TECHNO ECONOMIC EVALUATION OF DIFFERENT POST COMBUSTION CO_2 CAPTURE PROCESS FLOW SHEET MODIFICATIONS

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Key Messages

- Post combustion capture process improvements that are already well established such as intercooling in the absorber and improved heat integration with power plant, combined with improved solvents typical of those that are expected to become available by 2020, should substantially reduce the efficiency penalty on power plant.
- Current stage of process design improvements and improvements in solvent properties leads to reducing efficiency penalty from 9.8% to 6.11% for super critical pulverised coal (SCPC) fired power plant with amine based solvent CO_2 capture process base case.
- In natural gas combined cycle (NGCC) power plant with improved solvent properties CO_2 Capture process base case, reductions in efficiency penalty from 7.8% to 5.93% are achieved by flue gas recirculation, process design improvements.
- The overhead condenser (OHC) process modification was found to be having the lowest efficiency penalty of 5.84% for SCPC case, due to the reduction in steam extraction penalty, and for NGCC case was found to be 5.28%.
- The heat integrated stripper + OHC heat integration process modification was found to have the second lowest efficiency penalty for SCPC and NGCC case.
- The process modifications such as improved split flow process, OHC heat integration, vapour recompression + split flow and heat integrated stripper + OHC heat integration showed reduced CoE (cost of electricity) and lower CO_2 avoidance cost for both SCPC and NGCC case.
- Overall it can be noticed from this study that once all current improvements have been implemented in the solvent based post combustion capture process, different process modifications for SCPC and NGCC only bring slight improvements in the power plant efficiency penalty.
- The performance and cost of different post combustion capture process modifications depend on the type of solvent used. Therefore, for new solvents further evaluation for all process modifications will be required.

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Introduction

Post combustion CO_2 capture technology is one of the potential technologies which will most likely to be applied at large scale CO_2 capture facilities in power plants. One of the main concerns for the solvent based CO_2 post combustion capture (PCC) technology for power plant is the relatively large energy penalty. The energy required to regenerate the solvent and run the PCC process in a coal fired power plant is currently considered to be equivalent to a reduction in the thermal efficiency of about 20% (from roughly 44 -35% LHV) when around 90% CO_2 is captured[1]. A reduction in energy penalty for solvent based CO_2 post combustion capture process can be achieved by improving solvent properties, better integration with power plant as well as by improving process design.

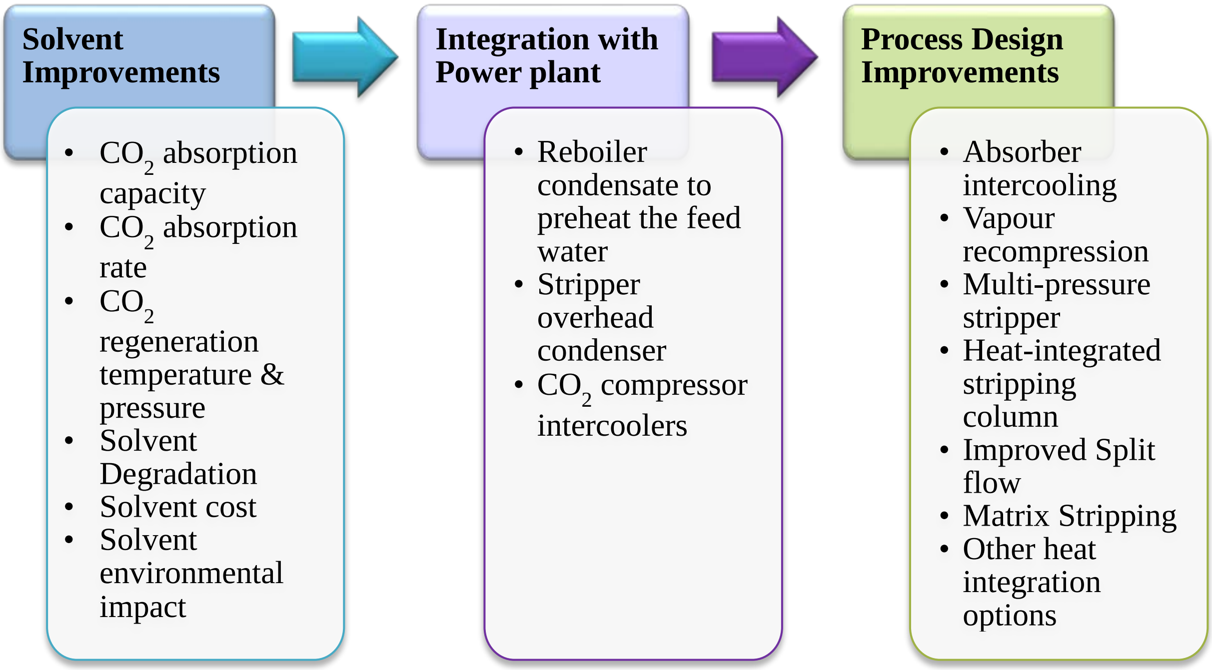


Figure 1 Improvements for amine based solvent CO_2 post combustion capture process

Regarding to the improvement in process design, different process flow sheet modifications have been reported in literature and patents for chemical solvent based CO_2 absorption processes[2]. These process modifications reduce the energy penalty imposed by the CO_2 post combustion capture plant. The proposed process flow sheet modifications are multi-

[1] Adams D., Davison J. 2007, Capturing CO_2, IEAGHG report.
[2] A. Cousinsa, L.T. Wardhaugh, P.H.M. Feron, 2011, A survey of process flow sheet modifications for energy efficient CO_2 capture from flue gases using chemical absorption, International Journal of Greenhouse Gas Control, 5, 605–619.

component column, inter-stage temperature control, heat integrated stripping column, split flow process, vapour recompression, matrix stripping and various heat integration options[2]. Comparison of these reported modifications was difficult as these were evaluated based on different solvent properties and process conditions. Also there are some process modifications more suitable for particular solvent than the others. In order to identify the suitable process modification for full scale PCC application it was necessary to evaluate further in detail these modifications on the same process condition for their energy savings, additional unit required and additional cost.

Therefore, there was a requirement to evaluate these process modifications on similar solvent and process conditions with a state of the art rate-based CO_2 absorption model. IEAGHG has commissioned this study to evaluate the feasibility of these different amine-based CO_2 post combustion capture process modifications for coal and natural gas based power plants.

Scope of the study

Following are the scope of this study:

- Technical evaluation of different process modifications shall be performed and issues related to operational, energy efficiency, process complexity and process control shall be identified.

- Economic evaluation of these process modification options shall be performed in order to find the trade-off between increased capital and lower operational cost.

- Identify major technical challenges and gaps for different process modification options.

Study Approach

In this study Super Critical Pulverised Coal (SCPC) fired power plant of 900MW gross power, with a net efficiency of 45.2% (LHV) without CO_2 capture and Natural Gas Combined Cycle (NGCC) power plant of 883MW gross power, with a net efficiency of 58.2% (LHV) without CO_2 capture are evaluated. The most suitable simulation tools for steady-state simulations were chosen; Ebsilon®Professional for the overall power plant and the CO_2 compression and Aspen Plus® for the CO_2 capture process. The CO_2 capture plant for SCPC and NGCC consists of two greenfield CO_2 capture trains. Moreover, current state of process improvement such as generic improved amine based solvent 'Solvent 2020', absorber intercooling and operating stripper at higher pressure (5Bar) was considered in this study. Solvent 2020 was an artificial solvent which has the same CO_2 absorption mechanisms as amines (carbamate and bicarbonate formation). The properties like density, viscosity and heat capacity were assumed to be similar to those of a solution with 7mol MDEA (Methyldiethanolamine) and 2mol PZ (Piperazine) per kg H_2O. Thus, the corresponding ASPEN Plus® property model was used for the simulations. The reaction kinetics of 'Solvent2020' were enhanced compared to 7MDEA/2PZ, which results in chemical reactions that are not kinetically hindered. This was the main property improvement compared to other solvents for 'Solvent 2020'. 'Solvent 2020' was assumed to be thermally stable up to approximately 150 °C, which was the same temperature as for PZ. Thus, thermal degradation was not expected to occur when operated at temperatures below this limit. Oxidative degradation was assumed to be negligible. In addition, 'Solvent 2020' was also assumed to be not corrosive in the chosen operating range.

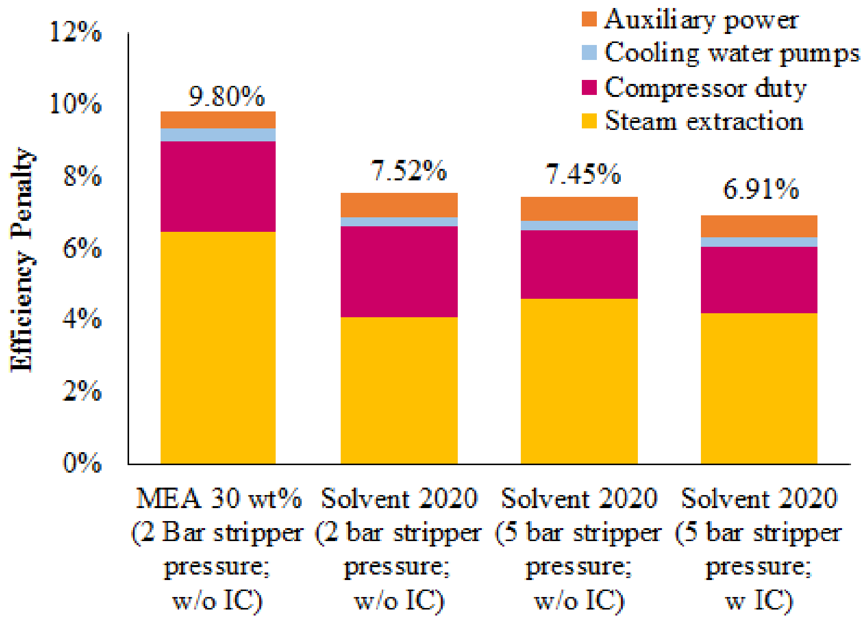


Figure 2, Effect of different improvements on SCPC CO_2 capture base case plant efficiency. [IC: Intercooling]

Figure 2 shows the impact of these CO_2 capture process improvements on the efficiency penalty for SCPC power plant. It can be noticed that the largest reduction on power plant efficiency penalty (from 9.8% to 7.52%) was achieved by using an improved solvent named 'Solvent 2020' when compared to conventional solvent 30wt% Monoethanolamine (MEA).

