed with the steam from the

medium pressure circuit, and reheated. Reheated steam is expanded in the medium pressure turbine and sent to the crossover tube, from which some quantity of steam at 2 bar is extracted and used in the AGR solvent regeneration process. Lastly, the remaining steam is expanded from 2 to 0.068 bar in the LP turbine, and discharged in the condenser. Both the high pressure and intermediate pressure turbines are preceded with a stop valve and a control valve to regulate the amount of steam admitted to the blades.

The steam condenser is a single pass heat exchanger with cooling water circulating inside the tubes, and it is divided into two separate sections that can be run independently in order to have on-line maintenance. Cooling water temperature is set to 16°C at the condenser inlet, and 27°C at the outlet. The system is equipped with extraction pumps for the removal of non condensable gases during turbine operation, or to create vacuum during startup before steam is admitted in the condenser. Condensate is extracted from the hotwell by a centrifugal pump with vertical axis, and it is sent to the deaerator tank through a series of preheaters.

The entire preheating line is supposed here to operate at the same pressure of the deaerator (2 bar) and no account is made of pressure losses or height differences.

Most of the water is preheated in a series of two shell and tube heat exchangers with residual heat from the low temperature syngas entering the AGR unit (see the appendix A for a detailed analysis of the LTHR section). However, since the amount of syngas is not sufficient to fully preheat all the water up to 100°C, a secondary preheating circuit has been also included in parallel. The secondary line recovers heat from the ZLD water treatment system.

Table 6: temperature, pressure, enthalpy and mass flow of the main streams of the bottoming steam cycle shown in Figure 17

	21	22	23	24	25	26	27	28
temperature (°C)	38,3	38,3	120,21	120,21	120,62	121,3	238,79	333,42
pressure (bar)	0,068	0,068	2	2	61	160	61	160
enthalpy (KJ/kg)	2381,512	161,11	504,677	504,677	510,577	520,477	1031,787	1544,736
mass flow (kg/sec)	294,21	297,31	21,92	306,59	21,92	306,59	21,92	306,59
	29	30	31	32	33	34	35	36
temperature (°C)	333,42	191,2	333,42	238,79	339	380	551,5	565
pressure (bar)	144	2	144	160	144	144	132,5	132,5
enthalpy	1544,736	2853,039	1544,736	1031,787	2628,681	2905,088	3472,748	3508,09
mass flow (kg/sec)	306,59	294,21	203,59	306,59	103	103	103	203,59
	37	38	39	40	41	42	43	44
temperature (°C)	560,6	359,6	359,6	563,9	191,2	244,19	238,79	15
pressure (bar)	132,5	36	36	33,12	2	36	36	1,01
enthalpy	3496,21	3125,817	3125,817	3598,2	2853,039	2802,484	1031,787	72,6
mass flow (kg/sec)	306,59	303,49	21,92	325,41	325,41	21,92	21,92	3,1
	45	46	47	48	49	50	51	
temperature (°C)	38,3	100	76,52	92,5	115,3	120,21	120,21	
pressure (bar)	2	2	2	2	2	2	2	
enthalpy	161,306	419,173	320,47	387,21	457,31	504,677	504,677	
mass flow (kg/sec)	297,31	201,5	95,81	297,31	297,31	297,31	328,51	

Preheated water enters the HRSG at a temperature close to 100°C, and it is brought to near-saturation conditions by a low temperature bank of tubes (ECO LP). Water is then contacted with steam in the upper part of the deaerator to eliminate incondensable gases.

It is assumed that the deaerator is mounted on the upper part of the HRSG and is provided with a dedicated bank of exchange tubes for the production of low pressure steam (2 bar). The reader is referred to the appendix A for a detailed analysis of the HRSG heat exchangers.

Feedwater pumps have the task of pumping water from the deaerator tank, across the economizers, up to the steam drums. In this analysis it is imagined that the steam drums are at 144 and 36 bar respectively, while the HP and MP economizers feature a pressure drop of 16 bar and 25 bar. Water from the MP drum is sent to the gasifier water walls, where it is vaporized. Medium pressure steam is then superheated in a dedicated section of the HRSG. A pressure drop of 8% is considered in all superheaters.

Water from the HP drum is instead split into two main different flows. About one third is vaporized and superheated in the HRSG, the rest is used to produce steam from the gasification train. In particular, HP water is mostly vaporized in the syngas coolers, however a fraction is vaporized by recovering heat from the gasification air. Some HP steam is also produced in the gasifier water walls.

Ultimately, all the steam produced in the gasification isle is superheated in a specific exchanger that cools the syngas from about 750°C to 350°C. (In appendix A it is provided a TQ diagram of the syngas coolers).

From the point of view of the steam turbine auxiliary systems present, it has to be specified that the steam turbine is connected to an H₂ cooled generator. The shaft bearings are assumed to be sealed with a gland sealing system in order to limit steam leakages or vacuum losses. Gland steam extracted from the bearings is condensed with cooling water in a dedicated exchanger, and recirculated to the condenser.

The bearings are lubricated with a closed loop oil-circulating system. Pressurized oil is cooled with water prior to entering the shaft bearings, then it trickles down to the lube oil tank. An emergency pump is mounted on the tank to secure a stable flow of oil in case of power failure.

Finally, the steam cycle is assumed to be provided with other miscellaneous pieces of equipment that are common to most modern steam power plants. For example the hot components of the steam cycle are supposed to be connected to the rest of the plant by means of expandable joints.

In case of ambiguity regarding the equipment considered, the reader is referred to case 1A of the 2019 NETL cost and performance baseline for further details.

It has however to be noted again that it has not been possible to gather a perfectly transparent account of the capital cost items included in the analysis even in the NETL cost and performance baseline report or the annexed documents.

Table 7: Main assumptions adopted to model the bottoming cycle

Item	Value	Item	Value
Electric generator efficiency %	98,7	Minimum HRSG temperature °C	115
Turbine mechanical efficiency %	99,5	HP turbine isentropic efficiency	0,922
Pumps hydraulic efficiency %	80	IP turbine isentropic efficiency	0,94
Minimum pinch point deltaT °C	10	LP turbine isentropic efficiency	0,88
Pressure losses HP/MP economizers, bar	16/25	Steam leakage from the HP turbine seals %	1
Pressure loss in superheaters %	8	Heat to the gasifier membrane walls, % of input coal LHV	2
Heat loss (% of transfer heat in the high temperature Syngas coolers)	0,7	Subcooling, deltaT °C	5

Table 8: auxiliary electric consumptions associated to the bottoming cycle

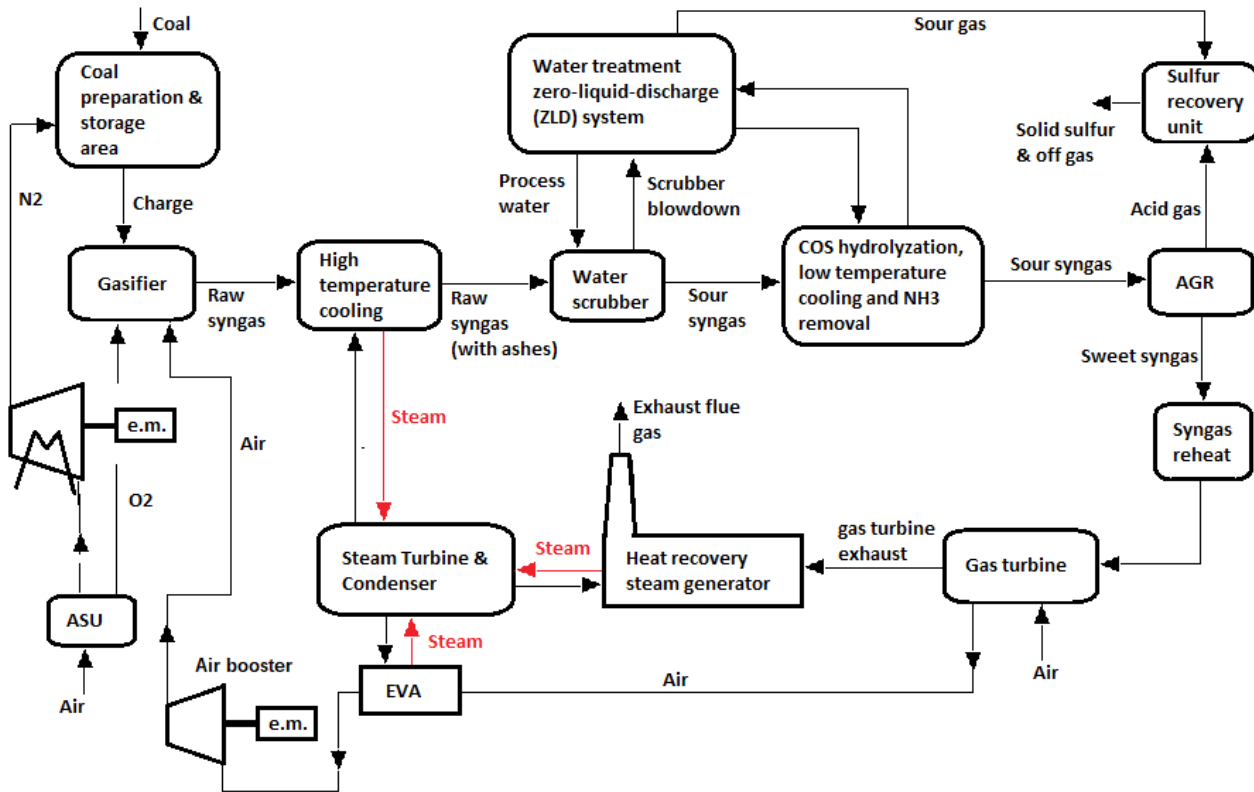
	Unit of measure	Value
Steam turbine auxiliaries	kW	332
Steam cycle pumps	kW	6450

3.3.3 Secondary sub-system

The series of transformations that have been described so far constitutes the main functional block of the plant. However the continuous operation of the power producing-processes would not be possible without the services offered by numerous other subsystems constantly running in parallel and exchanging mass/energy with the main block.

Figure 18 highlights the reciprocal interdependence between the most important functional units constituting the plant.

Figure 18: qualitative block flow diagram of the reference-case power plant



Some of the auxiliary processes have already been mentioned, the remaining ones are here listed and discussed:

- Circulating water system.** The role of the circulating water system is to provide cooling water to various equipments around the power plant. Cooling water is produced for the entire plant in a unique, mechanical draft cooling tower. The main condenser and the AGR unit, which are the units that reject the most heat, are directly connected to the circulating water system, while other users are connected indirectly through a secondary water circuit. The primary and secondary cooling water circuits are connected through a plate and frame heat exchanger. The cooling tower blowdown is discharged after being softened with a double stage reverse osmosis treatment.

Apart from a scale factor, in this study the circulating water system is imagined to be similar to the circulating water system that supplies the IGCC plants in the cases described in the NETL cost and performance baseline. The operating parameters are also supposed to be equal. In detail: it is considered a cooling water temperature of 27°C and 16°C entering and

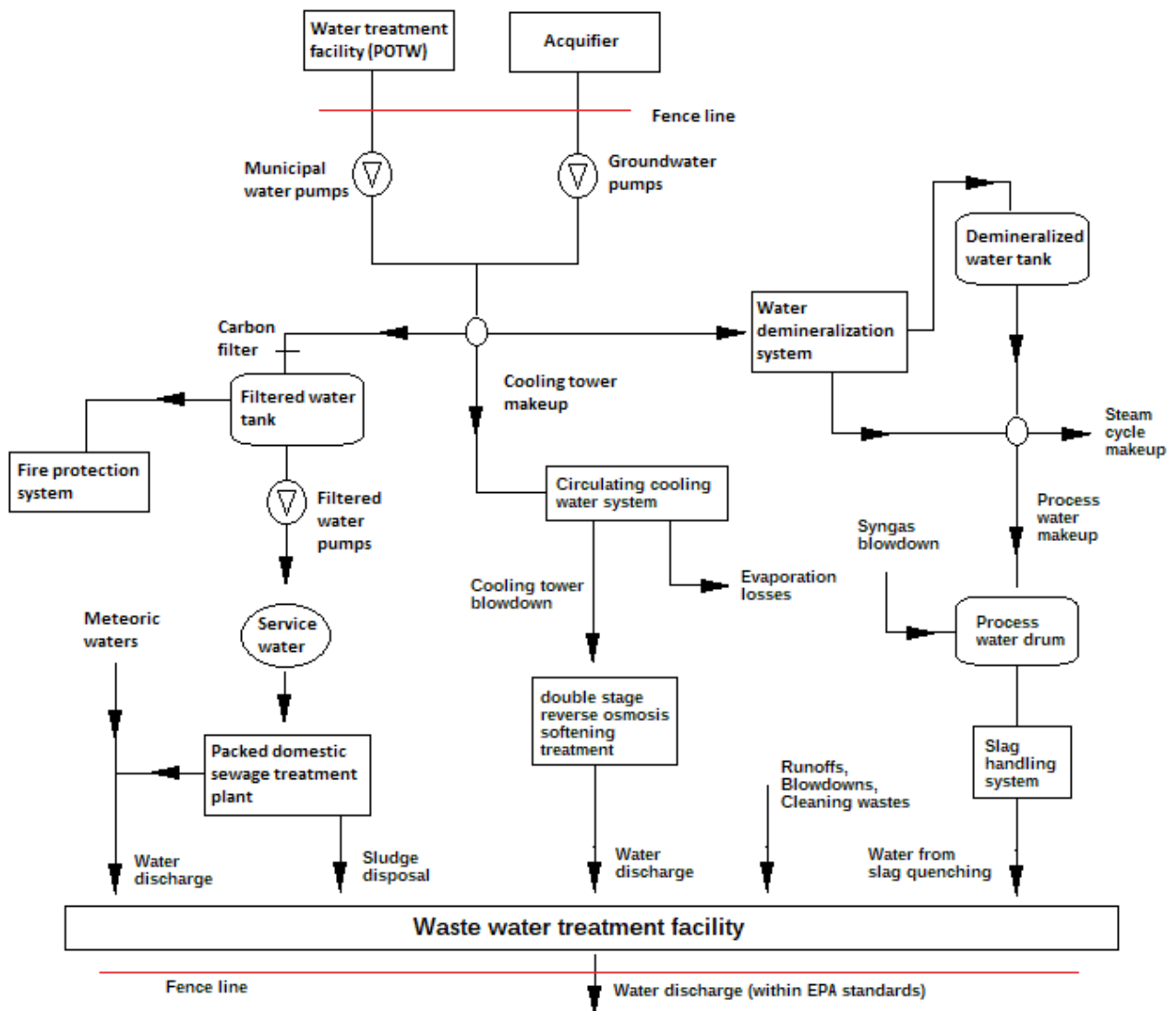
exiting the tower respectively. The specific electric auxiliary consumptions have been set to 11 kW per MW of thermal power rejected. The level of the evaporative losses is set to 1.6% of the circulating water, while the blowdown losses are imagined one third of the evaporative losses.

It's also worth to notice that, according to the NETL report, 90% of the cooling tower blowdown is supposed to reach the original water source after being discharged.

- Water supply** It is imagined that the water supply infrastructure included in the reference plant is similar to the NETL IGCC cases. The system is represented in Figure 19. Water is drawn partially from a POTW (publicly owned water treatment plant) and partially from an underground reservoir.

The overwhelming share of water is used, without any particular pre-treatment, to refill the circulating water system from the various losses of mass happening in the cooling tower. A relatively minor stream of water is used to refill the steam cycle and the process water. The water entering the steam cycle has to be demineralized to avoid plugging of the equipment. The pre-treatment is performed with activated carbon filters and ion exchangers in a specific module. Demineralized water is stored in a large tank from which it can be pumped to the condenser or the deaerator.

Figure 19: qualitative block flow diagram of the internal management of water within the plant



Potable water for in-site service consumption is stored in two separate tanks. Black waters are collected and treated in a modular waste treatment unit before being discharged. The tanks for potable water are connected to the fire protection equipment in case of emergency.

Finally, the plant is supposed to be geared with a waste water treatment facility able to process the streams of water exiting the plant in order to guarantee compliance with the EPA standards for suspended solids, oil and grease, pH, and miscellaneous metals. The water treatment facility is supposed to be constituted by the minimal equipment needed for the neutralization of water and the separation of oil residuals. Chemical pollutants might indeed be occasionally present in the sewage of the plant due to various non-routine conditions.

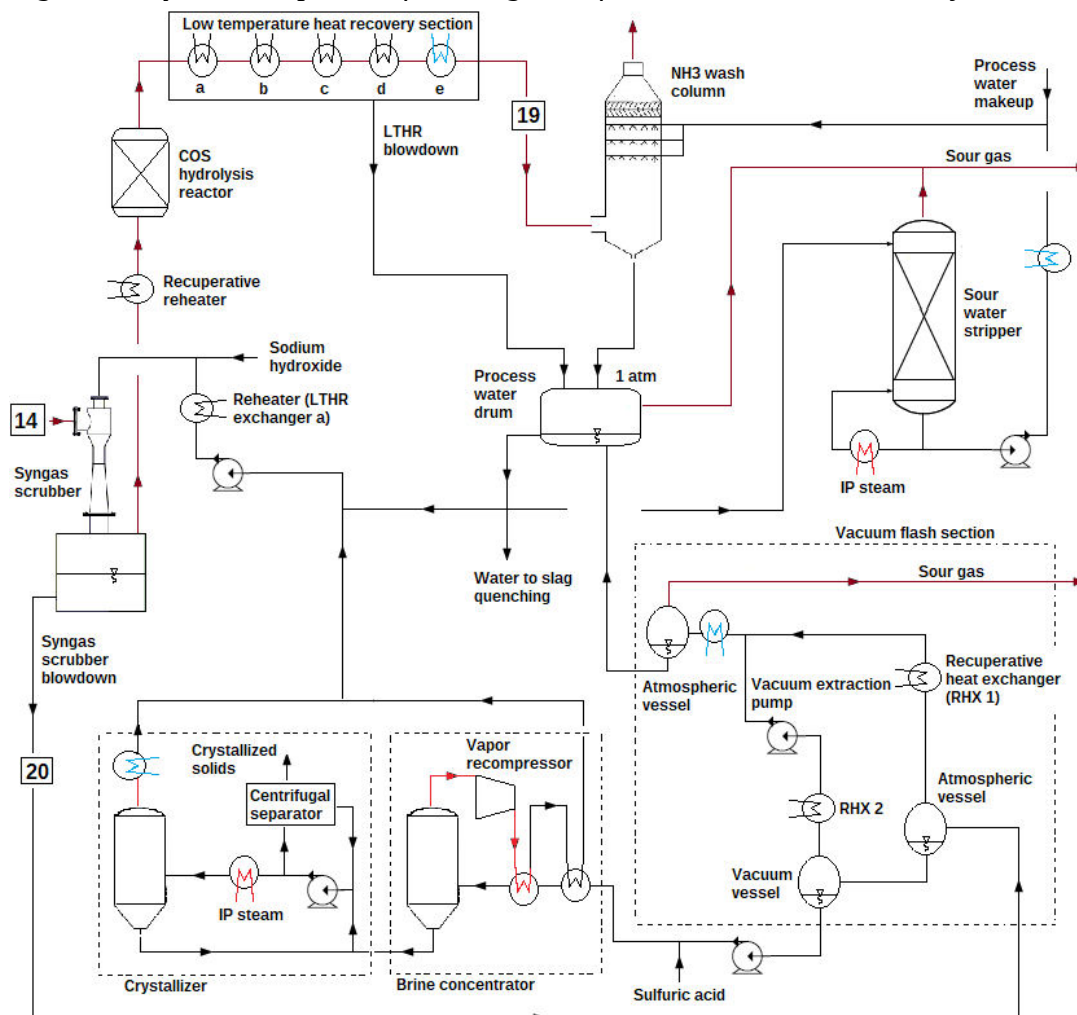
Since the cooling tower is by far the largest consumer of water in the plant, it has been imagined that the electric consumptions necessary to operate the water pumps are roughly proportional to the cooling tower load, with a proportionality coefficient of 0,69 kW per MJ of heat rejected. The coefficient was extrapolated from case 1A of the NETL 2019 report.

- **zero liquid discharge (ZLD) water treatment system**

The purpose of the ZLD system is to continuously collect and regenerate the process water needed for the operations of the syngas treatment train.

The ZLD unit is represented in Figure 20. Waters are collected from the bottom of the syngas scrubber and the NH₃ spray tower. Some condensate is also obtained by cooling the syngas at ambient temperature in the LTHR section.

Figure 20: qualitative process flow diagram of the ZLD water treatment system



The stream of water exiting the NH₃ washing tower is regenerated from dissolved gases in a stripping column. The syngas scrubber blowdown, after being flashed at sub-atmospheric pressure to release eventual dissolved gasses, is instead sent to a series of two evaporators (see section 4.2.5) aimed at crystallizing and separating the various non volatile fine pollutants removed from the syngas.

In particular, the first evaporator is an highly efficient vapor recompressing system that concentrates the solution until near saturated conditions. The second evaporator operates with a liquid-solid slurry, and requires the introduction of a small quantity of intermediate pressure steam in order to function.

The methodology that has been adopted to model the ZLD unit is a simplified one. It has been made the initial assumption that the system is qualitatively similar to the ZLD process that is described in case 1A of the NETL cost and performance baseline report[73]. Capital costs have been then calculated by modifying for a scale factor the capital cost figures provided by the NETL.

The various utilities (steam, electric power, cooling water) required by the ZLD unit are listed in Table 9.

Table 9: utilities that have been associated to the ZLD water treatment system

IP steam extraction for the NH ₃ stripper reboiler	MW(thermal)	7,2	0,522 MJ / cubic metre of syngas passing into the NH ₃ wash column
Cooling water NH ₃ stripper	MW(thermal)	7,2	
Electric power consumption	kW(el)	4870	74,4 kJ / kg of water blowdown from the Scrubber (flow 20)
IP steam crystallizer	MW(thermal)	3,75	0,096 MJ / kg of coal fed into the gasifier
Cooling water crystallizer	MW(thermal)	3,75	
Heat required for process water preheating	MW(thermal)	13,52	
Heat recovery potential from the ZLD unit	MW(thermal)	15,25	The waste heat can be used for water preheating up to 90°C

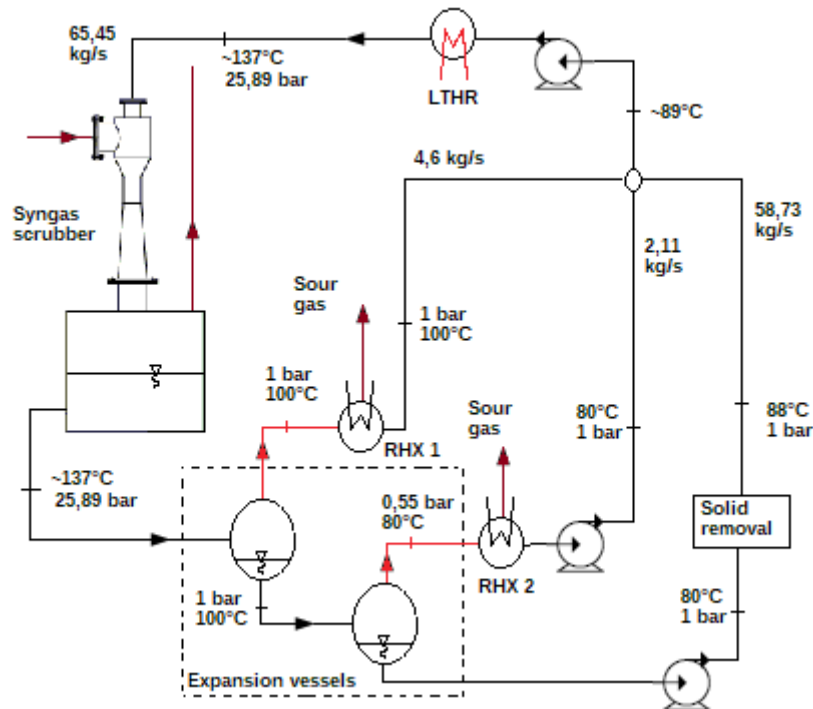
They have been extrapolated from case 1A of the NETL 2019 report through the following process:

1. The electric power consumption figure for water treatment has been imagined to be dominated by the electric power consumption of the vapor recompressor. The amount of vapor recompressed has been treated as proportional to the mass flow of water used in the syngas scrubber (flow 20). The quantity of water used in the syngas scrubber for each cubic metre of syngas has been taken as a constant from the NETL report.
2. The quantity of IP steam used in the crystallizer has been imagined proportional to the amount of crystallized solids produced, which has been in turn assumed proportional to the coal input of the gasifier. The pressure level of the steam extracted has been set to 4,48 bar, following the example offered by the NETL 2019 report.
3. The quantity of IP steam used in the sour water stripper has been taken proportional to the amount of water used in the NH₃ spray tower, which has been in turn assumed proportional to the volumetric flow of syngas (flow 19).
4. The amount of heat rejection through the cooling water circuit has been assumed to be equal to the heat introduced in the two heat exchangers discussed at point 2 and 3.
5. The transformation happening in the syngas scrubber has been modelled as an isoenthalpic humidification. As a consequence, it has been considered that the water entering the syngas scrubber needs to be preheated roughly to the temperature level of the syngas (~137°C).

In order to roughly estimate the amount of heat necessary for this purpose, as well as the potential for heat recovery, it has been built a simplified model of the water circulation around the scrubber. The process flow diagram is represented in Figure 21, and the most important quantitative figures are listed in Table 9. All the calculations have been performed assuming that the scrubbing solution is a stream of pure water.

As a final note, it has to be remembered that the NETL cost and performance baseline report does not assume an isenthalpic humidification transformation in the scrubber. On the contrary, the syngas is supposed to be contacted with relatively cold water. Therefore the hypothesis of similarity between the two systems does not appear to be completely justified. Despite the differences between certain pieces of equipment might not be well caught with a single-parameter model, it has been nonetheless decided to neglect such considerations and imagine that the capital cost of the ZLD unit is a unique, non-divisible figure which depends on the sheer amount of gas treated.

Figure 21: simplified model that has been adopted to represent the ZLD water treatment system

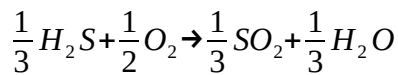


- Sulfur recovery unit.** The sulfur recovery unit is aimed at treating the stream rich in hydrogen sulfide that results from the processing of the syngas in the acid gas removal block. Some amounts of sour gases, primarily ammonia, are also dissolved in the process water and are sent to the sulfur recovery unit after being recovered in the water treatment unit.

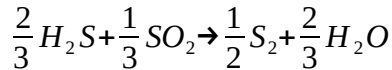
The sulfur recovery unit that has been considered for the reference plant is a three stage air blown Claus process followed by an hydrogenation reactor.

The system is represented in Figure 22.

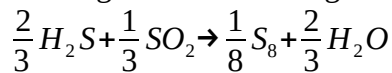
One third of the hydrogen sulfide is first fully oxydated to SO₂ in a furnace, while the rest of the gas is bypassed (the bypass is not shown in figure).



When the two streams are mixed at high temperature they react endothermically to form Disulfur (S₂), although the reaction extent is limited.



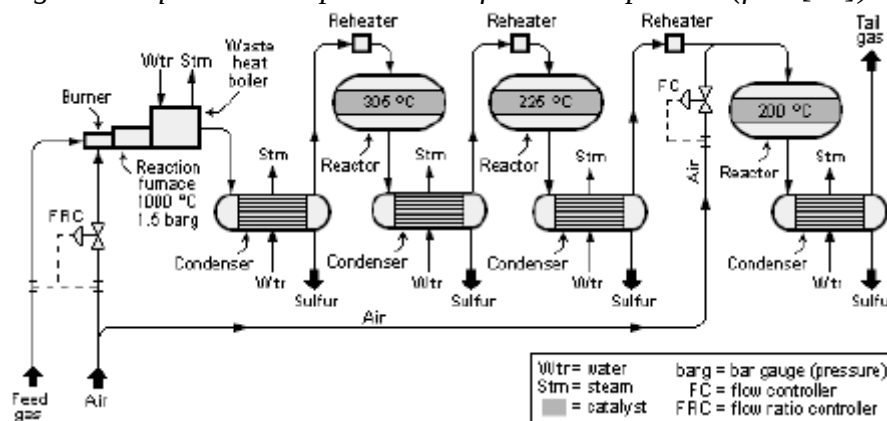
The stream is cooled down to separate liquid sulfur, which is a by-product of the Claus treatment of acid gases, then it is passed into a multistage catalytic reactor where hydrogen sulfide is almost completely converted to sulfur (S₈) in an exothermic way. In particular a three stage reactor should grant a conversion around 98%.



Sulfur is removed between subsequent stages of the reactor in order to keep the thermodynamics of the reaction favourable.

The tail gas exiting at the end of the multistage reactor is finally passed into a further catalytic bed that saturates with hydrogen all the remaining sulfur atoms (often found in the form of COS) to form H₂S. The final product is a stream of inert gases (mainly carbon dioxide, steam and nitrogen) containing also a small amount of hydrogen and hydrogen sulfide.

Figure 22: qualitative representation of the Claus process (from [73])



Being the overall process exothermic, medium pressure steam can be recovered. According to the NETL performance baseline report, it is possible to produce steam at 17 bar. Given the acid gas available it has been roughly calculated that about 8 MW of heat can be recovered from the Claus unit for each gasification train. Such heat is assumed to be used in the reference plant to satisfy the steam requirements of the water treatment system. It has been instead decided not to consider any low temperature heat recovery from the tail gas cooling, which would have been in any case of limited interest. More significantly, while in this analysis it is imagined that the tail gas is ultimately recompressed and recirculated to the syngas flow ahead of the AGR, an accurate study of such inner-loop is omitted. In other words the tail gas stream is ignored in the present discussion. The most important consequence is that there is a slight inaccuracy (about 1%) regarding the overall carbon balance of the plant.

- **Electric infrastructure.** The chief task of the electric equipment is to create an interface between the in-house generators and the external electric grid.

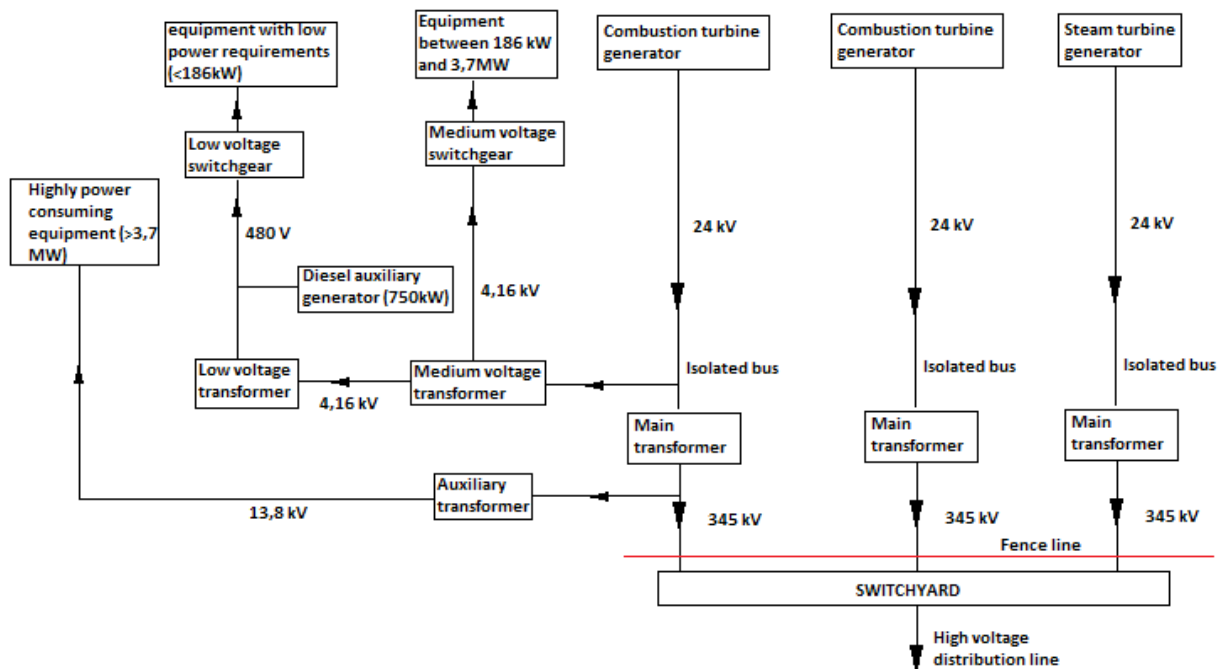
In this analysis it is imagined that each of the generators in the power plant is connected to a transformer raising the current voltage from the generation level (24kV) to the transmission level (345kV). Then it is imagined that a nearby switchyard station connects the plant with the outside grid.

Importantly, this analysis (together with the NETL reports) does not consider the switchyard station as included in the fence line in any of the cases object of study.

The electric infrastructure has also to provide a power supply to all the auxiliary systems requiring an electric connection. For this reason the generator of one of the combustion turbines is linked to a series of additional transformers that make electricity available for internal uses.

A small diesel generator is also installed to provide emergency power for all those critical systems that must be operational even in case of turbine failure (and subsequent power loss).

Figure 23: qualitative chart of the most important electric connections within the plant



- **Coal handling system.** In the cases studied in this thesis coal is supposed to be delivered to the power plant with a dedicated railway. Both the railway and the railway cars are considered external to the perimeter of the plant, and are therefore not accounted in the economic analysis. (On the contrary all truck roadways and unloading stations inside the fence area are supposed to be provided).

Coal is dumped from the bottom of the railcars into a receiving hopper that is purposely placed under the railroad tracks. From there coal is distributed on a belt conveyor that leads to the storage area, where coal is stacked into large outdoor piles by a particular crane called *stacker*.

When it is necessary, coal is loaded into an hopper and routed towards the gasifier with a second array of belt conveyors. Along the way it is also sampled, crushed into smaller grains, and magnetically separated from iron particles.

Ultimately, the conveyor loads the coal into three storage silos not far to the gasifier isle.

The particular piece of equipment used to unload the conveyors is called *tripper*. The flow of coal exiting at the bottom of the various hoppers is instead controlled by a machine called

coal feeder. In Appendix F are provided some images of the components constituting the coal handling system.

In table 10 the specific electric consumption figure that has been adopted for coal handling is listed, together with other subsystems that will be discussed in the next points.

Table 10: specific electric consumptions associated to miscellaneous functions

	Unit of measure	Value
Coal handling	KJ/kg coal	50
Slag handling	KJ/kg slag	100
BOP* auxiliaries	% of coal LHV	0,15

*Includes plant control systems, lighting, HVAC, and miscellaneous low voltage loads

- **Coal preparation & feed.** From the storage silos coal is loaded on a belt conveyor that brings it to the gasifier isle and drops it in a single large hopper with two outlets (one of each gasifier). Coal is subsequently pulverized and dried at the same time in a rotary mill. The drying is made possible by the extraction of a small stream of syngas before the CT. The syngas is burned with air in an incinerator, and the exhaust gases, before being vented in the atmosphere, are sent to the rotary mill to perform the drying operation. Pulverized coal is ultimately stored in a lock hopper with 2 hours of operation capacity, and sent to the alimentation system of the gasifier.

It is important to point out that the syngas extraction has been neglected in the analysis of the syngas production train. In case of an air blown gasification plant, a likely estimate of the required syngas extraction is around 2-4% of the total amount directed towards the CT.

- **Slag handling.** The purpose of the slag handling system is to recover and separate the spent material at the bottom of the gasifier. In this thesis it is assumed that the slag handling system employed in a MHI gasifier is qualitatively similar to the one employed in a Shell gasifier, which is described in the NETL baseline report.

In detail, fluid ashes slowly flow down the walls of the gasifier until they are quenched into a water bath in the lower part of the vessel. From there, batches of a slag / water slurry are extracted from the pressurized chamber by means of a lock hopper system. The slurry is then dewatered and the material is crushed into minor fragments. Slag is finally stored into a series of bins with a capacity of 72 hours of full load operation.

It's worth noticing that, according to the description of the process supplied by the NETL 2019 report, water employed in the slag handling block is assumed to be clarified and recycled, however no precise explanation is offered regarding the equipment necessary for this operation, or the energy consumption which is necessary. It is also not clear if such operation is performed in the waste water treatment facility.

In this analysis it has been decided to adopt the same approach employed in the NETL 2019 report, and not to associate any disposal or treatment cost to the water used for slag quenching.

- **Acid gas removal.** The function of the AGR unit in the case analyzed is to separate the sulfur compounds from the syngas ahead of the combustion turbine. As already mentioned, it is imagined that the system is constituted by an absorption and a regeneration column, and that the solvent employed is mono-diethanol-amine (MDEA). The capital cost of the AGR block has been evaluated using case 4A of the NETL 2019 report as a reference.

Table 11: utilities associated to the Acid gas removal unit

	Unit of measure	Value
AGR specific reboiler duty	MJ/kg H ₂ S	26,22
Total AGR reboiler duty	MW(th)	36,708
Heat rejection from the AGR	MW(th)	36,708
AGR Electric Auxiliaries	MJ/kg H ₂ S	1,11

The specific steam consumption in the reboiler has been set to 26,22 MJ of steam per kg of H₂S treated. It is also assumed that the totality of the heat introduced in the reboiler is then removed through the cooling water circuit.

Finally it needs to be noted that the composition of the flow of acid gases exiting the regenerator has been set according to the indications from [73] and [81]. A summary of all the significant figures is provided in Table 11.

- **Air separation unit.** The function of the air separation unit in the reference plant is to generate the nitrogen necessary to load the dry lock hoppers fueling the gasifier. In this analysis the ASU is supposed to be a cryogenic distillation system operating with two separate columns at different pressure. This technology is the one adopted in the NETL cost and performance baseline reports to model the ASU for all IGCC case studies, therefore the reader is referred to the NETL report[82] for a puntual description of the unit. Interestingly, the specific electric consumption of the ASU described in the NETL report is identical to the number used in the paper from Bonalumi et. al[75]. In other words there seems to be a particularly sound consistency between the main sources utilized when it comes to the modelling of the ASU. The NETL reports offer a slightly more complex analysis in that oxygen recovery rate is put at 98% instead of 100% (like assumed in [75]). For each kilogram of oxygen recovered the NETL report also takes into account process steam requirements roughly equal to 102 grams of steam at 17,2 bar. Little error is committed neglecting both of these non idealities.

In second place, since the overwhelming majority of the electric auxiliaries of the ASU is constituted by the main air compressor, which performs an intercooled compression, it has been decided to use the electric power consumption figure as a proxy for the cooling load required by the ASU.

Finally, nitrogen needs to be further compressed before entering the lock-hoppers, and the NETL report is ambiguous regarding which cost item the capital cost of the nitrogen

compressor belongs to. It has been made here the choice of treating the capital cost of the nitrogen compressor as accounted in “Asu & oxidant compression”.

The auxiliary electric consumption figure that has been adopted for this item was sourced from the article of Bonalumi et. al [75]. Again, being the compression intercooled, it is supposed that the amount of heat rejection required by the nitrogen compressor is equal to the electric power consumption. A summary of all the figures just discussed is presented in Table 12.