This reduction was due to the lower specific reboiler duty and cooling water requirement by using an improved solvent, 'Solvent 2020'. Further improvement was implemented by operating the stripper at a higher pressure of 5 bar, which shows that despite having a higher specific heat duty, the penalty imposed by compression duty was reduced which leads to lower efficiency penalty of 7.45%.

It can be noticed from Figure 2 that further process design improvement by implementing intercooling in the absorber, reduces the efficiency penalty to 6.91%. This was due to the increased solvent CO_2 absorption capacity, which resulted in a lower solvent circulation rate, leading to a lower steam extraction requirement.

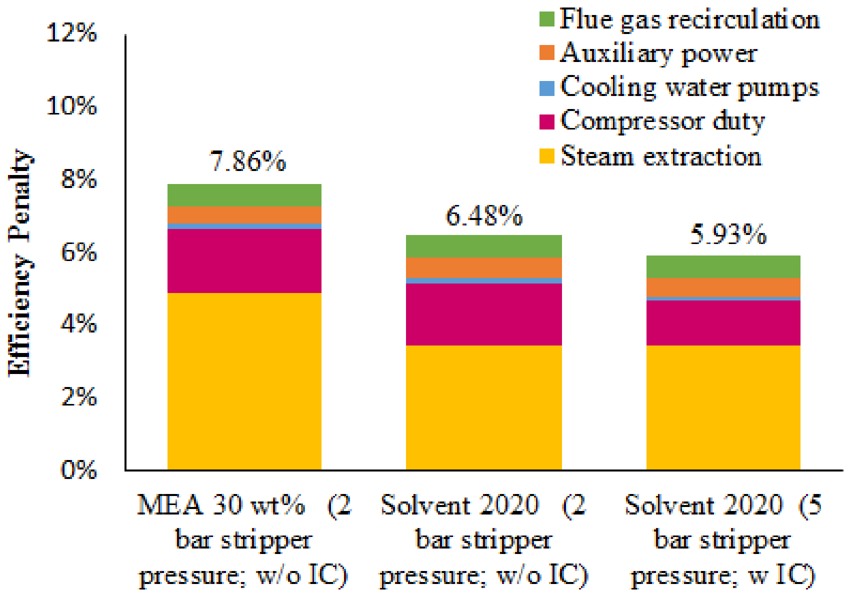


Figure 3, Effect of different improvements on NGCC CO_2 capture base case (with FGR) plant efficiency. [IC: Intercooling]

In the NGCC CO_2 capture base case, the CO_2 concentration in the flue gas was significantly lower. Therefore, in order to minimize the energy requirement of the CO_2 capture plant, flue gas recirculation (FGR) was considered which leads to a CO_2 concentration of 9.1vol% in the flue gas. Similar effect of improved solvent and improved process design was noticed for NGCC CO_2 capture base case (see Figure 3). It can be noticed that the improvements considered in this study reduce the NGCC efficiency penalty from 7.86% to 5.93%.

Effect of waste heat integration

For the SCPC CO_2 capture base case, basic heat integration with the power plant by returning reboiler condensate to the preheating route for the feed water was considered. Also advanced waste heat integration was performed by using heat available from the CO_2 compressor intercooler and stripper overhead condenser.

Table 1, Effect of waste heat integration on efficiency penalty for SCPC CO_2 capture base case.

SCPC Power plant	Base case with IC; w/o HI	Base case with IC; with HI
Steam extraction	4.16%	4.21%
Compressor duty	1.90%	2.06%
Cooling water pumps	0.23%	0.21%
Auxiliary power	0.62%	0.60%
Heat integration	-	-0.97%
Overall efficiency penalty	6.91%	6.11%

Note: IC: Intercooling, HI: Waste Heat Integration

Table 1 show that by implementing the above mentioned waste heat integration in the SCPC CO_2 capture base case, a reduction of 0.97% in efficiency penalty, resulting in a total efficiency penalty of 6.11% was achieved. In the NGCC case, basic integration was considered by injecting reboiler condensate into the superheated steam (spray attemperation) to reduce the temperature and prevent hot spots in the reboiler. The remaining reboiler condensate was partially returned to the water steam cycle upstream of the economiser of the heat recovery steam generator to increase the temperature to 60°C and thus prevent condensation of vapour in the flue gas. The rest of the condensate was returned downstream of the economiser. A more complex waste heat integration was not considered for the NGCC case as there was no available heat sink.

Therefore, the CO_2 base cases considered for SCPC and NGCC power plants in this study were taken at the current stage of process improvements and an improved amine based CO_2 solvent, representative of a future solvent, with generically improved CO_2 absorption properties probably available in the coming years.

Findings of the Study

Impact on efficiency penalty

Various process modifications were evaluated for SCPC and NGCC cases. This was based on energetic evaluation of the overall process, by looking at energy required/saved by steam extraction, compressor duty, cooling water pumps, auxiliary power and heat integration. Based on this, the overall efficiency penalty was estimated for each evaluated process modification (see Table 2). The overhead condenser SCPC case was found to have the lowest efficiency penalty, due to the reduction in steam extraction penalty. The heat integrated stripper+ OHC heat integration process modification was found to have the next lowest efficiency penalty. In the NGCC case the overhead condenser heat integration and the heat integrated stripper + OHC heat integration cases were found to have the lowest efficiency penalties. This was due to the reduced steam extraction, resulting in the lowest specific heat duty.

Table 2, Overall efficiency penalty for various process modifications

Different Process Modifications	SCPC case in %-points	NGCC case in %-points
Base case	6.11	5.93
Vapour recompression	6.09	5.86
Multi-pressure Stripper	6.25	5.86
Heat-integrated stripping column	6.18	5.92
Improved split flow process	5.99	5.46
Matrix stripping	6.41	6.04
Overhead condenser heat integration	5.84	5.28
Reboiler condensate heat integration	-	5.83
Vapour recompression + split flow	5.99	5.46
Heat-integrated stripper + OHC heat integration	5.88	5.34

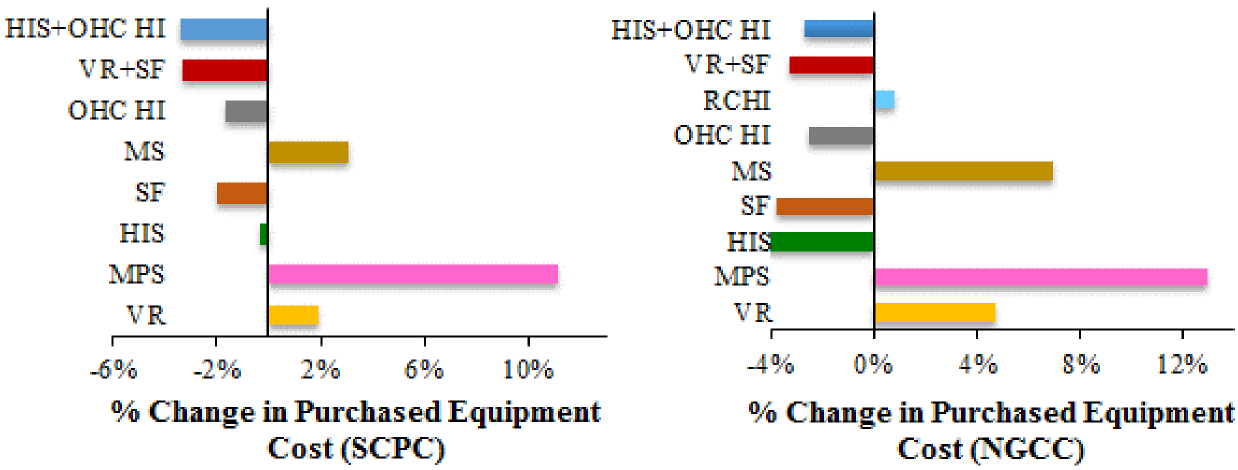
Moreover, it was also noticed that the combination of vapour recompression with split flow process modification was found to be having a slightly lower efficiency penalty when compared to that of the vapour recompression process modifications.

It can be noticed from these results that the matrix stripping process modification was found to be having a higher efficiency penalty than the base case for the SCPC and NGCC cases. This was due to the increased compressor duty by 0.41% points in the SCPC case compared to the base case, as well as the positive effect of advanced heat integration was reduced, since the temperature level, as well as available waste heat in the overhead condenser was reduced. In the SCPC case the multi-pressure stripping process modification was also found to be having a higher efficiency penalty. It showed that whereas the steam extraction penalty was reduced by 0.24%, the auxiliary power of the CO_2 capture plant was increased by 0.28% points. Also, the positive effect of heat integration was reduced by 0.10% points, since the temperature level of usable waste heat as well as the amount of heat was reduced.

Overall it can be noticed that different process modifications for SCPC and NGCC only bring slight improvements in the efficiency penalty.

Impact on required process equipment

Different process modifications will require additional equipment which will affect the capital investment cost of the unit. Figure 4 (a & b) shows the impact on percentage change in the purchased equipment cost (PEC) for different process modifications for SCPC and NGCC cases.



(a) **(b)**

Figure 4, Percentage change in purchased equipment cost for different process modifications compared to the base case. [VR: vapour recompression, MPS: multi-pressure stripper, HIS: heat integrated stripper, SF: split flow, MS: matrix stripping, OHC HI: overhead condenser heat integration, RCHI: reboiler condensate heat integration]

In this study some process modifications were found to be reducing the PEC when compared to that of the base case. Such as for SCPC case vapour recompression + split flow (VR+SF) and heat-integrated stripping column + overhead condenser heat integration (HIS +OHC HI) process modification were found to be lowering PEC when compared to that of SCPC base case. For VR+SF SCPC case the higher cost of an additional flash tank and flash vapour compressor was outweighed by the lower cost of different equipment such as rich solution pump, rich/lean heat exchanger, desorber overhead condenser, condensate return tank, reboiler, reclaimer, reboiler condensate pump and motor and filters required due to improved split flow process. Similarly for SCPC HIS+OHC HI case the additional heat exchanger and stripper heater cost will require smaller rich/lean heat exchanger (RLHX) as well as OHC HI also require smaller dimension for following equipment such as RLHX, desorber overhead condenser, reboiler and reclaimer.

For NGCC, the heat-integrated stripping column (HIS) case was found to have the most reduced PEC when compared to that of the NGCC base case. As in HIS case the RLHX was smaller in dimension and the rest of the equipment require smaller dimensions leading to lower PEC.

On the other hand, multi-pressure stripper (MPS) process modification showed the highest increase in PEC for SCPC (11%) and NGCC (13%) compared to the respective base cases, as this process modification requires additional two desorber columns and two centrifugal compressors to increase the pressure. For MPS, the SCPC case centrifugal compressors account for 7.4% of the total capture plant PEC and the desorber column accounts for 7.8% of the total capture plant PEC. Whereas for NGCC, the MPS case centrifugal compressors account for 9.8% of the total capture plant PEC and the desorber column accounts for 6.1% of the total capture plant PEC. The second highest increase in the PEC was found for the matrix stripping (MS) process modification; 3% for SCPC and 7% for NGCC when compared to the respective base cases. This was due to the required additional two desorber columns as well as additional two reboilers, reclaimer, overhead condenser and condensate return tank. Another widely evaluated process modification was vapour recompression, which was also found to be increasing the PEC for SCPC (2%) and NGCC (5%) cases when compared to the respective base cases. This was due to the requirement of an additional flash tank and flash vapour compressor.

Impact on Cost of electricity and CO_2 avoidance cost

An economic evaluation of various process flow sheet modifications was performed, based on the additional capital costs of the CO_2 capture plant and the changes in plant performance. The capital cost was estimated based on the major equipment items multiplied by factors to account for the related costs for instrumentation and controls, piping, electrical equipment, etc. The economic indicators which were calculated were the Cost of Electricity (CoE) in €/MWh and the cost of CO_2 avoidance in €/tCO_2 compared to a reference plant without CO_2 capture, using the same fuel. The results are summarised in Table 3. The process modifications such as improved split flow process, OHC heat integration, vapour recompression + split flow and heat integrated stripper + OHC heat integration shows the reduced CoE and lower CO_2 avoidance cost for both SCPC and NGCC case. This was due to the lower operational cost of these process modifications and in some cases also a better net efficiency which lead to lower CoE and CO_2 avoidance cost.

Table 3, Cost of electricity (CoE) and CO_2 avoidance cost for various process modifications.

Different Process Modifications	SCPC CoE €/MWh	SCPC relative change of CoE %	SCPC $C_{CO2, avoided}$ €/tCO_2	NGCC CoE €/MWh	NGCC relative change of CoE %	NGCC $C_{CO2, avoided}$ €/tCO_2
Base case, SCPC w/o CO_2 Capture (CoE_{Ref})	42.22	-	-	-	-	-
Base case, NGCC w/o CO_2 Capture (CoE_{Ref})	-	-	-	59.5	-	-
Base case, SCPC w CO_2 Capture	68.29	61.7%	38.32	-	-	-
Base case, NGCC w CO_2 Capture	-	-	-	76.82	29.1%	54.76
Vapour recompression	68.43	62.1%	38.54	76.99	29.4%	55.27
Multi-pressure stripper	69.53	64.7%	40.17	77.46	30.2%	56.76
Heat-integrated stripping column	68.39	62.0%	38.48	76.51	28.6%	53.77
Improved split flow process	67.87	60.7%	37.69	75.92	27.6%	51.85
Matrix stripping	68.95	63.3%	39.33	77.39	30.1%	56.57
OHC heat integration	67.65	60.2%	37.35	75.73	27.3%	51.21
Reboiler condensate integration	-	-	-	76.73	29.0%	54.46
Vapour recompression + split flow	67.78	60.5%	37.57	75.95	27.6%	51.94
Heat-integrated stripper + OHC heat integration	67.71	60.4%	37.45	75.8	27.4%	51.46

Note: Relative change of CoE was based on the % change when compared to CoE_{ref}.

It can be noticed that multi-pressure stripper and matrix stripper cases showed the highest increase in the cost of electricity and CO_2 avoidance cost. In the multi-pressure stripper case the increased capital cost and increased operation cost show that this modification was the most expensive among the other modification studied for both SCPC and NGCC cases. Similarly the matrix stripping modification was also found to be expensive.

Sensitivity analysis

Various aspects of the process modifications were evaluated:

- An increase in CO_2 capture percentage from 90% to 95% was expected to increase the heat duty requirement. Beside that the solvent mass flow rate and lean loading need to be manipulated due to the effect on rich loading.

- Increasing the size of power plant above 900MWe does not impose any limitation for the studied process modifications because additional trains of equipment can be built in parallel.

- The impact of solvent properties on process modification was mainly on the reboiler temperature, as it was limited by solvent degradation at higher stripper temperature and pressure. Therefore, process modifications such as vapour recompression, multi-pressure stripper, heat-integrated stripping column can show more positive improvement.

- During part load conditions the capture plant efficiency reduces and it was expected that vapour recompression and multi-pressure stripping will show higher loss in efficiency during part load. This was due to the reduction in fans' efficiency in part load operation.

- The requirement for process control rises with more complex process flow sheet modification. Matrix stripping was found to be the most complex and other modifications showed slight increases in the complexity.

- When considering retrofitting, issues like space, available utilities and IP/LP crossover pressure are of major importance. The multi-pressure stripper was found to be the most suitable for retrofit, as it shows the lowest temperature level in the reboiler.

- Retrofitting a CO_2 capture unit in a natural gas combined cycle (NGCC) power plant, the main issue will be the installation of flue gas recirculation to increase CO_2 concentration in the flue gas.

Expert reviewers' comments

In this study a generic improved solvent 'Solvent 2020' was considered. Some of the reviewers asked to explicitly show the improvements made by using this improved solvent on the power plant efficiency. Therefore, a further simulation was performed for a conventional solvent 30wt% MEA and at lower stripper pressure of 2bar. To compare the effect of generic improved solvent 'Solvent 2020', further simulation was performed at a lower stripper pressure of 2bar. Hence, such an evaluation makes it clear on the impact of different improvements in amine based solvent CO_2 absorption process. It was suggested by reviewers that the results from this study are very solvent specific. The focus of this study was to evaluate different process modifications based on the current state of improvements in process design, and by using a generic improved solvent. Hence, for a different solvent, the evaluation for each process modifications should be performed.

Conclusions

This study evaluated different post combustion capture process modifications for SCPC and NGCC power plant. The study also evaluated the current state of process design improvements such as absorber intercooling, operation at higher stripper pressure and an advanced level of waste heat integration for the SCPC case. In order to identify the effect of future improvements in the solvent; a generic improved amine based solvent 'Solvent 2020 was considered. Regarding to the different process modifications, matrix stripping was found to be having the highest efficiency penalty due to the increased energy requirement by compressors. Also the cost of electricity and cost of CO_2 avoided for this modification was found to be higher compared to other process modifications. Multi-pressure stripper was also found to be higher in power plant efficiency penalty as well as higher cost of electricity and cost of CO_2 avoided for SCPC and NGCC case. Other process modifications such as OHC heat integration, vapour recompression + split flow and heat integrated stripper + OHC heat integration show lower efficiency penalties, reduced cost of electricity and lower CO_2 avoidance cost when compared to for both SCPC and NGCC base cases. Hence, the evaluation shows that the major improvement in the efficiency penalty was already achieved by using an improved solvent for SCPC and NGCC case. Further process modifications only bring small change in the efficiency penalty.

Regarding to the other issues such as process control, multi-pressure stripping was the most complex, hence, will require a more complex process control system. When retrofitting these process modifications, multi pressure was found to be the more suitable for SCPC case. Whereas for NGCC case the flue gas recirculation was the main issue when considering retrofitting CO_2 capture process.

Recommendations to Executive Committee

This study has evaluated different process modifications and identified some potential process modifications for further evaluation. Further evaluating these identified potential process modifications for different potential solvents will provide very useful insights. Moreover, detailed analysis based on the different power plant load conditions, retrofitting, and process control could be performed. IEAGHG would also like to recommend the industry and researchers to evaluate these identified potential process modifications in a real pilot plant tests.

This study has identified that the improvements made in the solvent for CO_2 absorption characteristics was one of the important areas for improving CO_2 capture process efficiency. Hence, an improved solvent has to be tested in pilot plants and it was necessary to develop an exact property model of the solvent which describes the solvent with the effects of all process modifications. Also it was important to have improved solvent with a lower degradation and corrosion.