Table 12: utilities associated to the Air separation unit

	Unit of measure	Value
Asu specific electric consumption	kWh/ton O2	420
ASU electric consumption	MW(el)	10,5
Heat rejection needed in the ASU	MW(th)	10,5
Electric power for the nitrogen compressors	MW(el)	10,2
Heat rejection from the nitrogen compressors	MW(th)	10,2

- Air booster.** The function of the air booster is to elevate up to 33 bar the pressure of the gasification air from the CT compressor discharge. Lacking further indications, it is assumed also that the air booster is a centrifugal single-stage compressor. A total electric consumption of 14,9 MW for each gasification train was considered, based on the indications from [75]. The polytropic efficiency of the compression was set at the value of 90,5%. Preceding the air booster it is placed a natural circulation boiler where high pressure steam is produced cooling the air exiting the CT compressor. In detail, a potential for 6,8 MW of thermal recovery was estimated for each train.
- Other subsystems.** Besides what already discussed, the project is supposed to be granted with an integrated, microprocessor-based, plant-wide, distributed control system. The control system is designed to be available 99.5% of the time and it automatically monitors/controls all the major plant equipment. In the control room it is provided an interface for the operators to manually act on the process. The reader is referred to [73] for a more thorough description of the control equipment. In addition, for each gasification train it is taken into account the capital cost for a natural gas connection pipeline, a small secondary boiler, and a flare stack meant to depressurize the system in relative safety if needed. Because some small amounts of compressed air are needed in the plant for the functioning of various mechanical components (such as pneumatically actuated valves), some equipment for this purpose is also included. As a side note, compressed air requirements might be further divided into service air and instrument air requirements. *Instrument air* is a term meant to define compressed *dried* air, while *service air* does not need to be dried.

3.4 Performance indexes of the plant

While in the previous pages the different sub-units have been discussed in a relative separate way, in this section are briefly presented some important indicators to characterize the whole power plant system. In Table 13 it is presented a prospect for the total auxiliary electric consumptions.

Table 13: auxiliary electric consumptions summary

	MW el
Coal milling & handling	3,91
Slag handling	0,68
Steam cycle pumps	6,45
ASU auxiliaries	21
Cooling tower auxiliaries	8,769
Lock hopper compressor	20,4
Acid gas removal auxiliaries	2,95
BOP miscellaneous*	2,912
Air booster	29,8
Water treatment auxiliaries	4,87
Steam turbine auxiliaries	0,332
Groundwater pumps	0,525
CT generator loss & auxiliaries	4,6
Total	107,198

In Table 14 the demand for cooling water is broken down in its different components.

Table 14: summary of the cooling load required to the cooling tower

	MW(cooling water)
Condenser & Gland steam condenser	661,95
Acid gas removal	73,26
ASU Main compressor & Lock hopper compressor	41,4
LTHR	9,69
ZLD water treatment unit	10,95
Total	797,25

In Table 15 it is presented a summary of the performance of the plant in terms of net power production and electric efficiency.

Notably, the steam turbine gross power output is 3-4% lower than the figure provided in the paper [75] from Bonalumi et. al. This is mainly a consequence of the fact that it has been considered an higher level of the pressure in the condenser (0.068 vs 0.04 bar), with the consequence that the power output of the LP turbine is reduced.

Table 15: plant performance summary

	Value
Coal thermal input, MW	1941,39
CT expander (2 trains)	1049
CT compressor (2 trains)	495,8
Steam turbine gross power	485,89
Heat rejected at the Condenser	652,76
Gross electric power output, MW	1039,09
IGCC plant auxiliaries	107,198
CCS plant auxiliaries	0
Net power output	931,89
Net electric LHV efficiency %	48
Specific emissions kg co2 / Mwh	660,22

Finally, in Table 16 it is listed a series of parameters that have been used, as will be explained hereafter, to infer the capital cost of the different functional areas and sub-systems.

The parameters are listed both for the power plant object of study, and for case 1A from the NETL 2019 report, which was used as a benchmark.

Table 16: parameters used to characterize the reference power plant in comparison to the power plant represented in Case 1A of the NETL 2019 cost and performance baseline report

	Unit of measure	Case 1A NETL Oxygen Blown IGCC	air blown IGCC plant (reference plant)
Coal feed rate	kg coal /sec	54,86	78,2
Power exchanged in the evaporators of the gasification isle	MW th	201,53	220,68
Scrubbing water	kg H2O /sec	18,137	65,45
Gas flow in the scrubber	kg syngas /sec	99,75	344,4
AGR volumetric flow*	m3 syngas /sec	3,93	13,8
H2S to claus	kg H2S /sec	1,54	2,817
Slag production	kg Slag /sec	19,781	24,49
Boiler feedwater (HP circuit)	kg water /sec	197,458	306,59
HRSG heat load	MW th	557,77	719,82
Circulating water flow rate	Metric tons H2O /sec	11,58743	17,32
Raw water withdrawal	m3 H2O/ min	15,28	22,84
Process water discharge	kg H2O /sec	58	86,7
Cooling tower load	MW th	533,3	797,25
Gas flow to stack	kg exhaust /sec	1117,73	1330
Syngas flow to the CT unit	kg syngas /sec	94,18	161,2
Steam turbine power	MW th	301	485,89
CT net power	MW el	464	548,6
Condenser exchange area	m2	33568	46961,3
Total plant gross power	MW el	765	1039,09
Total auxiliary load	MW el	125	107,198
Net Power output	MW el	640	931,89

3.5 Economic evaluation: methodology and results of the reference case

3.5.1 Description of the financial framework of the project

The principal metric used in this report to characterize the economic competitiveness of the power plants investigated is the LCOE. The LCOE is calculated using the same set of hypothesis adopted in the 2019 NETL baseline report, which are briefly summed up in Table 17.

In particular it has been considered a 5 years period of construction (from January 2018 to December 2022) and an operational lifetime of 30 years. No inflation rate has been considered in the analysis. Moreover, it has been made the simplifying (although not very realistic) assumption in the calculations that the capacity factor of the plant assumes the constant value of 80% over its entire lifetime, and there is no performance deterioration or increase in operative costs.

Table 17: financial parameters used to develop the economic analysis

Real interest rate on equity	7,84%	Capital expenditure distribution	10%(2018), 30%,25%, 20%,15%(2022)
Real interest rate on debt	2,94%	Depreciated Total overnight capital	100,00%
Equity fraction	45,00%	Ammortization plan	20 years, 150% declining balance See [83] IRS 2016 Document 946 table A14
Debt fraction	55,00%	Capital expenditure period	5 years (2018-2022)
Effective Tax Rate	25,74%	Operational period	30 years (2023 onwards)
WACC	5,14%	Real inflation (or escalation) rate	0% for all the analysis
ATWACC	4,73%	FCR (fixed charge rate)	7,07%

Given the assumptions considered, the expression to calculate the LCOE can be simplified and reduced to the following formula:

$$LCOE = \frac{FCR \cdot TASC + OC_{FIX} + CF \cdot OC_{VAR}}{CF \cdot MWh_{YEAR}}$$

where:

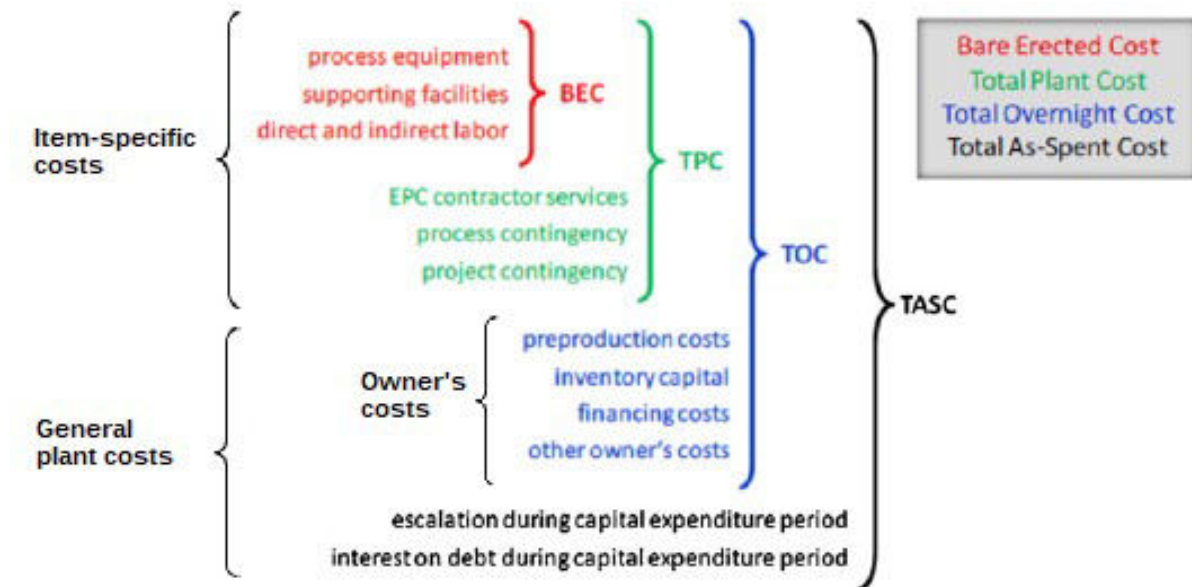
- TASC stands for Total As-Spent Capital. It equals the present value of all the costs that have to be sustained to build and launch the project.
- The FCR (Fixed Charge Rate) is a factor that converts the total capital value (TASC) to a uniform annual amount (annuity) that has to be paid flatly each year during the lifetime of the plant to compensate debtholders, shareholders and taxes (on profits). It has been calculated according to the assumptions described in Table 17 with the equations described in the document [74] of the NETL.
- CF stands for Capacity Factor.
- OC stands for Operative Costs. Operative costs may be variable or fixed. Variable costs include fuel cost, and they are proportional to the actual operational load of the plant, that is to the Capacity Factor.

In the next paragraphs it is detailed how the capital and operative costs have been estimated.

3.5.2 Capital costs

The Total-as-Spent-Capital cost (TASC) has been determined with a bottom-up method employing the same terminology adopted in the 2019 NETL baseline report. In Figure 24 it is provided a graphic illustration of the way in which the various cost items are organized and aggregated.

Figure 24: breakdown structure of the capital costs considered in the economic analysis (image from [68])



The starting point of the analysis consists in a list of components and equipments. The complete list of the equipments considered is provided in appendix D-1.

- Each piece of equipment is associated to a Process equipment cost. This figure stands for the gross amount to be paid to the manufacturer of the equipment, including all applicable sales taxes and delivery charges. It is the unitary and undecomposable block in function of which all other capital costs are then expressed.

The cost of equipment has been generally calculated by means of the following expression

$$\text{equipment cost} = \left(\frac{X_{\text{airblown}}}{X_{\text{ref}}} \right)^n \cdot (\text{cost}_{\text{ref}})$$

where $X_{\text{ref}}, X_{\text{airblown}}$ are values of a proper performance parameter that is specific to the equipment considered.

The performance parameters that have been chosen for the different equipments are shown in the first two columns of the Table D-1, which was placed in Appendix D due to space limitations.

The values $X_{\text{ref}}, X_{\text{airblown}}$ used to actually calculate the costs are listed in Table 16.

Case 1A from the NETL Cost and Performance Baseline Report has been used as a data point for most items so to obtain a reference value for the parameter X_{ref} . Values for the scaling exponent n have been gathered for the various items from the document [84] issued by the NETL as an annex to the Cost and Performance Baseline Report.

The ratio $\frac{X_{\text{airblown}}}{X_{\text{ref}}}$ is here called *adimensional scaling ratio* for the purpose of convenience.

The values calculated for the adimensional scaling ratio are listed in table D-1 in appendix D. The results of this methodology are instead presented in Table D-3.

An alternative approach has been adopted to estimate the equipment cost of some items, which are highlighted in Table D-1 in the appendix D. The method, which is customized for every item, is described in the appendix B.

In addition to the Process equipment cost, for each item in the list of the components of the plant it has been considered also a series of other item-specific costs. In this category there are:

- Costs necessary for the additional supporting facilities required to install the equipment. It is expressed as a linear function of the process equipment cost by means of a coefficient that is item-specific. The values for the coefficients are estimated from case 1A in the NETL 2021 report. For example the 2019 NETL report at pag 148 considers a cost of materials for account 5.3 which is equal to 19,5% of the process equipment cost.
- Labor costs directly or indirectly necessary to complete the installation of the equipment. It is expressed as a linear function of the process equipment cost with an item-specific coefficient.
- Engineering, procurement and construction fees to be paid to the contractor appointed for the realization of the plant. This capital cost item is expressed as a linear function of the so-called *Bare Erected Cost*, which is a term used to characterize the sum of the costs for process equipment, supporting facilities and construction labor. The value of the coefficient is 15% for all equipments.
- Process contingencies. Coefficients for process contingencies have been set to zero for the most part of the components of the plant, which are generally considered to be well known and already exploited at commercial level.
A process contingency of 5% of the BEC has been instead set for the Combustion Turbine Generator working with a low LHV syngas. A process contingency of 14% has also been set for the syngas coolers since the currently commercial coolers generally operate at lower temperatures than 1200°C.
- Project contingencies. Project contingency costs have been calculated for each equipment item as a fraction of Bare Erected Cost. The coefficients adopted are in the range 15-35% and they are the same employed in the 2019 NETL baseline report (Case 1A).

All the capital costs associated to the various items are finally aggregated in a sum that is defined *Total Plant Cost (TPC)*.

For the reference plant a detailed breakdown of the Total Plant Cost is provided in Table D-3 in Appendix D.

A different set of capital expenditures is constituted instead by the so called *Owner's costs*.

Owner's costs are costs that the owner has to pay in order to start the project. They include the cost for preliminary studies, the commissions paid to the financial intermediaries in order to obtain the financing, costs related to legal fees and permitings, land purchase, investments required to potentiate roads and railways nearby the power plant.

A certain sum has to be spent in advance to fill the inventories with fuel and spare parts for maintenance.

Owner's costs also include some months of labor and material costs which are required to train the plant operators in the period prior to the startup of production.

Finally, a fund has to be allocated to take into account possible delays and unplanned events under the responsibility of the owner (so called "owner's contingencies").

Owner's costs do not include the interconnection with the electric grid nor site improvements of unusual character. Taxes on capital are also not included.

Owner's costs have been determined according to the directive issued by the NETL in the document [74], and they are listed in Table 20 (on the left) at the end of the chapter in addition to the Total Plant Cost and other relevant metrics.

Owner's costs and Total Plant Cost (TPC) form the Total Overnight Cost (TOC), which equals the monetary amount that has to be paid for all the necessary components and activities for the plant.

In addition, being assumed that the project requires 5 years of construction before completion, it has to be also accounted an economic cost associated with the presence of fixed non-active capital during construction. This cost consists in a loss of owed financial interests on the immobilized capital, and it is equal to 15,4% of the TOC if it is calculated with the set of financial assumptions listed in Table 17.

The Total As-Spent Capital (TASC) is formed by the sum of the Total Overnight Cost (TOC) and these financial losses incurred during construction.

3.5.3 Variable and fixed costs

Fixed operative costs are constituted by the sum of property taxes and labor costs. Property taxes have been expressed as a function of the Total Plant Cost (TPC), while the labor cost has been estimated from the labor requirements of the power plant and the 2018 real salaries for a Midwest location. The assumptions and the calculations are reported in Table 18. The results are displayed in Table 20 (on the right).

Table 18: parameters used to determined the Fixed Operative Costs and the Cost associated to maintenance materials

hourly wage (generic midwest location)	38,5 \$/hour	Cost for maintenance labor	1,05% * TPC
operators in the power plant per each shift	16	Administrative labor	0,25*(operative +maintenance labor)
shift duration	8 hours	Property taxes	2% * TPC
number of shifts	3 shifts / day	Maintenance materials	1,95% * TPC
operating labor burden	30,00%		

Variable operative costs are incurred due to the purchase of coal and other production-related inputs, as well as the disposal of waste streams. The cost for maintenance materials has been also accounted among variable costs in this study. It is expressed as a function of the TPC (see Table 18).

Besides coal, the most relevant input of the reference power plant is water, which is consumed in large quantities in the cooling tower. Some chemicals are also used in more or less significant amounts. Table 19 illustrates the various consumables that are employed.

Table 19 also shows the disposal costs, where the most significant cost items are related to the disposal of the slag and the solids from the crystallizer. Other waste streams are of secondary relevance.

Apart from coal, all input and waste streams have been quantified using a scaling approach. For each stream an appropriate scaling parameter has been chosen, and the researched mass flow has been then estimated using case 1A as a benchmark.

For example it has been judged that the amount of sodium hydroxide added into the scrubbing water to control the PH should be more or less proportional to the amount of coal that is fed into the gasifier. The reason is that coal typically contains small quantities of Chlorine that tend to be then dissolved in the scrubbing water. Therefore it is recommended by the NETL reports the mixing of sodium hydroxide in the scrubbing water in order to avoid pitting of the metal at the bottom of the scrubber.

Being the coal input in the power plant here analyzed 42,5% higher comparatively to case 1A, the amount of sodium hydroxide necessary for one day of full load operation of the reference plant has been supposed to be also 42,5% greater, as shown in Table 19.

Table 19: summary of the variable costs (maintenance materials excluded). From top to bottom: costs associated to miscellaneous consumables, fuel costs, disposal costs

	price	scaling Variable	scaling Parameter	unit of measure	quantity employed in the CASE 1A NETL every day at full load	Quantity employed every day in the air blown IGCC plant	Cost \$ /year
Water consumption	1,9 \$/10 ³ gallons	raw water Withdrawal	1,495	10 ³ gallons	2971,44	4442,3028	2464589,5934
Makeup and Wastewater treatment chemicals	550 \$/Ston	raw water Withdrawal	1,495	Short ton	8,85	13,23075	2124858,45
Sodium Hydroxide (50%wt solution)	600 \$/Ston	coal feed rate	1,425	Short ton	18	25,65	4493880
COS hydrolysis catalyst	1300 \$/ft ³	coal feed rate	1,425	ft ³	0,923	1,315275	499278,39
MDEA solution	2,8 \$/gallon	coal feed rate	1,425	gallon	38,3	54,5775	44622,564
Claus catalyst	48 \$/ft ³	h2s to claus	1,822	ft ³	1,81	3,29782	46222,24512
Sulfuric acid (98%wt solution)	210 \$/Ston	coal feed rate	1,425	Short ton	0,32	0,456	27961,92
							9701413,1626

	price	scaling Variable		unit of measure		Quantity employed every day in the air blown IGCC plant	Cost \$ /year
Coal	45,83	coal feed rate		Short ton coal	-	7447,667904	99667373,052

	unitary cost of disposal	Scaling variable	unit of measure	scaling Parameter	Quantity produced in the case 1A NETL every day at full load	Quantity produced in the IGCC Air blown	Cost \$ /year
slag disposal	38 \$/Ston	slag output	Short ton	1,238	523	647,474	7184371,504
crystallizer sol	38 \$/Ston	coal feed rate	Short ton	1,425	34	48,45	537601,2
claus catalyst	2,5 \$/ft ³	h2s to claus	ft ³	1,822	1,81	3,29782	2407,4086
mdea solution	0,35 \$/gal	coal feed rate	gallon	1,425	38,3	54,5775	5577,8205
cos hydrolysis	2,5 \$/ft ³	coal feed rate	ft ³	1,425	0,923	1,315275	960,15075
							7730918,0839

Disposal costs and unitary prices for all input excluded coal have been kept constant with respect to the figures used in the 2019 version of the NETL cost and performance baseline report. Coal price has been adjusted with the respect to the NETL report volume 1 (Bituminous coal) in order to take into account the different properties of coal. In particular in this report it is considered a quality of coal with a relatively high content of sulfur, which is generally regarded as an undesired species; therefore the price of coal includes a discount factor. The adjustment has been calculated according to the NETL directives issued in the annex documents [85][86].

Table 20: Owner's costs (on the left), Fixed Operative Costs (top right), LCOE results (down right), summary of the capital costs (down on the left)

Description	k\$
Pre-Production Costs	
6 Months All Labor	29585,29
1 Month Maintenance Materials	7800,28
1 Month Non-Fuel Consumables	1010,57
1 Month Waste Disposal	805,31
25% of 1 Months Fuel Cost at 100% CF	2595,51
2% of TPC	76802,75
Total	118599,71
Inventory Capital	
60-day supply of fuel and consumables at 100% CF	22473,04
0.5% of TPC (spare parts)	19200,69
Total	41673,73
Other Costs	
Initial Cost for Catalyst and Chemicals	2618,2
Land	900
Other Owner's Costs	576020,6
Financing Costs	103683,71
Total	683222,51
Total Plant Cost	3840137,29
Total Owner's Costs	843495,91
Total Overnight Costs (TOC)	4683633,2
TASC Multiplier (IOU, 35 year)	1,154
Total As-Spent Cost (TASC)	5404912,71

Description	k\$/year
FIXED OPERATING COSTS	
Annual Operating Labor	7015,01
Maintenance Labor	40321,45
Administrative & Support Labor	11834,12
Property Taxes and Insurance	76802,75
Total	135973,33

Maintenance Materials	74882,68
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Component	Value, \$/MWh
Capital	58,51
Fixed	20,83
Variable	14,14
Fuel	15,27
Total LCOE	108,75
CO2 T&S	0
Total LCOE (Including T&S)	108,75

3.5.4 Discussion of the results of the economic analysis on the reference case

In Table 20 it is represented the LCOE figure determined for the reference plant. In Figure 25 and Figure 26 the results are broken down and compared with other cases (without carbon capture) provided by the NETL in order to contextualize them and facilitate the reading.

It can be noticed, first of all, that the cost of electricity is largely constituted by financial interests that need to be set aside to pay back the capital used to build the plant. Then, it can be noticed how capital costs (TASC) are constituted by a large number of items, the most relevant ones being the cost of the gasification reactor, the financial losses during construction and the so-called *other owner's costs* needed to modernize the local infrastructure.

Figure 25: breakdown of the LCOE in its components (left), results of the reference plant compared with the LCOE of various alternative technologies (right)

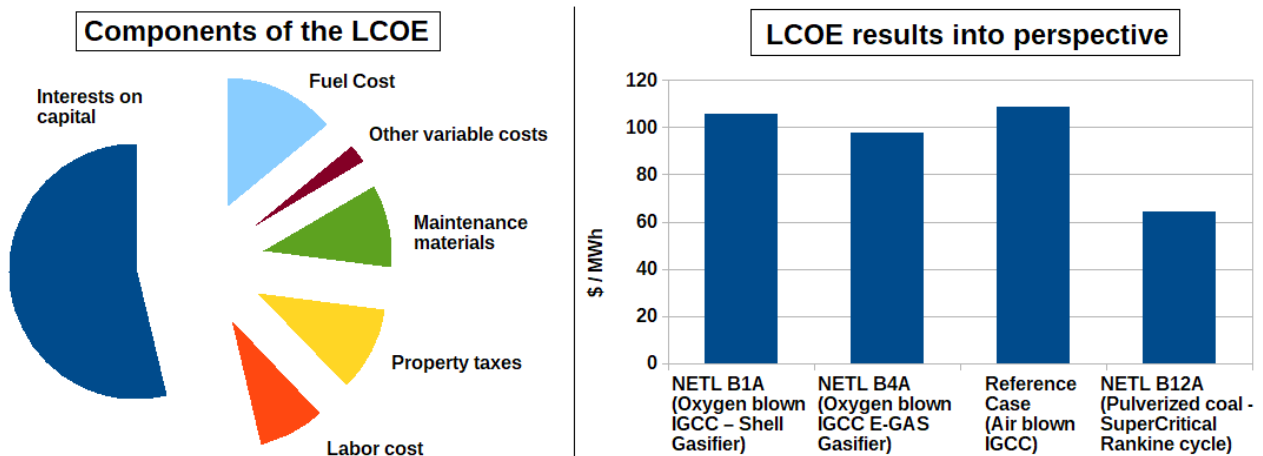
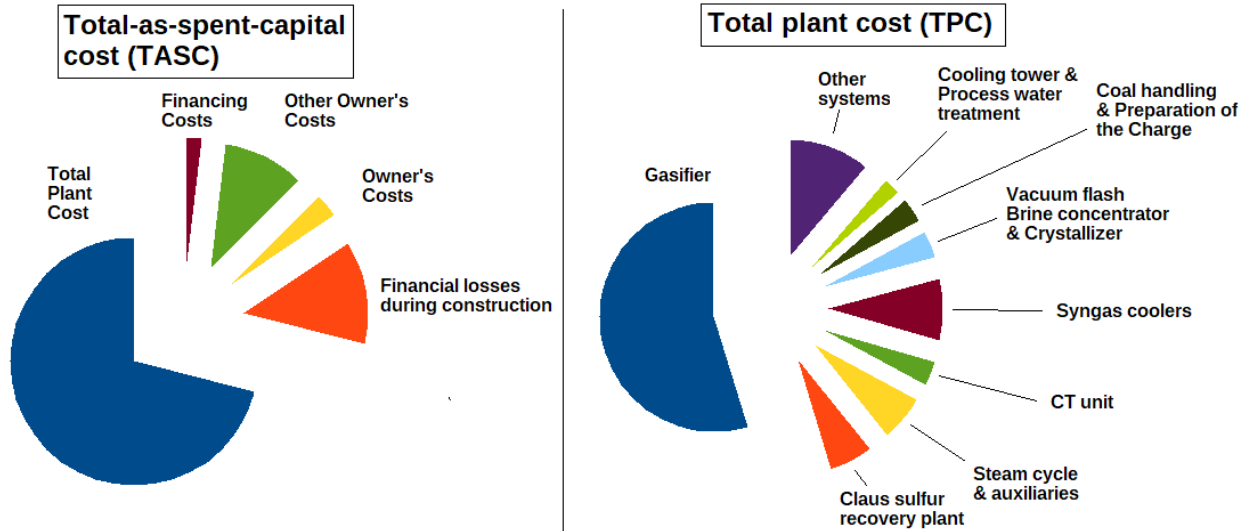


Figure 26: breakdown respectively of the TASC (on the left) and the TPC (on the right)



A comparison of the reference plant with the best available technology (Pulverized SC coal) shows that the cost of electricity produced with the air-blown IGCC plant is not competitive.

Furthermore, the reference plant appears less convenient from an economic point of view if it is compared with two other IGCC plants based on oxygen-blown gasification technologies.

In particular, while the air blown gasification technology is more efficient from a thermodynamic perspective due to a drastic reduction in the auxiliary consumptions required from the ASU, it is at the same time associated to a considerable increase in the capital costs necessary to build the

gasification train, and, more specifically, the gasification reactor. This is a consequence of the fact that the gasification train must be able to produce and treat a far bigger volumetric flow of syngas.

The modellization developed in this study indicates that the efficiency gains, which are in the order of some percentage points, are outweighed by the additional capital costs that the reference plant requires in comparison with the other solutions. However this conclusion should be taken with caution given some of the limits of the methodology adopted. More in detail:

- The various plants considered in Figure 25 (on the right) are smaller than the reference plant, and therefore they feature a lower amount of scale economies.
- The modellization of reference plant introduces some approximations and neglects some cost items which are instead considered in the NETL studies. This is the case for example of the auxiliary electric consumptions required from the Claus unit, or the extraction of syngas required from the coal dryer.
- The single most important capital cost item in the reference plant is constituted by the air-blown gasification reactor, which accounts for more than half of the equipment cost according to the model developed. Unfortunately however the model for the gasifier is highly dependent on the choice of the parameter n used to describe the scale economies for the price of metal needed to create the reactor vessel (see appendix B). As a consequence the analysis is not very robust and there is a non-negligible change in the results if slightly different assumptions are employed. For example the LCOE figure estimated would increase by 5% if it was assigned the value of 0.7 (instead of 0.6) to the n parameter of the gasifier. It is in particular this instability problem that severely affects the comparisons of the reference plant with systems involving a different gasification technology.

Given the uncertainty regarding the real cost of the air-blown gasifier, the accuracy (in absolute terms) of the present study is compatible with a Class 4 / Class 5 analysis according to the AACE classification, with the implication that a series of further investigations is needed specifically on the gasification reactor if the accuracy of the cost estimate has to be improved.

On the other hand, the gasification train is a piece of equipment which is shared without differences in all the configurations involving carbon capture that are studied in Chapter 4. Therefore, despite its shortcomings, the model developed for the gasifier does not affect the kind of technological comparison between different CO₂ capture systems which is the main objective of this work.

4 DESCRIPTION OF THE POWER PLANT CONFIGURATIONS WITH CARBON CAPTURE

The purpose of this section is to present the different power plant configurations involving carbon capture that have been studied in this thesis. It is also reported the methodological framework through which the cases with carbon capture have been analyzed and modelled, both from a physical and economical point of view.

The chapter is split into two parts: in the first section it is provided an high-level picture of the main characteristics and differences between the various cases and the associated assumptions.

In the second part the sub-systems constituting the CO₂ capture unit (or strictly integrated to it) are discussed in a separate manner, highlighting the most relevant performance indicators, as well as other aspects of technical interest.

4.1.1 General classification of the cases with carbon capture

A total of seven different configurations of power plant are discussed. For the sake of simplicity, they will be sometimes referred to as numbers from zero to six.

The first arrangement (Case number 0) is the air blown IGCC reference plant previously discussed. The two most important features of the reference plant, in comparison to the other cases, are the fact that it doesn't involve carbon capture, and that it is characterized by the presence of a Claus-type sulfur recovery unit in which hydrogen sulfide removed from the syngas is converted into solid sulfur.

The remaining configurations all involve the coupling of an air blown IGCC plant with a post combustion CO₂ capture unit. Three different carbon capture technologies have been taken into consideration. A detailed discussion of such technologies will be provided in the next section. The capture rate has been set at 90% for all the cases, and carbon dioxide, after being separated from the exhausts, is supposed to be compressed to supercritical pressure and outsourced to a third party storage facility which is imagined to exist in relative proximity.

It is then possible to group the cases analyzed also according to a second criterion: the treatment of the hydrogen sulfide produced in the AGR block.

A first cluster of cases can be identified which is similar to the reference plant, in that hydrogen sulfide is converted to sulfur in a Claus Unit and sold to the market.

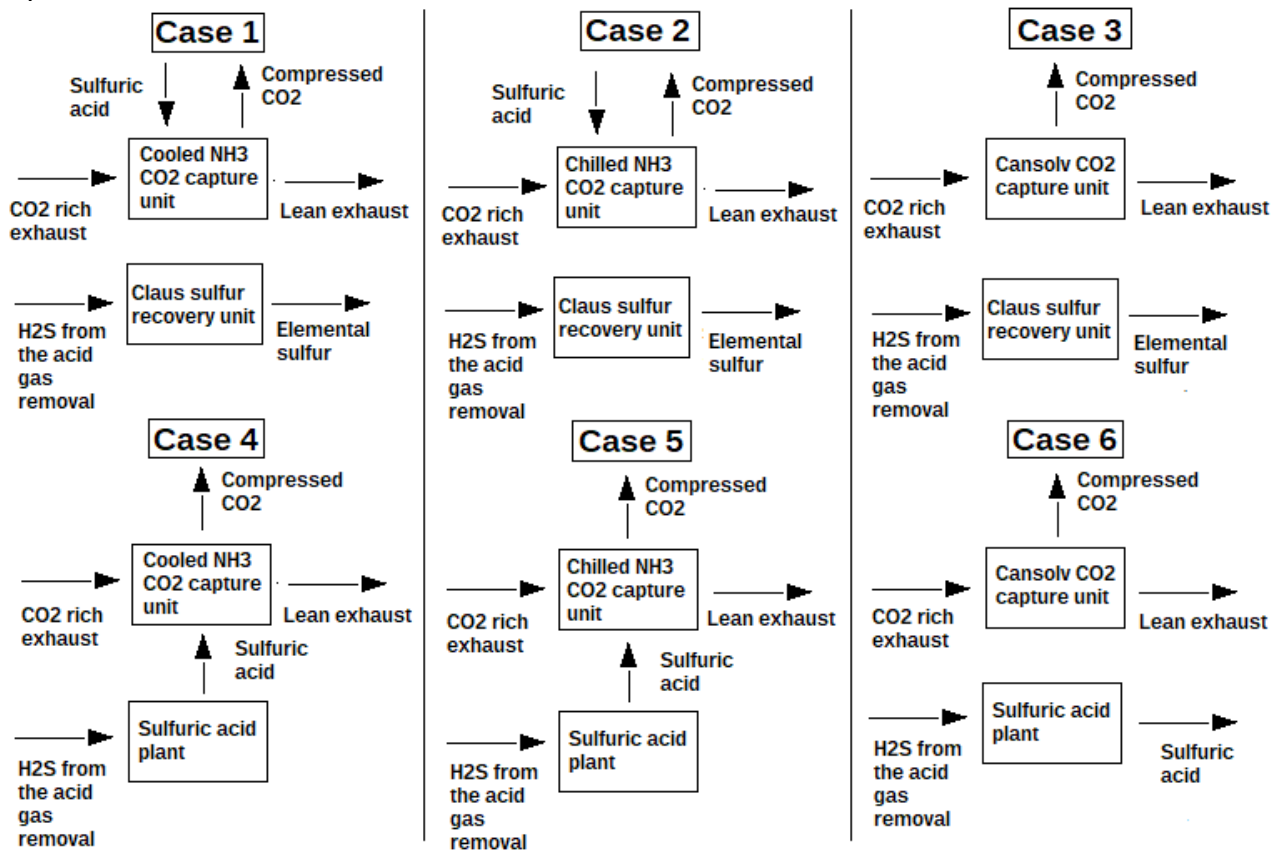
Since sulfuric acid is required for the functioning of some of the post combustion CO₂ capture units, a second group of cases has also been investigated in which hydrogen sulfide is converted into sulfuric acid in a specific isle called *sulfuric acid plant*. Sulfuric acid is ultimately employed in the process or sold as a by-product.

These various features distinguishing the capture cases are schematically represented in Figure 27.

- In the Cases 1 and 2 sulfuric acid is purchased from the market and employed in the CO₂ capture section for the operation of acid wash (that will be discussed in the next chapters). The by-products of the acid wash operation are sold to the market; at the same time elemental sulfur, which is produced in the Claus-type sulfur recovery unit, is also sold to the market.
- In Case 3 there is no need for sulfuric acid in the CO₂ capture unit. Elemental sulfur is produced in the Claus unit and sold to the market.
- In the Cases 4 and 5 sulfuric acid is produced in the sulfur recovery unit and employed within the CO₂ capture block for the acid wash operation. The by-products of the acid wash operation are then sold to the market. Sulfuric acid might be also sold, if it is in excessive supply.

- Case 6 is similar to Case 3, with the difference that sulfuric acid is sold to the market instead of elemental sulfur.

Figure 27: summary of the main features distinguishing the power plants investigated with carbon capture



4.1.2 Gasification island

For what concerns the gasification island of the power plant, no particular changes have been taken into consideration between the various cases involving carbon capture. Some noteworthy difference have been instead considered in the carbon capture cases with respect to the reference case.

In particular, since high pressure carbon dioxide is a product of the capture section, nitrogen can be replaced by carbon dioxide as inert gas employed in the lock hoppers.

For this reason the Air Separation Unit has been removed altogether in the capture cases, while it has been considered a greater extraction of compressed air from the CT unit towards the gasifier in order to offset the lack of oxygen enrichment.

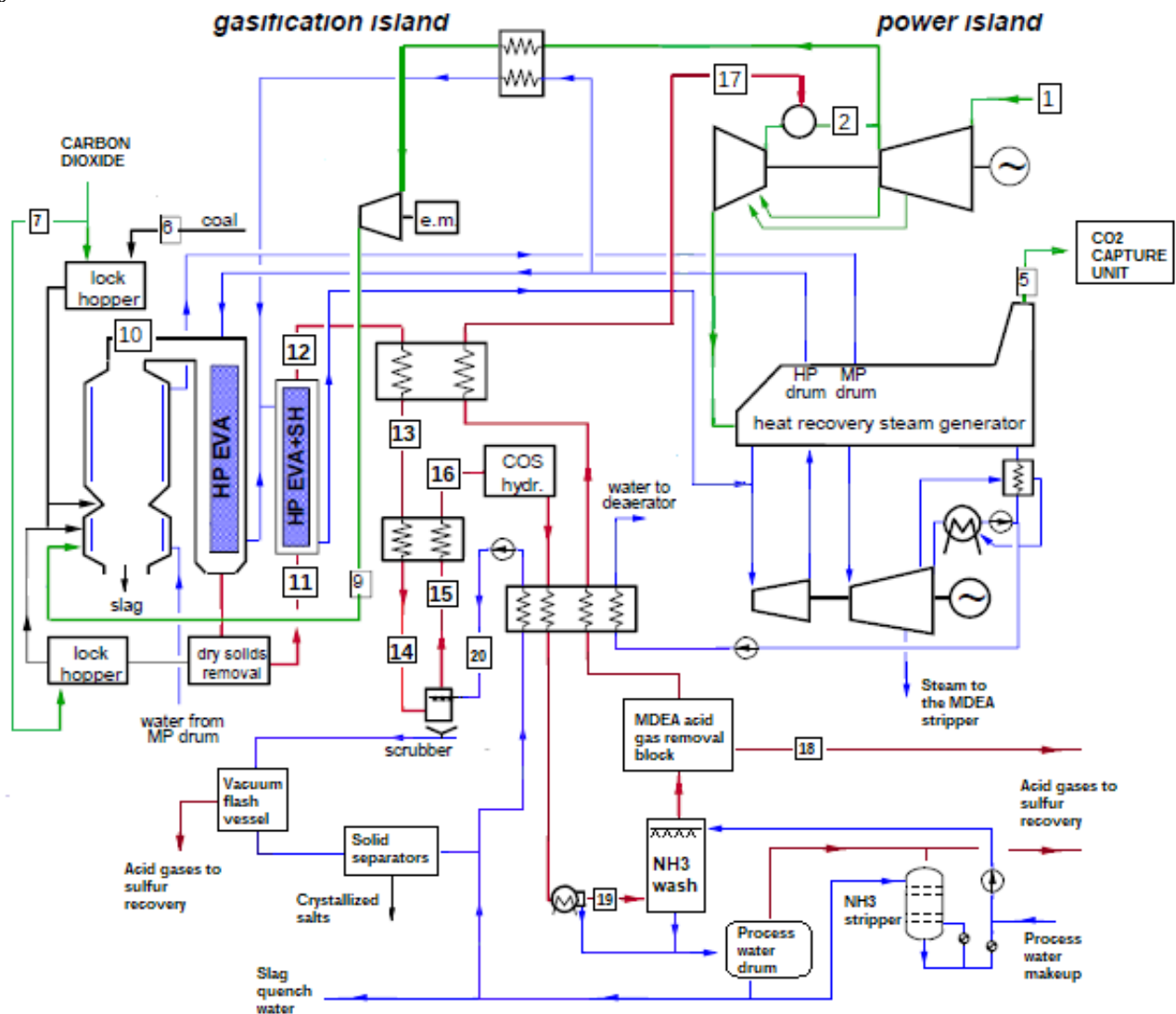
The mass flow and the composition of the syngas at the gasifier outlet have been also set to different values than the reference case. This is a consequence of the fact that a non negligible additional quantity of nitrogen enters in the system with the gasification air.

An important implication is the fact that the net power output of the combustion turbine is slightly lower, while a greater quantity of heat can be recovered in the syngas coolers. The LHV of the syngas at the combustion turbine inlet is also slightly decreased, being the syngas relatively more diluted with nitrogen.

The overall electric consumption from the auxiliaries tends to diminish in the carbon capture cases, since the increase in the load of the air boosters is more than compensated by the elimination of the ASU and the lock hopper compressors.

The gasification island and CT unit of the capture cases have been modelled from a quantitative point of view according to the information provided in the article from Bonalumi et. al [75]. In detail, the data on the performance indicators of the CT unit and the air booster have been directly resumed from that source. For what concerns other physical variables, such as the composition of the syngas, the aforementioned article has been used as a track, but it has been necessary to add some new hypothesis in order to obtain a result.