Techno Economic Evaluation of different Post Combustion CO_2 Capture Process Flow Sheet Modifications

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Report

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Hamburg, July 2014

Contents

Abbreviations

BC	base case
CAPEX	capital expenditure
CCS	carbon capture and storage
CECPI	Chemical Engineering Chemical Plant Index
CoE	Cost of Electricity
CoT&S	cost of transport & storage
DCC	direct contact cooler
ELECNRTL	electrolyte non-random two liquid
ESP	electrostatic precipitator
FGD	flue gas desulphurisation
FGR	flue gas recirculation
GHG	Greenhouse gas
HI	heat integration
HIS	heat-integrated stripper
HP	high pressure
HRSG	heat recovery steam generator
ID	induced draft
IP	intermediate pressure
IPCC	Intergovernmental Panel on Climate Change
L/G	liquid gas ratio
LHV	lower heating value
LMTD	log mean temperature difference
LP	low pressure
MDEA	methyl diethanolamine
MPS	multi-pressure stripper
MS	matrix stripping

NGCC	natural gas combined cycle
OHC	overhead condenser
OPEX	operational expenditure
PCC	post-combustion capture
PEC	purchased equipment costs
PZ	piperazine
q_{cool}	specific cooling duty
q_{hi}	specific waste heat
q_{reb}	specific reboiler heat duty
RC	reboiler condensate
RID	relative interheater duty
RLHX	rich-lean heat exchanger
SA	spray attemperation
SCPC	supercritical pulverised coal
SF	split flow
TCR	total capital requirement
TPC	total plant costs
VR	vapour recompression
w_{aux}	specific auxiliary duty
VC	vapour compressor

1 Introduction

1.1 Background

As commonly agreed, climate change will be a serious economic and ecologic challenge in the next decades. To limit the global temperature rise to 2 °C, a reduction of greenhouse gas (GHG) emissions by 80%, compared to 1990, until 2050 is recommend by the IPCC (Intergovernmental Panel on Climate Change) [1]. The emissions from fossil-fuelled power plants can be reduced by increasing the energy conversion efficiency or by separating and withholding carbon dioxide (CO_2), commonly referred to as carbon capture and storage (CCS). The post-combustion capture (PCC) technology is a promising possibility to reduce CO_2 emissions from fossil fuel fired power plants. One of the main concerns for the PCC is the rather large efficiency penalty. A reduction in efficiency penalty for solvent based PCC can be achieved by improving the solvent properties as well as by improving the process design.

The solvent determines the process behaviour and the efficiency penalty. A lot of solvents have been modelled and tested in pilot plants [2]. Important interface quantities for the overall process are the specific reboiler heat duty and the reboiler temperature, which strongly depend on the solvent CO_2 absorption characteristics.

There are different process flow sheet modifications with an improvement in process design reported in various literature [3, 4, 5, 6]. These process modifications potentially can reduce the efficiency penalty of the overall process. Some of the promising process flow sheet modifications are multicomponent column, inter-stage temperature control, heat integrated stripping column, split flow process, vapour recompression, matrix stripping and various heat integration options.

1.2 Aim and scope

A detailed comparison of the overall efficiency for different process flow sheet modifications with an improved solvent is necessary because most evaluations of these processes in literature are based on different boundary conditions and different solvents. Therefore, there is a requirement to evaluate these process modifications on similar solvent and process conditions. For this study a supercritical pulverised coal fired power plant (SCPC) and a natural gas combined cycle power plant (NGCC) were chosen to be evaluated for different CO_2 capture process modifications.

In this study, first, a process description of the capture unit and a solvent selection are done. The modelling approach for the capture plant, the power plants and the CO_2 compressor are presented. A technical evaluation of the different process flow sheet modifications is subsequently performed and additional aspects of interest are worked out in a qualitative analysis. In an economic evaluation, different process flow sheet modifications are compared. Major gaps are identified and recommendations are made. A summary concludes the study.

2 Process description

A schematic flow diagram of a typical plant for post-combustion CO_2 capture by chemical absorption is shown in Figure 1. To improve the CO_2 absorption process, the flue gas is first cooled before entering the absorber column at the bottom. As the flue gas rises in the column, the CO_2 is absorbed by a chemical solvent in aqueous solution in a counter-current flow. The column is filled with random or structured packing to increase the interfacial area between gas and liquid phase. A washing section at the top of the absorber reduces the slip of solvent to the environment by contacting the outgoing treated flue gas with cold water. An induced draft (ID) fan is required to overcome the additional pressure losses in the flue gas cooler and the absorber. The treated flue gas at the top of the absorber is released to the atmosphere. At the bottom of the absorber, the CO_2-rich solution is gathered and pumped to the desorber, passing a rich-lean heat exchanger (RLHX) where it is preheated to a temperature close to desorber temperature.

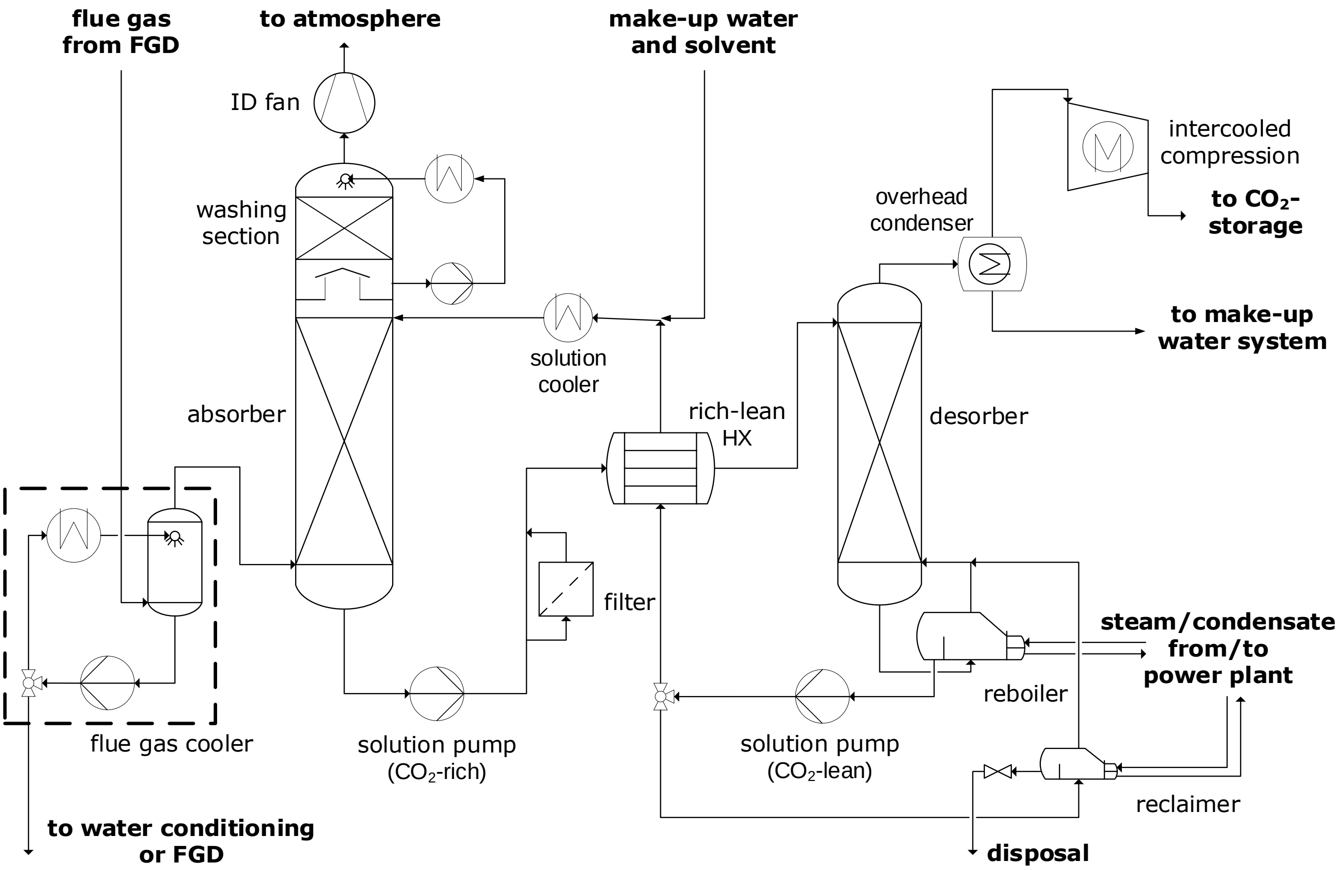


Figure 1: Process flow sketch [7]

In the desorber, the absorbed CO_2 is stripped from the rich solution at high temperature and the solvent is regenerated. The rich solution flows downwards and releases the captured CO_2. The necessary driving force (partial pressure difference) and sensible heat as well as heat for the separation of CO_2 from the solvent is delivered by a counter-current flow of vapour (stripping steam), consisting mainly of steam and CO_2. The required heat duty is provided by the reboiler, in which steam from the power plant is condensed and vapour (stripping steam) is generated.

At the head of the desorber, the gas is led to the overhead condenser (OHC) where the CO_2-rich gas stream is cooled and part of the water vapour is condensed. The remaining gas stream can be compressed and is then ready for transportation to a storage site. An additional washer downstream the OHC might in practice be necessary to reduce the amine content in the CO_2, but is not incorporated in this study. The CO_2-lean solution is gathered at the bottom of the reboiler and is returned to the absorber, passing the RLHX and another heat exchanger (solution cooler), in which the temperature is lowered to the desired absorber temperature. The lean solution is dispersed at the top of the absorber column, closing the process cycle.

3 Solvent Selection

In the chemical solvent based post-combustion CO_2 capture process, the solvent determines the process behaviour. A lot of work has been done in the field of solvent development and there are various solvent type available for CO_2 absorption. The characteristics of some of these with respect to the key factors relevant for CO_2 capture are listed in Table 1. The heat of absorption and the CO_2 absorption capacity are important factors relevant for the energy requirement of the capture process. While a high CO_2 capacity is generally beneficial for the process, the working range of the solvent, the difference between the effectively reached lean and rich CO_2 loadings have a higher impact on the process. This is due to the fact that the working range determines the required solution mass flow. The absorption rate affects the absorber design, since a solvent with low absorption rate would require a long hold-up time and thus a higher absorber or a packing with a higher specific area in order to reach CO_2 loadings close to equilibrium. A low degradation tendency of the solvent is essential, since solvent loss has to be as low as possible for an economic operation of a CO_2 capture plant.

Table 1: Simplified overview of solvent properties

	heat of absorption*	absorption rate	CO_2 capacity	degradation tendency
MEA	●	●	◑	●
DEA	●	◑	◑	◑
MDEA	◑	○	●	○
AMP	●	◑	●	○
PZ	●	●	◑	◑
K_2CO_3	○	○	●	○
NH_3	◑	◑	●	○

● = high; ◑ = medium; ○ = low;

* Note that the heat of absorption represents only a fraction of the total energy requirement for the regeneration of the solution.

MEA: monoethanolamine; DEA: diethanolamine; MDEA: methyldiethanolamine; AMP: 2-amino-2-methyl-1-propanol; PZ: piperazine; K_2CO_3: potash; NH_3: ammonia

It can be seen in Table 1 that no existing solvent excels the others in all properties. The tertiary amine methyl diethanolamine (MDEA), for example, has a low degradation tendency and high CO_2 capacity, but the absorption rate is low. A promising approach is therefore to blend different solvents in order to combine the positive properties of both solvents. One of these blends for example is a mixture of MDEA with the polyamine piperazine (PZ), which has higher rates of absorption in the absorber compared to MDEA, while maintaining its low heat of regeneration in the desorber [8].

In accordance with the technical specification of this project, the absorption process shall use a generic improved solvent, representing a future solvent, with generically improved CO_2 absorption properties probably available in the coming years. Improvements are possible for the above mentioned CO_2 absorption properties, as well as for the solvent corrosion behaviour, the vapour pressure and the viscosity. It is not reasonable, though, to design a solvent with better values compared to all existing solvents for all above mentioned properties.

Therefore a solvent for this study called Solvent2020 was developed. It is an artificial solvent which has the same CO_2 absorption mechanisms as amines (carbamate and bicarbonate formation). The properties like density, viscosity or heat capacity are assumed to be similar to those of a solution with 7 mol MDEA and 2 mol PZ per kg H_2O. Thus, the corresponding ASPEN Plus® property model is used for the simulations [9].

The reaction kinetics of Solvent2020 are enhanced compared to 7MDEA/2PZ, though, which results in chemical reactions that are not kinetically hindered. This is the main property improvement compared to other solvents for solvent 2020. This assumption is used for modelling of desorbers with state-of-the-art solvents, as well. Due to the high temperatures, which catalyse the chemical reactions of CO_2 desorption, this is found to be a reasonable approach. The absorber is generally not assumed to be in chemical equilibrium, though. Despite the chemical equilibrium, the columns are not in total equilibrium, since mass and heat transfer are calculated by rate based modelling. This approach would overestimate the absorption rate of a slower solvent but is assumed to be reasonable for fast solvents.

The CO_2 absorption loading of the solvent is an important parameter for the process design and is shown in Figure 2 where the CO_2 partial pressure is plotted against the CO_2 loading of the aqueous amine based solution for different temperatures. The CO_2 loading range of this solvent for a typical process condition is between 0.2 and 0.4 mol CO_2/mol amine. The heat of absorption differs for relevant temperatures and loadings ranging between 60 and 75 kJ/mol CO_2.

Solvent2020 is assumed to be thermally stable up to approximately 150 °C, which is the same temperature as for PZ. Thus, thermal degradation is not expected to occur when operated at temperatures below this limit. Oxidative degradation is assumed to be negligible, as well. In addition, Solvent2020 is assumed to be not corrosive in the chosen operating range.

The results obtained with this solvent are solvent specific, as for all other solvents. The conclusions drawn from this are thus not generally valid for all solvents, but give a good idea of the possible performance of future solvents.

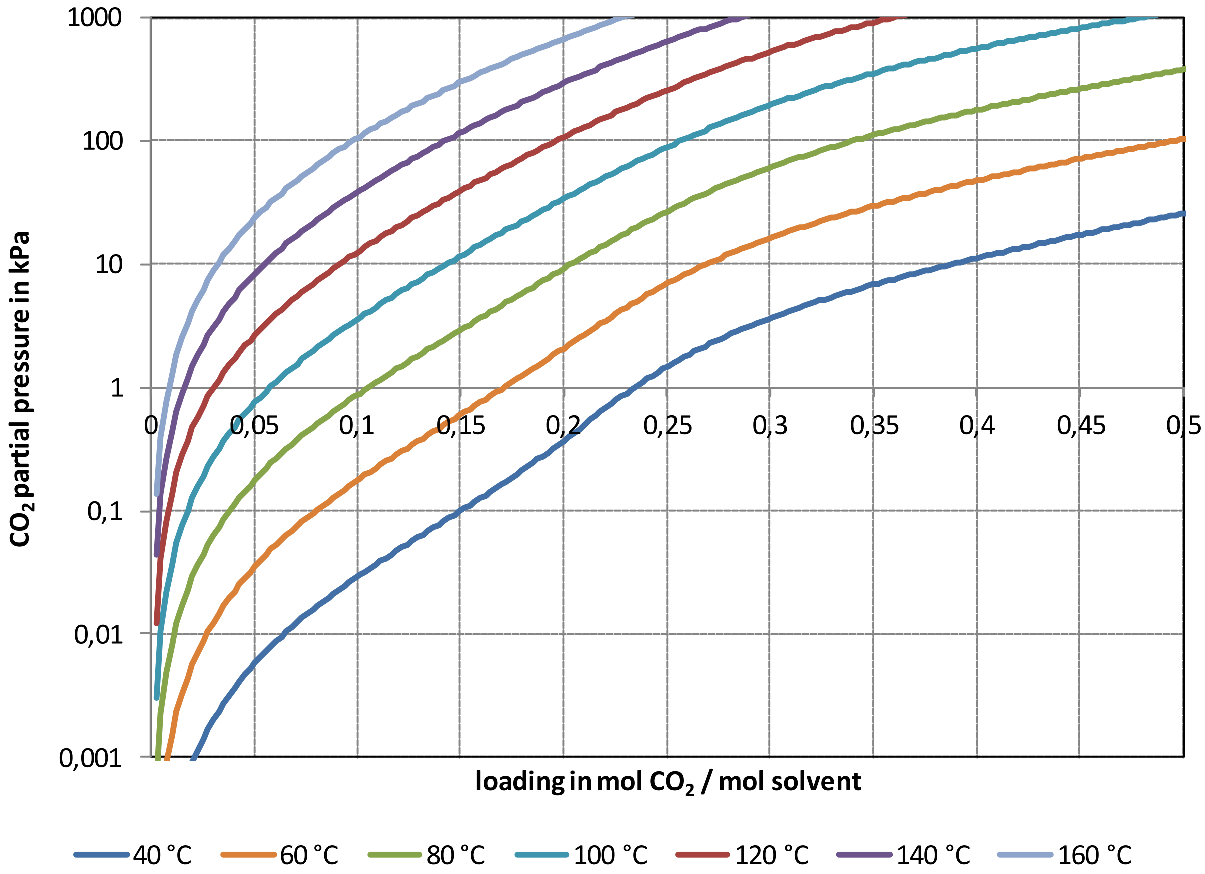


Figure 2: CO_2 partial pressure against CO_2 loading of Solvent2020 for different temperatures

4 Modelling approach

The overall process consists of the power plant, the PCC process and the CO_2 compression. For each sub-process, the most suitable simulation tools for steady-state simulations are chosen, Ebsilon®Professional for the overall power plant and the CO_2 compression and Aspen Plus® for the CO_2 capture process. Between the simulation tools, interface quantities are defined and used to analyse the overall process performance.

The process flow schemes for the CO_2 capture processes are established on the basis of the scheme shown in Figure 1. The liquid properties are computed using the electrolyte non-random two liquid (ELECNRTL) method, the vapour properties are computed using the Redlich-Kwong equation of state; CO_2, N_2, O_2, CO and H_2 are selected as Henry-components.

The mass and heat transfer in the columns is calculated by rate-based modelling with differential mass and energy balances at the phase boundary between liquid and vapour phase. The diffusion resistance is hereby assumed to occure solely in a film between the two phases, while the rest of the respective phase is in equilibrium. The film is divided into a liquid film and a gaseous film [10].

The mass transfer coefficients and the interfacial area are calculated using the correlation of Bravo et al. [11]. The heat transfer coefficients are calculated using the Chilton-Colburn method [12].

The columns are filled with structured packing Sulzer MELLAPAK 250.Y. The effective packing surface area is thereby fixed to approx. 250 m^2/m^3. The retention time and the pressure drop for the packing are calculated using vendorspecific correlations. Different packing materials could be used, possibly leading to smaller columns. Since the focus of this study is on the process flowsheet modifications, there would not be any benefit.