Figure 28: schematic of the IGCC plant when it is integrated with the post combustion CO2 capture system



In particular, while the composition of the syngas should theoretically be calculated as a result of the conditions within the gasification reactor, in this analysis it has been evaluated through an heuristic method:

- The ratio between the molecules of Hydrogen and Carbon monoxide at the gasifier outlet has been fixed to an arbitrary value, which has been chosen in order not to be too dissimilar

from the reference case. The same procedure has been applied to the ratio between the molecules of Methane and Carbon monoxide.

- Given the above-mentioned hypothesis, the composition of the syngas has been determined by fixing all the remaining molar compositions in such a way that the various material and energy balances mentioned in the article are respected.

A similar method introduces some error in the calculation of the specific heat capacity of the syngas, however it has been concluded that the error is negligible for the purposes of the present work.

It has been determined that, based on the auxiliary load of the air booster, the mass flow of the syngas must increase roughly 10% with respect to the reference case.

Finally, it has been concluded that the potential for heat recovery increases of 10% and 30% respectively in the high temperature syngas coolers and in the air-driven boiler. This increase in the heat recovery finds its counterpart in the following facts:

- the coal input is 1% higher in the capture cases
- the cold gas efficiency of the gasifier is slightly lower (from 73,03% to 72,5%)
- more work is spent to compress the air entering the gasifier

Drawing on all such considerations, it has been possible to produce a process flow diagram (represented in Figure 28 and Table 21) of the gasification island for the carbon capture cases.

Table 21: Temperature, pressure, flow rate and composition of the main streams of the plant shown in Figure 28

	Temperature °C	Pressure Bar	Mass flow (kg/sec)	Ar	CO	CO2	H2	H2O	H2S	N2	O2	CH4
1	15	1,01	628,64	0,92	0	0,03	0	1,03	0	77,28	20,73	0
2	417,6	18,16	363,64	0,92	0	0,03	0	1,03	0	77,28	20,73	0
5	115	1,01	665	0,89	0	9,97	0	4,36	0	75,1	9,68	0
6	15		39,5	Coal as received (%wt: 61.27 C, 4.69H, 8.83 O, 1.1 N, 3.41 S, 12 moisture, 8.7 ash)								
7	80	56,13	10,5	0	0	100	0	0	0	0	0	0
9	477,4	33,02	140,1	0,92	0	0,03	0	1,03	0	77,28	20,73	0
10	1200	28,06	187,67	0,62	24,1	6,63	10,92	4,21	0,58	52,44	0	0,5
11	900	28,06	185,27	0,62	24,1	6,63	10,92	4,21	0,58	52,44	0	0,5
12	350	27,5	185,27	0,62	24,1	6,63	10,92	4,21	0,58	52,44	0	0,5
13	206,6	26,95	185,27	0,62	24,1	6,63	10,92	4,21	0,58	52,44	0	0,5
14	137	26,41	185,27	0,62	24,1	6,63	10,92	4,21	0,58	52,44	0	0,5
15	112	25,89	185,27	0,6	23,7	6,51	10,74	5,82	0,57	51,56	0	0,5
16	180	25,37	185,27	0,6	23,7	6,51	10,74	5,82	0,57	51,56	0	0,5
17	250	23,16	176,46	0,64	25,41	6,13	11,51	0,46	0	55,2	0	0,53
18*	45	1,01	3,93	0	0	58,36	0	0	41,64	0	0	0
19*	35	~24,5	180,4	0,63	26,33	3,77	12,4	0,23	0,63	55,4	0	0,58
20*	137	25,89	72	0	0	0	0	100	0	0	0	0

4.1.3 Secondary sub-system

Within the scope of the secondary sub-systems the following changes have been considered with respect to the reference case:

- The flow of water used in the syngas scrubber increases proportionally with the syngas flow (+10%). The electric auxiliaries of the ZLD water treatment system tend to increase proportionally to the flow of water used in the syngas scrubber. No changes have been considered concerning the quantity of steam needed in the crystallizer.
- It has been considered an increase of 10% in the amount of heat necessary to preheat the scrubbing water. A 10% increase has also been considered regarding the potential for heat recovery in the exchangers of the vacuum flash unit.

- The quantity of water necessary to wash the syngas prior to the AGR has been supposed to increase proportionally to the syngas (+10%). As a consequence, it has been assumed that a similar variation must also affect the quantity of steam necessary for the NH₃ stripping column.
- It has been taken into account a 1% increase in the auxiliaries associated to the operations of coal handling, slag handling, and acid gas removal. A 1% increase has been also considered regarding the BOP auxiliaries.
- The auxiliaries associated to the operations of water supply and heat rejection from the cooling tower have been supposed to increase proportionally to the cooling tower load, which is different in each of the capture cases.

4.1.4 Steam cycle

The recuperative steam cycle has been modelled in a different way in each of six the capture cases analyzed in this work. The precise process flow diagrams of the various cases are all reported in appendix C, which the reader is referred to.

In the next paragraphs it is provided a brief description of the general characteristics and the main differences among the cases.

Preheated water enters the HRSG at a pressure of 2 bar and a temperature variable between 85°C and 105°C. After being brought near saturation, water is compressed by the feedwater pumps and split into a medium and an high pressure circuit.

Water from the medium pressure circuit, after being preheated in the HRSG economizers, is sent to a series of boilers that are dependent on the specific case analyzed:

- In the Cases 1, 2 and 3 medium pressure water is sent to the gasifier water walls, where it is evaporated. The steam is then superheated in the HRSG.
- In the Cases 4, 5 and 6 water from the medium pressure drum is sent also to the boiler of the sulfuric acid plant. Consequently, more water flows in the medium pressure circuit in these cases.

Similarly to the reference case, water from the high pressure circuit is mostly sent to the gasification island to be evaporated and superheated, although some water is also evaporated and superheated within the HRSG, according to the amount of remaining heat available.

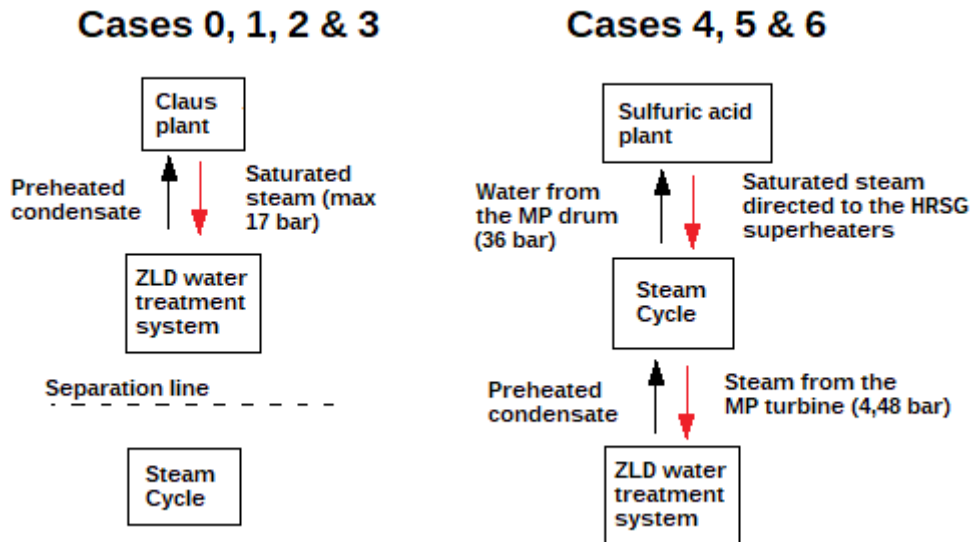
As already mentioned in the previous pages, the potential for heat recovery from the gasification island is roughly 10% higher in the capture cases.

Steam is expanded in the various turbines in a similar fashion to the reference case, although the number and the entity of the steam extractions is different:

- In all the carbon capture cases steam is extracted in great quantity to regenerate the solvent for the CO₂ capture operation. The pressure of these steam extractions is dependent on the specific CO₂ capture technology employed.
- In all the cases steam is extracted at 2 bar to regenerate the solvent for the acid gas removal operation. In the cases 3 and 6 the quantity of steam necessary is slightly lower, since it is possible to recuperate some steam at the appropriate pressure by expanding the flow of condensate from the reboiler of the CO₂ capture unit.
- In the Cases 1 and 4 a secondary steam extraction is necessary from the LP turbine in order to satisfy the process requirements of the CO₂ capture unit.
- In the Cases 4, 5 and 6 a steam extraction at the pressure of 4,48 bar is necessary from the medium pressure turbine to satisfy the process requirements of the ZLD water treatment system.

The ZLD system requires a certain quantity of steam in all the cases analyzed in this work, including the reference case, however in the Cases 0, 1, 2 and 3 the requirements are completely covered by the Claus unit, without the need of extractions from the bottoming cycle. In Figure 29 the differences between the various configurations are graphically illustrated.

Figure 29: different types of connection with the sulfur recovery unit



At the end of the turbine expansion steam is discharged in the condenser, whose size is relatively small since a great quantity of steam is sent to the CO₂ capture unit and it is condensed in the reboiler.

In this analysis it has been made the hypothesis that there are different pumps and different preheating lines for the streams of condensate exiting the main condenser and the reboiler of the CO₂ capture unit. The different flows are then supposed to be mixed in the de-aerator vessel or prior to their entrance in the HRSG.

The stream of water exiting the main condenser is supposed to be preheated up to roughly 90°C being passed in the LTHR syngas coolers and the recuperative heat exchangers from the vacuum flash section. It has been calculated that the quantity of heat available for this purpose is far greater than the amount actually needed.

The condensate from the reboiler of the CO₂ capture unit does not need to be preheated before entering the HRSG, made exception for the Cases 2 and 5. In these cases is imagined that there is a process of preheating from 73°C up to nearly 110°C in the upper section of the LTHR syngas coolers. Notably, in the other cases (Case 1, 3, 4 and 6) it has been calculated that there is about 12,5 MW of unexploited potential for heat recovery in an interesting temperature range within the LTHR syngas coolers.

In the Cases 4, 5 and 6 it has been considered in the calculations that the stream of condensate from the ZLD unit needs to be expanded from 4,48 bar to 2 bar before being dumped in the de-aerator vessel. Since the water is near saturation, a small quantity of vapor tends to form during the expansion. It has been imagined that vapor is condensed in a specific heat exchanger.

4.1.5 Notes on the methodology employed for the economic analysis

As a general note, the economic analysis has been performed in the capture cases with the same set of assumptions and criteria adopted in the reference case. In the next lines a couple of noteworthy differences are highlighted:

- the capital cost of most of the equipments (the complete list is present in Table D-6 in appendix D) has been extrapolated from the NETL report case 1A with the same scaling expression discussed for the reference case. $equipment\ cost = \left(\frac{x_{CCS}}{x_{NETL}}\right)^n \cdot (cost_{NETL})$

It is however important to point out that the adimensional scaling ratio $\frac{x_{CCS}}{x_{NETL}}$ tends to

change significantly in the capture cases for some items, like the syngas coolers or the main condenser for example.

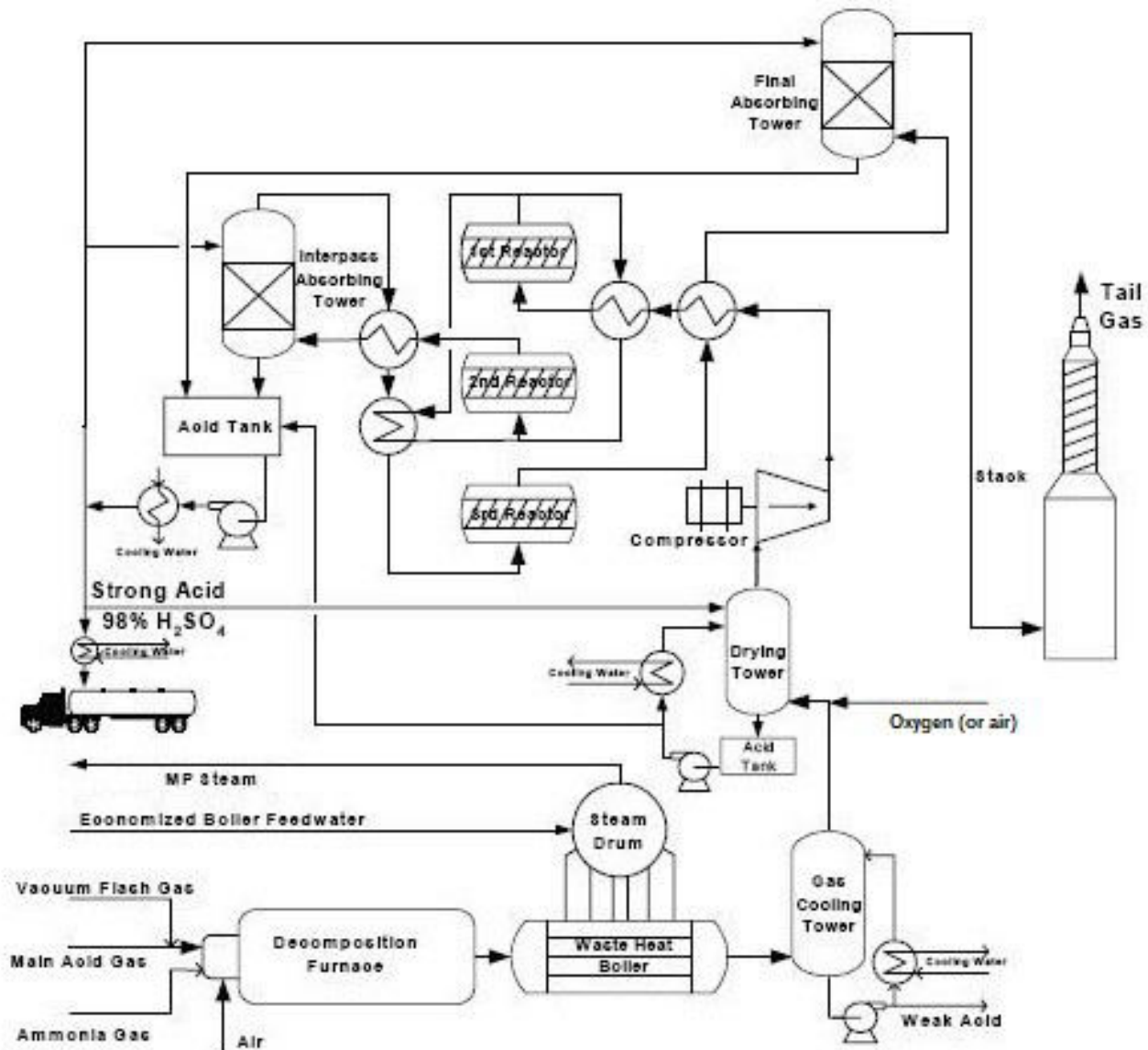
- No differences have been considered with respect to the reference case regarding the capital cost of the air blown gasifier and the LTHR syngas cooling heat exchangers. It has been instead considered a 10% increase in the cost of the syngas-driven superheaters, and the regenerative heat exchangers of the gasification train. The equipment cost of the air booster has been increased by 30%, reflecting the change in the load required to this component.
- Among the capital and variable costs some additional items must be considered which are associated to the new sections of the power plant. These costs are described in the next pages.
- The cost of operative labor in the capture cases has been calculated considering that the number of operators-per-shift increases from 16 to 17. No differences in the cost of operating labor have been considered between the configurations with carbon capture. In Chapter 5 this point will be raised again and it will be shown that, even if there are small differences in the labor force required in the different cases, the impact of labor costs is negligible in comparison with other factors.

4.2 Process areas typical of the CO₂ capture cases analyzed

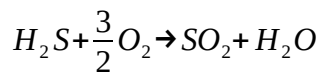
4.2.1 Sulfuric acid plant

The sulfuric acid plant is a system whose purpose is to treat the sour gases produced by the AGR and ZLD blocks, and produce liquid sulfuric acid out of the hydrogen sulfide available. It has been possible to identify a least one case study (Tampa electric IGCC[87][89]) of existing large scale IGCC (oxygen blown) power plant operating with a sulfuric acid process. The process flow diagram of the Tampa power plant is represented in Figure 30 and commented in the next paragraphs. In the next pages it will be also described, both on a qualitative and quantitative basis, how the configuration of the Tampa plant might be adapted to the air blown IGCC plant studied in this thesis.

Figure 30: qualitative schematic of the sulfuric acid plant in the Tampa IGCC plant (from [87])



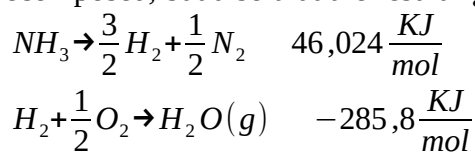
The first step of the sulfuric acid process is a refractory-lined incineration furnace where hydrogen sulfide is oxidized to sulfur dioxide at near atmospheric pressure through the introduction of air:



The incinerator is adiabatic, therefore the temperature at the outlet of the incinerator is regulated by the amount of inert species admitted in the reaction chamber. In particular, since the temperature of the incinerator has to be as high as 1200°C in order to decompose ammonia, it has been calculated that an excess of air equal to 40% should fit the characteristics of the flow from the AGR unit.

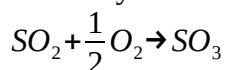
Hot gases are then cooled down in a water tube boiler to produce medium pressure (36 bar) steam. The amount of steam that can be produced in this way depends on several factors, the most important of whom is the temperature of the gas at the outlet of the tube boiler. In this work it has been considered a temperature of 320°C, which allows an heat recovery of 16,5MW for each train based on the amount of hydrogen sulfide available from coal.

Another factor that might affect the overall result is the potential for heat recovery from ammonia decomposition. According to the technical report of the Tampa station ammonia is indeed converted in the furnace to nitrogen and water vapour, implying as a consequence not only that ammonia is decomposed, but also that the resulting hydrogen is then oxydized to water vapour.



Since the overall reaction is highly exothermic, according to the coal characteristics it could be possible to significantly increase the heat recovery if all ammonia was to be converted according to this reaction path. However no precise indication in this direction was found in the surveyed literature, therefore any quantitative description of the behaviour of ammonia in the decomposition furnace has been neglected in this work.

Sulfur dioxide at the outlet of the incinerator, in order to be converted into sulfuric acid, needs to be further oxydized to sulfur trioxide

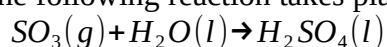


This transformation is accomplished in a fixed bed catalytic reactor through which the sulfur dioxide-rich stream is circulated after being heated to the temperature of 400°C. The catalyst employed is Vanadium Pentoxide.

According to the data gathered from the academic literature [88], the molar concentration of sulfur dioxide in the stream entering the reactor should be around 9-10%, while oxygen should be supplied with a 30-40% excess with respect to the stoichiometric minimum. It has been verified that, based on the coal quality considered for this report, a similar composition of the acid gas might be easily achieved without the need of oxygen enrichment.

Being the reaction exothermic, there is a tendency for the driving force of the transformation to progressively disappear as the temperature increase, and the reactants are converted into products. As a consequence, in the Tampa plant the reactor is split into three beds with intermediate cooling in order to keep the temperature within an optimized range.

Sulfur trioxide is ultimately transformed into liquid sulfuric acid in a bubble column reactor where the following reaction takes place:



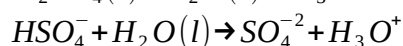
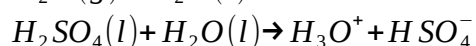
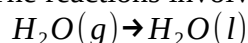
In detail, the column is filled with an aqueous solution of sulfuric acid at 98% weight concentration of acid. The gaseous SO₃-rich stream coming from the catalytic reactor is then

insufflated from the bottom of the column, which generates turbulence in the liquid and facilitates the interphase contact.

Water is continuously added to keep the concentration of the solution constant, while non-reacted gases are extracted at the top of the column.

Two bubble columns are usually employed. The first column is placed between the second and the third catalytic bed in order to remove sulfur trioxide and maximize the conversion in the last section of the catalytic reactor. The second column is placed at the outlet of the last catalytic bed. Residual gases (tail gases) are ultimately emitted to the atmosphere by means of a secondary stack. According to the data sheet of the Tampa plant, such system such grant a conversion of sulfur dioxide into sulfuric acid equal to 99,5%.

Before entering the catalyst bed, a fan is included to let the gas flow into the subsequent components. Moreover, since water is poisonous to the catalyst, it is necessary to pass the flow of acid gas and air into a drying stage before entering the multistage reactor. This is done in a drying column where sulfuric acid at 98% (wt) is employed as dehydrating agent. The reactions involved in the column are:



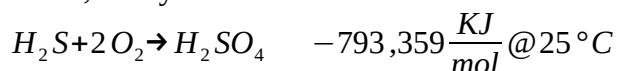
Additionally, the process flow diagram of the Tampa plant depicted in Figure 30 features a contact cooler placed before the drying and compression stages.

Considering that cooling implies an exergy loss, and that the gas needs to be reheated to 400°C before entering the reactor, the role of the contact cooler is not completely clear. It is believed that a stage of cooling was introduced to facilitate the operations of drying, which is exothermic, and compression.

Furthermore, it is not clear why a contact cooler was preferred on a more conventional shell and tube heat exchanger. It is believed that this choice was related to reliability and control considerations. On the other hand, the use of a contact cooler implies the production of a weakly acidic solution that is effectively a waste stream. It has been decided not to count this stream in the disposal costs since little information is available, and there seems to be a potential for avoiding this cost by employing a different type of exchanger.

In Figure 31 it is represented a preliminary block diagram for the sulfuric acid plant that might be adopted in the capture cases studied.

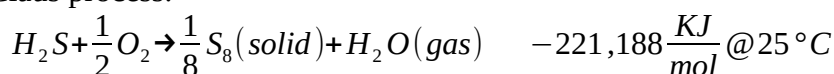
Overall, the system realizes the transformation:



It should be noted that no additional water needs to be added to the system other than the amount necessary to keep the outlet concentration of sulfuric acid at 98% wt.

It has been evaluated that a total of 32,6 MW of thermal power is produced by the exothermic reactions realized in the various components, based on the amount of hydrogen sulfide supplied by one gasification train (ammonia decomposition is ignored, as already discussed).

This figure is more than three times higher than the amount of heat recoverable from an equivalent Claus process:



It has been verified that, in the specific case investigated, a sulfuric acid process allows an heat recovery of 16,5MW downstream to the incineration furnace; the steam is produced at 36 bar.

Conversely, a Claus plant allows an heat recovery of 7,79 MW at the relatively modest pressure of 17 bar, which cannot be easily employed for power production.

A remaining amount of heat equal to 14,72MW can be recovered from the sulfuric acid tanks at lower temperature. Although no specific quantitative analysis were carried on to sustain such hypothesis, it is believed this amount of heat might be suitable to preheat water in the 30-100°C temperature range.

In this work it has been imagined that cooling water is employed to remove heat from the sulfuric acid.

If compared with the Claus-type sulfur recovery plant studied in the reference case, the system represented in Figure 31 has an important difference: the desulphurized tail gas is emitted from a secondary stack instead of being recompressed and reinjected into the syngas.

Figure 31: simplified block flow diagram used to model the sulfuric acid plant

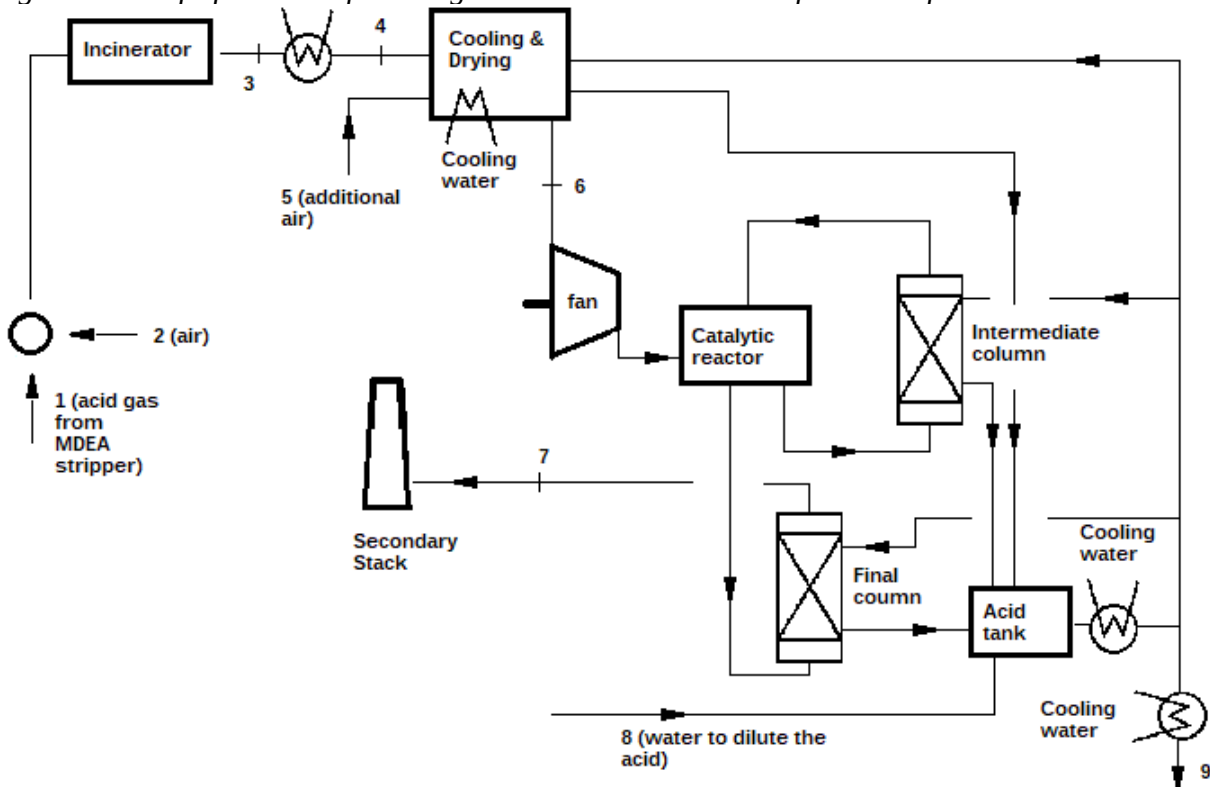


Table 22: Temperature, flow rate and composition of the main streams of the sulfuric acid plant shown in Figure 31 (the pressure of the streams has to be considered closed to the atmospheric one)

	Temperature °C	Mass flow (kg/sec)	O2	N2	H2S	CO2	H2O	SO2	SO3	H2SO4
Stream 1	45	3,93	0	0	41,64	58,36	0	0	0	0
Stream 2	15	11,85	21	79	0	0	0	0	0	0
Stream 3	1210	15,78	5,04	66,38	0	11,78	8,4	8,4	0	0
Stream 4	320	15,78	5,04	66,38	0	11,78	8,4	8,4	0	0
Stream 5	15	0,42	21	79	0	0	0	0	0	0
Stream 6		16,2	5,51	66,75	0	11,43	8,16	8,16	0	0
Stream 7	115	12,18	1,78	83,85	0	14,35	0	0	0	0
Stream 8	15	0,08	0	0	0	0	100	0	0	0
Stream 9	60	4,11	0	0	0	0	10	0	0	90

This is relevant because it means that the post combustion CO₂ capture unit is effectively bypassed by a fraction of the carbon dioxide, therefore lowering the carbon capture rate a couple percentage points below 90%.

It might be possible to recirculate the tail gas towards the CO₂ capture unit in order to emit it from the primary stack, however this option is not considered in this work. The interaction between the CO₂ capture solvent and the sulfur dioxide in the tail gas should be first studied in particular.

In Table 23 the most important utilities associated to the sulfuric acid plant are shown for the cases investigated in this report. The catalyst consumption and the auxiliary electric consumption have been scaled according to the indications provided in a 2013 article [88] where a similar (although not identical) sulfuric acid plant to the one sketched in Figure 31 is thoroughly described.

Table 23: inputs and outputs associated to the sulfuric acid plant

	unit of measure	value	comments
Auxiliary electric Consumption	kW(el)	0 (1779,47)	60 kWh/ton(H ₂ SO ₄) From [88]
Water consumption (1 train)	kg/sec	0,082	water is only used to dilute the acid at 98% weight conc
Cooling Load (1 train)	MW(th)	14,722	
Catalyst initial fill	Metric tonnes	937,06	1,316 tons of catalyst per each ton of acid produced In a day (from [88])
Catalyst consumption	Metric Tonnes/year	46,853	5% of Initial fill (from [88])
Sulfuric acid production (98% wt) (2 gasification trains)	Metric Tonnes/day	711,791	

CAPITAL COST & OTHER COSTS

The capital cost of the sulfuric acid plant has been estimated according to the empirical equation

$$C_2 = C_1 \left(\frac{X_2}{X_1} \right)^{0.6}$$

where:

C₁ = 7 000 000 \$(1989) ; X₁=300 short tonnes of sulfuric acid production per day (100% wt) while C₂ and X₂ are the capital cost in 1989 dollars and the daily production in short tonnes respectively. The value of X₂ was set at 784,6 short tonnes / day. It was therefore imagined that there is a unique sulfuric acid plant servicing both the gasification trains.

The equation has been sourced from Sinnott, Coulson & Richardson's Chemical Engineering Volume 6 (Chemical Engineering Design), 1999. The result has been then adjusted to 2019 dollars according to the CEPCI index.

The Bare erected cost has been calculated by multiplying the equipment cost by 1.48, which is the same coefficient used in the NETL report for the Claus plant. It has been imagined that the fees for the contractor are 15% of the bare erected cost, while the project contingencies are 50%, since there are some unresolved uncertainties regarding the scope of the system.

It has been considered a market price of 5 \$/lb for the catalyst[90].

4.2.2 Cansolv CO₂ capture unit

The Shell-Cansolv process is a term used in reference to a family of solvents for post combustion capture developed by Cansolv Technologies Inc. (controlled by Shell since 2008) exploiting as CO₂ capture agent a nondisclosed mixture of water and amines. In particular the solvent DC-103 has been first developed and tested from 2004 to 2008, while the solvent DC-201 is second generation solvent with improved properties, and has been developed from 2009 to 2012.

Notably, a solvent developed by Cansolv has been adopted in the BoundaryDam project [91], which is the largest post combustion CO₂ capture project currently in operation.

In Figure 32 it is represented the basic configuration of the process as it is described from the producer [91].

In general the absorber is supposed to be placed in series to a pre-scrubber unit whose task is to drastically reduce the sulfur dioxide content in the gas. Sulfur dioxide in particular has a tendency to react with the Cansolv solvent and degrade it over time creating heat stable salts that do not participate in the CO₂ capture cycle and cannot be regenerated [92].

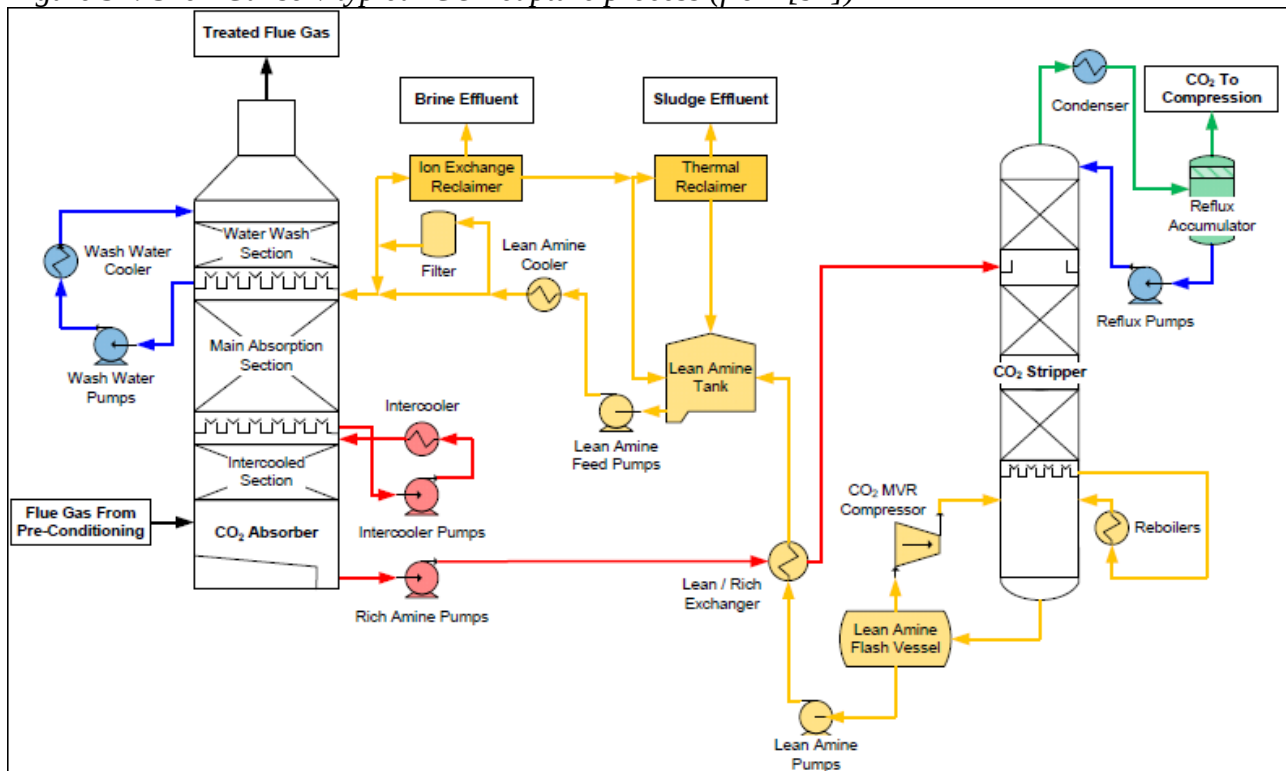
According to the information available, sulfur dioxide needs to be around 2 ppmv at the absorber inlet [73]. A pre-scrubber solvent that might be used to this end is caustic (NaOH).

The BoundaryDam project, which operates with a coal-fired boiler, actually involves not only a pre-scrubber unit for the absorption of SO₂, but also a stripper column that regenerate the solvent separating the sulfur dioxide [93].

Another important role of the pre-scrubber is to quench the flue gas stream before it enters the absorber, where it would otherwise evaporate excessive quantities of water over time.

It might be possible instead to eliminate the pre-scrubber and substitute it with a direct contact cooler if the initial quantity of sulfur dioxide in the exhausts was particularly low, like in the case of exhaust gases obtained burning natural gas or syngas [73].

Figure 32: Shell Cansolv typical CO₂ capture process (from [91])



The absorber is a packed column divided into two or three absorption stages with intermediate cooling to better control the temperature profile. A water wash stage is placed at the end of the absorption section of the column. Its function is to eliminate (or drastically reduce) the losses of solvent to the environment due to carry-over or evaporation. In particular it is reported that amine emissions are at sub ppm level for a gas fired boiler in a system operating with the DC-103 solvent[92].

The heat for the regeneration of the solvent is supplied not only from a reboiler, but also from a mechanical vapour compressor. The lean solvent at the bottom of the regenerator is expanded to atmospheric pressure, which causes the formation of a water vapour phase in the lean ammine flash vessel. The vapour is then extracted from the top of the vessel by a vapour compressor and sent to the regenerator in order to minimize the reboiler duty.

The Cansolv process is equipped with a series of components aimed at preserving over time the properties of the ammine solvent used in the cycle by eliminating the non regenerable products from the circulating flow. A possible pathway to achieve this objective is to filter the solvent with selective ion-exchanging membranes [94]. Another option is to periodically distillate batches of solvent under vacuum conditions, which is needed to avoid solvent thermal degradation, in order to separate the non-volatile salts, which are then disposed. This methodology is referred to as *thermal reclaiming*, and it is usually employed in the Cansolv process[92].

The precise design and choice of these sub-units is a function of the contaminants that are expected in the flue gas stream[91]. The degraded solvent is supposed to be periodically refilled.

The Cansolv solvent is constituted a mixture of amines in aqueous solution, and it incurs in the corrosion problems typical of this family of compounds. The solution that is adopted in the Cansolv process to handle the solvent consists in employing stainless-steel as material of choice for the equipment. In particular the stripping column is a stainless steel column operating at 2 bar and containing structured packings of stainless steel.

Considering that it would be very expensive to build the CO₂ absorber entirely with stainless steel, the absorption column is instead a rectangular concrete structure, lined with an acid resistant material, and containing stainless-steel packings[73].

Given that limited information is publicly available regarding the operating parameters of the Cansolv process, the entire unit has been modelled in this thesis as a black-box, without characterization of the single components. In Figure 33 it is depicted the simplified block diagram that has been used to represent the system, while the most important parameters of external characterization are gathered in Table 25. Such parameters have been generally chosen in order to be consistent with cases 12B and 31B from the NETL cost and performance baseline report, in which two power plants implementing the Cansolv process are broadly described.

The data provided in the NETL report are declared to be the result of direct counselling between CTI and the Department of Energy, although they do not appear to be officially confirmed by CTI. Regarding the operating parameters of the process, an interesting parameter is in particular the pressure at which the steam is extracted from the steam cycle to feed the reboiler. In the NETL cost and performance baseline report it is assumed that steam is extracted at 5.1 bar at the cross-over between the IP and LP turbines, while the condensate exiting the reboiler is at 150°C. Given that the report considers an approach temperature of 3°C between the hot and cold fluid in the reboiler, this implies that the boiling point of the solvent is as high as 147°C. It has not been possible to find any confirmation of this piece of data in the open literature. In this report it is assumed that the Cansolv process requires steam at precisely 5.1 bar. If it turned out that the steam pressure doesn't need to be

so high, it would imply that the performances calculated in this report for the Cansolv process are slightly underestimated.

The amount of heat necessary to feed the regeneration process has been instead set to a value (2,33 MJ/kg CO₂) directly issued by CTI in a 2014 report as a research and development target[91]. The value is anyway not very distant from the data published in the NETL report (2,4 MJ/kg).

No data were available regarding the auxiliary electric consumptions of the process, therefore it has been decided to modify the figure provided in the NETL report (27300 kW) by making the hypothesis that auxiliaries depends linearly from the amount of CO₂ captured, and the volumetric flow of gas treated. In particular it has been assigned a weight of 0.6 and 0.4 to these parameters respectively. The expression with which the final figure of the electric auxiliary consumptions have been estimated is therefore:

$$Aux = 27.3 \text{ MW} * (0.4 \frac{\dot{V}_2}{\dot{V}_1} + 0.6 \frac{\dot{m}_2}{\dot{m}_1})$$

where V₂ and m₂ indicate the volumetric flow of gas in the absorber and the captured mass flow of CO₂ in the investigated cases (air blown igcc with carbon capture) while V₁ and m₁ are referred to the case B12 in the netl report. The volumetric flow of gas entering the absorber has been estimated with the hypothesis that the gas is at 50°C.