In this model the chemical reactions take place only in the liquid phase and are not kinetically hindered. Thus, they are modelled with an equilibrium reaction model. This approach is chosen for the desorber as well as for the absorber since the absorption is assumed to be very fast, as stated earlier.

The absorber and desorber column diameter is adjusted for each design point to reach an optimal loading, which is at 70% of the maximal loading. The maximal loading is achieved 5 – 10% below the flooding point and the optimal operation range is between 50% and 80% below the flooding point. Columns with very large diameters are not reasonable, though. The maximum diameter is thus set to 18 meters [13]. It is possible that multiple serial capture units are needed to ensure that this limit is not exceeded. The solution from each absorber is regenerated in a separate desorber.

The flue gas from the power plant is first cooled down in the flue gas cooler, which is modelled using a flash unit. A water stream is cooled to 24 °C and led to the flue gas in counter–current flow. The water mass flow is adjusted to reach the desired flue gas temperature of 40 °C. The water leaving the flash unit is pumped back to the cooler. A fraction of the water is removed to ensure a stable water balance and to prevent enrichment of particles.

Solution mass flow to the absorber is adjusted to reach a CO_2 capture rate of 90%. The CO_2 capture rate is the ratio between the CO_2 mass flow absorbed in the absorber and the CO_2 mass flow in the flue gas.

The rich solution from the absorber and the lean solution from the stripper are cross heat exchanged in the rich-lean heat exchanger (RLHX). The log mean temperature difference (LMTD) is set to 5 K in order to allow for good heat exchange. The following definition of LMTD is used:

$$LMTD = \frac{\Delta T_H - \Delta T_C}{\ln\left(\frac{\Delta T_H}{\Delta T_C}\right)} \tag{1}$$

with the temperature difference at the hot side $\Delta T_H = T_{hot\ lean\ solution} - T_{hot\ rich\ solution}$ and the temperature difference at the cold side $\Delta T_C = T_{cold\ rich\ solution} - T_{cold\ lean\ solution}$.

The desorber is equipped with kettle type reboilers, since this reboiler type is operated with a low temperature approach and is very reliable [14].

The absorber is modelled using a RadFrac unit. Differing from the process configuration shown in Figure 1, the absorber is designed as an intercooled absorber. The solution is withdrawn at half height of the absorber and cooled down to 40 °C, which is the inlet temperature at the absorber head, as well. The cooled solution is fed back directly downstream of the extraction. The absorber height is optimized for the base case, for the other cases it was kept constant. This simplification is expected to be of no relevance for the comparison of the different flow sheet modifications, since most process modifications do not affect the absorber.

The washing section on the flue gas side downstream the absorber is modelled using a RadFrac unit and has the same diameter as the absorber. As for the flue gas cooler, a water cycle is used to model the cooling and pumping of the washing water. A fraction of the washing water is removed and led to the absorber bottom.

The pumps and blowers in the capture process are modelled using the isentropic and mechanical efficiencies given in Table 2. The blower has to overcome the additional pressure drop in the capture plant. The pumps have to overcome the pressure drop in the heat exchangers and pipes. In addition, the hydrostatic pressure due to the height difference between pump outlet and column inlet has to be taken into account. The pressure drop in the columns is calculated by vendor specific correlations. The pressure drop in the heat exchangers is assumed to be 0.5 bar and includes the pressure drop in the pipes.

Table 2: Capture process boundary conditions

CO_2 capture rate	90%
Absorber inlet temperature flue gas and solvent	40 °C
RLHX LMTD	5 K
Isentropic/mechanical efficiency of the pumps	85%/99.5%
Isentropic/mechanical efficiency of the blowers	83%/99.5%

5 Power Plants and CO_2 Compression

The heat for solvent regeneration is commonly provided by extracting low-pressure steam from the water-steam-cycle of the power plant. The magnitude of the efficiency penalty is not only determined by the amount of extracted steam (quantity) but also by the quality of extracted steam (steam pressure) [15, 16]. When optimising process parameters of the CO_2 capture unit such as the solution circulation rate or the desorber pressure, the variation of these parameters can have opposite effects on the required steam quantity and quality:

- Solution circulation rate: An increase in solution circulation rate can enlarge the amount of stripping steam, because more sensible heat is required to heat up the solution from absorber to desorber temperature (heat quantity ↑). If the CO_2 capture rate is assumed to be constant, an increased solution circulation leads to a higher lean loading, meaning a smaller degree of regeneration in the desorber. The higher lean loading is achieved at lower temperatures. Therefore, steam at a lower pressure level can be used for solvent regeneration (heat quality ↓).
- Desorber pressure: For solvents with a high heat of absorption such as MEA, an increase in desorber pressure and reboiler temperature leads to a smaller amount of water vapour at the desorber head [17]. Therefore, less heat and less steam must be provided in the reboiler (heat quantity ↓). The increase in desorber pressure and the corresponding increase in reboiler temperature, however, is equivalent to a need for steam with higher pressure (heat quality ↑).

These two examples illustrate that an overall process optimisation requires the consideration of the impact of process parameters not only of the CO_2 capture unit in an isolated manner, but of the overall process in a holistic approach. Therefore, adequate models of the power plant and the compression train are essential.

5.1 SCPC Model

To facilitate comparisons with currently planned power plant projects, the model used in this work is based on a state-of-the-art supercritical pulverised coal power plant. The power plant is modelled with the commercial software tool EBSILON®*Professional*. The coal-fired power plant with high-pressure and high-temperature steam (295 bar, 600 °C) has a gross electrical power output of 900 MW. At its design point (full load operation without CO_2 capture), the net efficiency is 45.2%, related to the LHV. The schematic flow diagram of the reference power plant is shown in Figure 3. The ambient air, which is taken from the inside of the boiler building, is split into primary air and secondary air. While the secondary air is sent directly to the boiler, the primary air is used for preheating a feed water bypass and then used as mill air. A steam preheater is foreseen to increase the air temperature at the air preheater inlet and thus also to increase the flue gas temperature and thereby avoiding local passing below the dew point. The flue gas cleaning consists of the common three cleaning steps:

- DeNOx,
- electrostatic precipitator (ESP),
- flue gas desulphurisation (FGD).

The preheating train consists of five LP-preheaters, the feed water tank and three HP-preheaters. Just before entering the boiler unit, the feed water is heated to 300 °C. The cooling system is based on a natural draught cooling tower which supplies cooling water at 16 °C. With a temperature gain in the condenser of 10 K and a temperature approach of 3 K the condenser pressure is determined to be 40 mbar.

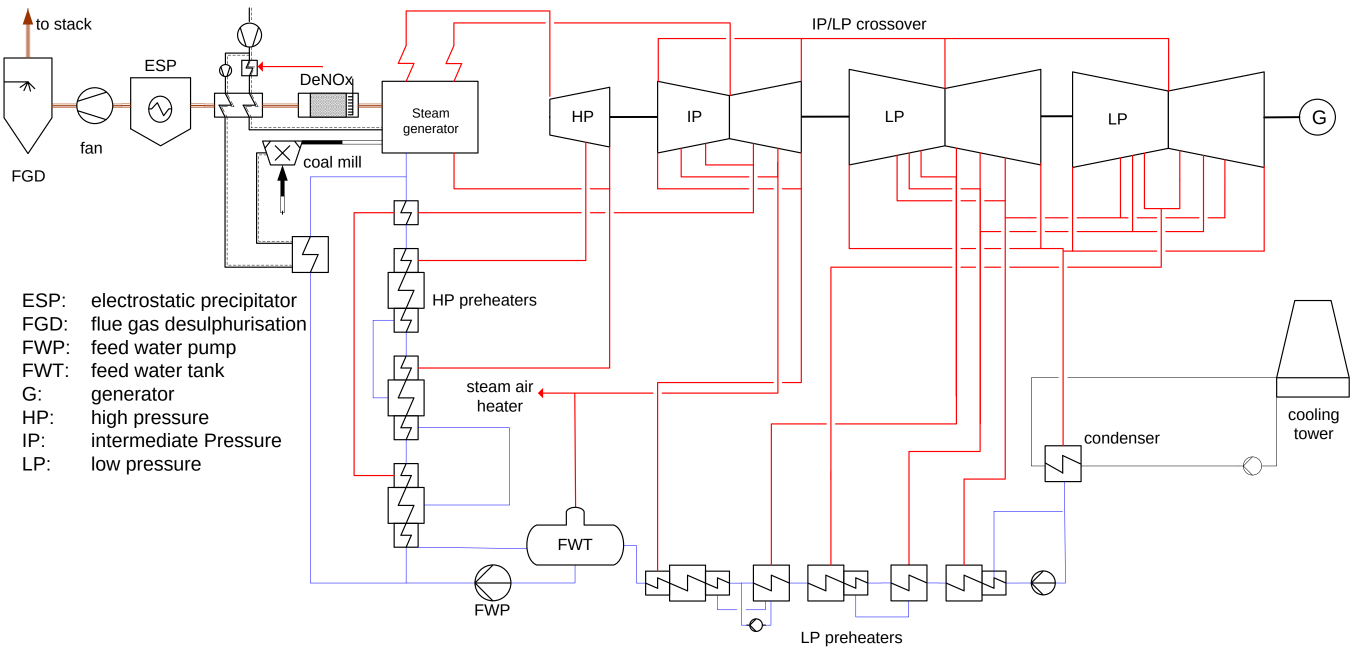


Figure 3: Flow sheet of the SCPC plant without CO_2 capture

The major characteristics of the SCPC model are summarised in Table 3. The flue gas data downstream of the FGD unit serve as interface quantities between the power plant and the capture plant models.