Figure 33: simplified representation used to model the Cansolv CO₂ capture unit

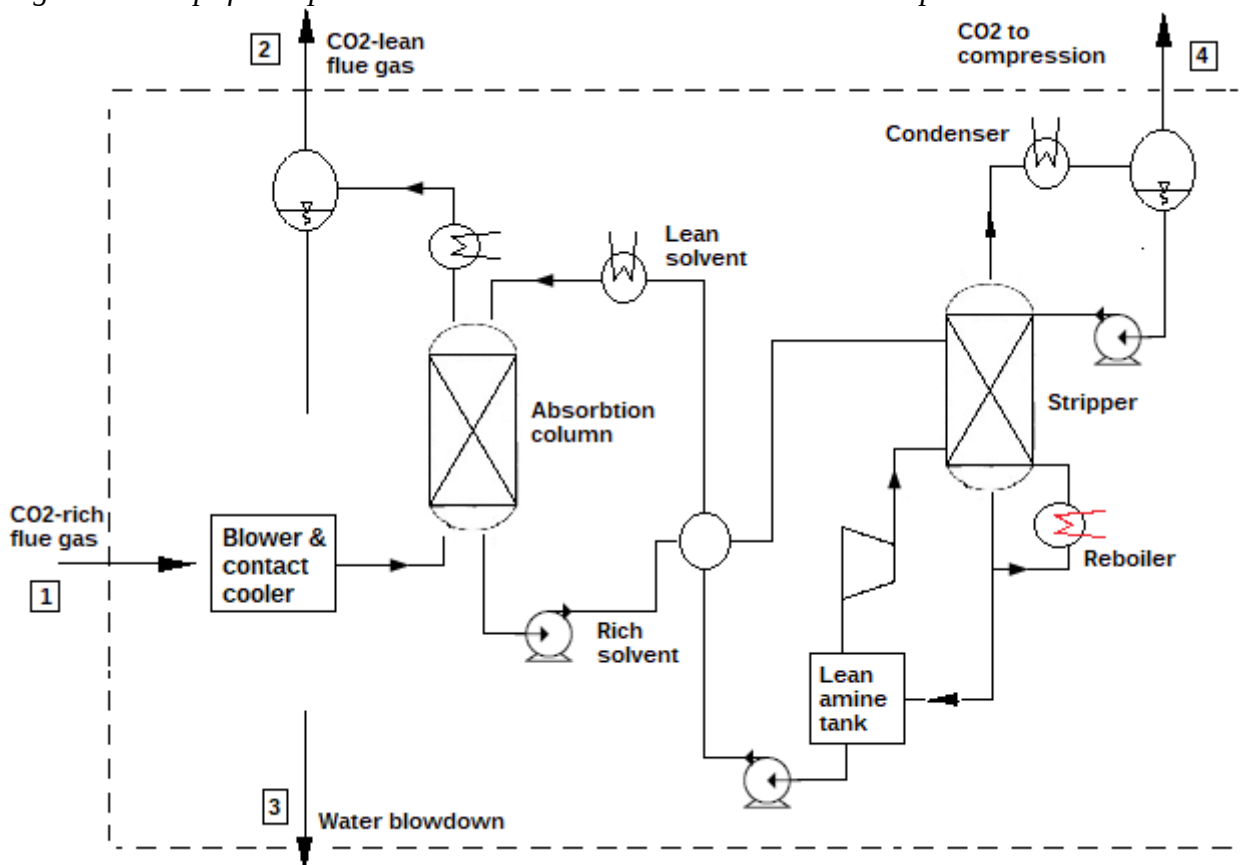


Table 24: Temperature, pressure, flow rate and composition of the main streams of the Cansolv-type CCS plant shown in Figure 33

	Temperature °C	Pressure Bar	Mass flow (kg/sec)	Ar	CO2	H2O	N2	O2
Stream 1	115	1,01	665	0,89	9,97	4,36	75,1	9,68
Stream 2	30	1,01	574,45	0,98	1,1	4,26	82,97	10,7
Stream 3	30	1,01	2,03	0	0	100	0	0
Stream 4	30	2	88,52	0	100	0	0	0

CAPITAL AND VARIABLE COSTS

The capital cost (Bare Erected Cost) associated with the Cansolv unit in the investigated capture cases has been estimated by modifying the cost figure provided by the 2019 NETL report in order to account for the differences in load that the capture block equipment must be able to satisfy. The specific expression that has been adopted is:

$$BEC_{IGCC}^{airblown} = BEC_{12B} * (0.4 \left(\frac{\dot{V}2}{V1} \right)^{0.6} + 0.6 \left(\frac{\dot{m}2}{\dot{m}1} \right)^{0.6})$$

where BEC(12B) is the Bare Erected Cost associated to the Cansolv unit in the case 12B (Coal boiler). It has been verified that the result does not change meaningfully if the data point provided in case 31B (NG boiler) is used to initiate the calculation.

The costs associated to the engineering fees have set to 17,5% of the Bare erected cost (BEC), while the project and process contingencies have been set to the values suggested by the NETL, that is 23,5% and 17% of the BEC respectively.

The variable costs for solvent replacement and disposal of the reclaimers solids have been estimated from case 12B of the NETL report by assuming that they increase proportionally with the mass flow of exhaust gas treated.

For what concerns the disposal of the blowdown from the direct contact cooler it has been decided not to account any cost since the major pollutants, made exception for the NO_x, are supposed to be eliminated from the syngas prior to the CT unit.

Table 25: main parameters and utilities associated to the Cansolv CO₂ capture system

	unit of Measure	value	source of the data
Reboiler Heat Duty	MJ(th)/kg co2 captured	2,33	Direct data from [91]
Auxiliary electric * Consumption	kWh/kg co2 captured	0,0523	Modified data from Case 12B in [73]
Total heat removed by the cooling water in the Various exchangers	MW(thermal)	269,011	Calculated with an energy balance around the dashed line in Figure 33
CO2 discharge* Pressure	bar	2	Data directly employed
Steam pressure * In the reboiler	bar	5,1	From case 12B in [73]
Thermal Reclaimer Solids	Stons / kg of exhaust treated	4,31*10 ⁻⁸	Extrapolated from case 12B
Cost for the replacement of the Cansolv solvent	\$/kg of exhaust treated	3,6*10 ⁻⁴	Extrapolated from case 12B

4.2.3 Ammonia-based CO₂ capture unit

Aqueous ammonia is an innovative solvent for carbon capture that has raised significant interest over recent years. The advantages offered by ammonia over the more conventional amine-based solvents are the following:

- Ammonia is a relatively cheap and largely available chemical
- Ammonia has an high CO₂-loading capacity[95], meaning that high capture ratios should be possible
- In contrast with amines, ammonia does not face problems of degradation if it is contacted with oxygen[96], which is the main contaminant normally present in a stream of flue gases. Additionally, ammonia is more thermally stable than conventional solvents[95], which means that an higher pressure can be achieved in the regenerator column. High pressure regeneration, in turn, is advantageous because it reduces the amount of work needed for the subsequent compression of the carbon dioxide output stream. On the contrary, ammonia is reported to face problems of degradation and formation of heat stable salts in the presence of sulfur oxides[96].
- An aqueous solution of ammonia is reportedly [96] less corrosive and easier to handle than conventional amine-based solvents.

Several small-to-medium scale trials have been conducted with the objective of demonstrating the feasibility of the concept, with the most notable ones being managed by AlstomPower and Delta Electricity. In particular Alstom conducted from 2006 to 2013 a series of pilot projects (Pleasant Prairie, Mountaineer coal plant, Mongstad refinery) capturing the emissions from different typologies of thermal plants for a maximum capacity of 54 MW [97]. The specific design tested (and patented) by Alstom is the so called Chilled-Ammonia-Process (CAP), which according to the literature featured an inlet temperature of the solvent in the absorber within the range 0-10°C [98]. The pilot project managed by Delta Electricity at the Munmorah coal power station in Australia was instead operating at near ambient conditions (15-20°C) [99].

In general terms, the projects carried on have proven the feasibility of concept achieving carbon capture ratios as high as 90% [96], however two important criticalities have been encountered. The first issue, known as ammonia slip, consists in the evaporation and loss of a consistent fraction of ammonia from the top of the absorption and regeneration columns [100]. Ammonia slip from the absorber in particular is problematic not only because it rapidly depletes the solvent, but also because it produces a non tolerable amount of hazardous emissions.

Three main strategies have been proposed by Jilvero et. al [101] to limit ammonia slip:

- A first option consists in reducing the evaporation of ammonia within acceptable limits by employing a low temperature (0-7°C) of the solvent in the absorber. This is effectively one of the points of strength of the CAP design proposed by Alstom.
- A second alternative is to wash the gas exiting the absorber with low temperature (<10°C) water in an absorption column. This strategy is known as *water wash*, and it exploits the high solubility of ammonia in water to reduce the concentration of ammonia in the flue gases sent to the stack.

The main drawback however is that, if ammonia is to be recovered and reintegrated in the CO₂ capture cycle, water and ammonia need to be separated, which can be done in a stripping column at the expense of an additional amount of steam extracted from the LP turbine. Overall, the water wash strategy is a powerful tool to reduce the ammonia slip, but it increases the energy burden of the CO₂ capture unit, possibly jeopardizing the benefits of ammonia with respect to other solvents in terms of thermal load required for solvent regeneration[101].

- A third alternative, usually referred to as *acid wash*, consists in washing the flue gases with a liquid solvent that chemically reacts with ammonia. The solvent usually proposed is sulfuric acid, which reacts with ammonia into nitrogen sulfate, a valuable fertilizer. This strategy is very effective in achieving a low concentration of ammonia in the flue gases sent to the stack, but it has the drawback of consuming both ammonia and sulfuric acid. Therefore an acid wash is proposed by some authors as an after-treatment in combination, and not in alternative, to one of the other aforementioned methods for ammonia control.

A second relevant problem that has been encountered during the pilot trials is the formation of solid precipitates [102], which might potentially result in clogging of the equipment, especially for what concerns the narrow packings of the columns.

It is not in the intentions of this work to supply an in depth discussion of all the conditions that may lead to the formation of solids (the reader is referred to [95]), however it is possible to indicate an high concentration of ammonia in the aqueous mixture as a central factor in the occurrence of this phenomenon[99]. Low temperatures are also a contributing factor for the precipitation of solids. The net effect is that there is a constraint on the operating parameters in which the system can operate during both normal and off-design conditions. This is remarkable because low temperature and high ammonia concentration are exactly those conditions that would minimize respectively the ammonia slip from the absorber and the thermal load required to regenerate the solvent.

An alternative pathway offering some prospects in light of the problem of solids formation is the possibility of adopting, in some parts of the system, an equipment appropriate to handle a liquid-solid slurry.

In particular packed towers would need to be substituted with tray towers, which are more tolerant to the presence of suspended solids, but have less satisfactory properties of mass exchange per unit of volume. The pumps for the solvent would need to be designed with particular care in order to sustain wear from the continuous contact with solid particles, and an hydrocyclone would need to be employed to separate the liquid solvent from the solids if the design of the unit entails a recycle-loop around the absorber.

Overall, there is a successful precedent to draw experience from in order to realize a technical solution of this kind, since sulfur dioxide wet scrubbers employed in large scale coal plants usually adopt a similar equipment[103].

On the other hand the capital costs for such a plant would be higher with respect to a more conventional CO₂ capture plant adopting an entirely liquid solvent.

In conclusion, ammonia can be labelled as a promising solvent in alternative to conventional amines, although its advantages have not still been convincingly demonstrated on a full scale plant (e.g. 250 MW) over a sustained period of time. At the same time, the choice of the best operating parameters to maximize the potential of the solvent appears to be a crucial problem of non-obvious solution, where several factors have to be carefully weighted in order to find the optimal trade-off. In the following paragraphs two different plant schemes based on ammonia are briefly presented, being both sourced from the article [75] from Bonalumi et. al, where a set of results of the aforementioned process of optimization of the operating parameters is presented.

The process flow diagrams of the plants are shown in the Figures 34 and 35, with particular emphasis on the different functional blocks constituting the system. The most important operating parameters are instead listed in Table 28. The two plant configurations are labelled respectively *cooled* or *chilled* according to the temperature set for the streams entering the absorber.

The general scheme of the process is similar:

- the stream of exhausts from the HRSG is cooled down up to the absorber inlet temperature

in a series of contact coolers, releasing a noticeable amount of moisture. The exhaust is also slightly compressed with a fan in order to be able to win the pressure losses in the components that come after. The contact coolers in the cooled and chilled case are different both in number and in the amount of cooling supplied to the gas.

Figure 34: schematic representation of the cooled-ammonia CO₂ capture unit (from [75])

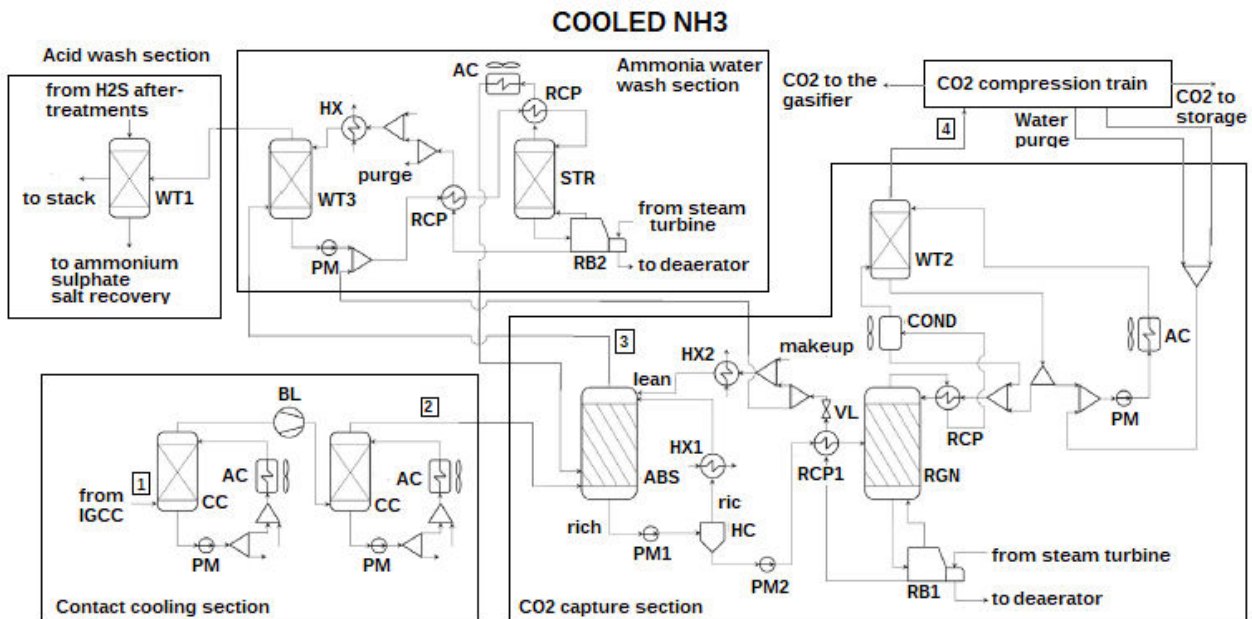


Figure 26: Temperature, pressure, flow rate and composition of the main streams of the CCS plant shown in Figure 34

	Temperature °C	Pressure Bar	Mass flow (kg/sec)	Ar	CO ₂	H ₂ O	N ₂	O ₂
Stream 1	115	1,01	665	0,89	9,97	4,36	75,1	9,68
Stream 2	20	1,09	656,6	0,91	10,18	2,34	76,68	9,88
Stream 3	20	1,06	568,08	1,01	1,14	1	85,8	11,05
Stream 4	25	5	88,52	0	100	0	0	0

- The solvent is continuously circulated between absorber and regenerator in the typical arrangement of chemical absorption cycles, with a certain fraction of rich solvent (see Table 28) being recycled around the absorber.
- The CO₂-lean exhaust gas exiting from the top of the absorber is passed in a series of components aimed at reducing its content in ammonia. In the chilled case, being the ammonia slip relatively contained due to the colder temperatures, an acid wash reactor is sufficient, given the amount of sulfuric acid available from the sulfur recovery unit. On the contrary, according to the article [75], in the cooled case the ammonia slip is so relevant that the in-site production of sulfuric acid is not sufficient for a complete neutralization of ammonia in the exhausts. Therefore the acid wash reactor is placed in series to a water wash column performing a pre-abatement of the NH₃ particles. A fraction of the ammonia dissolved in the water is then recovered from the top of the NH₃ stripping column, and reinjected in the absorber.
- A water wash column for ammonia abatement is placed upstream to the CO₂ compression train in both the configurations. The stream of concentrated carbon dioxide is then compressed to supercritical pressure according to the modalities described in the section 4.2.4.

The parameters in Table 28 are very relevant to define the best conditions in which the CO₂ capture plant should operate, however the key parameters for the purposes of this analysis are those defining the performance level of the CO₂ capture block from the perspective of the entire power plant system. Borrowing from the language adopted for electric circuits, these indicators might be called *external parameters*.

Figure 35: schematic representation of the chilled-ammonia CO₂ capture unit (from [75])

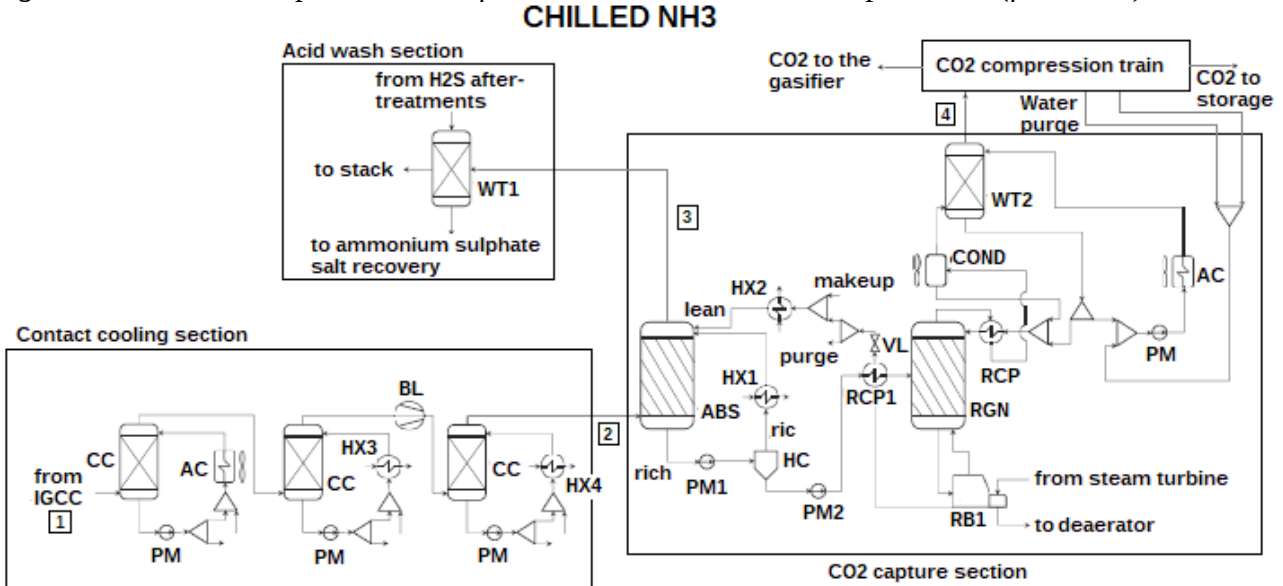


Table 27: Temperature, pressure, flow rate and composition of the main streams of the CCS plant shown in Figure 35

	Temperature °C	Pressure Bar	Mass flow (kg/sec)	Ar	CO ₂	H ₂ O	N ₂	O ₂
Stream 1	115	1,01	665	0,89	9,97	4,36	75,1	9,68
Stream 2	7	1,09	651,3	0,92	10,32	0,02	77,73	10,02
Stream 3	16	1,06	562,78	1,01	1,14	1	85,8	11,05
Stream 4	25	5	88,52	0	100	0	0	0

Table 28: optimal operating parameters for the ammonia CO₂ capture process (from [75]):

	Unit of measure	Chilled	Cooled
Temperature absorber inlet streams	°C	7	20
CO ₂ capture ratio	-	90,00%	90,00%
Ammonia initial wt concentration (upon entering the absorber)	-	20,00%	10,00%
NH ₃ /CO ₂ ratio	-	5	4,75
Regeneration pressure	bar	5	5
Solvent recycle ratio (around the absorber)	-	80,00%	10,00%

Where:

$$\text{Ammonia concentration} \stackrel{\text{def}}{=} \frac{m_{\text{NH}_3}}{m_{\text{NH}_3} + m_{\text{H}_2\text{O}}|_{\text{lean}}} \quad \text{NH}_3 \text{ to CO}_2 \text{ ratio} \stackrel{\text{def}}{=} \frac{n_{\text{NH}_3}|_{\text{lean}}}{n_{\text{CO}_2}|_{\text{gas-inlet}}}$$

The external parameters adopted in the economic analysis to model the ammonia CO₂ capture block are listed in Table 29 and 30.

As a general note, the values adopted have been directly picked from article [75], or calculated using the data in article [75] as a starting point for the calculations.

In particular:

- Temperature and reboiler heat duty have been directly reported from the article [75] without modification. The same holds also for the auxiliary electric consumption, which does not include the auxiliary consumption of the CO₂ compression section. It must be noted that the chilled design features a significantly higher electric consumption due to the need of refrigerating the solvent at 7°C. The cooled design instead requires only a limited amount of refrigerating power due to the necessity of cooling to 7°C the water directed to the water wash column.
- The amount of water condensate from the contact coolers has been estimated by modelling each of the contact coolers as an heat exchanger followed by an equilibrium stage. Water properties have been calculated with the Redlich-Kwong equation of state. No disposal cost has been associated to the blowdown of the contact coolers.
- The amount of total heat rejected from the system (compression excluded) has been evaluated as the sum between the heat exchanged in the contact coolers and the heat introduced in the reboilers.

In order to facilitate the comparison with the cansolv-type CO₂ capture plant, it has been likewise imagined that all heat is ultimately removed from the system by means of the circulating cooling water, without the use of air-cooled heat exchangers. This assumption is at odds with the article [75] from which the data on auxiliary consumptions have been sourced, however it has been judged that the difference in the specific electric energy consumption for heat rejection between air-cooled and water-cooled exchangers is small enough to be neglected.

If anything, since more electric power is necessary to reject heat with an air-cooled exchanger, the current figure for the electric auxiliary consumptions for heat rejection should be viewed as slightly higher than the real figure that would be achieved with a perfectly coherent set of assumptions. This is particularly the case for the cooled configuration, where a greater amount of heat need to be rejected.

Table 29: main parameters used to characterize the capture section

	Unit of Measure	Chilled	Cooled
Regenerator reboiler duty	MJ/kg CO ₂	2,188	2,504
Regenerator reboiler temperature (solvent side)	°C	67,4	106,5
NH ₃ stripper reboiler duty	MJ/kg CO ₂	-	0,212
NH ₃ stripper reboiler temperature	°C	-	97,9
Auxiliary electric consumption (compression excluded)	MW(el)	51,6	13,71
Contact coolers water blowdown	kg H ₂ O/sec	13,68	8,33
Heat rejection from the contact coolers	MW(th)	112,52	91,05
Total Heat to be rejected	MW(th)	306,22	331,55

Table 30: paramters used to characterize the NH3 wash section (if present)

	Unit of Measure	Chilled	Cooled	Comments
Ammonia slip from the absorber	ppmv	3050	6190	
Ammonia concentration in the gas entering the acid wash	ppmv	3050	4100	
Ammonia discharged in the purged water	kg NH3 /sec	-	0,25*(0)	*these results have been obtained with the ideal mixture model for the H2O-NH3 solution. These results are not actually used in the economic analysis. The figures that are used in the calculations are those in the brackets
Water blowdown from the WW unit	kg H2O /sec	-	300,25*(0)	
Ammonia concentration in the blowdown	ppmv	-	880,9*(0)	
Total NH3 consumption	kg NH3 /sec	1,023	1,629*(1,4)	

- The ammonia slip from the absorber has been estimated with an heuristic method. Based on the amount of sulfur available from coal, it has been evaluated that a concentration of ammonia in the flue gases greater than 4100 ppmv would not be reducible in the acid wash section.

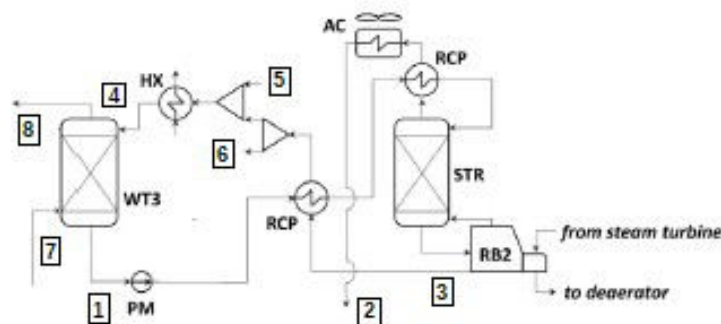
According to the article [75], the ammonia slip in the chilled configuration is low enough to be eliminated in the acid wash section, while a water wash section is also needed in the cooled configuration. Drawing on this data, the ammonia slip for the chilled case has been arbitrarily set to the value of 3050. At the same time, considering that going from 7°C to 20°C the vapour pressure of pure ammonia grows more than 50%, the ammonia slip exiting the absorber in the cooled case has been set at the doubled value of 6100 ppmv. Because of the strong approximation and uncertainty of this method a sensitivity analysis has been performed in order to assess the relevance of the costs for the replacement of ammonia on the various cases investigated. The discussion of the results is presented in Chapter 5.

- In the cooled configuration, the water wash absorbion column has been treated as a discrete series of 50 (meaning infinite) equilibrium stages. The flows exchanged between the different stages have been estimated by solving the MESH equations [69] under the hypothesis that both the gaseous and liquid phase are ideal mixtures. The system has been considered as a mixture of water, nitrogen and ammonia, neglecting all other species. The properties of water and ammonia as pure compounds have been evaluated with the RKS equation of state.

The NH3 stripper has been solved under the same set of hypothesis with the Mccabe Thiele method.

It has been therefore generated a solution (represented in Figure 36 and Table 31) of the flowsheet under the hypothesis of ideal mixtures.

Figure 36: detail of the NH3 wash section



According to the results, a considerable amount of water (300,6 kg/second) needs to be continuously substituted in order to purify the water from ammonia to a sufficient degree (740 ppmv) for the absorption operation. It has been also estimated that about two third of the ammonia dissolved in the water (stream 1) are recovered (stream 2), while the remaining quantity is eliminated in the purge (stream 6). If this result was a realistic description of the performance of the system, it would mean that an unacceptably high volumetric flow rate of water mixed to ammonia would be produced as a waste stream.

However it has been judged that the discharge of water mixed to ammonia is likely much smaller if one takes into account also the non-ideal effects of the interactions between the molecules of water and ammonia.

In particular, being both ammonia and water polar molecules, they have a tendency to form weak bonds (dipole-dipole interactions). In the water wash absorber the expected consequence is that a minor (or less pure) amount of water will be needed to absorb the ammonia. On the contrary, in the stripping column it is expected that the weak bonds between water and ammonia will be the easiest ones to break, therefore enhancing greatly the evaporation of ammonia with respect to a situation where the liquid phase is supposed to be ideal.

Both these considerations point towards a reduction of the quantity of ammonia discharged in the purge. For this reason, it has been in the end decided to ignore in the calculations the loss of ammonia from the purge of the water wash unit, supposing it negligible if compared with the loss of ammonia in the acid wash column.

Table 31: Results of the simulation on the NH3 wash unit

	Mass flow (Kg / sec)	N2	H2O	NH3
Stream 7	571,67	0,9677	0,02611	0,00619
Stream 8	576,24	0,9698	0,0261	0,0041
Stream 1	1956,03	0	0,99912	0,00088
Stream 2	8,3	0	0,93643	0,06357
Stream 3	1947,62	0	0,99901	0,00099
Stream 5	312,25	0	1	0
Stream 6	300,5	0	0,99912	0,00088
Stream 4	1959,37	0	0,99926	0,00074

CAPITAL & OTHER COSTS

The capital cost of the ammonia co2 capture plant has been estimated with a bottom up method, that is by aggregating the capital cost of the single items constituting the system.

Such costs have been obtained by modifying quotations from the existing literature for similar pieces of equipment. In particular, the cost quotations used as a baseline have been sourced from the paper [104] from Bonalumi and Valenti, where the techno-economic analysis of a co2 capture plant based on ammonia is presented. Remarkably, the plant there discussed is very similar to the cooled co2 capture unit object of this work.

The capital cost figures for the various pieces of equipment have been determined with the following scaling expression:

$$\text{Capital cost} = \text{reference cost figure} * \left(\frac{\text{value of the scaling parameter}}{\text{reference scaling parameter}} \right)^{0.6}$$

where the reference cost figures are sourced from [104], and proper scaling parameters are chosen to represent the likely dependence of the capital costs on a scale-factor.

For example it has been judged that the volumetric flow rate of gas entering the absorber is the most significant parameter influencing the dimension of the absorption column. Therefore the capital cost of the absorber in the cooled and chilled configurations has been determined by making a comparison of the value actually assumed by this parameter with respect to the case reported in the article [104].

In order to generate some of the cost estimates it has been sometimes necessary to aggregate different pieces of equipment. For example the heat exchangers in the contact cooling section have been treated as a unique item whose cost depends on the total heat exchanged in the contact cooling section.

For three relatively minor cost items no data were available to attribute a value to the scaling parameter chosen, therefore the cost figure from the article [104] has been directly utilized without any scaling or modification factor.

The scaling parameters adopted are listed in Table 32, together with the cost evaluations obtained for the investigated plant configurations. All the figures are expressed in 2011 dollars and need to be converted in 2019 dollars with the CEPCI index (1.037 is the conversion factor).

It should be added that the capital cost of the chilling plant for chilled configuration has been calculated also with an alternative methodology that produced a very similar result. In particular it was imagined that the refrigerating effect was produced by a NH₃-based vapour compression cycle with a COP equal to 5. Then the cost of the chilling plant was determined as the sum of the pieces of equipment necessary to realize the cycle.

The figure for the total Bare erected cost of the CO₂ capture unit has been obtained multiplying by 2.34 the total equipment cost, which is the value adopted in the NETL report for the Cansolv post combustion technology (Case 12B).

Engineering fees, project contingencies and process contingencies have been set respectively to 17.5%, 23.5% and 17% of the bare erected cost. These coefficients are the same used in the NETL report for the Cansolv CO₂ capture unit. The process contingencies of the chilled ammonia plant have been set to 50% of the bare erected cost, in order to take into account the various difficulties related to a lower kinetics in the absorber and the necessity of handling solids.

The same identical coefficients have been used also for the acid wash column.

Finally, it needs to be specified that the cost of ammonia replacement has been estimated considering a market price of 400 \$/Ton of ammonia (MetricTon) [105].

Table 32: detail of the capital costs associated to the CO2 capture plant with ammonia. In the second column it is provided the list of the parameters used to scale the quotations from [118]

	Scaling parameter	Unit of measure	Original Quotation (k\$ 2011)	Scaled cost for the Cooled Configuration	Scaled cost for the Chilled Configuration
CONTACT COOLING SECTION					
Contact Cooler n.1	Volumetric flow in the CC1	m3/sec	1363	1394,7	1394,8
Contact Cooler n.2	Volumetric flow in the CC2	m3/sec	1430	1419,6	1475,1
Contact Cooler n.3	Volumetric flow in the CC3	m3/sec	-	-	1304,6
Heat exchanger n.1	Power exchanged in the contact cooling section	MW	3882	2634,6	2991,5
Heat exchanger n.2					
Heat exchanger n.3					
Pump n.1	Volumetric flow in the CC1	m3/sec	141	144,2	144,2
Pump n.2	Volumetric flow in the CC2	m3/sec	42	41	43
Pump n.3	Volumetric flow in the CC3	m3/sec	-	-	38
Fan	Mass flow of gas to the absorber	kg/sec	393	385,9	384
Subtotal			7251	6020	7775,2
CO2 CAPTURE AREA					
Absorber	Volumetric gas flow to the absorber	m3/sec	18741	18587,4	17944
Washing tower for the co2 product stream	CO2 separated	kgCO2/sec	3635	2990,1	2990,1
Regenerator	CO2 separated	kgCO2/sec	7561	5272,4	5272,4
Regenerator reboiler	Heat exchanged in the reboiler	MW	9007	7838,2	7228,3
Recuperative HX	Solvent flow * (ΔT between absorber inlet and reboiler)	kg*Kelvin/s	13624	13344,1	5031,4
Heat exchanger n.1	Heat exchanged in the reboiler	MW	4499	3915,2	3610,5
Heat exchanger n.2					
Regenerator condenser					
Pump n.1	Solvent recycled around the absorber	kg/sec	260	86,9	348,2
Pump n.2	Mass flow of loaded solvent entering the stripper	kg/sec	795	767,9	215,4
Washing tower pump	CO2 captured	kgCO2/sec	25	20,5	20,5
Subtotal			58147	52822,7	42660,8
	Scaling parameter	Unit of measure	Original Quotation (k\$ 2011)	Scaled cost for the Cooled Configuration	Scaled cost for the Chilled Configuration
AMMONIA WATER WASH SECTION					
WW absorber	Volumetric flow of gas entering	m3/sec	153	156,6	-
NH3 stripper	-	-	442	442	-
Stripper reboiler	Heat exchanged in the reboiler	MW	1896	828	-
Recuperative heat exchangers	-	-	2242	2242	-
Water Pump	-	-	18	18	-
Other heat exchangers	Heat exchanged in the reboiler	MW	517	225,8	-
Subtotal			5268	3912,4	0
CHILLING PLANT					
Chilling plant	Refrigerating power demanded	MW	979	979	13725,42
Total			71645	63734,1	64161,42

4.2.4 CO₂ compression train

The CO₂ compression unit is a subsystem whose purpose is to compress to supercritical pressure the flow of concentrated carbon dioxide exiting at the top of the solvent regeneration column. From a theoretical point of view, the most efficient thermodynamic pathway to realize this transformation is by means of a perfectly isothermal compression. The temperature of the compression should be, ideally, the minimum possible in order to reach supercritical conditions while at the same time avoiding the vapor-to-liquid transition.

Ideal isothermals are usually realized in practice with a series of intercooled adiabatic compression stages, with an ideal isothermal process being equivalent to an infinite number of intercooled stages.

Provided that the compressors for carbon dioxide used for large scale CCS are highly application-specific and represent a relatively niche-market[67], integral-gear centrifugal compressors [106] are the leading technology currently proposed for this purpose [107].

In particular integral gear centrifugal compressors have been adopted in the two largest CCS projects developed so far (BoundaryDam[108] and PetraNova[109]). The number of intercooled compression stages usually considered is eight[67], which allows to achieve a significant efficiency. While this solution offers undoubtedly many advantages, some authors [110] observe that integrally geared compressors are less reliable and require significant additional maintenance if compared to single shaft centrifugal compressors. This is due to the presence of the gearbox and the high number of bearings and gas seals. As a consequence, in order to guarantee full operability it would be perhaps necessary to always have a spare compressor, which would ultimately increase the capital costs associated to this solution.

As a possible alternative, supersonic CO₂ compressors are an advanced technology that offers some prospects and that is in the process of being commercialized by Siemens, with partnership of the US DOE[111].

Supersonic compressors in particular allow a much greater compression ratio for each stage, with the effect that a total of two compression stages is sufficient to achieve supercritical conditions. This solution is interesting because is very compact in size and it allows to avoid many of the disadvantages described earlier. Moreover, supersonic compressors are interesting for their possibilities of heat recovery. In particular a study of the DOE [112] shows that heat recovery from high pressure-ratio compressors might be used to reduce the amount of medium pressure steam used for water preheating in a USC power plant, with small improvements in the overall efficiency. The main drawback is that supersonic compressors are inevitably less efficient than conventional compressors of factor as high as 25% (according to a 2015 study[110] of the DOE). This is both a consequence of the intrinsic nature of super-sonic flows and the higher average temperature of the gas.

A secondary but important function that must be satisfied in a CO₂ compression system is that of progressively drying the flow of compressed gas. This is necessary because if water was not removed there would be formation of acid condensates within the carbon steel pipes that deliver the flow of supercritical carbon dioxide from the capture facility to the site of injection (in case of geological sequestration)[55]. Another reason for drying the stream of gas is to prevent or limit the risk of hydrates formation within the compressor, which might severely damage the internals[113]. Additional stages of purification (that are not discussed here) might be also present with the aim of removing the eventual incondensable gases, oxygen in particular, to meet specific requirements imposed by the reservoir in which the flow is injected.

In light of the above reasons, the design of the compressor usually involves water knockouts placed in between some of the intercooled stages. Moreover, to obtain a dehydration in the order of the

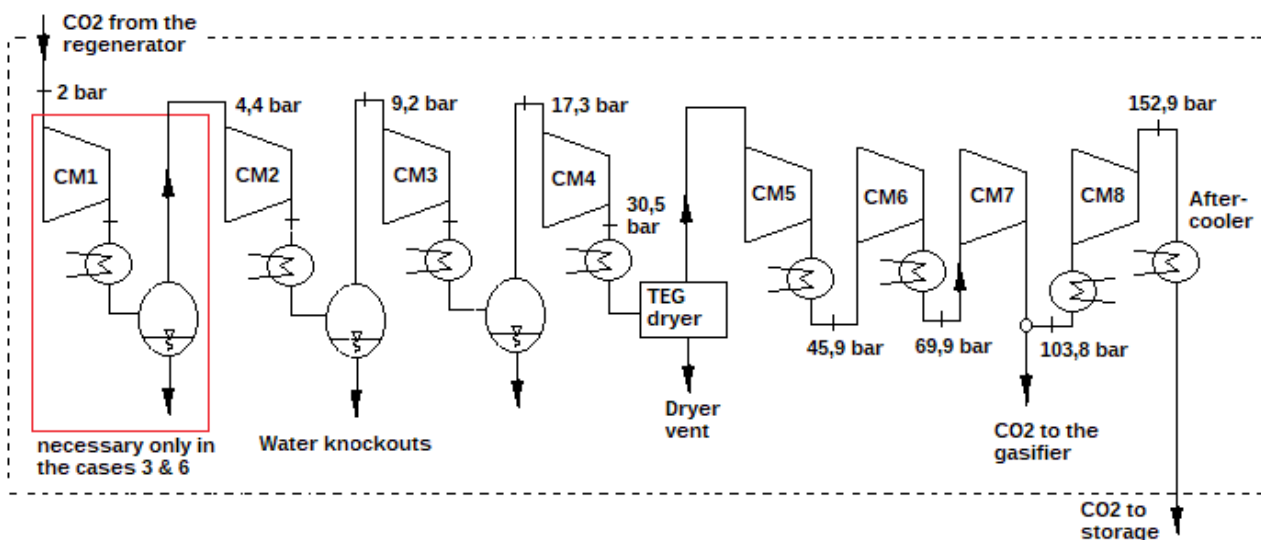
hundreds of ppm of water, it is usually included in the process also a component called *dryer*, which is specifically meant to remove water as much as possible. The process most often mentioned for this purpose is the dehydration with tri-ethylene glycol (TEG)[113]. It consists in a pressurized packed column where the water-rich carbon dioxide flow is contacted with tri-ethylene glycol, a physical solvent with strong affinity with water. The solvent is then regenerated through depressurization and heating in a stripping column, requiring a negligible quantity of steam in relative terms.

Once the flow of CO₂ has reached the dense phase, it can be pumped across a potentially long pipeline with relatively inexpensive and non-critical centrifugal pumps [107], possibly located in a series of booster stations with all the necessary controls and redundancies.

A series of images representing some of the typical components in the CO₂ compression process is reported in Appendix F.

For what concerns the power plant configurations with carbon capture investigated in this analysis, the compression train has been imagined to be constituted by a multistage, intercooled, integrally geared machine, equivalent to that adopted for case 12B of the 2019 NETL cost and performance baseline report. In particular the number of stages has been imagined to be eight for the cases involving a Cansolv-type CO₂ capture unit, while just seven stages have been deemed sufficient in the other configurations since the regenerator pressure is higher and less compression is needed. The system considered is represented in Figure 37, and the most significant operating parameters have been listed in Table 33.

Figure 37: schematic representation of the carbon dioxide compression process



Notably, the specific power consumption of the cases with ammonia is lower since the compression load in the first stage has been neglected. In the calculations it has been assumed a 100% carbon dioxide flow without taking into account water and other impurities (the mass flow of gas is shown in Table 33).

No heat recovery has been considered, and, in particular, the amount (in terms of MWs) of cooling water necessary to operate the intercoolers has been supposed to be equal, in first approximation, to the electric power spent to drive the compressor. No work reduction has been included to account the fact that a fraction of gas is sent to the gasifier instead of walking through the last part of the compression.

Table 33: list of the parameters used to model the CO2 compressor

	Unit of Measure	Case 12B NETL	Cases 3 and 6	Cases 1,2,4 and 5
Number of compression Trains	-	2*50%	2*50%	2*50%
CO2 compressed	kgco2/sec	161,4	177,04	177,04
Specific electric energy necessary for the compression	kJ(el)/kgco2	274,85	274,85	236,94
Electric power Consumption	MW(el)	44,36	48,65	41,94
Cooling tower load	MW(thermal)	-	48,65	41,94
TEG consumption	gallons/kg Co2 processed	3,9*10 ⁻⁵	3,9*10 ⁻⁵	3,9*10 ⁻⁵

CAPITAL AND VARIABLE COSTS

The capital cost for the co2 compression unit has been estimated from case 12B by means of the expression:

$$EC_2 = EC_1 \left(\frac{\dot{P}_2}{\dot{P}_1} \right)^n$$

where EC indicates the equipment capital cost, while P indicates the electric power consumption of the compression unit, which has been chosen as scaling parameters. The scaling exponent n has been set to the value 0.61 according to the NETL scaling guidelines[84]. The capital cost EC1 used as reference is, in detail, the sum of accounts 5.4 and 5.5 in the 2019 document of case 12B.

Engineering fees and project contingencies have been set to 17,5% of the bare erected cost, which in turn has been calculated as 1.48 times the equipment cost, which is the coefficient employed in the NETL report (case 12B).

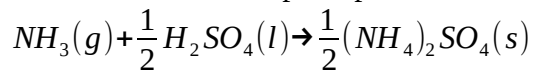
It has been considered a cost associated to the consumption and disposal of the tri-ethylene glycol spent in the process. In particular for all the cases it has been imagined that the consumption of tri-ethylene glycol per unit of CO2 processed is equal to Case 12B from the NETL report.

4.2.5 Acid wash plant

The acid wash column is a reactor in which the flue gases from the CO₂ capture plant are stripped of their ammonia content through a scrubbing operation performed with an aqueous solution of sulfuric acid.

More in general, the acid wash column is a component of a broader subsystem, here referred to as *acid wash plant*, whose task is to prepare the scrubbing solution, collect it at the reactor outlet, and finally process it in order to isolate the valuable ammonium sulphate crystals.

Overall, the acid wash plant performs the following transformation:



In contrast to the other subsystems constituting the power plant, it has not been possible to find proposals in the surveyed literature on ammonia-based CO₂ capture regarding the specific design of the acid wash section. Therefore, in the spirit of providing a comprehensive picture of the power plant configurations investigated, it has been decided to put forward a tentative design for the acid wash unit based on the indications gathered from the broader chemistry literature.

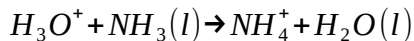
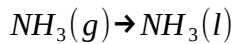
In this respect, high attention has been paid to a passage from a recent chemistry textbook[114] where the author (J. Speight) describes the manufacturing process for ammonium sulphate. The passage is here recalled:

“..a mixture of ammonia gas and water vapor is introduced into a reactor that contains a saturated solution of ammonium sulfate and about 2%–4% (v/v) of free sulfuric acid at 60°C; the heat of the reaction maintains the desired temperature... Ammonium sulfate crystals are formed by circulating the solution through an evaporator where it thickens. A centrifuge separates the crystals from the mother liquor. The crystals contain 1%–2.5% moisture and are dried in a fluidized bed or rotary drum dryer. Dryer exhaust gases are sent to a particulate collection system (e.g., wet scrubber) to control emissions and recover residual product. Coarse and fine granules are separated by screening before they are stored or shipped.”

In the next paragraphs a brief discussion is provided on the applicability of the process suggested by Speight to the cases of object of this study, then it is proposed a tentative flow diagram for the process, complemented with some order-of-magnitude estimates concerning the quantities involved.

1. The main objective of the acid wash column is to obtain a very high degree of removal (conversion) of ammonia from the gaseous phase.
If it is taken into account that ammonia molecules are dispersed in a volume of gas in the hundreds of cubic metres, it jumps to the eye that the amount of sulfuric acid in the scrubbing solution entering the reactor needs to be much greater than the stoichiometric minimum in order for a deep removal of ammonia to be achieved.
In view of this consideration, the range of sulfuric acid concentrations suggested by Speight (2-4% molar) appears fairly reasonable.
The use of a superabundant quantity of acid has in turn an important repercussion on the process flow diagram, since it implies that the amount of non-reacted sulfuric acid needs to be continuously recycled to the top of the column in a closed loop.
2. It has been verified that, if concentrated sulfuric acid is mixed with water up to 4% molar concentration, the heat of dilution should be roughly enough to raise the temperature of the solution to the suggested level of 60°C.

On the other hand there appear to be some uncertainties regarding the temperature of the gas entering the reactor. In particular, the reactor features the occurrence of the following exothermic reactions:



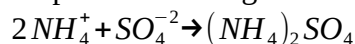
However, given the relatively small quantity of ammonia available (< 1%), the reaction heat doesn't seem sufficient to maintain the temperature around the desired level of 60°C if the flow of flue gases enters the reactor in cooled or chilled conditions (~ 7-20°C).

Therefore the questions are raised here of which might be an acceptable gas inlet temperature, and if it might be necessary to reheat the stream of flue gases prior to the acid wash reactor, possibly with a steam extraction from the LP turbine.

3. In this analysis the acid wash column has been supposed to be a packed column, coherently with what is considered by Bonalumi et. al[104].

A consequence of this assumption is that the acid mixture should be not too close to saturation in order to avoid the formation of solid ammonium sulphate in the midst of the column packings.

4. The crystals of ammonium sulphate, which is a non-volatile species, can be recovered, as suggested by Speight, by evaporating a fraction of the acidic solution. Solids tend to then precipitate according to the following reaction:



The operation of dewatering is typically done in the process industry through a compression of the liquid, followed by heating. The hot and compressed liquid is then expanded in a large vessel, at the top of which water vapor is extracted. The concentrated liquid is instead collected at the bottom, and partially recirculated.

In particular it is possible to distinguish the systems for evaporation between brine concentrators and crystallizers.

Figure 38 represent a possible design for a crystallizer (on the left) and an evaporator (on the right).

The brine concentrators are used to concentrate a solution until near-saturation conditions, and they are not geared to handle the formation of solids. Crystallizers instead are specifically intended to produce precipitates, and they employ particular slowly rotating slurry pumps. A fraction of the slurry is sent to a centrifugal separator to isolate the crystals.

For their functioning both the systems require the introduction of an external energy input, which rapidly increases when the quantities of water to evaporate become significant. A possible strategy to contain the operative costs is to limit the steam temperature in the heat exchanger, while at the same time expanding the liquid up to vacuum pressures in the vessel. In this way the system might be designed to operate with waste heat in the temperature range 100-150°C from other processes.

An alternative approach is the so called *vapor recompression*, which consists in exploiting the latent heat of the vapor produced as an heat source to evaporate new steam. In order to have a driving force in the heat exchanger all is needed is to slightly compress the vapor at the top of the vessel. With such an arrangement it is then possible to drastically reduce, or completely eliminate, the need for external steam, at the same time without replacing it with excessive electric consumptions.

An example of vapor recompressing system is represented on the right in Figure 38.

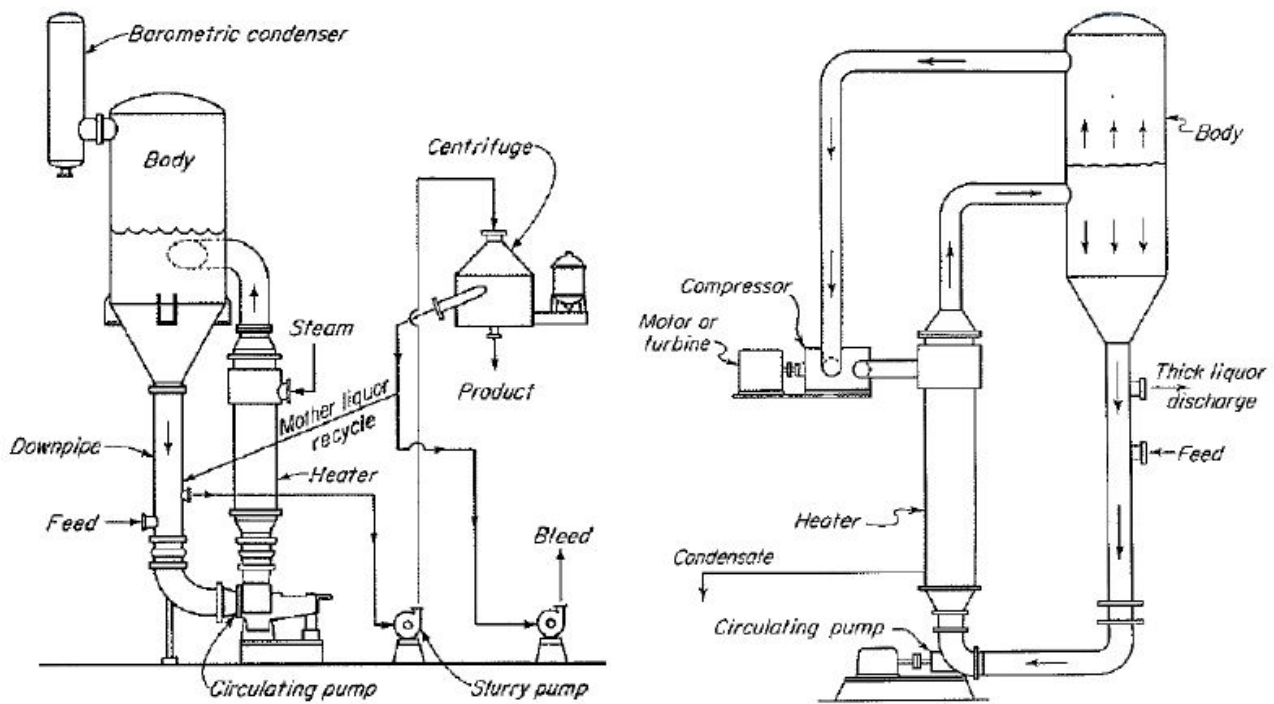


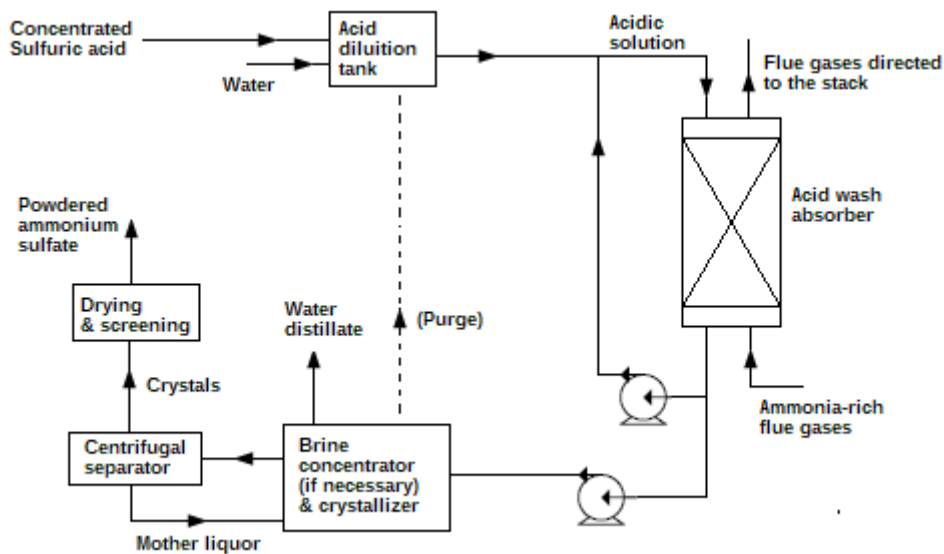
Figure 38:

steam fuelled crystallizer coupled with a centrifuge for the separation of the solids (on the left), vapor recompressing evaporator (on the right)

In the cases investigated it has been judged that a crystallizer is always necessary, while the need for a brine concentrator is ultimately dependent on the choice of the operating parameters of the column, with particular emphasis on the distance of the solvent from saturation conditions.

- The operations of drying and screening of the crystals do not appear to be particularly energy consuming, therefore they are not explored in this work.

Figure 39: block diagram proposed for the acid wash plant

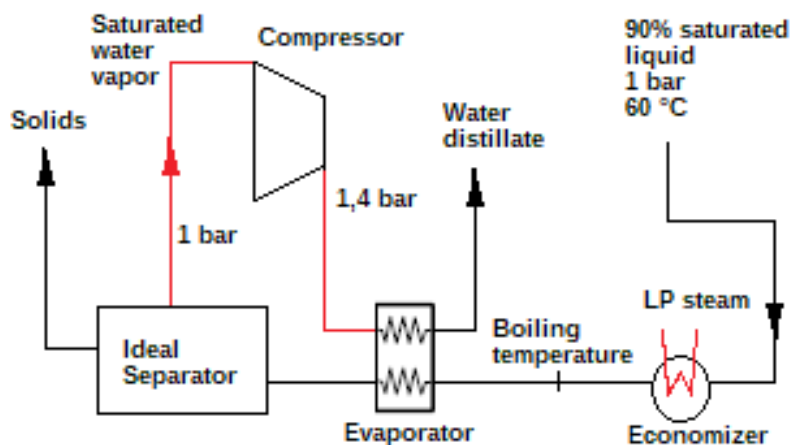


Based on the above, a block diagram that appears appropriate for the acid wash plant is represented in Figure 39. A purge stream is included in addition to all the streams discussed to avoid an excessive build-up of sulfuric acid in the crystallization unit.

A gross estimate of the electric and steam consumptions of the equipments for evaporation has been determined with the simplified model depicted in Figure 40. Essentially, it has been imagined that salts of ammonium sulphate are separated from water in an ideal separator after all the water present in the inlet stream is evaporated. The inlet stream is supposed to be at 60°C, and the amount of salts dissolved in the water is equal to 90% of the saturation quantity (the data on solubility of ammonium sulphate in water have been gathered from Perry Chemical Engineering Handbook[115]).

In order to minimize the quantity of external heat necessary, it has been imagined that LP steam is used to preheat the mixture up to its boiling point, while the latent heat of evaporation is supplied through a vapor recompression system.

Figure 40: simplified model for the salt recovery section



Additionally, it has been imagined that sulfuric acid is not present, that the boiling point elevation effect is negligible, and that the specific heat of water is a good proxy for the specific heat of the whole solution.

The results found are the following:

- It has been determined that the electric consumption of the compressor is very limited. In particular it has been found a figure of 377 kW and 511 kW for each train in the chilled and cooled cases respectively. The isentropic efficiency of the compressor was set at 85%.
- The amount of heat needed to preheat the solution up to its boiling point (here assumed to be 100°C) is also limited. The amount was estimated in 1,49 MW and 2,05 MW for each train in the chilled and cooled cases respectively.

These results tend to characterize the process of separation of the crystals from the liquid as non-critical in terms of energy consumption. However it should be remembered that the boiling point elevation effect was not considered. If the boiling temperature of the saturated solution of water and ammonium sulphate was significantly higher than 100°C, all the figures related to the energy intensity of the process (compression work, thermal load for preheating, steam pressure in the preheater) would also increase, potentially changing the picture.

In conclusion, due to the lack of data on the colligative properties of the water-ammonium sulphate system, the quantitative evaluation of the acid wash plant has been omitted from the economic

analysis, made exception for the capital cost of the acid wash column, and the operative costs for sulfuric acid consumption. In Chapter 5 it is also discussed the degree to which the revenues related to the sale of ammonium-sulphate may potentially influence the LCOE of the plant if the process of separation of the salts proves to be relatively inexpensive.

The various figures that have been included or omitted are represented in Table 34.

CAPITAL COST

The capital cost of the column has been extrapolated from the cost figure presented in the article [104] from Bonalumi et. al through a two-factor scaling expression:

$$Capital\ cost = (0,6 * TC * (\frac{SP\ 1}{RP\ 1})) + (0,4 * TC * (\frac{SP\ 2}{RP\ 2})^{0.5})$$

where:

- TC is the capital cost figure from the above-mentioned article
- SP1 and RP1 (which stand for *scaling parameter n.1* and *reference parameter n.1*) are the volumetric flow rate of gas in the acid wash column in the cases analyzed in this thesis and in the article, respectively.
- SP2 and RP2 are the initial concentration of ammonia in the flue gases entering the column.

This formula has been adopted to try to express the dependence of the cost of the column from the quantity of ammonia that is captured in it, in addition to the volumetric flow of gas treated.

In the cases in which sulfuric acid needs to be purchased from the market it has been considered a price of 210 \$ / short ton of 100% pure sulfuric acid [116].

Table 34: main parameters associated to the acid wash plant

	Unit of Measure	Chilled	Cooled
Ammonia concentration in the gas entering the acid wash	ppmv	3050	4100
Ammonia at the stack	ppmv	10	10
Sulfuric acid consumption	kg h2so4 /sec	2,936	4,025
Ammonium Sulphate production	Kg / sec	3,95	5,42
Steam extraction from the LP turbine	MW(thermal)	omitted	omitted
Auxiliary electric consumptions	MW(el)	omitted	omitted
Capital cost acid wash column	k\$	11358	12754
Capital cost acid tank, separation equipment, drying and screening equipment	k\$	omitted	omitted
Revenues for ammonium sulphate sale	k\$	omitted	omitted

5 RESULTS & DISCUSSION

In this section the main results of the techno-economic analysis are presented and discussed. In Table 35 the most significant indexes characterizing the thermodynamic performance of the various power plant configurations are listed. The following results are in particular worth of attention.

The power production from the steam turbine decreases in the carbon-capture cases with respect to the reference situation due to the need of extracting steam for solvent regeneration. The reduction is relatively limited in the cases operating with chilled ammonia, while it is much more severe in the other cases.

Predictably, the increased possibilities for thermal recovery from the syngas in the carbon capture cases, which were discussed in section 4.1.2, do not compensate for the amount of steam that need to be extracted for the regeneration of the solvent.

Among the different plant configurations, those using cooled-ammonia as CO₂ capture solvent show the greatest overall thermodynamic efficiency, with Case 4 displaying the highest value. The plant configurations integrated with the Cansolv-type CO₂ capture unit show instead the poorest performance, due to a very significant reduction in the power produced from the steam turbine. The chilled-ammonia configurations are in an intermediate position: while on the one hand the power production from the steam turbine is elevated, on the other hand the electric consumptions of the CCS auxiliaries are also very significant.

Table 35: summary of the performance of the power plants investigated. The results from the previous literature [75][118] are shown in brackets to facilitate the comparison

	Reference Case (without Capture)	Case 1 claus Cooled	Case 2 claus Chilled	Case 3 claus Cansolv	Case 4 sulfur Cooled	Case 5 sulfur Chilled	Case 6 cansolv Sulfur
Steam turbine gross power	485,89	424,98 (433,1)	473,48 (474,1)	401,17	432,86 (450,1)	481,43	409,46
Heat rejected at the condenser	652,76	228,18 (285,8)	304,72 (337,8)	327,95	242,38 (331,6)	319,03	342,73
Gross electric power output, MW	1039,09	956,18	1004,68	932,37	964,06	1012,63	940,66
IGCC plant Auxiliaries	107,202754	78,71	78,81	78,25	79,2	79,65	78,75
CCS plant Auxiliaries	0	69,36 (66,4)	145,14 (142,6)	81,99	69,36 (66,46)	145,14	81,99
Net power output	931,88	808,12 (817,7)	780,73 (781,9)	772,14	815,5 (834,1)	787,85	779,93
Net electric LHV Efficiency %	48,00	41,21	39,81	39,37	41,59	40,17	39,77
Specific emissions Kg CO ₂ / Mwh	660,22	87,64	90,71	91,72	86,85	89,89	90,81
SPECCA MJ/kg CO ₂	-	2,15	2,7	2,89	2,02	2,55	2,72

The power plant configurations that are integrated with a sulfuric acid plant show a power production from the steam turbine which is around 8-9 MW higher than the performance achieved with a Claus-type sulfur recovery unit. This figure is roughly 50% lower than the one determined by Bonalumi et. al[118].

It has been judged that this discrepancy can be partially explained by considering that in the current analysis some steam produced in the sulfuric acid plant is supposed to be used for the ZLD auxiliaries. As a consequence, the amount of steam available for power production is diminished of 3-4 MWs(net-electric) than it would otherwise be.

The IGCC-related auxiliary electric consumptions of the carbon-capture cases are significantly lower than the reference case. This is mainly a consequence of the elimination of the Air separation unit and the lock hoppers nitrogen compressors. The increase of work of the air boosters does not compensate for the elimination of such components.

On the contrary, made exception for small variations in the cooling tower auxiliaries, there are no noteworthy differences between the capture cases for what concerns the IGCC-related auxiliaries.

In general some of the results obtained concerning Cases 1, 2 and 4 are not in perfect agreement with the results found by Bonalumi et al. in the articles [75] and [118], which are represented in brackets within Table 35.

In particular the value of the steam turbine total power production in Case 1 has been found 8 MWs lower than expected. It has been verified numerically that this can be explained from the fact that different values of the pressure in the condenser (0,068 vs 0,04 bar) have been considered.

On the other hand, such a worsening of the performance has not been observed in Case 2, meaning that, according to the analysis performed, the figure for the steam turbine gross power production of Case 2 would have been higher than predicted by Bonalumi if the condensing pressure chosen for the calculations had been 0,04 instead of 0,068 bar.

The results have been found different also in regard to the amount of heat rejected from the condenser, which is significantly lower than expected. This point is difficult to explain in particular since it means that there are not only differences in the way in which the energy has been distributed between the turbine and the condenser, but also that the amount of energy entering the cycle has been itself undercounted.

Results in terms of LCOE

Given the uncertainty regarding the value to attribute to some parameters, the final results of the techno-economic analysis have been obtained by exploring different scenarios in which the uncertain parameters are floated around a relatively large interval of values.

In Table 36 it is presented a detail of the LCOE determined for the different power plant configurations under the simplified hypothesis that there are no differences between the various cases for what concerns the capital cost of the CO₂ capture plant and the sulfur recovery unit.

Table 36: LCOE results in the hypothesis that there are no differences in the capital cost of the CO₂ capture unit and the sulfur recovery plant between the various cases

	Case 1 claus Cooled	Case 2 claus Chilled	Case 3 claus Cansolv	Case 4 sulfur Cooled	Case 5 sulfur Chilled	Case 6 cansolv Sulfur
Capital \$/Mwh	79,88	82,97	83,47	80,04	83,19	83,78
Fixed	28,15	29,26	29,51	28,21	29,32	29,55
Variable	32,81	30,29	22,52	24,42	23,99	22,61
Fuel	17,77	18,39	18,6	17,61	18,23	18,41
Total LCOE	158,62	160,92	154,11	150,29	154,74	154,37

It can be easily recognized that the LCOE tends to drastically diminish for the ammonia-based carbon capture configurations (Cases 1-4 and 2-5) when the integration with the sulfuric acid plant is taken into account. Such reduction in the LCOE is a consequence of the possibility, offered by the presence of the sulfuric acid plant, to eliminate the purchasing of sulfuric acid from an external supplier.

At the same time this simplification is useful because it allows to verify that, regarding from the differences in the capital cost between the various components, Case 4 still emerges as the most

interesting option. This is essentially a consequence of the fact that Case 4 has a better electric efficiency than the other alternatives, although variable costs for ammonia replacement are higher.

In Table 37 the detail for the LCOE is represented without any simplification regarding the sulfur recovery plant. In particular the capital cost of the sulfuric acid plant is set to a value which is roughly equal to 38% of the capital cost of the Claus unit, reflecting the results of the capital cost calculations presented in the methodology section.

Table 37: LCOE results in the hypothesis that there are no differences in the capital cost of the CO2 capture unit between the various cases

	Case 1 claus Cooled	Case 2 claus Chilled	Case 3 claus Cansolv	Case 4 sulfur Cooled	Case 5 sulfur Chilled	Case 6 cansolv Sulfur
Capital \$/Mwh	79,88	82,97	83,47	76,72	79,75	80,31
Fixed	28,15	29,26	29,51	27,09	28,17	28,39
Variable	32,81	30,29	22,52	23,77	23,31	21,93
Fuel	17,77	18,39	18,6	17,61	18,23	18,41
Total LCOE	158,62	160,92	154,11	145,21	149,47	149,05

As a consequence, it can be appreciated a significant reduction in the LCOE of Cases 4-5-6 with respect to the situation described in the previous paragraph.

Finally, in Table 38 it is presented the result of the economic analysis as it was found, without any simplification hypothesis.

In particular, according to the methodology presented in section 4.2.3, the capital costs of the Cansolv unit and the chilled NH3-based unit have been set to a value which is respectively 40% and 20% higher than that of the cooled NH3-based CO2 capture equipment.

As a result, the cooled-ammonia configurations (Case 1 and Case 4) show a lower LCOE than the alternatives.

It is possible to notice that the cooled-ammonia configuration clearly emerges as the best solution only if it is considered the integration with a sulfuric acid plant. On the contrary, if it is considered the integration with a Claus-type sulfur recovery unit, the advantage of cooled ammonia with respect to the Cansolv solvent is minimum.

Table 38: LCOE results (no simplification hypothesis)

	Case 1 claus Cooled	Case 2 claus Chilled	Case 3 claus Cansolv	Case 4 sulfur Cooled	Case 5 sulfur Chilled	Case 6 cansolv Sulfur
Capital \$/Mwh	79,88	85,05	87,88	76,72	81,82	84,67
Fixed	28,15	29,96	30,99	27,09	28,86	29,85
Variable	32,81	30,7	23,39	23,77	23,72	22,79
Fuel	17,77	18,39	18,6	17,61	18,23	18,41
Total LCOE	158,62	164,12	160,87	145,21	152,65	155,74

These results can be justified from a physical point of view considering that the Cansolv solvent is based on traditional amines, which are reportedly corrosive and difficult to handle. In addition, the Cansolv technology employs a lower regeneration pressure than the CO2 capture units based on ammonia, meaning that the regeneration column needs to be of greater volume.

Both of these considerations point towards the conclusion that the Cansolv unit effectively features a more expensive equipment than the cooled-ammonia CO₂ capture system.

It is more difficult to assess the correctness of the relative cost difference that has been found between the equipment of the Cansolv unit and the chilled-ammonia CO₂ capture system.

In this analysis it was determined that the equipment of the chilled-ammonia system is less expensive than the equipment of the Cansolv plant. It is believed that this result should be taken with particular caution due to the low accuracy of the methodologies employed to obtain the capital cost quotation of the chilled ammonia system.

A similar set of considerations has to be made also for what concerns the difference in the capital costs that has been determined between the Claus unit and the sulfuric acid plant. It was determined that the Claus plant is almost three times more expensive than the sulfuric acid plant. It has been reasoned that a similar deviation in the capital costs might find a possible explanation in the fact that there is a long technical experience with the production of sulfuric acid, but this result still does not seem completely convincing.

Although the full results have not been reported here, since the ammonia slip in the Cases 1,2,4,5 has been calculated with a very approximated method, it has been determined how the LCOE of these Cases would change if the cost for ammonia replacement was to be increased of 50%. In detail, it is found that the LCOE of Cases 1 and 4 would increase of 2,5 \$/Mwh, and the LCOE of Cases 2 and 5 would increase of about 1,85 \$/Mwh. It can be therefore concluded that the consumption of ammonia due to the ammonia slip effect is a serious issue, particularly for what concerns the cases 1 and 4, but there seems to be sufficient margin to say that Case 4 would still be more economically attractive than Case 6 even if the cost of ammonia substitution was higher than expected. To this point it has to be also remembered that the price of ammonia has been assumed in this analysis at 400 \$/ton, which is relatively high if compared to the values chosen by other authors.

Another parameter that has been investigated is the pressure at which the flow of steam for the regeneration of the solvent is extracted from the LP turbine. While it has been generally assumed in this study that steam is extracted from the LP turbine without pressure losses at respectively 1.5 bar (in the Cases 1 and 4) and 0.34 bar (in the Cases 2 and 5), it has been verified how the performance would change if the steam was extracted at 2 bar from the crossover pipe connecting the IP and LP sections of the steam turbine. In particular it has been found that the LCOE of Cases 1 and 4 would decrease of about 2 \$/Mwh, therefore evidencing a certain robustness of the results previously obtained.

On the other hand the Cases 2 and 5 would run into a much more significant loss in terms of performance (10 \$/Mwh) if the steam was not extracted at the minimum possible pressure.

In all the cases analyzed the interest on capital has been found as the most important component of the LCOE, with the CO₂ capture unit accounting typically for 15-20% of the total capital investment (TPC).

It has been found moreover that, besides coal purchasing, the other most relevant cost items are the miscellaneous materials for maintenance, the property taxes and the costs for solvent replacement. These costs constitute alone nearly 90% of all costs in each configuration.

On an opposite note, the results of the economic analysis indicate that the labor cost has a very limited incidence; in particular it has been verified that if the cost of labor of a certain power plant configuration was to be doubled, the LCOE would increase of only 1,5 \$/Mwh.

In other words, while it may be argued for example that the operation of recovery and handling of the solids in the fertilizer plant (in the Cases 4 and 5) requires a more conspicuous labor force than

Case 6, it seems unlikely that the results might be significantly affected by a small difference (5-10 operators per turn) in the number of power plant operators between the different cases analyzed.

The costs for water withdrawal and slag disposal have been found also not particularly relevant in the overall economic account, although it needs to be remembered that the presence of a stable supply of water is a *conditio-sine-qua-non* for the operation of the plant, regardless of the price of water.

To this purpose it has to be noticed that the power plant configurations featuring carbon capture require a 35-40% increase in the water withdrawal with respect to the reference case, as it is shown in Table 39.

Table 39: raw water withdrawal necessary for the functioning of each power plant configuration

	Unit of measure	Reference Case (without Capture)	Case 1 claus Cooled	Case 2 claus Chilled	Case 3 claus Cansolv	Case 4 sulfur Cooled	Case 5 sulfur Chilled	Case 6 cansolv Sulfur
Raw water withdrawal	1000 Gallons /day	4442,11	5912,11	6011,9	5768,37	6149,82	6251,06	6004,02

This is a consequence of the fact that the cooling tower load increases considerably due to the need of rejecting large quantities of heat from the CO₂ capture unit. The cooling tower load, in turn, implies larger evaporative and blowdown water losses, that need to be refilled.

All the cases analyzed in this study involve a non-negligible amount of process water discharge. In particular the overwhelming share of the process water is discharged from the cooling tower in order to control the build-up of salts in the cooling water. The cooling tower blowdown is supposed to be re-integrated to the water supply after being subject to reverse osmosis and unspecified further treatments in the water treatment facility.

Other sources of potentially polluted process water, although in much smaller volumes, are the operations of quenching of the slag and contact cooling of the exhausts downstream to the HRSG, both of which produce an aqueous blowdown. Finally, Cases 1 and 4 involve the discharge of a certain quantity of water mixed to ammonia, which, while it is believed to be small, unfortunately has not been precisely quantified in this study.

In conclusion, for what concerns the production of polluted process water, it appears that there are no dramatically relevant reasons to prefer one of the power plants investigated over the others. However this point should be revisited after a more precise assessment of the blowdown from the CO₂ capture unit in the Cases 1 and 4.

To understand to possible impact on the LCOE of the revenues obtained by selling the by-products of the plant it has been determined that a conservative estimate for the historical market prices of sulfuric acid and ammonium sulphate is around 150 \$/ton for both the commodities [117]. A market value of 150 \$/ton seems to be also acceptable for what concerns elemental sulfur, which is produced when the plant is integrated with a Claus-type sulfur recovery unit.

If these prices are hold true, it should be possible to obtain a revenue of 41 mln \$ /year from the sale of ammonium sulphate in Case 4, as shown in Table 40.

If only sulfuric acid is produced, as it happens in Case 6, a revenue around 31 mln \$ /year should be produced from the sale of the acid.

Table 40: summary of the revenues obtainable through the sale of by-products

	Unit of measure	Case 1 claus Cooled	Case 2 claus Chilled	Case 3 claus Cansolv	Case 4 sulfur Cooled	Case 5 sulfur Chilled	Case 6 cansolv Sulfur
Production of Elemental Sulfur	Metric Ton/day	227,664	227,664	227,664	-	-	-
Gross production of Sulfuric acid	Metric Ton/day	-	-	-	696	696	696
Sulfuric acid consumption	Metric Ton/day	696	507	~ 0	696	507	~ 0
Net production of sulfuric acid (available for sale)	Metric Ton/day	-	-	-	0	189	696
Ammonium Sulphate production	Metric Ton/day	-	-	-	938	683,6	0
Price	\$ / ton	150					
Income	Mln \$ / year	9,97	9,97	9,97	41,08	38,21	30,48
LCOE before all incomes	\$/Mwh	158,62	164,12	160,87	145,21	152,65	155,74
Income-per-unit of energy produced	\$/Mwh	1,76	1,82	1,84	7,18	6,92	5,57
LCOE after all incomes are included	\$/Mwh	156,85	162,29	159,02	138,02	145,72	150,16

In comparison, based on the quantity of sulfur available in the coal, it should not be possible to obtain a revenue much higher than 10 mln \$ /year from the sale of elemental sulfur alone.

In order to contextualize these numbers it might be useful to consider that 44 mln \$ / year would be the cost for carbon storage and sequestration if it was assumed, as it is done in the NETL reports, that the service of carbon sequestration was sold from an external supplier at the price of 10 \$ / ton of carbon dioxide.

In other words, if the revenues associated with the sale of sulfuric acid and ammonium sulphate are considered, there seems to be excellent prospects of further improving the economic performance for the power plant configurations that are integrated with a sulfuric acid plant, as it is shown in Table 40.

In conclusion, it might be guessed with good confidence that, although the capital and operative cost for the recovery of ammonium sulphate have not been precisely determined in the section dedicated to the analysis of the acid wash column, the revenues generated from the sale of ammonium sulphate should easily offset any cost related to the recovery of the salts, even in the eventuality that the recovery had to prove expensive in relative terms.

As a final point, in addition to all the cases presented so far, in Table 41 it is presented a detail of the LCOE of all the various configurations in the hypothesis of a cold external climate.

In a simplified way, it is imagined that the only effect of a cold climate on the power plant is to reduce substantially the auxiliary electric consumptions used for refrigeration in the chilled-ammonia configurations. In detail, it is assumed that the auxiliaries of the Case 2 and Case 5 are reduced of 50 megawatts.

Table 41: LCOE results in the hypothesis of cold climate

	Case 1 claus Cooled	Case 2 claus Chilled	Case 3 claus Cansolv	Case 4 sulfur Cooled	Case 5 sulfur Chilled	Case 6 cansolv Sulfur
Capital \$/Mwh	79,88	79,81	87,88	76,72	76,82	84,67
Fixed	28,15	28,12	30,99	27,09	27,1	29,85
Variable	32,81	28,83	23,39	23,77	22,28	22,79
Fuel	17,77	17,29	18,6	17,61	17,14	18,41
Total LCOE	158,62	154,06	160,87	145,21	143,35	155,74

Looking at the results presented in Table 41 it can be concluded that the chilled-ammonia system should be more carefully considered for all those situations in which a stream of cold water can be pumped from the surrounding environment to the power plant to produce low-cost refrigeration.

6 CONCLUSIONS

During the course of this work it has been provided a description and a techno-economic analysis of a series of air blown IGCC power plants featuring different possible strategies for sulfur recovery and post combustion CO₂ capture.

In detail, three different technologies for carbon capture have been compared, ranging from a solvent based on amines, which have been traditionally used in the petrochemical industry for the removal of acid gases, to more innovative concepts.

Six main different scenarios have been tested, indicating that the least expensive route to reduce the carbon emissions for the studied reference plant is to adopt a cooled-ammonia system integrated with a sulfuric acid plant.

In this configuration sulfuric acid is produced out of the stream of hydrogen sulfide separated from the syngas in the AGR unit. Then it is employed for the so-called acid wash operation, in which the flue gases from the CO₂ capture block are purified from ammonia before leaving the system from the stack. A by-product of this operation is ammonium-sulphate, which is a salt that can be used as a fertilizer.

During the analysis, some uncertainties have arisen regarding the correct value to associate to certain parameters or capital costs. In particular, due to the approximate methodology adopted some perplexities have arisen regarding the quantification of the mixture of ammonia and water purged from the cooled-ammonia CCS plant configurations.

It has been also calculated a figure for the capital cost of the sulfuric acid plant which is suspiciously low if compared with a similar Claus-type sulfur recovery unit.

As a general note, a final word on these matters is left to future more accurate studies, however a series of sensitivity tests has shown that the conclusions of the economic analysis determined in this work are quite robust, since they do not change in an appreciable way if the parameters affected by uncertainty are varied within a relatively wide range.

Set aside all the considerations related to the thermodynamic efficiency, a factor that seems very relevant in determining the economic attractiveness of CO₂ capture with cooled ammonia is the lower cost of equipment that is associated to this solution with respect to other solvents based on amines. More in detail, it seems that a central element of the economic case for cooled ammonia is the reported possibility of employing carbon steel instead of stainless steel to realize large portions of the equipment.

The proposed integration between the AGR unit and the CO₂ capture unit, which represents a peculiarity specific to some of the power plant configurations studied in this work, has been commented with reference to the design of an existing oxygen-blown IGCC power plant (Tampa IGCC plant), showing that there are successful technical precedents among commercial plants for a design similar to the one studied in this work.

At the same time it has been proposed a tentative design regarding the section of the plant dedicated to the operation of acid wash and recovery of the ammonium-sulphate crystals.

The proposed process has not been characterized in a quantitative manner due to the lack of data regarding the properties of the mother liquor.

However it is believed that the revenues associated to the sale of ammonium sulphate should largely compensate the costs related to the recovery of the crystals, with a likely positive impact on the balance sheet of the plant.

Finally, it needs to be remembered that this analysis is intended to be a technological assessment, and therefore has been performed by neglecting a series of cost elements that were deemed non-essential for the purpose of making a comparison between different CO₂ capture systems. For example, in this analysis it has been provided no account of the auxiliary electric consumptions associated to the sulfur recovery unit, and no syngas extraction has been considered for the operation of coal drying. Moreover, very little characterization has been provided of the water treatment units that are supposed to recycle the streams of process water exiting the plant. It is believed that the performance of the plant would likely decrease of some percentage points if these factors were properly taken into account.

At the same time some of the elements neglected might be associated to a small improvement of the efficiency of the plant if they were considered.

For example it might be conceivable to imagine an increase in the potential of high temperature heat recovery from the decomposition furnace boiler when the presence of ammonia in the sour gases from the ZLD unit is included in the calculations.

A second aspect might be related to the potentiality for heat recovery of the plant, which, according to the present analysis, is not completely exploited. In particular it appears that there are unexploited possibilities for heat recovery near 100°C from the LTHR section of the syngas coolers, and, to a lesser extent, from the CO₂ compressor and the sulfuric acid plant.

A potential recipient of the quantities of unrecovered heat might be the acid wash unit. It has been indeed noticed that a reheat of the exhaust gases up to around 50°C might be necessary prior to the acid wash column.

Another option might be that of recovering the waste heat with an absorption chiller in order to reduce the electricity consumptions needed for refrigeration purposes.