Table 3: Characteristics of the SCPC model without CO_2 capture

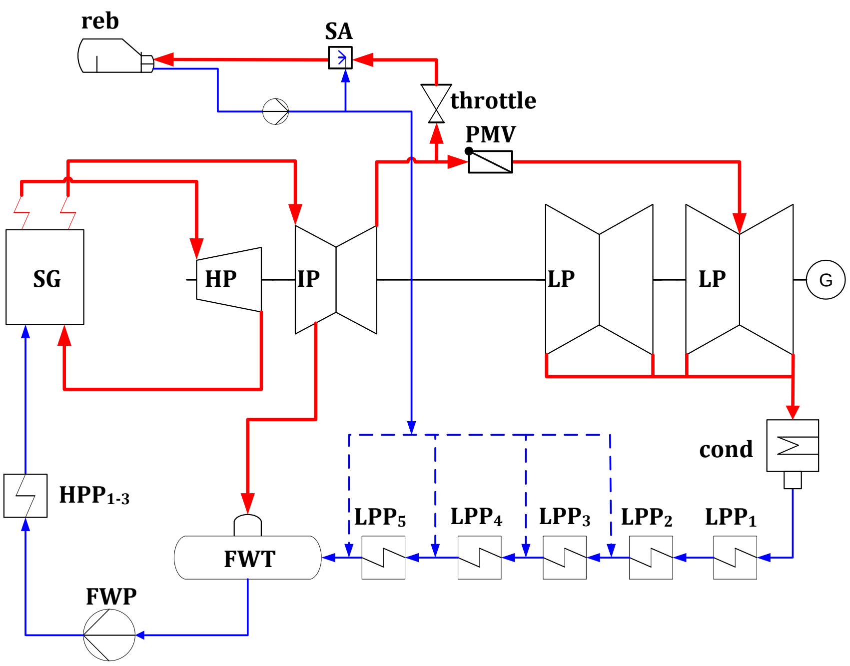
Heat input	1835.44 MW_{th}
Net output	830.48 MW_{el}
Gross output	900.00 MW_{el}
Net efficiency	45.2%
Gross efficiency	49.0%
Specific CO_2 emissions	769 g/kWh
Live steam temperature	600 °C
Live steam pressure	295 bar
Hot reheat temperature	620 °C
Hot reheat pressure	55 bar
Condenser pressure	40 mbar
Flue gas downstream of FGD	
Mass flow	869.64 kg/s
Pressure	1.018 bar
Temperature	50 °C
CO_2	13.5 Vol%
H_2O	12.0 Vol%
N_2	70.2 Vol%
O_2	3.5 Vol%
Ar, SOx, NOx	0.8 Vol%

5.1.1 Basic Integration

For the overall process evaluation, the Greenfield case will be taken into consideration. In this case the power plant is designed for the operation with CO_2 capture. A retrofit of an existing power plant would be very site specific and could influence different flow sheet modifications in different ways, making a comparison of the modifications impossible. The water-steam-cycle is adapted so that the steam pressure in the IP/LP crossover matches the extraction pressure required for CO_2 capture at full-load operation. This eliminates the losses induced by steam conditioning measures such as a throttle or a pressure maintaining valve that occur in the retrofit integration case [16]. Note that a perfect match of IP/LP steam pressure and extraction pressure required for CO_2 capture is only valid for one operational point. As soon as the power plant load or the process parameters of the capture unit are changed, the throttle or the pressure maintaining valve must be activated leading to an additional energy penalty. The pressure drop Δp_{ext} in the steam pipe between IP/LP crossover and reboiler is assumed to be 0.3 bar. Furthermore, the mean temperature difference ΔT_{reb} in the reboiler is assumed to be 10 K. The extraction pressure p_{ext} can be calculated with equation (2).

$$p_{ext} = p_{sat}(T_{reb} + \Delta T_{reb}) + \Delta p_{ext} \tag{2}$$

The simplified flow sheet of the basic reboiler integration is shown in Figure 4. The reboiler condensate is returned to the preheating route where the feed water shows the closest temperature. To avoid hot spots in the reboiler which could lead to thermal degradation of the solvent or increased fouling in the reboiler, the steam for solvent regeneration has to be almost saturated (superheated steam 15 K above boiling temperature). This is realised by recycling and injecting reboiler condensate into the superheated steam (spray attemperation).



cond: condenser
FWT: feed water tank
FWP: feed water pump
G: generator
HP: high pressure
HPP: high pressure preheater
IP: intermediate pressure
LP: low pressure
LPP: low pressure preheater
PMV: pressure maintaining valve
reb: reboiler
SA: spray attemperation
SG: steam generator
WH: waste heat

Figure 4: Basic integration for the SCPC case [18]

The pressure levels of the steam tappings for the preheating train are optimised using a nested one-dimensional iterative solution method. For each desired IP/LP crossover pressure the pressure levels of the steam tappings are adapted to ensure an equivalent comparison among different reboiler temperatures. The boiler island is not affected by the CO_2 capture unit and is thus identical to the case without CO_2 capture.

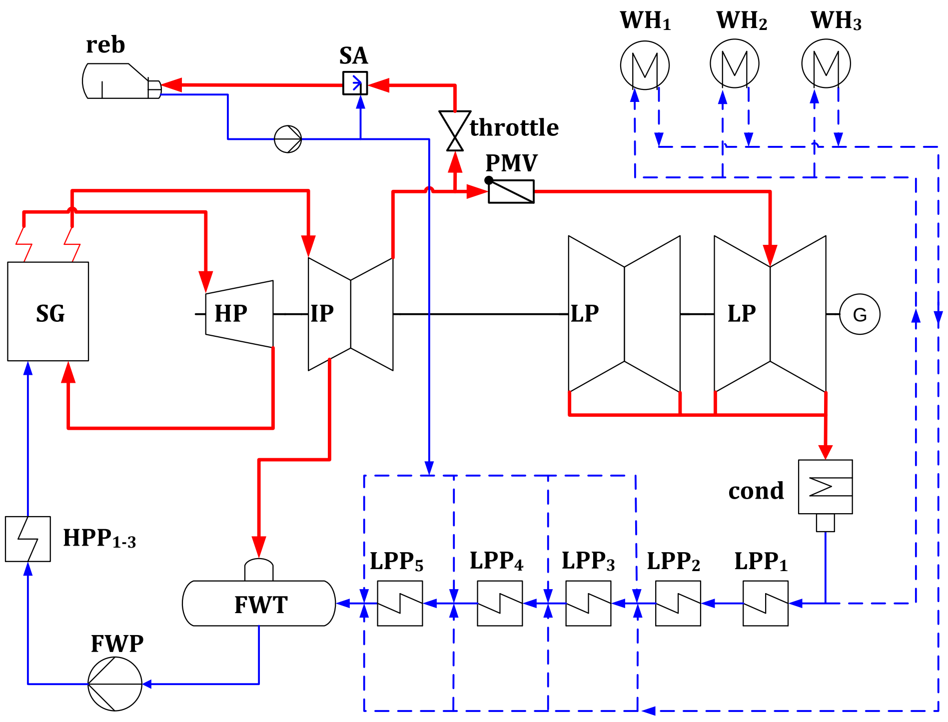
5.1.2 Heat Integration

Besides the steam extraction and optimised integration of the reboiler condensate (Basic integration), waste heat sources are identified and used for feed water preheating. A typical waste heat source within the capture process is the overhead condenser, where the CO_2 stream is cooled to condense remaining steam. In this case the temperature level is around 10-20 K below the reboiler temperature.

Another reasonable waste heat source is the intercooling of CO_2 compression. The temperature level depends on the number and position of intercoolers and can hence be directly influenced. The higher the temperature level, the more efficient the waste heat can be integrated in the power plant. However, a higher temperature level leads to an increased electrical power duty of the engine drive. The energetic optimum of these two opposing effects lies between the two extreme cases (minimal electrical power duty and maximal temperature level). Therefore, both effects have to be included into the overall process optimisation.

As heat sinks the combustion air and the water steam cycle of the power plant are available. Preheating of the combustion air is realised by air preheaters where sensible heat from the flue gas is transferred to the combustion air. Furthermore, a steam preheater is provided to increase the air temperature at the air preheater inlet and thus also the flue gas temperature to avoid local passing below dew point. Even if waste heat integration could (from the energetic point of view) substitute the steam preheater, the control of the flue gas temperature at the preheater outlet still requires a steam preheater. To maximise the effect of waste heat integration for combustion air preheating, enormous capital expenditures are required. Thus, the combustion air does not represent a realistic heat sink for waste heat integration [19].

Another heat sink is the preheating route of the water steam cycle (see Figure 5). The low pressure condensate has a pressure of less than 20 bar and can (as a parallel stream) be transported to the waste heat sources. The amount of waste heat, which can be integrated in the preheating train, strongly depends on the available condensate mass flow. Therefore, a high heat duty of the capture process leads to a limited potential of waste heat integration. The temperature level is limited by the feed water tank. An undercooling of 5 – 20 K is required to ensure degasification in the feed water tank. Further approaches for heat integration (e. g. district heating) are classified to be very special and are thus neglected in this study.



cond: condenser
FWT: feed water tank
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HP: high pressure
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PMV: pressure maintaining vavle
reb: reboiler
SA: spray attemperation
SG: steam generator
WH: waste heat

Figure 5: Waste heat integration for the SCPC case [18]

Several waste heat sources are concurring as the heat sinks are limited. To find the best waste heat utilisation the following issues should be considered:

- waste heat, which is available without additional energetic effort, should be preferred;
- waste heat at a high temperature level should be preferred.