A precise analysis of these aspects is left to a future and more accurate study, however the most important question that should be answered revolves around a more precise estimate of the cost of the air blown MHI gasifier.

As a matter of fact, Case 4 emerges as the best alternative among those studied in this work, however, due to the uncertainty in the cost of the gasifier, the results should be taken with great care when they are read as absolute numbers.

APPENDIX A

In this section it is provided a representation of the TQ diagrams for three heat exchangers of the reference plant that has been deemed particularly important. The blue line represents the temperature of the hot fluid. The inferior line represents the temperature of the cold fluid. The intermediate line is the maximum temperature achievable given a pinch temperature of 10°C.

Figure A-1: T-Q diagram of the five heat exchangers constituting the Low Temperature Heat Recovery section of the syngas cooling train

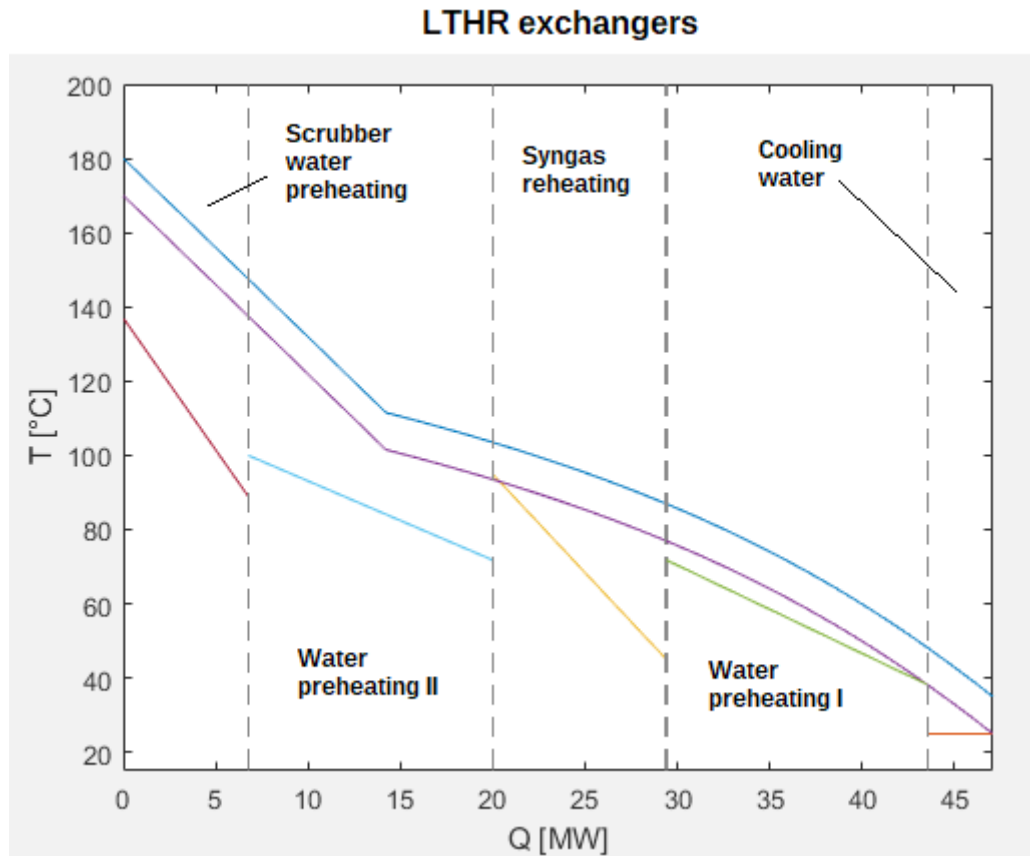


Figure A-2: T-Q diagram of the HRSG
HRSG pipes

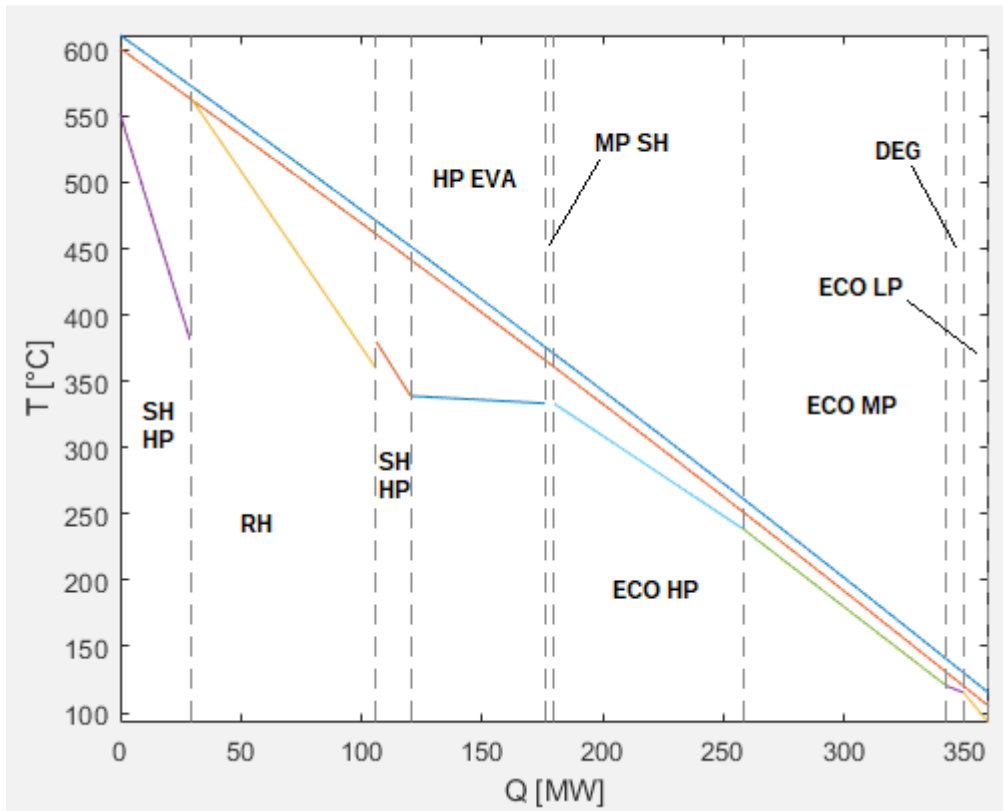
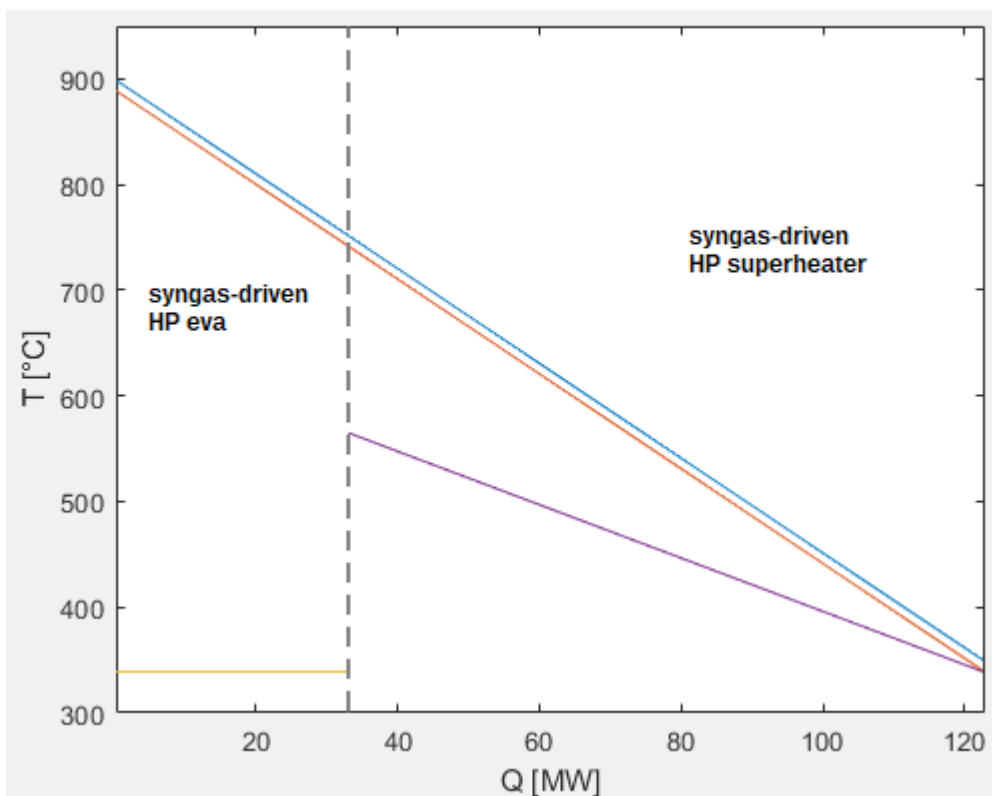


Figure A-3: T-Q diagram of the high temperature syngas cooling exchangers
SYNGAS COOLERS (from 900°C to 350°C)



APPENDIX B.1

According to the modellization provided in the NETL guidelines [73] the coal gasifier is the most expensive piece of equipment involved in an IGCC power plant, accounting for about 40-50% of the total equipment cost of the plant (without capture). For this reason it has been decided to develop a model of the MHI gasifier in order to extrapolate its cost from available estimates of other gasification technologies presenting similar technical details.

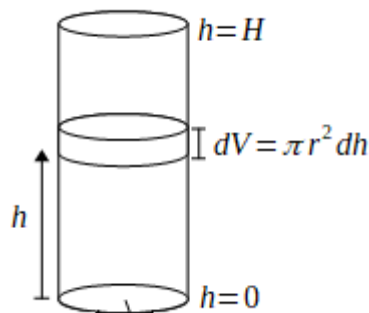
In particular the Shell oxygen blown gasifier has been chosen as a reference benchmark since it features some similar characteristics to the MHI technology. The most similar features between the two systems are the fact that they both adopt a dry lock hopper alimentation system and a water walls internal jacket [76][121].

Gasifier model simulation

The gasifier is modelled as a cylinder having H, r, t as height, radius and wall thickness respectively. The flow is supposed to be a gaseous mixture with suspended entrained coal particles going from bottom to top of the cylinder.

All the variable are supposed to be function only of the h coordinate. The cylinder is supposed to be high ($H \gg 4r$).

Figure B-1: gasifier model



If we consider dV as the volume of flow that goes through a section of the cylinder in an infinitesimal time $d\tau$, by definition of volumetric flow we can write:

$$\dot{Q} = \frac{dV}{d\tau}$$

or, alternatively, we can reorganize the equation so to express the time $d\tau$ necessary for the flow to go through the volume dV

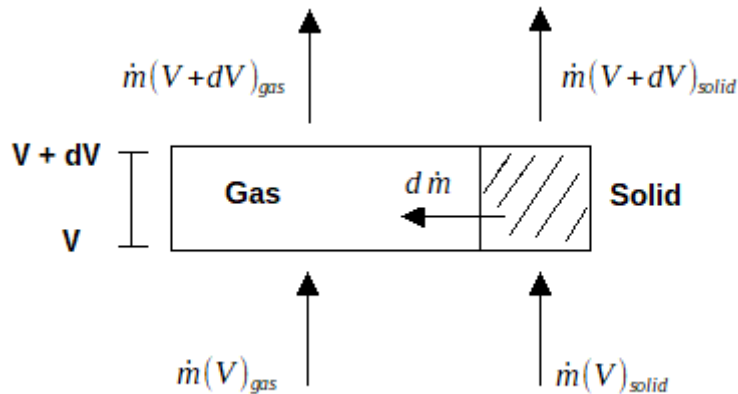
$$d\tau = \frac{dV}{\dot{Q}}$$

Integrating from bottom to top of the cylinder we obtain the total residence time spent by the mixture in the gasifier:

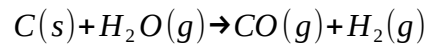
$$\tau_{tot} = \int_0^H \frac{dV}{\dot{Q}}$$

Now, as shown in Figure B-2, the solid-gas flow inside the reactor can be modelled as the juxtaposition of two single phase parallel flows (so called *pseudo-phases*) reciprocally exchanging mass and energy.

Figure B-2: mass balance on an infinitesimal volume dV

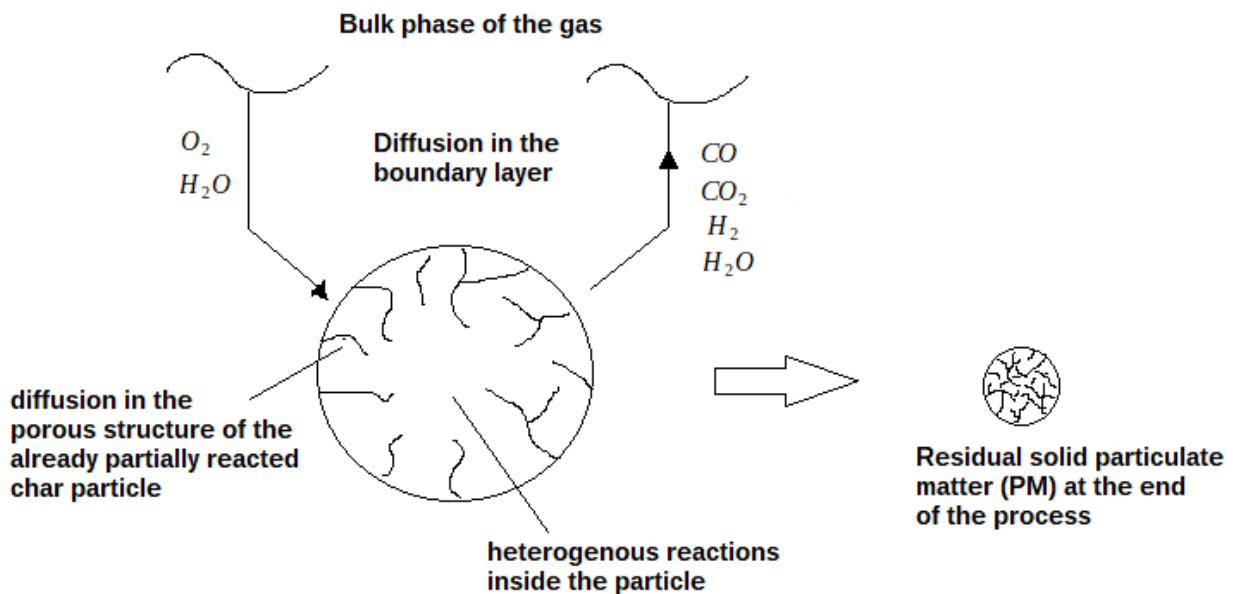


In particular it is imagined that, while the inter-phase flow is always bi-directional, in every section of the reactor there is ultimately a net transfer of mass from the solid to the gaseous phase. The solid carbon particles are progressively converted into gaseous species along the vertical axis of the cylinder due to the following reactions.



At the top section of the gasifier the solid phase is supposed to be almost completely disappeared, made exception for small entrained PM particles. The process of shrinkage of the particles is shown in Figure B-3.

Figure B-3: qualitative description of the mass exchange process around a coal particle



Being the density of a gas extremely smaller than that of a solid, the solid phase can be treated as occupying a negligible amount of space in the cylinder if compared to the gas. In other words, for each section of the cylinder it is true that:

$$\dot{Q}(V)_{tot} \simeq \dot{Q}(V)_{gas}$$

Therefore, if the equation of continuity $\dot{Q}(h)_{gas} = \frac{\dot{m}(h)_{gas}}{\rho(h)_{gas}}$ is taken into consideration, and the density of the gaseous phase is expressed with the ideal gas law, the following expression for the total residence time of the flow in the gasifier is obtained:

$$\tau_{tot} = \frac{\pi r^2}{R} \int_0^H \frac{P(h) * MM(h)_{mix}}{T(h) * \dot{m}(h)_{gas}} dh \quad (\text{with } R=8,314 \text{ kJ/kmol*K})$$

Since the MHI and Shell gasifiers are both entrained flow reactors we are going to assume that the total residence time is designed to be more or less the same for the two technologies.

$$\tau_{tot}(MHI) \simeq \tau_{tot}(shell)$$

that is to say:

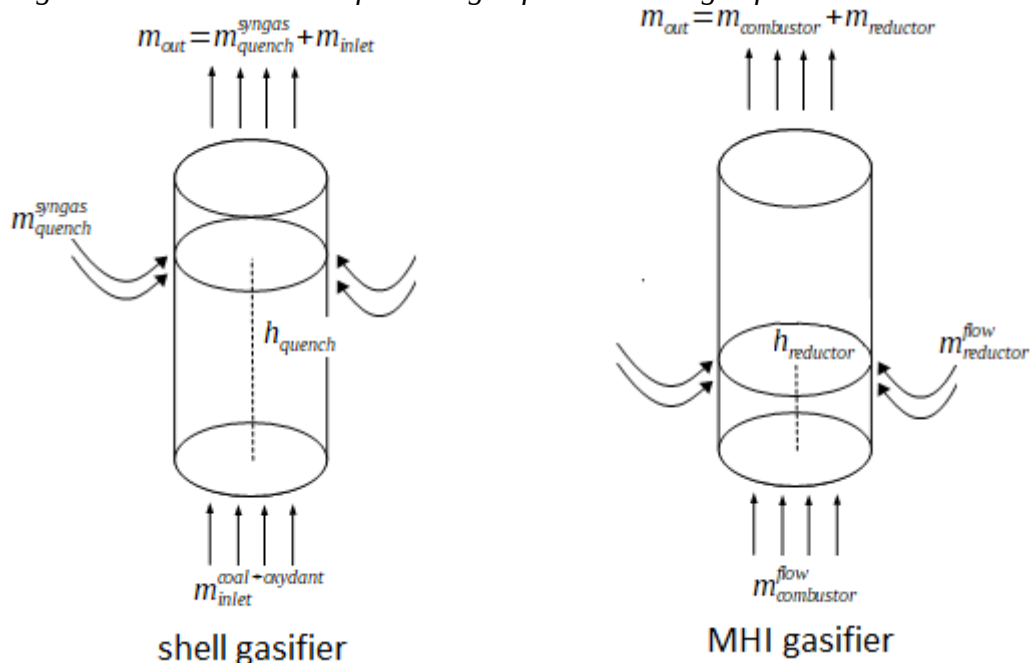
$$\frac{\pi r_{MHI}^2}{\pi r_{shell}^2} \simeq \frac{\int_0^H \frac{P(h) * MM(h)_{mix}}{T(h) * \dot{m}(h)_{gas}} \Big|_{shell} dh}{\int_0^H \frac{P(h) * MM(h)_{mix}}{T(h) * \dot{m}(h)_{gas}} \Big|_{MHI} dh} \quad (1)$$

This equation allows us to evaluate the relative volume of the cylinder that is necessary for a MHI gasifier with respect to a Shell gasifier to have the same permanence time (provided that the expression of temperature, pressure, molecular mass and mass flow is known along the reactor). In the following paragraphs the expected behavior each of these functions is commented.

Mass flow

The mass flow of gas within the MHI gasifier and the Shell gasifier is not constant. As shown in the images 1 and 2 in each technology there are different points of the reactor where additional mass is introduced.

Figure B-4: modelization of a Shell gasifier vs a MHI gasifier



Regarding the MHI gasifier, it is here assumed (the same way of Lozza et al [122]) that 60% of the coal feed is introduced in the reductor. It has not been possible to obtain numerical data about the relative space occupied by the reduction and combustion stages, however a sensitivity analysis (here not reported for sake of brevity) has shown that small variations of this parameter don't affect significantly the residence time. Hence it is assumed $h_{reductor} = 0,38 H$ with H being the total height of the cylinder.

In the Shell gasifier all the feed is supposed to be introduced at the bottom of the column, however a flow of colder syngas is recirculated and introduced near the gasifier outlet in order to obtain a temperature compatible with the subsequent stage of cooling in an heat exchanger.

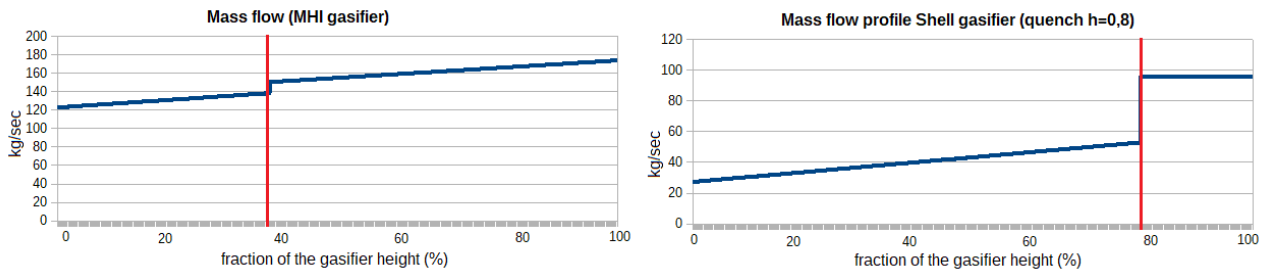
The position of the quench in the reactor is not known. Having a sensitivity analysis has suggested that variations of this parameter do not dramatically change the overall result, it has been chosen to perform the calculation with the hypothesis that the quench stage is in the last fifth of the reactor.

$$h_{low}^{quench} = 0,8 H$$

Additionally, the mass of the gaseous phase flowing in the reactor changes along the vertical coordinate due to the progressive conversion of the solid phase. In this analysis it is made the simple hypothesis that coal is completely converted and the conversion happens linearly along the axial coordinate.

The mass flow profile of the various configurations analyzed are shown in Figure B-5.

Figure B-5: mass flow profile vs height coordinate for a MHI and Shell gasifier (according to the simulation performed)



Pressure

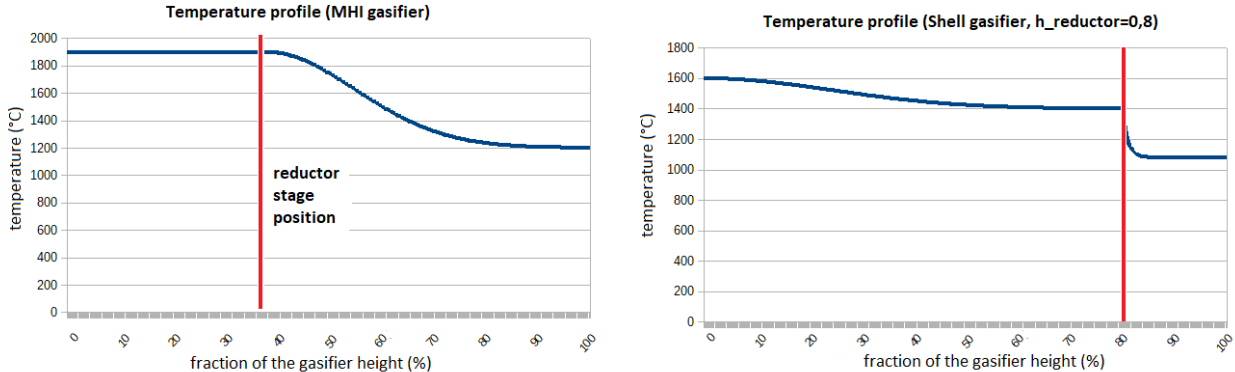
The pressure value adopted in the calculations for the MHI gasifier is $p = 28,06$ bar. The value for the Shell technology is 42,4 bar. It is assumed that there are no pressure losses and there is a constant value across the cylinder.

Temperature

Drawing on the publicly available data [121], the temperature profile of the Shell gasifier is supposed to be an exponential distribution ranging from the inlet value of 1600°C to the outlet value of 1400°C. After being quenched the syngas exits the system with a temperature of about 1080°C.

The temperature profile adopted for the MHI gasifier is the one proposed by the NETL [76]: the temperature in the combustor is 1900°C, and it gradually decreases until 1200°C in the reductor.

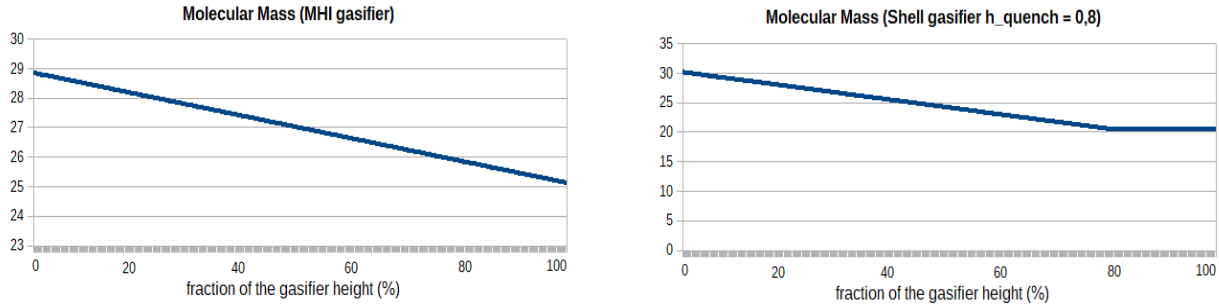
Figure B-6: temperature profile vs height coordinate for a MHI and Shell gasifier



Molecular mass

Since the concentration profile of the different chemical species within the reactor is not known, it has been decided to simplify the problem by making the hypothesis that the average molecular mass of the gaseous phase varies linearly from the inlet to the outlet of the reactor.

Figure B-7: molecular mass of the syngas in the gasifier vs height coordinate



Result

With the hypothesis just discussed, it can be easily determined that

$$\frac{\pi r_{MHI}^2}{\pi r_{shell}^2} \approx \frac{\int_0^H \frac{P(h) * MM(h)_{mix}}{T(h) * \dot{m}(h)_{gas}} \Big|_{shell} dh}{\int_0^H \frac{P(h) * MM(h)_{mix}}{T(h) * \dot{m}(h)_{gas}} \Big|_{MHI} dh} = 5,02 \quad \text{or alternatively} \quad \pi r_{MHI}^2 = 5,02 \pi r_{shell}^2$$

which means that $V_{MHI} = 5,02 V_{shell}$ and $S_{MHI} = 2,24 S_{shell}$.

Wall thickness

The wall thickness t necessary for a cylinder of diameter D to bear an internal pressure of P (barg) is estimated with the formula (employed from [69]).

$$t = \frac{(P+1)D}{2 * E * S - 1,2(P+1)} + CA$$

where:

CA = 0,00315 m - corrosion allowance

E = 0,9 weld efficiency

S = 944 bar – maximum allowable stress for carbon steel

It is important to point out that this expression is accurate only when the walls of the vessel are operating at a temperature equal or lower than 300°C (condition that is supposed to be verified here).

The results obtained are:

$$\text{if } r_{MHI} = 2,24 r_{shell} \rightarrow t_{MHI} = 1,41 t_{shell}$$

which means that the total steel (indicated with M) employed to manufacture the MHI gasifier can be estimated in the following manner:

$$M_{MHI} = S_{MHI} * t_{MHI} = 2,24 S_{shell} * 1,41 t_{shell} = 3,17 M_{shell}$$

Cost of the MHI gasifier

In the previous pages a rough estimate was found for the amount of metal that a Mitsubishi air blown gasifier requires for the construction of its walls in comparison with an oxygen blown gasifier adopting the Shell design.

Now, if metal is used as a scaling parameter, the formula to adopt in order to obtain the N-th of a kind (NOAK) capital cost of the MHI gasifier is:

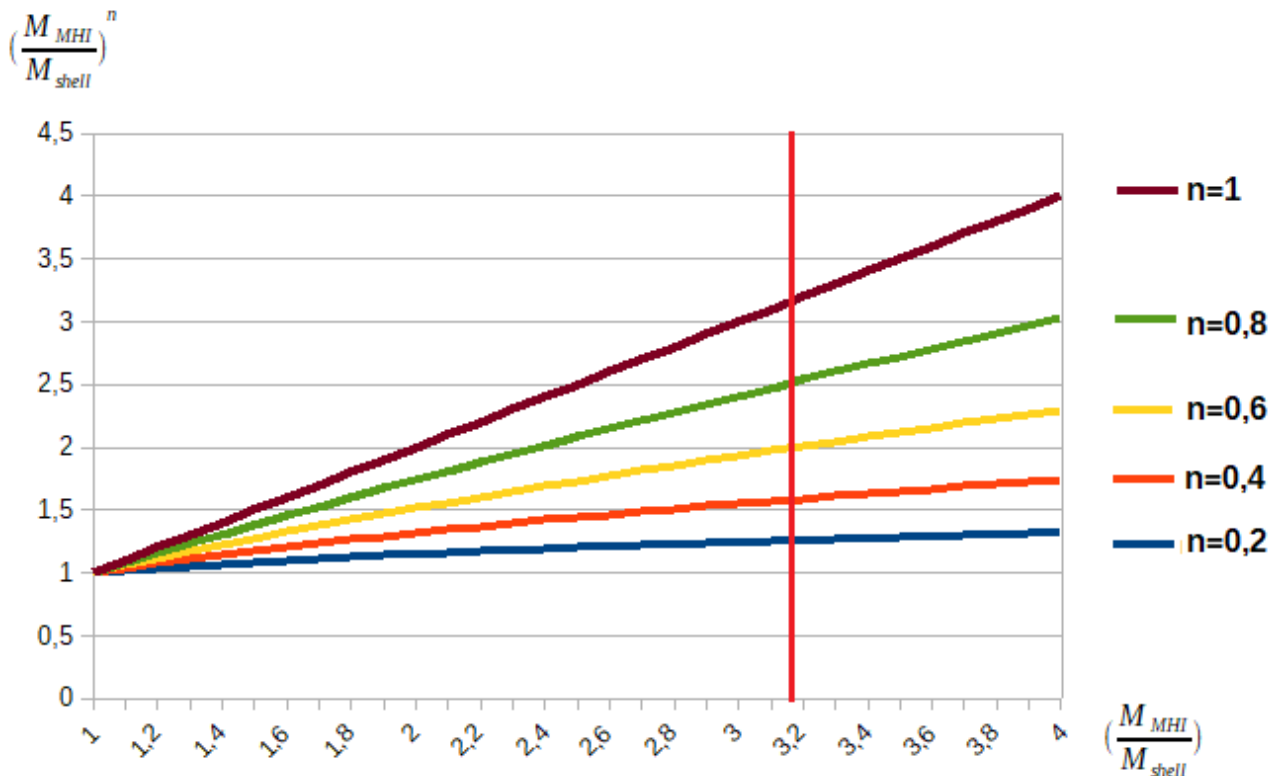
$$cost_{MHI} = cost_{shell} * \left(\frac{M_{MHI}}{M_{shell}} \right)^n$$

where the ratio in the brackets equals 3.17, while the scaling exponent n is unknown. The NOAK capital cost of the equipment for the shell technology is 544 706 k\$.

As evident from the Figure B-8, the choice of the scaling exponent n has a critical influence over the result of the present calculation. In other words a slight variation in the choice of the parameter n might easily result in a variation of hundreds of millions of \$ in the capital cost estimated.

Lacking additional data, it has been put $n = 0,6$ according to the so-called "two-thirds rule" [69], however the final result should be taken more as a general tendency of the system rather than an accurate economic evaluation. In order to obtain a greater accuracy of the result a more complicated modellization is ultimately called for in future works.

Figure B-8: graphic representation of the influence of the scaling parameter n on the analysis on the cost of the gasifier:



APPENDIX B.2

In this section it is explained the procedure through which the bare erected cost has been estimated for various items constituting the reference case power plant.

Air Booster

In the reference case the air booster is a compressor of 14,9 MW net power output (for each train). The capital cost estimate for this item has been developed by modelling the compressor as a parallel of 5 single-stage centrifugal compressors erogating a net power of 2980 kW each. The capital cost C_p^0 of each compressor has been estimated with the empirical equation (from Turton[69]).

$$\log_{10}(C_p^0) = 2,2897 + 1,3604 * \log_{10}(A) - 0,1027 * \log_{10}^2(A)$$

where A is the net power transferred to the fluid, which equals to 2980 kW for each item. The figure obtained in this way is expressed in 2001 dollars and needs to be converted into 2019 dollars. This is done by means of the CEPCI index

$$C_p^0(2019) = \left(\frac{CEPCI(2019)}{CEPCI(2001)} \right) * C_p^0(2001) = 1,54 * C_p^0(2001)$$

The Bare Erected Cost is found with the following empirical expression (from Turton[69])

$$C_{BM} = C_p^0(2019) * (B_1 + B_2 * F_M * F_p) \quad \text{with } B_1=1,89 \quad B_2=1,35 \quad F_M=1 \quad F_p=1$$

Syngas-fuelled steam superheater & syngas regenerative heat exchangers

The cost of the syngas-driven steam superheaters (1 for each gasification train) has been estimated with the empirical expression for fixed tube HXs [69]:

$$\log_{10}(C_p^0) = 4,3247 - 0,303 * \log_{10}(A) + 0,1634 * \log_{10}^2(A)$$

where A is the required exchange area expressed in square metres. The heat exchange area has been determined with the expression:

$$A = \frac{\dot{Q}}{U * LMTD} \quad \text{with } \dot{Q} = 89,5 \text{ MW}, \quad LMTD = 62,714^\circ \text{C}$$

The heat transfer coefficient U has been set at $30 \frac{W}{m^2 \cdot ^\circ C}$ according to an heuristic expression[69]

for gas-gas heat exchangers. The exchange area determined is 47 580 square metres. Being the empirical equation valid only within a limited range of values for A, the cost of the heat exchanger has been determined by making the hypothesis, for the calculations, that several identical heat exchangers are set in parallel. In particular the equipment cost C_p^0 has been found by supposing that 48 heat exchangers with 991,25 square metres of surface area are set in parallel.

The bare erected cost has been calculated with the expression

$$C_{BM} = C_p^0(2019) * (B_1 + B_2 * F_M * F_p) \quad \text{with } B_1=1,63; \quad B_2=1,66; \quad F_M=2,7; \quad F_p=1,1$$

where the pressure and material factors have been set by keeping into consideration an internal steam pressure of 144 bar and a tube bundle in stainless steel on both sides [69]. The final result has been adjusted for inflation with the CEPCI index.

Figure B-9: syngas fuelled SH

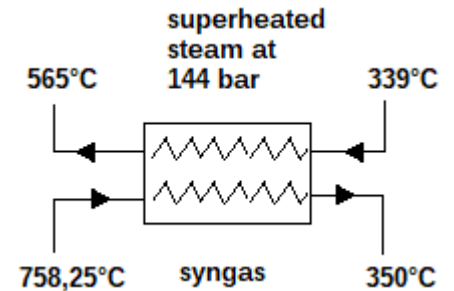
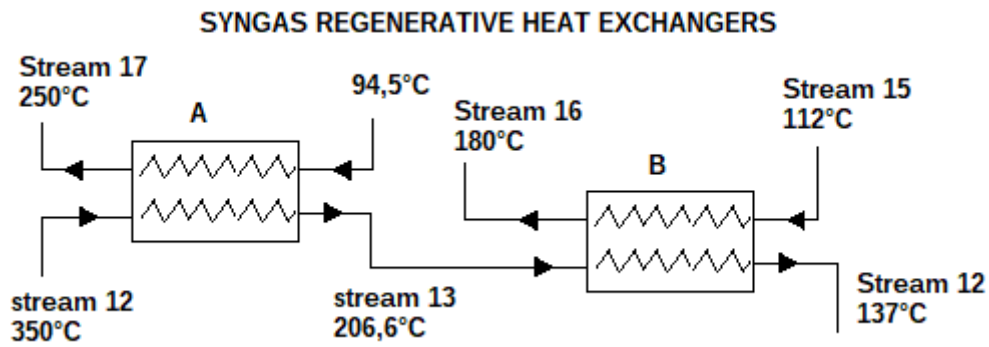


Figure B-10



The cost of the two **syngas regenerative heat exchangers** (shown in the above figure) has been also determined with the same procedure. The pressure factor F_p has been set to 1 since none of the two exchangers need to bear a significant pressure. The material factor F_M has instead been set to 2,7 considering a stainless steel tube bundle.

Low temperature heat recovery exchangers

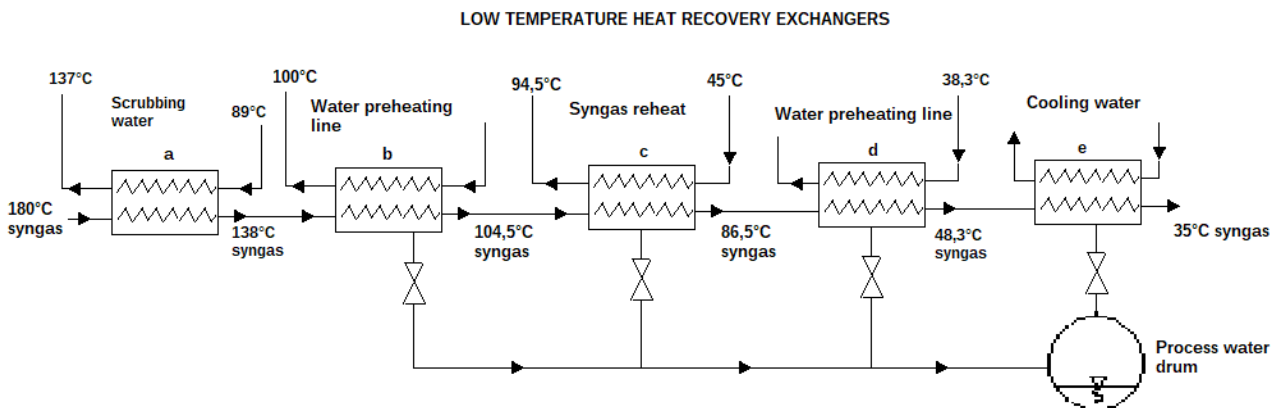
The LTHR section of the gasification train is represented in Figure B-11. Medium temperature syngas from the COS hydrolysis reactor is used to preheat the condensate and reheat the syngas exiting from the AGR section. Since since water from the condenser is at 29°C, while the syngas need to have a temperature of 35°C entering the AGR block, circulating cooling water is also used as cold fluid in the last part of the process.

The cost of the LTHR section has been estimated as the sum of the cost of 5 different heat exchangers (called a / b / c / d / e).

The cost of these exchangers was estimated with the expression[69]:

$$\log_{10}(C_p^0) = 4,3247 - 0,303 * \log_{10}(A) + 0,1634 * \log_{10}^2(A)$$

Figure B-11: schematic representation of the LTHR section of the syngas cooling train



Where A is the heat exchange area expressed in square metres.

According to the already mentioned heuristics, the heat transfer coefficients used for this part of the analysis were originally:

$$U_1 = 60; U_2 = 60; U_3 = 30; U_4 = 60; U_5 = 60 \frac{W}{m^2 \cdot C}$$

However the heat transfer coefficient has been multiplied by two in order to take into consideration the presence of condensing phenomena in the syngas side of the exchangers.
The pressure factor has been set equal to 1, while the material factor has been set to 2 (stainless steel on the hot syngas side).
All costs have been finally converted from 2001 to 2019 US dollars by means of the CEPCI index.

APPENDIX C

In this section are presented the process flow diagrams of the recuperative steam cycle for the cases with carbon capture.

Figure C-1: schematic representation of the bottoming steam cycle for Case 1 – cooled ammonia with a Claus-type sulfur recovery unit

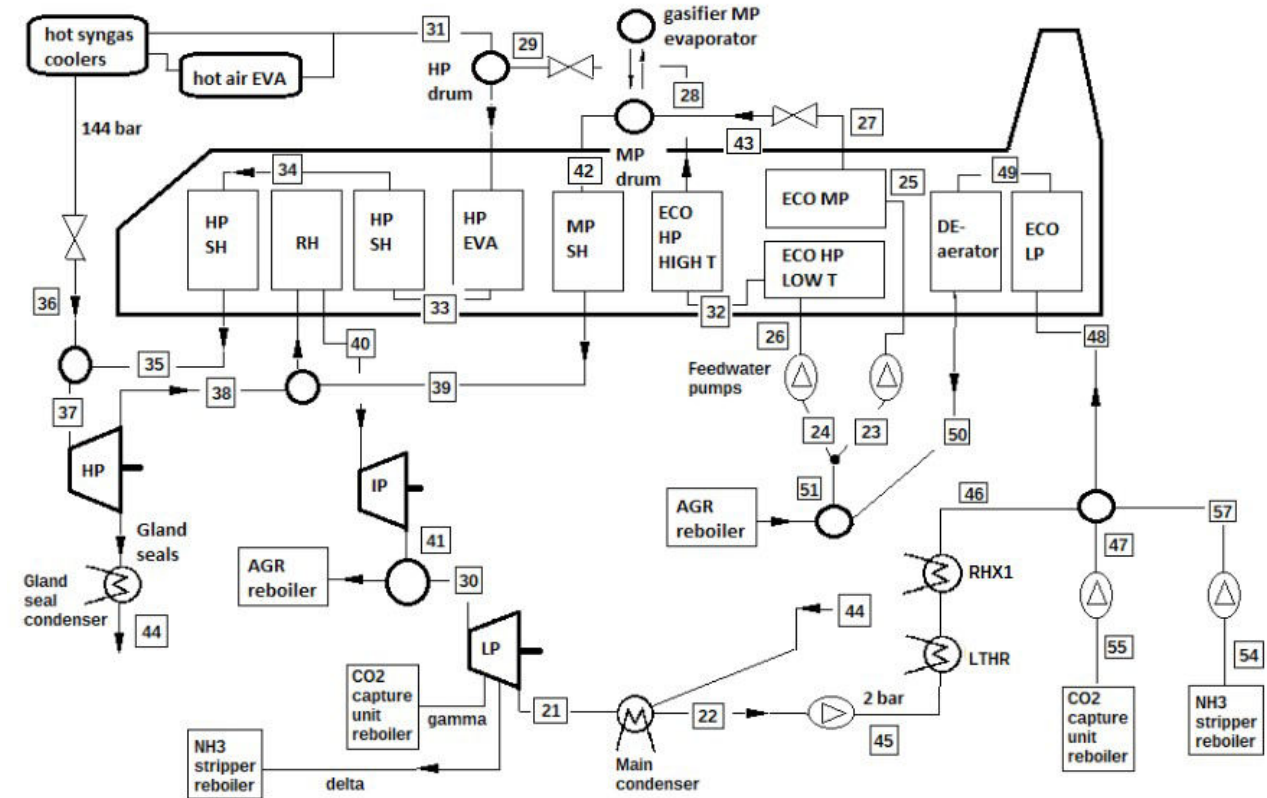


Table C-1: temperature, pressure, enthalpy and mass flow of the main streams of the bottoming steam cycle shown in Figure C-1

	21	22	23	24	25	26	27	28	29
temperature (°C)	38,3	38,3	120,21	120,21	120,62	121,3	238,79	333,42	333,42
pressure (bar)	0,068	0,068	2	2	61	160	61	160	144
enthalpy (KJ/kg)	2381,51	161,11	504,67	504,67	510,577	520,477	1031,787	1544,736	1544,736
mass flow (kg/sec)	102,76	105,86	21,92	321,29	21,92	321,295	21,92	321,295	321,295

	30	31	32	33	34	35	36	37	38
temperature (°C)	191,2	333,42	238,79	339	380	551,5	565	560,6	359,6
pressure (bar)	2	144	160	144	144	132,5	132,5	132,5	36
enthalpy	2853,03	1544,73	1031,78	2628,68	2905,08	3472,74	3508,09	3497,26	3125,81
mass flow (kg/sec)	308,91	222,93	321,29	98,36	98,36	98,36	222,93	321,29	318,19

	39	40	41	42	43	45	46	47	48
temperature (°C)	359,6	563,9	191,2	244,19	238,79	38,3	90	111,3	103,7
pressure (bar)	36	33,12	2	36	36	2	2	2	2
enthalpy	3125,817	3598,2	2853,039	2802,484	1031,787	161,306	375,73	467,368	434,42
mass flow (kg/sec)	21,92	340,11	340,11	21,92	21,92	105,86	105,86	189,95	156

	49	50	51	54	55	57	gamma	delta
temperature (°C)	119,5	120,21	120,21	102,8	111,3	102,8	164,02	138,17
pressure (bar)	2	2	2	1,12	1,5	2	1,5	1,12
enthalpy	501,22	504,677	504,677	430,918	466,868	431,798	2801,16	2752,04
mass flow (kg/sec)	312,015	312,015	343,215	16,2	189,95	16,2	189,95	16,2

Figure C-2: schematic representation of the bottoming steam cycle for Case 2 – chilled ammonia with a Claus-type sulfur recovery unit

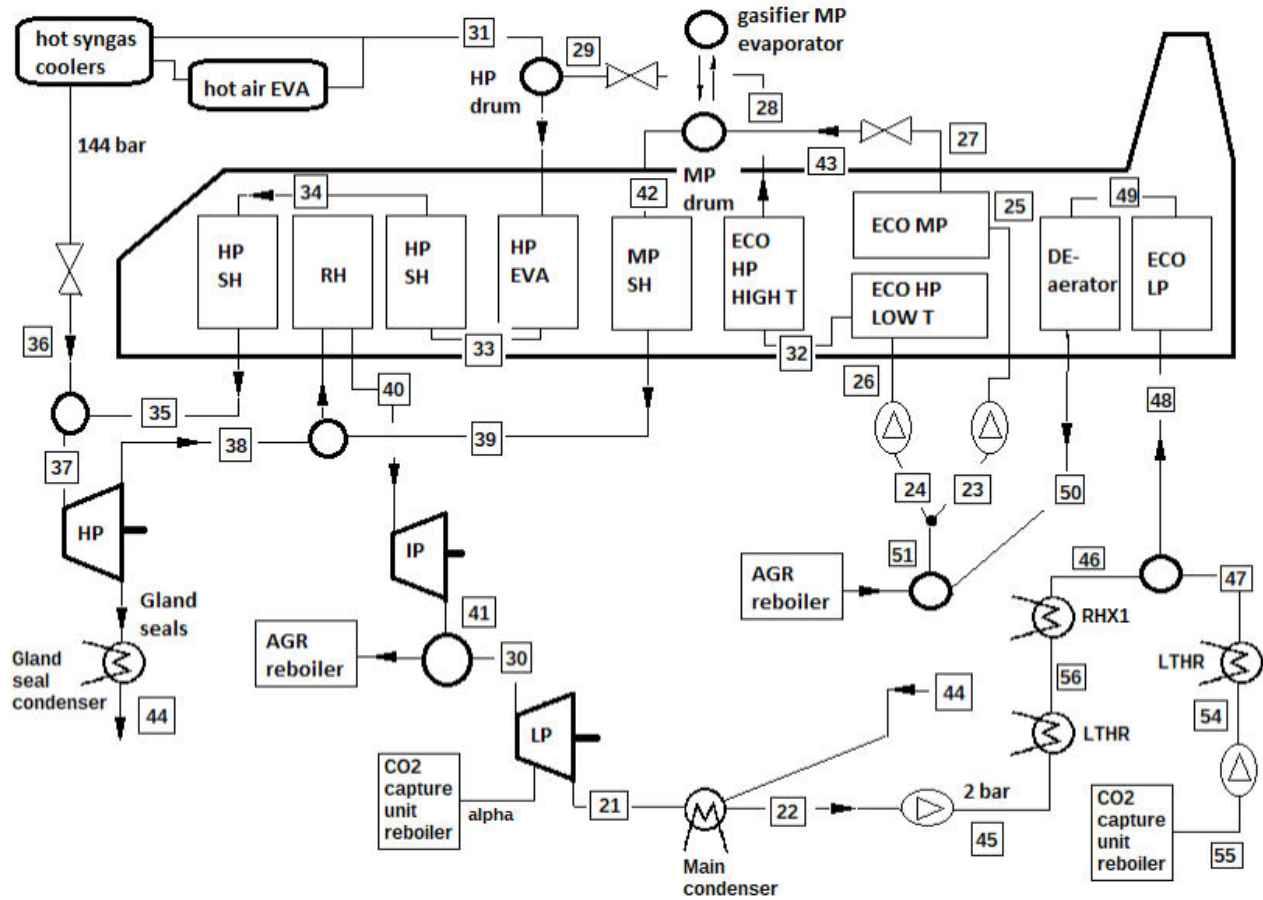


Table C-2: temperature, pressure, enthalpy and mass flow of the main streams of the bottoming steam cycle shown in Figure C-2

	21	22	23	24	25	26	27	28	29
temperature (°C)	38,3	38,3	120,21	120,21	120,62	121,3	238,79	333,42	333,42
pressure (bar)	0,068	0,068	2	2	61	160	61	160	144
enthalpy (KJ/kg)	2381,51	161,11	504,677	504,677	510,577	520,477	1031,787	1544,736	1544,736
mass flow (kg/sec)	137,45	140,55	21,92	319,86	21,92	319,86	21,92	319,86	319,86

	30	31	32	33	34	35	36	37	38
temperature (°C)	191,2	333,42	238,79	339	380	551,5	565	560,6	359,6
pressure (bar)	2	144	160	144	144	132,5	132,5	132,5	36
enthalpy	2853,03	1544,73	1031,78	2628,68	2905,08	3472,74	3508,09	3497,37	3125,81
mass flow (kg/sec)	307,48	222,93	319,86	96,93	96,93	96,93	222,93	319,86	316,76

	39	40	41	42	43	44	45	46	47
temperature (°C)	359,6	563,9	191,2	244,19	238,79	15	38,3	90	107,75
pressure (bar)	36	33,12	2	36	36	1,01	2	2	2
enthalpy	3125,81	3598,2	2853,03	2802,48	1031,78	72,6	161,3	378,93	451,24
mass flow (kg/sec)	21,92	338,68	338,68	21,92	21,92	3,1	140,55	140,55	170,02

	48	49	50	51	54	55	56	alpha
temperature (°C)	99,8	115,75	120,21	120,21	72,51	72,51	71,25	72,51
pressure (bar)	2	2	2	2	2	0,3474	2	0,3474
enthalpy	418,52	485,619	504,677	504,677	303,618	303,53	298,258	2582,02
mass flow (kg/sec)	310,58	310,58	310,58	341,78	170,024	170,024	140,55	170,024

Figure C-3: schematic representation of the bottoming steam cycle for Case 3 – Cansolv CO₂ capture with a Claus-type sulfur recovery unit

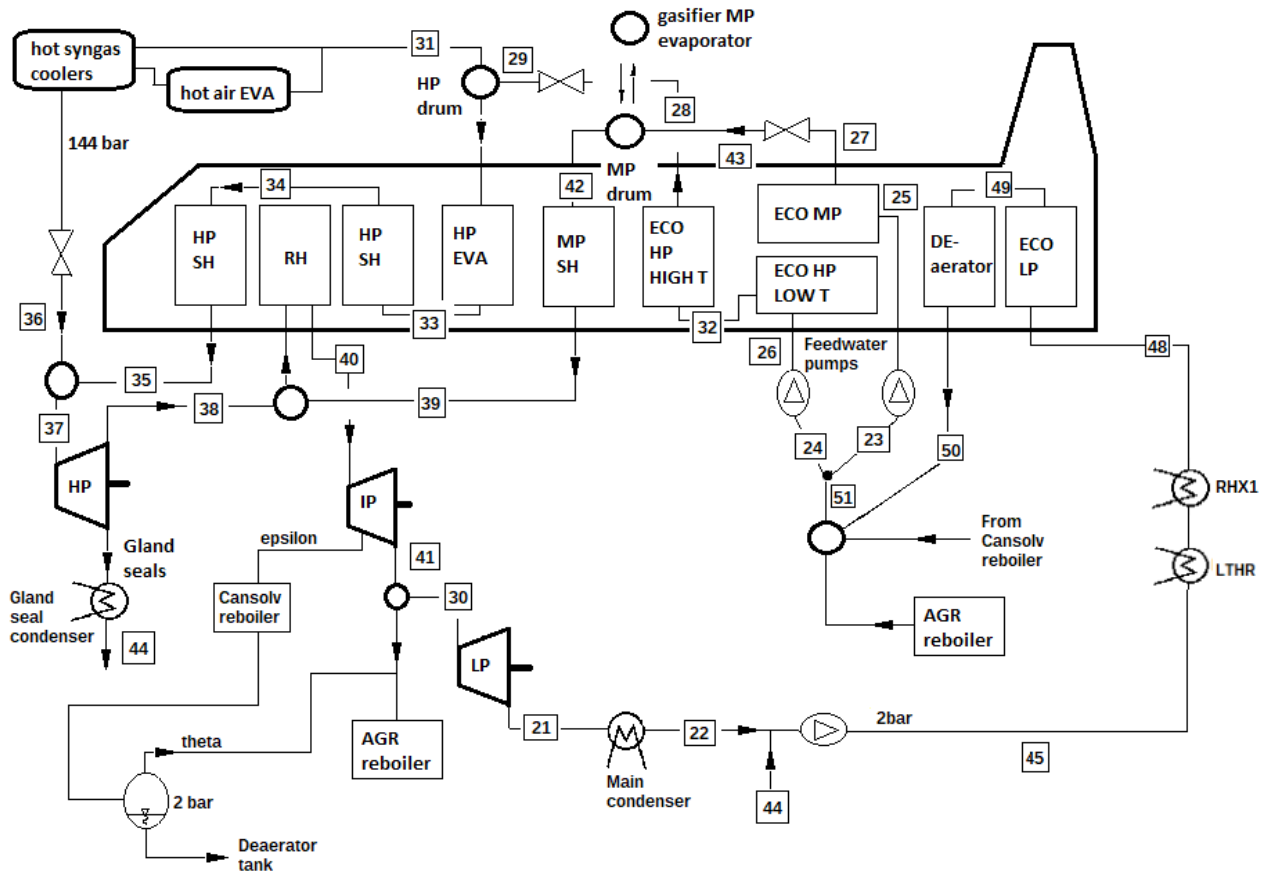


Table C-3: temperature, pressure, enthalpy and mass flow of the main streams of the bottoming steam cycle shown in Figure C-3

	21	22	23	24	25	26	27	28
temperature (°C)	38,3	38,3	120,21	120,21	120,62	121,3	238,79	333,42
pressure (bar)	0,068	0,068	2	2	61	160	61	160
enthalpy (KJ/kg)	2381,51	161,11	504,677	504,677	510,577	520,477	1031,78	1544,73
mass flow (kg/sec)	147,69	150,79	21,92	321,29	21,92	321,295	21,92	321,295

	29	30	31	32	33	34	35	36
temperature (°C)	333,42	191,2	333,42	238,79	339	380	551,5	565
pressure (bar)	144	2	144	160	144	144	132,5	132,5
enthalpy	1544,73	2853,03	1544,73	1031,78	2628,681	2905,088	3472,74	3508,09
mass flow (kg/sec)	321,295	147,69	222,93	321,295	98,365	98,365	98,365	222,93

	37	38	39	40	41	42	43	44
temperature (°C)	560,6	359,6	359,6	563,9	191,2	244,19	238,79	15
pressure (bar)	132,5	36	36	33,12	2	36	36	1,01
enthalpy	3497,26	3125,81	3125,81	3598,2	2853,03	2802,48	1031,78	72,6
mass flow (kg/sec)	321,295	318,195	21,92	340,115	168,795	21,92	21,92	3,1

	45	48	49	50	51	epsilon	theta
temperature (°C)	38,3	85,84	118,5	120,21	120,21	293,6	120,21
pressure (bar)	2	2	2	2	2	5,1	2
enthalpy	1544,73	359,58	497,78	504,677	504,677	3051,16	2706,25
mass flow (kg/sec)	150,79	150,79	150,79	150,79	150,79	171,32	10,79

Figure C-4: schematic representation of the bottoming steam cycle for Case 4– cooled ammonia capture integrated with a sulfuric acid plant

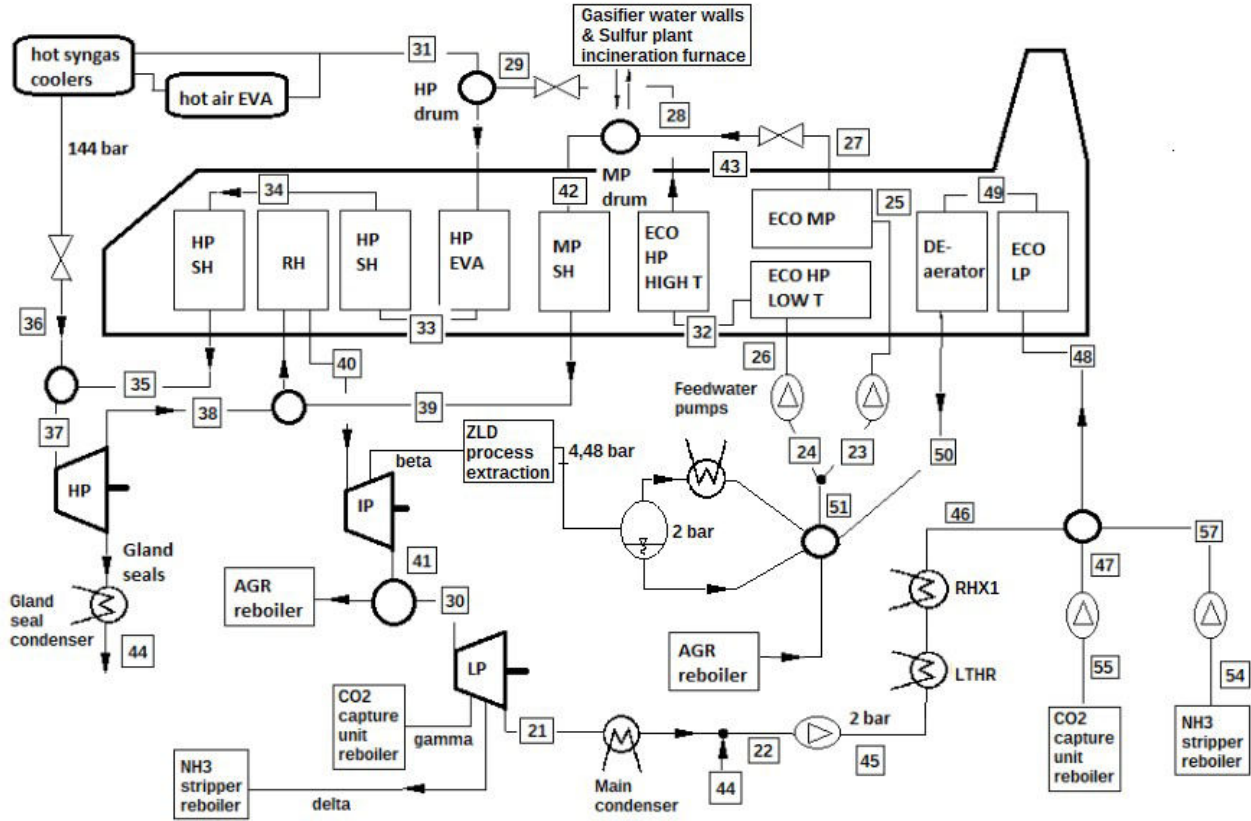


Table C-4: temperature, pressure, enthalpy and mass flow of the main streams of the bottoming steam cycle shown in Figure C-4

	21	22	23	24	25	26	27	28	29
temperature (°C)	38,3	38,3	120,21	120,21	120,62	121,3	238,79	333,42	333,42
pressure (bar)	0,068	0,068	2	2	61	160	61	160	144
enthalpy (KJ/kg)	2381,51	161,11	504,677	504,677	510,577	520,47	1031,78	1544,73	1544,73
mass flow (kg/sec)	109,15	112,25	40,55	313,92	40,55	313,9202	40,55	313,92	313,92

	30	31	32	33	34	35	36	37	38
temperature (°C)	191,2	333,42	238,79	339	380	551,5	565	560,6	359,6
pressure (bar)	2	144	160	144	144	132,5	132,5	132,5	36
enthalpy	2853,03	1544,73	1031,78	2628,68	2905,08	3472,74	3508,09	3497,84	3125,81
mass flow (kg/sec)	315,3	222,93	313,92	90,99	90,99	90,99	222,93	313,92	310,82

	39	40	41	42	43	45	46	47	48
temperature (°C)	359,6	563,9	191,2	244,19	238,79	38,3	90	111,3	103,5
pressure (bar)	36	33,12	2	36	36	2	2	2	2
enthalpy	3125,81	3598,2	2853,03	2802,48	1031,78	161,306	375,73	467,36	433,814
mass flow (kg/sec)	40,55	351,37	346,5	21,92	21,92	112,25	112,25	189,95	318,4

	49	50	51	54	55	57	gamma	delta	beta
temperature (°C)	119	120,21	120,21	102,8	111,3	102,8	164,02	138,17	278,36
pressure (bar)	2	2	2	1,12	1,5	2	1,5	1,12	4,48
enthalpy	499,265	504,67	504,677	430,91	466,868	431,798	2801,16	2752,04	3021,38
mass flow (kg/sec)	318,4	318,4	318,4	16,2	189,95	16,2	189,95	16,2	4,86

Figure C-5: schematic representation of the bottoming steam cycle for Case 5 – Chilled ammonia capture integrated with a sulfuric acid plant

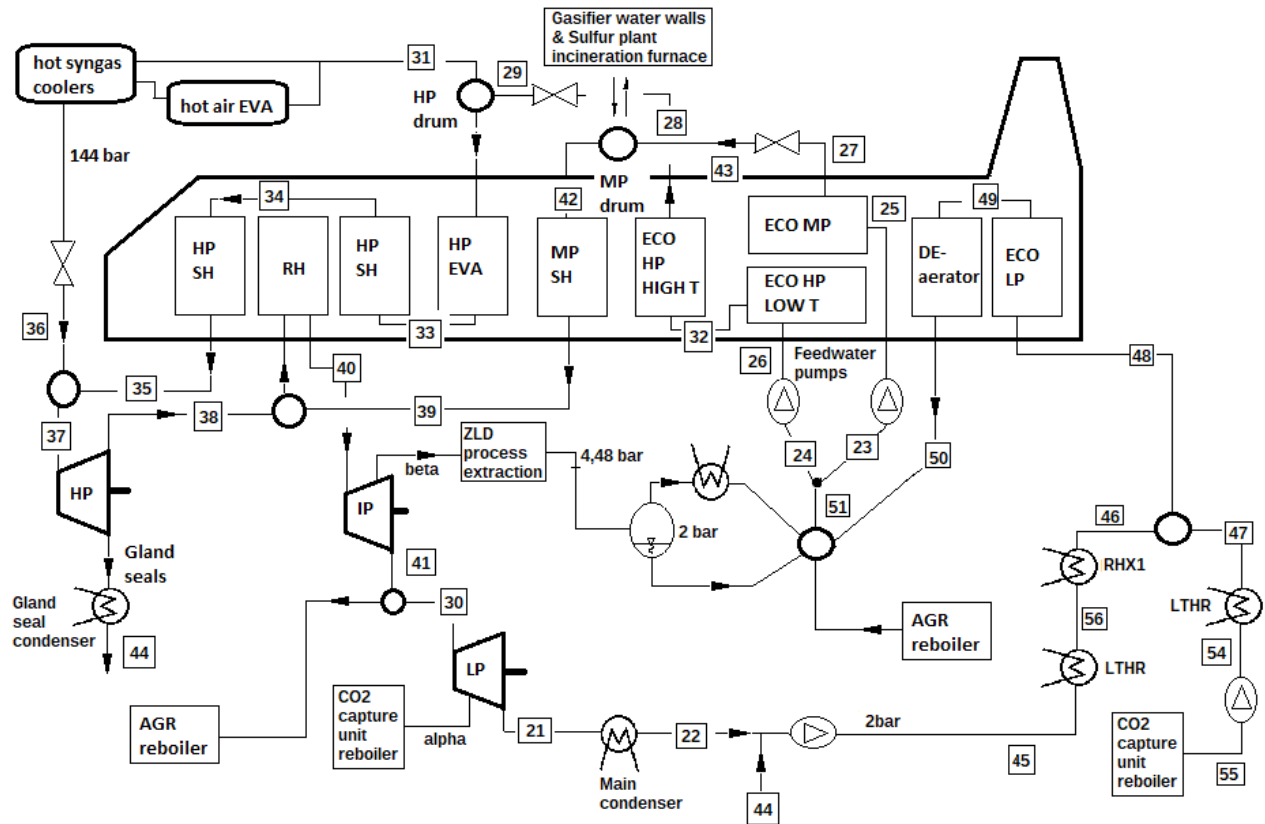


Table C-5: temperature, pressure, enthalpy and mass flow of the main streams of the bottoming steam cycle shown in Figure C-5

	21	22	23	24	25	26	27	28	29
temperature (°C)	38,3	38,3	120,21	120,21	120,62	121,3	238,79	333,42	333,42
pressure (bar)	0,068	0,068	2	2	61	160	61	160	144
enthalpy (KJ/kg)	2381,512	161,11	504,677	504,677	510,577	520,477	1031,787	1544,736	1544,736
mass flow (kg/sec)	143,902	147,002	40,55	312,54	40,55	312,54	40,55	312,54	312,54

	30	31	32	33	34	35	36	37	38
temperature (°C)	191,2	333,42	238,79	339	380	551,5	565	560,6	359,6
pressure (bar)	2	144	160	144	144	132,5	132,5	132,5	36
enthalpy	2853,039	1544,736	1031,787	2628,681	2905,088	3472,748	3508,09	3497,956	3125,817
mass flow (kg/sec)	313,926	222,93	312,54	89,61	89,61	89,61	222,93	312,54	309,44

	39	40	41	42	43	44	45	46	47
temperature (°C)	359,6	563,9	191,2	244,19	238,79	15	38,3	90	108
pressure (bar)	36	33,12	2	36	36	1,01	2	2	2
enthalpy	3125,817	3598,2	2853,039	2802,484	1031,787	72,6	161,306	378,922	452,809
mass flow (kg/sec)	40,55	349,99	345,126	21,92	21,92	3,1	147,002	147,002	170,024

	48	49	50	51	54	55	56	alpha	beta
temperature (°C)	99,85	115,5	120,21	120,21	72,51	72,51	72,08	72,51	278,36
pressure (bar)	2	2	2	2	2	0,3474	2	0,3474	4,48
enthalpy	418,548	484,284	504,677	504,677	305,183	303,53	301,78	2582,02	3021,38
mass flow (kg/sec)	317,026	317,026	317,026	353,09	170,024	170,024	147,002	170,024	4,864

Figure C-6: schematic representation of the bottoming steam cycle for Case 6 – Cansolv CO2 capture

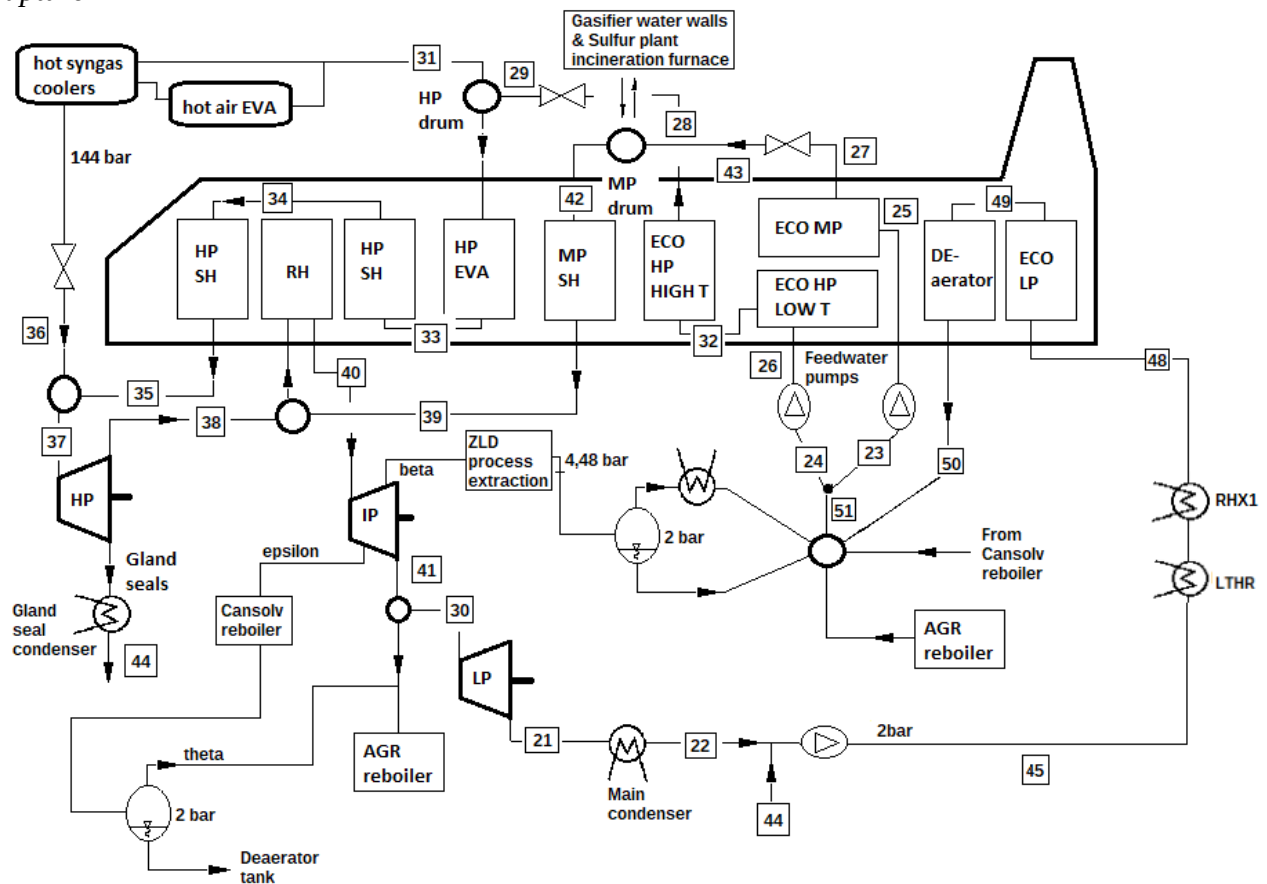


Table C-6: temperature, pressure, enthalpy and mass flow of the main streams of the bottoming steam cycle shown in Figure C-6

	21	22	23	24	25	26	27	28
temperature (°C)	38,3	38,3	120,21	120,21	120,62	121,3	238,79	333,42
pressure (bar)	0,068	0,068	2	2	61	160	61	160
enthalpy (KJ/kg)	2381,51	161,11	504,67	504,67	510,57	520,47	1031,78	1544,73
mass flow (kg/sec)	154,35	157,45	40,55	314,18	40,55	314,188	40,55	314,188

	29	30	31	32	33	34	35	36
temperature (°C)	333,42	191,2	333,42	238,79	339	380	551,5	565
pressure (bar)	144	2	144	160	144	144	132,5	132,5
enthalpy	1544,73	2853,03	1544,73	1031,78	2628,68	2905,08	3472,74	3508,09
mass flow (kg/sec)	314,18	154,35	222,93	314,18	91,25	91,25	91,25	222,93

	37	38	39	40	41	42	43	44
temperature (°C)	560,6	359,6	359,6	563,9	191,2	244,19	238,79	15
pressure (bar)	132,5	36	36	33,12	2	36	36	1,01
enthalpy	3497,82	3125,81	3125,81	3598,2	2853,03	2802,48	1031,787	72,6
mass flow (kg/sec)	314,18	311,08	40,55	351,63	175,45	21,92	21,92	3,1

	45	48	49	50	51	epsilon	theta	beta
temperature (°C)	38,3	85,84	118,5	120,21	120,21	293,6	120,21	278,36
pressure (bar)	2	2	2	2	2	5,1	2	4,48
enthalpy	1544,73	359,58	497,78	504,67	504,67	3051,16	2706,25	3021,38
mass flow (kg/sec)	157,45	157,45	157,45	157,45	157,45	171,32	10,79	4,86

APPENDIX D.1

In this section are listed the different pieces of equipment that have been considered within the reference plant. The results of the methodology used to determine the total capital cost of the plant are also shown.

Table D-1: list of the scaling parameters used to calculate the equipment cost for the different items in the reference case

Item description	scaling parameter	Scaling exponent	Cost of the Reference equipment	Adimensional scaling Ratio
COAL HANDLING				
Coal Receive & Unload	Coal feed rate	0,62	929	1,425
Coal Stackout & Reclaim	Coal feed rate	0,62	3037	1,425
Coal Conveyors & Yard Crush	Coal feed rate	0,62	28971	1,425
Other Coal Handling	Coal feed rate	0,62	4512	1,425
Coal & Sorbent Handling Foundations	Coal feed rate	0,62	81	1,425
COAL PREPARATION & FEED				
Coal Crushing & Drying	Coal feed rate	0,66	2242	1,425
Prepared Coal Storage & Feed	Coal feed rate	0,66	6890	1,425
Dry Coal Injection System	Coal feed rate	0,66	8792	1,425
Miscellaneous Coal Preparation & Feed	Coal feed rate	0,66	680	1,425
Coal & Sorbent Feed Foundation	Coal feed rate	0,66	1655	1,425
FEEDWATER & MISCELLANEOUS BOP SYSTEMS				
Feedwater System	Feedwater flow to the HP drum	0,71	2213	1,552
Water Makeup & Pretreating	Raw water withdrawal	0,71	4489	1,495
Other Feedwater Subsystems	Feedwater flow to the HP drum	0,71	1144	1,552
Service Water Systems	Raw water withdrawal	0,71	1342	1,495
Other Boiler Plant Systems	Feedwater flow to the HP drum	0,73	297	1,552
Natural Gas Pipeline and Start-Up System	Coal feed rate	0,24	7076	1,425
Waste Water Treatment Equipment	Process water discharge	0,71	6552	1,495
Vacuum Flash, Brine Concentrator, & Crystallizer	Syngas scrubber blowdown	0,76	22582	3,6
Miscellaneous Plant Equipment	Coal feed rate	0,24	15068	1,425
GASIFIER, ASU & ACCESSORIES				
Gasifier & Auxiliaries (MHI)	See appendix B			
Syngas Cooler	Power exchanged in the evaporation section of the syngas coolers	0,33	51993	1,15
Air Separation Unit/Oxidant Compression	O ₂ production & ASU main compressor	0,7	52343	0,2529
Miscellaneous Gasification Equipment	Coal feed rate	0,5	4021	1,425
Low Temperature Heat Recovery	See appendix B			
Flare Stack System	Coal feed rate	0,5	1835	1,425
Major Component Rigging	Coal feed rate	0,5	224	1,425
Gasification Foundations	Coal feed rate	0,5	454	1,425
Syngas driven superheater & regenerative heat exchangers	See appendix B			
Air booster	See appendix B			
SYNGAS CLEANUP				
MDEA	volumetric flow of syngas to the AGR unit	0,7	5162	3,51
Elemental Sulfur Plant	H ₂ S flow to be treated in the Claus unit	0,67	45308	1,829
Carbonyl Sulfide (COS) Hydrolysis	Coal feed rate	0,6	7887	1,425
Fuel Gas Piping	Syngas flow to the CT	0,72	2686	1,182
Gas Cleanup Foundations	H ₂ S flow to be treated in the Claus unit	0,79	209	1,829
COMBUSTION TURBINE & ACCESSORIES				
Combustion Turbine Generator	Syngas flow to the CT	0,66	74945	1,182
Combustion Turbine Accessories	Syngas flow to the CT	0,66	2687	1,182
Compressed Air Piping	Syngas flow to the CT	0,66	510	1,182
Combustion Turbine Foundations	Syngas flow to the CT	0,66	216	1,182
HRSG, DUCTWORK & STACK				
Heat Recovery Steam Generator	HRSG heat duty	0,7	35544	1,29
Heat Recovery Steam Generator Accessories	HRSG heat duty	0,7	12691	1,29
Ductwork	Gas flow to stack (kg/sec)	0,7	1068	1,19
Stack	Gas flow to stack (kg/sec)	0,7	9083	1,19
Heat Recovery Steam Generator, Ductwork & Stack Foundations	Gas flow to stack (kg/sec)	0,7	226	1,19

Table D-2: list of the scaling parameters used to calculate the equipment cost for the different items in the reference case

STEAM TURBINE & ACCESSORIES				
Steam Turbine Generator & Accessories	Steam turbine gross power	0,7	39671	1,614
Steam Turbine Plant Auxiliaries	Steam turbine gross power	0,71	1928	1,614
Condenser & Auxiliaries	Heat exchange area	1	7137	1,399
Steam Piping	Feedwater flow to the HP drum	0,72	7401	1,55
Turbine Generator Foundations	Steam turbine gross power	0,72	301	1,614

COOLING WATER SYSTEM				
Cooling Towers	Cooling tower load	0,72	10457	1,495
Circulating Water Pumps	Circulating water flow rate	0,72	1377	1,495
Circulating Water System Auxiliaries	Circulating water flow rate	0,67	9483	1,495
Circulating Water Piping	Circulating water flow rate	0,61	5355	1,495
Make-up Water System	Raw water withdrawal	0,63	545	1,495
Component Cooling Water System	Circulating water flow rate	0,64	193	1,495
Circulating Water System Foundations	Circulating water flow rate	0,59	443	1,495

SLAG RECOVERY & HANDLING				
Slag Dewatering & Cooling	Slag output	0,64	1848	1,238
Gasifier Ash Depressurization	Slag output	0,64	1047	1,238
Cleanup Ash Depressurization	Slag output	0,64	471	1,238
Ash Storage Silos	Slag output	0,64	1061	1,238
Ash Transport & Feed Equipment	Slag output	0,64	409	1,238
Miscellaneous Ash Handling Equipment	Slag output	0,64	59	1,238
Ash/Spent Sorbent Foundation	Slag output	0,64	416	1,238

Item description	scaling parameter	Scaling exponent	Cost of the Reference equipment	Adimensional scaling Ratio
ACCESSORY ELECTRIC PLANT				
Generator Equipment	Steam turbine gross power	0,54	2843	1,614
Station Service Equipment	Auxiliary electric consumption	0,45	3566	0,857
Switchgear & Motor Control	Auxiliary electric consumption	0,45	21516	0,857
Conduit & Cable Tray	Auxiliary electric consumption	0,45	95	0,857
Wire & Cable	Auxiliary electric consumption	0,45	1305	0,857
Protective Equipment	Auxiliary electric consumption	0	241	0,857
Standby Equipment	Total plant gross power	0,48	865	1,352
Main Power Transformers	Total plant gross power	0,71	6569	1,352
Electrical Foundations	Total plant gross power	0,7	76	1,352

INSTRUMENTATION & CONTROL				
Integrated Gasification and Combined Cycle Control Equipment	Auxiliary electric consumption	0,13	660	0,857
Combustion Turbine Control Equipment	Auxiliary electric consumption	0,13	652	0,857
Steam Turbine Control Equipment	Auxiliary electric consumption	0,13	602	0,857
Other Major Component Control Equipment	Auxiliary electric consumption	0,13	1165	0,857
Signal Processing Equipment	Auxiliary electric consumption	0,13	904	0,857
Control Boards, Panels & Racks	Auxiliary electric consumption	0,13	262	0,857
Distributed Control System Equipment	Auxiliary electric consumption	0,13	9484	0,857
Instrument Wiring & Tubing	Auxiliary electric consumption	0,13	472	0,857
Other Instrumentation & Controls Equipment	Auxiliary electric consumption	0,13	1059	0,857

IMPROVEMENTS TO SITE				
Site Preparation	Bare Erected Cost sections 1-12	0,08	416	1,651
Site Improvements	Bare Erected Cost sections 1-12	0,08	1888	1,651
Site Facilities	Bare Erected Cost sections 1-12	0,08	2947	1,651

BUILDINGS & STRUCTURES				
Combustion Turbine Area	CT gross power	0	314	1,182
Steam Turbine Building	Steam turbine gross power	0,06	2784	1,614
Administration Building	Steam turbine gross power	0,04	886	1,614
Circulation Water Pumphouse	Circulating water flow rate	0,46	135	1,495
Water Treatment Buildings	Raw water withdrawal	0,71	297	1,495
Machine Shop	Steam turbine gross power	0,02	486	1,614
Warehouse	Steam turbine gross power	0,02	379	1,614
Other Buildings & Structures	Steam turbine gross power	0,02	278	1,614

Table D-3: breakdown of the Total plant cost for the reference case

Description	Equipment Cost	Material Cost	Direct labor	Bare Erected Cost	Fees for engineering Services	Process Contingencies	Project Contingencies	Total capital cost
1-COAL HANDLING								
Coal Receive & Unload	1157,13	0	558,02	1715,14	257,28	0	394,49	2366,9
Coal Stackout & Reclaim	3782,78	0	904,28	4687,05	703,06	0	1078,03	6468,13
Coal Conveyors & Yard Crush	36085,17	0	9186,02	45271,19	6790,68	0	10412,38	62474,24
Other Coal Handling	5619,98	0	1265,5	6885,47	1032,82	0	1583,66	9501,95
Coal & Sorbent Handling Foundations	0	100,9	264,06	364,95	54,75	0	83,94	503,64
Subtotal				k\$ 58558,84			k\$	81314,83
2-COAL PREPARATION & FEED								
Coal Crushing & Drying	2832,4	170,55	406,8	3409,74	511,47	0	784,24	4705,44
Prepared Coal Storage & Feed	8704,37	2090,82	1345,45	12140,63	1821,1	0	2792,35	16754,07
Dry Coal Injection System	11107,23	127,6	1016,99	12251,81	1837,78	0	2817,92	16907,49
Miscellaneous Coal Preparation & Feed	859,07	627,88	1849,52	3336,47	500,47	0	767,39	4604,32
Coal & Sorbent Feed Foundation	0	2090,82	1793,94	3884,75	582,72	0	893,5	5360,96
Subtotal				k\$ 35023,4			k\$	48332,28
3-FEEDWATER & MISCELLANEOUS BOP SYSTEMS								
Feedwater System	3023,54	5183,59	2591,8	10798,92	1619,84	0	2483,76	14902,51
Water Makeup & Pretreating	5972,35	597,37	3384,64	9954,36	1493,16	0	3434,26	14881,76
Other Feedwater Subsystems	1563,01	512,35	486,39	2561,74	384,27	0	589,2	3535,2
Service Water Systems	1785,46	3407,26	11033,34	16226,05	2433,91	0	5597,99	24257,94
Other Boiler Plant Systems	409,37	148,86	372,15	930,37	139,56	0	213,99	1283,91
Natural Gas Pipeline and Start-Up System	7703,78	330,98	248,23	8282,97	1242,45	0	1905,09	11430,5
Waste Water Treatment Equipment	8717,05	0	5393,61	14110,65	2116,6	0	4868,18	21095,42
Vacuum Flash, Brine Concentrator, & Crystallizer	59779,51	0	36984,3	96763,81	14514,58	0	33383,52	144661,89
Miscellaneous Plant Equipment	16404,81	2151,31	8336,32	26892,44	4033,87	0	9277,89	40204,19
Subtotal				k\$ 186521,27			k\$	276253,29
4-GASIFIER, ASU & ACCESSORIES								
Gasifier & Auxiliaries(MHI)		-		1560240,31	234036,05	0	301906,5	2096182,85
Syngas Cooler	54447,16	0	23601,83	78048,99	11707,35	10926,86	15102,48	115785,67
Air Separation Unit/Oxidant Compression	19995,07	0	7596,47	27591,53	4138,73	0	4759,54	36489,8
Miscellaneous Gasification Equipment	4800,01	0	2080,68	6880,69	1032,11	0	1186,92	9099,71
Low Temperature Heat Recovery		-		49648	7447,2	0	11419,04	68514,24
Flare Stack System	2190,51	0	386,77	2577,28	386,6	0	592,78	3556,64
Major Component Rigging	267,4	0	115,8	383,19	57,48	0	65,53	506,2
Gasification Foundations	0	541,96	323,51	865,46	129,82	0	248,39	1243,67
Syngas driven superheater & regenerative heat exchangers		-		114949	17242,35	0	22242,64	154433,99
Air booster		-		29830	4474,5	0	5145,68	39450,18
Subtotal				k\$ 1871014,45			k\$	2525262,95
5-SYNGAS CLEANUP								
MDEA	12431,78	0	10476,22	22907,99	3436,2	0	5268,84	31613,02
Elemental Sulfur Plant	67898,18	13235,56	86999,22	168132,95	25219,95	0	38670,58	232023,47
Carbonyl Sulfide (COS) Hydrolysis	9754,41	0	12652,16	22406,57	3360,99	0	5153,51	30921,06
Fuel Gas Piping	0	3029,64	1980,66	5010,3	751,55	0	1152,37	6914,21
Gas Cleanup Foundations	0	336,75	227,18	563,92	84,59	0	0,35	648,86
Subtotal				k\$ 219021,73			k\$	302120,62
6-COMBUSTION TURBINE & ACCESSORIES								
Combustion Turbine Generator	83689,36	0	6028,94	89718,29	13457,75	4485,92	16149,3	123811,24
Combustion Turbine Accessories	3000,52	0	183,14	3183,65	477,55	0	549,18	4210,38
Compressed Air Piping	0	569,51	371,86	941,36	141,21	0	216,52	1299,08
Combustion Turbine Foundations	0	241,21	279,17	520,38	78,06	0	179,53	777,96
Subtotal				k\$ 94363,68			k\$	130098,66
7-HRSG, DUCTWORK & STACK								
Heat Recovery Steam Generator	42479,48	0	8226,04	50705,51	7605,83	0	8746,7	67058,04
Heat Recovery Steam Generator Accessories	15167,32	0	2936,42	18103,74	2715,56	0	3122,9	23942,19
Ductwork	0	1206,3	844,86	2051,16	307,68	0	471,77	2830,6
Stack	10259,17	0	3827,85	14087,02	2113,06	0	2430,02	18630,08
Heat Recovery Steam Generator, Ductwork & Stack Foundations	0	255,27	256,4	511,66	76,75	0	176,53	764,94
Subtotal				k\$ 85459,09			k\$	113225,85

Table D-4: breakdown of the Total plant cost for the reference case

8-STEAM TURBINE & ACCESSORIES								
Description	Equipment Cost	Material Cost	Direct labor	Bare Erected Cost	Fees for engineering Services	Process Contingencies	Project Contingencies	Total capital cost
Steam Turbine Generator & Accessories	55463,29	0	8512,92	63976,21	9596,44	0	11035,9	84608,54
Steam Turbine Plant Auxiliaries	2708,44	0	6169,84	8878,28	1331,75	0	1531,51	11741,52
Condenser & Auxiliaries	9984,67	0	5093,95	15078,61	2261,8	0	2601,06	19941,46
Steam Piping	10146,81	0	4400,93	14547,73	2182,16	0	4182,48	20912,36
Turbine Generator Foundations	0	424,88	749,53	1174,4	176,16	0	405,17	1755,72
Subtotal	k\$ 103655,23				k\$ 138959,6			
9-COOLING WATER SYSTEM								
Cooling Towers	13968,47	0	4217,13	18185,6	2727,84	0	3137,02	24050,45
Circulating Water Pumps	1839,4	0	125,57	1964,97	294,75	0	338,96	2598,67
Circulating Water System Auxiliaries	12415,25	0	1739,95	14155,19	2123,28	0	2441,77	18720,24
Circulating Water Piping	0	6843,7	6198,31	13042	1956,3	0	2999,66	17997,96
Make-up Water System	702,14	0	963,67	1665,8	249,87	0	383,14	2298,81
Component Cooling Water System	249,65	298,81	205,67	754,12	113,12	0	173,45	1040,69
Circulating Water System Foundations	0	561,62	999	1560,62	234,1	0	538,42	2333,13
Subtotal	k\$ 51328,3				k\$ 69039,95			
10-SLAG RECOVERY & HANDLING								
Slag Dewatering & Cooling	2118,58	0	1037,51	3156,08	473,42	0	544,43	4173,92
Gasifier Ash Depressurization	1200,3	0	588,12	1788,41	268,27	0	308,5	2365,17
Cleanup Ash Depressurization	539,97	0	264,83	804,79	120,72	0	138,83	1064,33
Ash Storage Silos	1216,35	0	1313,79	2530,14	379,53	0	436,45	3346,11
Ash Transport & Feed Equipment	468,89	0	108,91	577,8	86,67	0	99,67	764,14
Miscellaneous Ash Handling Equipment	67,64	82,55	24,08	174,26	26,14	0	30,06	230,46
Ash/Spent Sorbent Foundation	0	476,91	630,53	1107,44	166,12	0	382,07	1655,62
Subtotal	k\$ 10138,92				k\$ 13599,75			
11-ACCESSORY ELECTRIC PLANT								
Generator Equipment	3681,67	0	2777,77	6459,44	968,92	0	1114,26	8542,6
Station Service Equipment	3326,77	0	285,48	3612,25	541,84	0	623,12	4777,19
Switchgear & Motor Control	20072,57	0	3482,57	23555,14	3533,27	0	4063,27	31151,67
Conduit & Cable Tray	0	88,63	256,56	345,18	51,78	0	98,38	495,34
Wire & Cable	0	1217,46	2176,49	3393,94	509,1	0	967,28	4870,31
Protective Equipment	241	0	837	1078	161,7	0	185,96	1425,66
Standby Equipment	999,74	0	922,3	1922,04	288,31	0	331,56	2541,89
Main Power Transformers	8137,54	0	166	8303,53	1245,53	0	1432,36	10981,42
Electrical Foundations	0	93,87	238,37	332,23	49,84	0	114,62	496,69
Subtotal	k\$ 49001,75				k\$ 65282,77			

Table D-5: breakdown of the Total plant cost for the reference case

12-INSTRUMENTATION & CONTROL								
Description	Equipment Cost	Material Cost	Direct labor	Bare Erected Cost	Fees for engineering Services	Process Contingencies	Project Contingencies	Total capital cost
Integrated Gasification and Combined Cycle Control Equipment	646,9	0	280,32	927,22	139,09	0	159,49	1225,78
Combustion Turbine Control Equipment	639,06	0	46,07	685,12	102,77	0	117,85	905,73
Steam Turbine Control Equipment	590,05	0	90,18	680,22	102,04	0	117	899,25
Other Major Component Control Equipment	1141,87	0	777,25	1919,12	287,87	0	345,45	2552,42
Signal Processing Equipment	886,05	0	28,43	914,47	137,18	0	157,75	1209,39
Control Boards, Panels & Racks	256,8	0	168,59	425,39	63,81	0	102,1	591,28
Distributed Control System Equipment	9295,64	0	302,87	9598,5	1439,78	0	1727,73	12766,01
Instrument Wiring & Tubing	462,63	370,5	1480,01	2313,13	346,97	0	693,94	3354,04
Other Instrumentation & Controls Equipment	1037,97	0	514,58	1552,54	232,89	0	279,46	2064,88
Subtotal				k\$ 19015,71				k\$ 25568,78
13-IMPROVEMENTS TO SITE								
Site Preparation	0	433,03	9888,8	10321,83	1548,28	0	3561,03	15431,13
Site Improvements	0	1965,27	2778,24	4743,5	711,53	0	1636,51	7091,54
Site Facilities	3067,61	0	3444,43	6512,04	976,81	0	2246,66	9735,49
Subtotal				k\$ 21577,37				k\$ 32258,16
14-BUILDINGS & STRUCTURES								
Combustion Turbine Area	0	314	177	491	73,65	0	84,7	649,35
Steam Turbine Building	0	2865,13	4079,51	6944,64	1041,7	0	1197,95	9184,28
Administration Building	0	903,13	655,44	1558,57	233,79	0	268,86	2061,2
Circulation Water Pumphouse	0	162,44	85,43	247,86	37,18	0	42,76	327,8
Water Treatment Buildings	0	395,15	384,5	779,64	116,95	0	134,49	1031,08
Machine Shop	0	490,68	336,21	826,88	124,04	0	142,64	1093,55
Warehouse	0	382,65	246,35	629	94,35	0	108,51	831,85
Other Buildings & Structures	0	280,68	218,08	498,76	74,82	0	86,04	659,61
Waste Treating Building & Structures	0	774,54	1479,6	2254,13	338,12	0	388,84	2981,08
Subtotal				k\$ 14230,48				k\$ 18819,8
TOTAL				2818910,22				3840137,29

APPENDIX D.2

In this section it is listed the series of the equipments that have been considered as part of the power plant analyzed in Case 4 (cooled NH₃ – sulfuric acid plant). Moreover, it is shown the list of capital costs accounted to determine the Total-as-Spent-Capital cost (TASC). Finally, it is presented a breakdown of the LCOE and the variable costs associated to Case 4.

Table D-6: breakdown of the Total plant cost for Case 4

Description	Equipment Cost	Material Cost	Direct labor	Bare Erected Cost	Fees for engineering Services	Process Contingencies	Project Contingencies	Total capital Cost
1-COAL HANDLING								
Coal Receive & Unload	1164,4	0	561,5	1726	259,2	0	397,3	2382,6
Coal Stackout & Reclaim	3806,8	0	910	4716,8	707,9	0	1084,7	6509,5
Coal Conveyors & Yard Crush	36314,8	0	9244,4	45559,3	6833,7	0	10478,6	62871,7
Other Coal Handling	5655,7	0	1273,5	6929,2	1038,9	0	1593,1	9561,3
Coal & Sorbent Handling Foundations	0	101,5	265,7	367,2	55,2	0	85,2	507,7
Subtotal				k\$ 59298,5				k\$ 81832,8
2-COAL PREPARATION & FEED								
Coal Crushing & Drying	2851,5	171,7	409,5	3432,8	514,6	0	790,2	4737,7
Prepared Coal Storage & Feed	8763,3	2104,9	1354,5	12222,9	1833,5	0	2811	16867,5
Dry Coal Injection System	11182,4	128,4	1023,8	12334,8	1849,8	0	2836,6	17021,3
Miscellaneous Coal Preparation & Feed	864,8	632,1	1862	3359	503,4	0	772,7	4635,2
Coal & Sorbent Feed Foundation	0	2104,9	1806	3911	586,4	0	899,7	5397,3
Subtotal				k\$ 35260,5				k\$ 48659
3-FEEDWATER & MISCELLANEOUS BOP SYSTEMS								
Feedwater System	3073,2	5268,7	2634,3	10976,3	1646,5	0	2524,5	15147,4
Water Makeup & Pretreating	7523	752,4	4263,4	12538,9	1881,1	0	4326,1	18746,3
Other Feedwater Subsystems	1588,6	520,7	494,3	2603,8	390,3	0	598,8	3593
Service Water Systems	2249	4291,9	13898,1	20439,1	3065,5	0	7050,6	30555,2
Other Boiler Plant Systems	416,2	151,3	378,4	946	141,8	0	217,2	1305,1
Natural Gas Pipeline and Start-Up System	7722,7	331,7	248,8	8303,3	1245,6	0	1909,7	11458,7
Waste Water Treatment Equipment	10980,4	0	6794	17774,4	2665,8	0	6131,1	26571,5
Vacuum Flash, Brine Concentrator, & Crystallizer	64387,6	0	39835,2	104222,8	15633,4	0	35956,1	155812,4
Miscellaneous Plant Equipment	16445,1	2156,5	8356,8	26958,5	4044,2	0	9300,9	40303,7
Subtotal				k\$ 204763,1				k\$ 303493,3
Description	Equipment Cost	Material Cost	Direct labor	Bare Erected Cost	Fees for engineering Services	Process Contingencies	Project Contingencies	Total capital Cost
4-GASIFIER, ASU & ACCESSORIES								
Gasifier & Auxiliaries (MHI)				1561658	234248,7	0	302180,8	2098087,5
Syngas Cooler	55202,7	0	23929,3	79132,1	11869,8	11078,6	15312	117392,7
Miscellaneous Gasification Equipment	4824,6	0	2091,3	6915,9	1037,8	0	1193	9146,8
Low Temperature Heat Recovery				49648	7447,2	0	11419	68514,2
Flare Stack System	2201,7	0	388,7	2590,4	387,9	0	595,2	3573,6
Major Component Rigging	268,7	0	116,3	385,1	58	0	66,1	509,2
Gasification Foundations	0	544,7	325,1	869,8	130,8	0	250,1	1250,9
Syngas driven superheater & regenerative heat exchangers				126443,9	18966,5	0	24466,8	169877,3
Air booster				39039,2	5855,8	0	6734,2	51629,4
Subtotal				k\$ 1866682,4				k\$ 2519981,6

Table D-7: breakdown of the Total plant cost for Case 4

5-SYNGAS CLEANUP								
Description	Equipment Cost	Material Cost	Direct labor	Bare Erected Cost	Fees for engineering Services	Process Contingencies	Project Contingencies	Total capital Cost
MDEA	13293,3	0	11202,2	24495,5	3674,3	0	5634	33804
Sulfuric acid plant	21637,7	4327,5	27696,2	53661,5	8049,3	0	26830,7	88541,6
Carbonyl Sulfide (COS) Hydrolysis	9814,4	0	12730	22544,5	3382,2	0	5185,3	31112,1
Fuel Gas Piping	0	4236	2769,3	7005,4	1050,2	0	1611,5	9667,1
Gas Cleanup Foundations	0	335,7	226,5	562,3	84,8	0	194	841,2
CO2 Capture unit			132212,6	309492	54161,1	52613,6	72845,1	489111,8
Acid wash			26597,3	62260,8	10895,6	10584,3	14654,3	98395
CO2 compression			40001	59373,6	10390,3	0	10390,3	80154,2
Subtotal				k\$ 539395,6			k\$ 831627	
6-COMBUSTION TURBINE & ACCESSORIES								
Combustion Turbine Generator	81903,7	0	5900,3	87804	13170,1	4390	15805	121169,2
Combustion Turbine Accessories	2936,4	0	179,2	3115,7	467,7	0	537,6	4121,1
Compressed Air Piping	0	557,3	363,9	921,2	137,6	0	212	1270,9
Combustion Turbine Foundations	0	236	273,2	509,2	76,4	0	175,9	761,7
Subtotal				k\$ 92350,1			k\$ 127322,9	
7-HRSG, DUCTWORK & STACK								
Heat Recovery Steam Generator	42491,7	0	8228,4	50720,1	7607,9	0	8749,6	67077,7
Heat Recovery Steam Generator Accessories	15171,6	0	2937,2	18108,9	2716,1	0	3123,7	23948,8
Ductwork	0	1080,3	756,6	1837	275,1	0	422,8	2535
Stack	9188,2	0	3428,2	12616,5	1892,6	0	2175,9	16685,1
Heat Recovery Steam Generator, Ductwork & Stack Foundations	0	228,6	229,6	458,2	68,7	0	157,8	684,8
Subtotal				k\$ 83740,7			k\$ 110931,4	
8-STEAM TURBINE & ACCESSORIES								
Steam Turbine Generator & Accessories	51157,9	0	7852	59010	8851,5	0	10179,7	78041,2
Steam Turbine Plant Auxiliaries	2495,3	0	5684,3	8179,6	1226,9	0	1410,7	10817,3
Condenser & Auxiliaries	4482,7	0	2529,3	7012,1	1052	0	1209,7	9273,9
Steam Piping	10325,4	0	4478,4	14803,8	2221	0	4256,5	21281,5
Turbine Generator Foundations	0	390,9	689,7	1080,7	162,3	0	372,8	1615,9
Subtotal				k\$ 90086,2			k\$ 121029,8	
9-COOLING WATER SYSTEM								
Cooling Towers	17654,3	0	5329,9	22984,2	3447,4	0	3964	30395,8
Circulating Water Pumps	2324,7	0	158,6	2483,4	373,1	0	428,8	3285,3
Circulating Water System Auxiliaries	15438,1	0	2163,5	17601,7	2640,5	0	3036,1	23278,5
Circulating Water Piping	0	8345,5	7558,5	15904,1	2386	0	3657,7	21947,8
Make-up Water System	861,7	0	1182,7	2044,4	306,7	0	469,6	2820,8
Component Cooling Water System	307,4	367,9	253,2	928,6	138,5	0	213,4	1280,6
Circulating Water System Foundations	0	680,4	1210,3	1890,7	284,1	0	652,7	2827,6
Subtotal				k\$ 63837,1			k\$ 85836,4	

Table D-8: breakdown of the Total plant cost for Case 4

10-SLAG RECOVERY & HANDLING								
Slag Dewatering & Cooling	2132,1	0	1044,1	3176,3	476,5	0	548	4200,8
Gasifier Ash Depressurization	1207,9	0	591,8	1799,8	269,9	0	310,3	2380,2
Cleanup Ash Depressurization	543,4	0	266,5	809,9	121,1	0	139,6	1070,6
Ash Storage Silos	1224,1	0	1322,2	2546,3	381,8	0	439,5	3367,8
Ash Transport & Feed Equipment	471,8	0	109,6	581,5	87,6	0	100,3	769,5
Miscellaneous Ash Handling Equipment	68	83	24,2	175,3	26,5	0	29,9	231,9
Ash/Spent Sorbent Foundation	0	479,9	634,5	1114,5	167,2	0	384,2	1666
Subtotal			k\$	10203,6			k\$	13686,8
Description	Equipment Cost	Material Cost	Direct labor	Bare Erected Cost	Fees for engineering Services	Process Contingencies	Project Contingencies	Total capital Cost
11-ACCESSORY ELECTRIC PLANT								
Generator Equipment	3459,1	0	2609,8	6069	910,1	0	1046,3	8025,5
Station Service Equipment	3847,4	0	330,1	4177,5	626,8	0	720,7	5525,1
Switchgear & Motor Control	23214	0	4027,6	27241,6	4085,8	0	4699,7	36027,2
Conduit & Cable Tray	0	102,4	296,7	399,1	60,4	0	114,3	573,9
Wire & Cable	0	1407,9	2517,1	3925,1	584,2	0	1119,3	5628,6
Protective Equipment	241	0	837	1078	162	0	186	1426
Standby Equipment	966,5	0	891,6	1858,2	278,2	0	320,6	2457,1
Main Power Transformers	7741,2	0	157,9	7899,1	1185,5	0	1362,2	10446,9
Electrical Foundations	0	89,3	226,9	316,2	47	0	109,3	472,6
Subtotal			k\$	52963,8			k\$	70582,9
12-INSTRUMENTATION & CONTROL								
Integrated Gasification and Combined Cycle Control Equipment	674,6	0	292,3	966,9	145,1	0	166,6	1278,7
Combustion Turbine Control Equipment	666,4	0	48	714,5	107,3	0	123,6	945,5
Steam Turbine Control Equipment	615,3	0	94	709,3	106,3	0	122,6	938,3
Other Major Component Control Equipment	1190,8	0	810,5	2001,4	300,5	100,1	359,8	2761,9
Signal Processing Equipment	924	0	29,6	953,6	143,1	0	164,5	1261,3
Control Boards, Panels & Racks	267,8	0	175,8	443,6	66,4	22,4	106,3	638,8
Distributed Control System Equipment	9694,4	0	315,8	10010,2	1501,5	500,8	1802,1	13814,8
Instrument Wiring & Tubing	482,4	386,3	1543,5	2412,3	361,8	120,6	723,7	3618,5
Other Instrumentation & Controls Equipment	1082,4	0	536,6	1619,1	243,2	80,7	291,3	2234,4
Subtotal			k\$	19830,9			k\$	27492,2
Description	Equipment Cost	Material Cost	Direct labor	Bare Erected Cost	Fees for engineering Services	Process Contingencies	Project Contingencies	Total capital Cost
13-IMPROVEMENTS TO SITE								
Site Preparation	0	437,1	9983,3	10420,5	1562,6	0	3595	15578,3
Site Improvements	0	1984	2804,8	4788,8	718,8	0	1651,9	7159,6
Site Facilities	3096,9	0	3477,3	6574,3	985,7	0	2267,8	9827,8
Subtotal			k\$	21783,6			k\$	32565,7
14-BUILDINGS & STRUCTURES								
Combustion Turbine Area	0	314	177	491	73,9	0	85	650
Steam Turbine Building	0	2845,3	4051,3	6896,6	1034,3	0	1189,6	9120,6
Administration Building	0	898,9	652,4	1551,3	232,3	0	267,8	2051,5
Circulation Water Pumphouse	0	188,6	99,2	287,8	43,3	0	48,9	380
Water Treatment Buildings	0	497,7	484,3	982	147,4	0	169,2	1298,8
Machine Shop	0	489,5	335,4	824,9	123,8	0	142	1090,8
Warehouse	0	381,7	245,7	627,5	94,6	0	108,7	831
Other Buildings & Structures	0	280	217,5	497,6	74,5	0	85,6	657,7
Waste Treating Building & Structures	0	797,5	1523,5	2321	348	0	400,3	3069,4
Subtotal			k\$	14479,7			k\$	19149,8
Tot			k\$	3154675,8			k\$	4394191,6

Table D-9: CASE 4 - Owner's costs (on the left), Fixed Operative Costs (top right), Electric auxiliaries (down right), summary of the capital costs (down on the left)

Description	k\$
Pre-Production Costs	
6 Months All Labor	33495,342747
1 Month Maintenance Materials	8925,7192133
1 Month Non-Fuel Consumables	4315,5472185
1 Month Waste Disposal	910,44709767
25% of 1 Months Fuel Cost at 100% CF	2621,3799132
2% of TPC	87884,004561
Total	138152,44075
Inventory Capital	
60-day supply of fuel and consumables at 100% CF	29196,625061
0.5% of TPC (spare parts)	21971,00114
Total	51167,626201
Other Costs	
Initial Cost for Catalyst and Chemicals	12973,602587
Land	900
Other Owner's Costs	659130,03421
Financing Costs	118643,40616
Total	791647,04296
Total Plant Cost	4394191,6
Total Owner's Costs	980967,10991
Total Overnight Costs (TOC)	5375158,7099
TASC Multiplier (IOU, 35 year)	1,154
Total As-Spent Cost (TASC)	6202933,1512

Description	k\$/year
FIXED OPERATING COSTS	
Annual Operating Labor	7453446
Maintenance Labor	46138518,585
Administrative & Support Labor	13397991,146
Property Taxes and Insurance	87884004,561
Total	154873960,29
Maintenance Materials	85686904,447

Electric Auxiliaries	MW
Coal milling & handling	3,9491
Slag handling	0,6868
Steam cycle pumps	6,66
Cooling tower auxiliaries	12,141
Acid gas removal auxiliaries	2,95
BOP miscellaneous*	2,94
Air booster	39
Water treatment auxiliaries	5,357
Steam turbine auxiliaries	0,291
Groundwater pumps	0,767
CT auxiliaries	4,4
CO2 compression	41,94
CO2 capture unit	27,42
Total auxiliaries	148,5019

Table D-10: CASE 4 - summary of the variable costs (maintenance materials excluded). From top to bottom: costs associated to miscellaneous consumables, fuel costs, disposal costs

	Price	Unit of measure	Quantity employed every day at full capacity	Cost \$ /year
Water consumption	1,9 \$/10 ³ gallons	10 ³ gallons	6149,0017411	3411466,166
Makeup and Wastewater treatment chemicals	550 \$/Ston	Short ton	18,313903498	2941212,9017
Sodium Hydroxide (50%wt solution)	600 \$/Ston	Short ton	25,913831759	4540103,3241
COS hydrolysis catalyst	1300 \$/ft ³	ft ³	1,3288037063	504413,88691
MDEA solution	2,8 \$/gallon	gallon	55,138875353	45081,544489
Sulfuric acid (98%wt solution)	0 \$/Ston	Short ton	0,4606903424	0
Triethylene Glycol (gal):	6,8 \$/gallon	gallons	596,64721508	1184702,7103
Ammonia Substitution	400 \$/ton	Metric ton	242,17304919	28285812,145
Sulfuric acid (acid wash)	0 \$/Ston	Metric ton	695,63487379	0
V2O5 Substitution	11023 \$/ton	Metric ton	0,1283643836	516460,619
				41429253,297

	Price	Unit of measure	Quantity employed every day at full capacity	Cost \$ /year
Coal	45,83	Short ton coal	7521,9160645	100660988,67

	Price	Unit of measure	Quantity employed every day at full capacity	Cost \$ /year
Slag disposal	38 \$/Ston	Short ton	653,97870178	7256547,675
Crystallizer solids	38 \$/Ston	Short ton	48,948348877	543130,87914
MDEA Solution	0,35 \$/gal	gallons	55,138875353	5635,1930611
COS hydrolysis Catalyst	2,5 \$/ft ³	ft ³	1,3288037063	970,02670559
Triethylene Glycol (gal):	0,35 \$/gal	gallons	596,64721508	60977,345381
				7867261,1193

APPENDIX E

Table E-1: summary of capital cost estimating Classifications (from [69])

<p>Order of magnitude (also known as Ratio or Feasibility) estimate Data: This type of estimate typically relies on cost information for a complete process taken from previously built plants. This cost information is then adjusted using appropriate scaling factors, for capacity, and for inflation to provide the estimated capital cost. Diagrams: Normally requires only a block flow diagram.</p>
<p>Study (also known as Major Equipment or Factored) estimate Data: This type of estimate utilizes a list of the major equipment found in the process. This includes all pumps, compressors, and turbines, columns and vessels, fired heaters, and exchangers. Each piece of equipment is roughly sized and the approximate cost determined. The total cost of equipment is then factored to give the estimated capital cost. Diagrams: Based on PFD. Cost from generalized charts. Note: Most individual student designs are in this category.</p>
<p>Preliminary design (also known as Scope) estimate Data: This type of estimate requires more accurate sizing of equipment than is used in the study estimate. In addition, approximate layout of equipment is made along with estimates of piping, instrumentation, and electrical requirements. Utilities are estimated. Diagrams: Based on PFD. Includes vessel sketches for major equipment, preliminary plot plan, and elevation diagrams Note: Most large student group designs are in this category.</p>
<p>Definitive (also known as Project Control) estimate Data: This type of estimate requires preliminary specifications for all the equipment, utilities, instrumentation, electrical, and off-sites. Diagrams: Final PFD, vessel sketches, pilot plan, and elevation diagrams, utility balances, and preliminary P&ID</p>
<p>Detailed (also known as Firm or Contractor's) estimate Data: This type of estimate requires complete engineering of the process and all related off-sites and utilities. Vendor quotes for all expensive items will have been obtained. At the end of a detailed estimate, the plant is ready to go to the construction stage. Diagrams: Final PFD and P&ID, vessel sketches, utility balances, plot plan And elevation diagrams, and piping isometrics. All diagrams are required to complete the construction of the plant if it is built.</p>

APPENDIX F

In this section it is provided a collection of Figures describing some of the equipments encountered in the power plants analyzed. (see the list of images for the source of the images)

Figure A: discharge of the coal from the train. Figure B: coal stacker. Figure C: discharge from the belt conveyor with the tripper. Figure D: coal conveyor belt and transfer towers

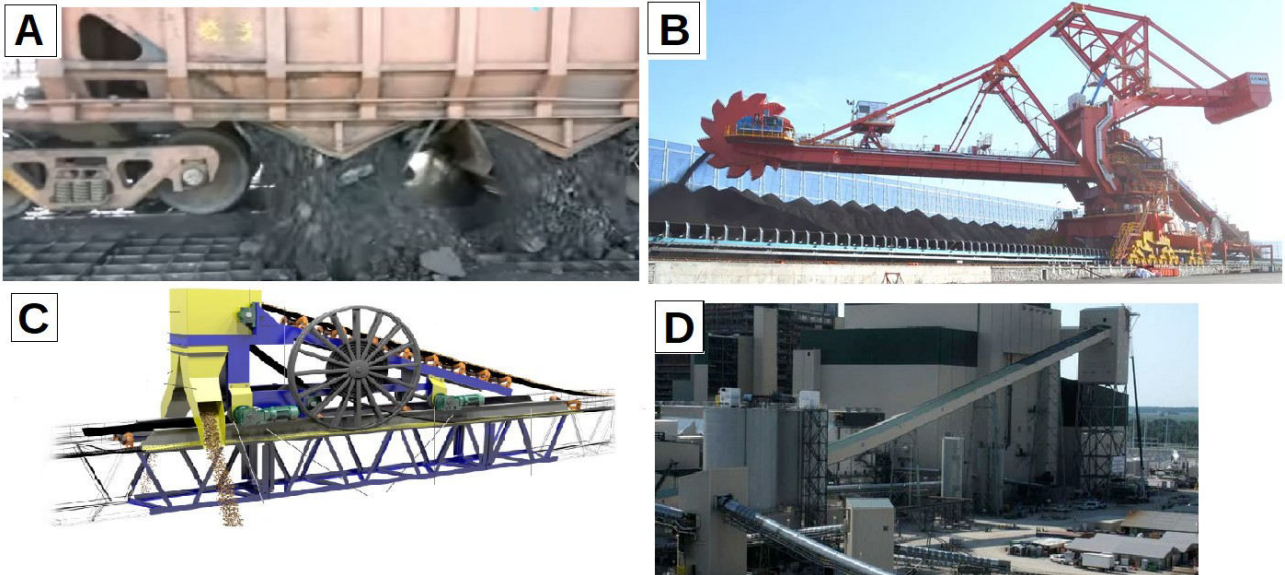


Figure E: coal crusher. Figure F: coal receiving hoppers. Figure G: hopper-feeder connection. Figure H: externals of a coal feeder. Figure I: internals of a coal feeder.

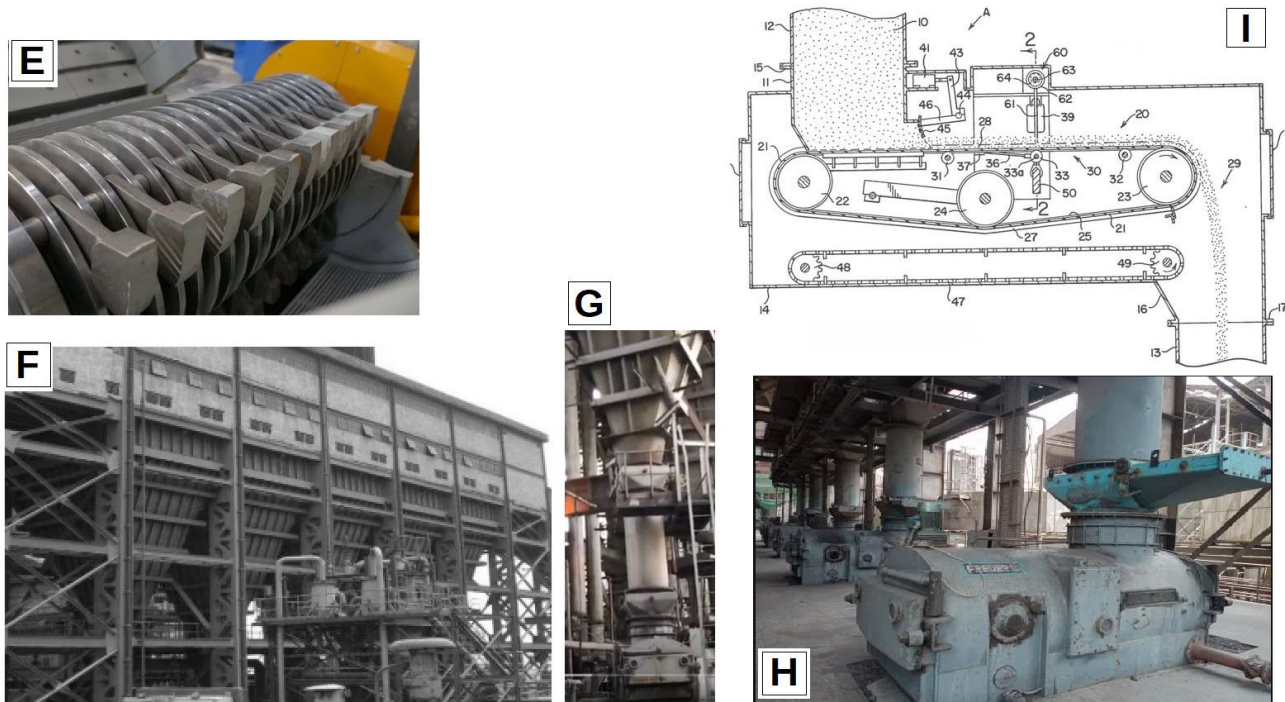


Figure L: representation of the CO₂ interstage heat exchangers in an existing CO₂ compression facility (Coffeyville). Figure M: TEG drying column in an existing CO₂ compression facility. Figure N: integrally geared CO₂ compressor (MHI)



Figure O: aerial view of the CO₂ capture unit in the Petra Nova CCS project

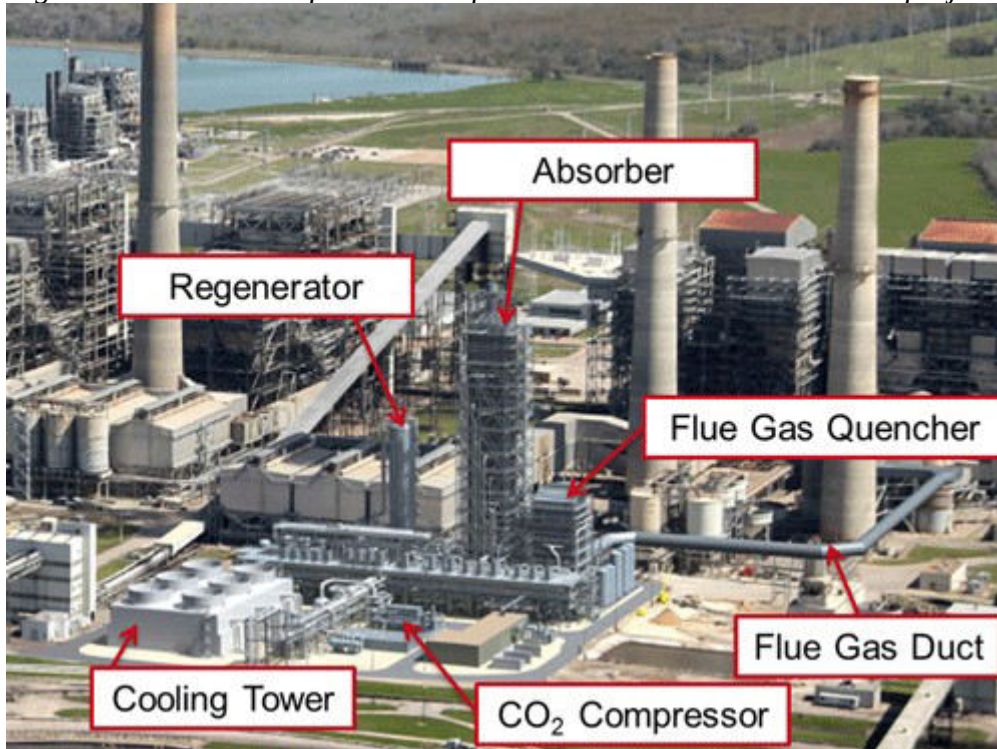


Figure P: carbon dioxide boosting station



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