These two issues lead to an optimisation algorithm for the integrated waste heat $q_{\mathrm{hi,used}}$ from a waste heat source i (see equation (3)). As a precondition, the waste heat sources have to be sorted following the two issues above. That means that the waste heat source on the highest temperature level, which is available without additional effort, gets the index i = 1.

$$q_{\mathrm{hi,used}_i} = \max\left\{0, \left[\min\left(q_{\mathrm{hi,max}_i} - \sum_{k=1}^{i-1} q_{\mathrm{hi,used}_k}, q_{\mathrm{hi}_i}\right)\right]\right\} \quad (3)$$

q_{hi} is the available waste heat, $q_{\mathrm{hi,max}}$ is the maximal integrable waste heat. This algorithm enables the optimal utilisation of several waste heat sources. Waste heat, which cannot be integrated, is added to the cooling duty.

The potential of waste heat integration with regards to the net efficiency is shown in Figure 6. As described above, a higher heat duty leads to a reduced condensate mass flow and thus to a lower maximal net efficiency increase. The temperature level of the waste heat does not only affect the exergy ratio in the waste heat, but also the amount of integrable waste heat. Low temperature levels of the waste heat lead to steam extraction for the preheating route, which lowers the available condensate mass flow [18].

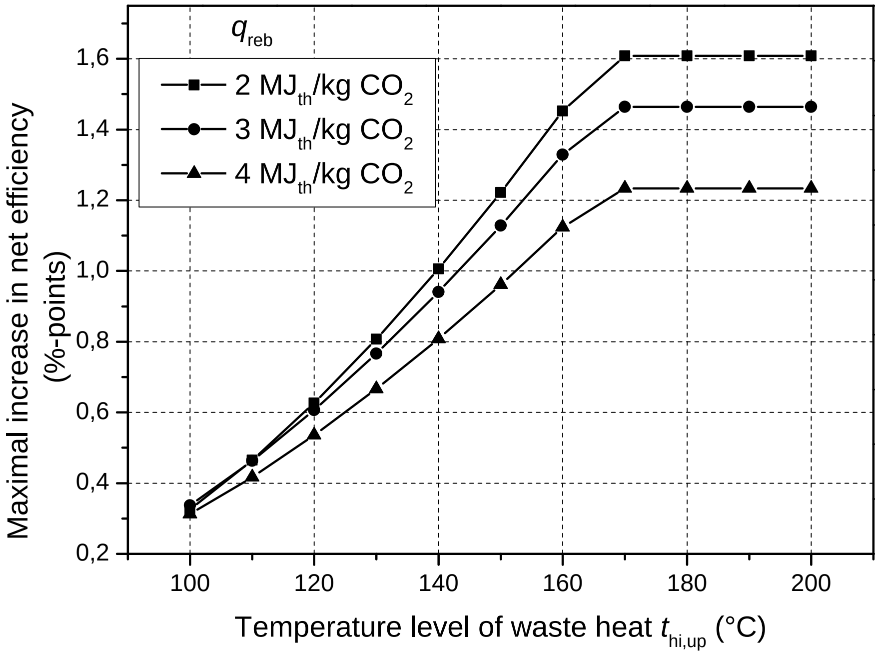

Figure 6: Maximal increase in net efficiency through heat integration for different heat duties for solvent regeneration [18]

The main drawback of a highly integrated process is the increased process complexity. Adding more equipment and piping increases the number of control variables making process control more complex. In addition, the number of components that can potentially fail is increased reducing plant availability.

5.2 NGCC Model

The model used in this work is based on a state-of-the-art natural gas combined cycle plant (NGCC). The power plant is modelled with the commercial software tool EBSILON®*Professional*. The plant consists of two gas turbines, each of which is equipped with a heat recovery steam generator (HRSG) to use the heat of the flue gas downstream the gas turbine. The steam produced in the two HRSGs is lead to a common steam turbine. The whole plant has a gross electrical output of 883 MW, consisting of 278 MW from each of the gas turbines and 327 MW from the steam turbine. The net efficiency of the power plant in full load operation without CO_2 capture is 58.2%, related to the LHV. The schematic flow diagram of the reference power plant is shown in Figure 7.

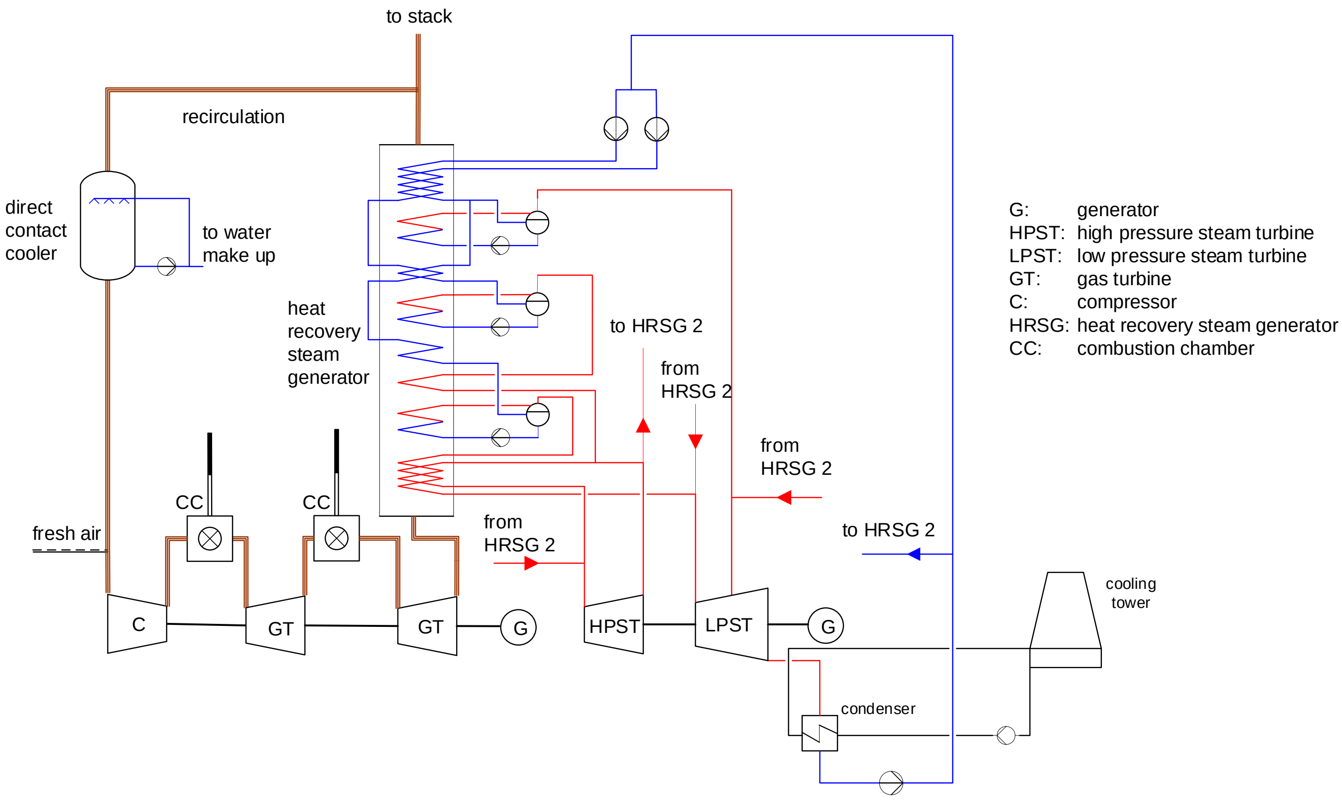

Figure 7: Flow sheet of one gas turbine, its HRSG and the steam turbine of the NGCC plant without CO_2 capture

The gas turbine is a sequential combustion gas turbine delivering high flue gas temperature for the subsequent HRSG. The water-steam cycle is a three pressure level process (live steam 585 °C, 159 bar) with a reheat (585 °C, 40 bar). The cooling system is based on a mechanical draught cooling tower which supplies cooling water at 19 °C. With a temperature gain in the condenser of 11 K and a temperature approach of 3 K the condenser pressure is determined to be 45 mbar.

The CO_2 concentration in the flue gas of an NGCC plant is very low compared to the flue gas of an SCPC plant (4.2 Vol.-% for NGCC, 13.5 Vol.-% for SCPC). This is equivalent to a reduced partial pressure of CO_2 which increases the energy requirement for the capture plant. In order to minimize this energy requirement for the capture plant, flue gas recirculation (FGR) is used. Part of the flue gas downstream the HRSG is recirculated, cooled down in a direct contact cooler and led back to the compressor inlet, where it is mixed with fresh air. At a recirculation rate of 0.54 (ratio of recirculated flue gas to flue gas leaving the HRSG), the CO_2 concentration in the flue gas is increased to 9.1 Vol.%. Higher recirculation rates are not reasonable, since the O_2 concentration in the combustion chamber would be too low to ensure stable combustion conditions [20].

The major characteristics of the NGCC model with and without FGR are summarised in Table 4. The flue gas data downstream of the HRSG unit serve as interface quantities between the power plant and the capture plant models.

Table 4: Characteristics of the NGCC model without CO_2 capture

	NGCC plant with FGR	NGCC plant w/o FGR
Heat input	1520.79 MW_{th}	1504.49 MW_{th}
Net output	874.00 MW_{el}	874.00 MW_{el}
Gross output	884.34 MW_{el}	883.85 MW_{el}
Net efficiency	57.47%	58.09%
Gross efficiency	58.15%	58.75%
Specific CO_2 emissions	356 g/kWh	356 g/kWh
Compressor pressure ratio	34	34
Gas turbine exhaust temperature	619 °C	619 °C
Live steam temperature	585 °C	585 °C
Live steam pressure	159 bar	159 bar
Hot reheat temperature	585 °C	585 °C
Hot reheat pressure	40 bar	40 bar
Condenser pressure	45 mbar	45 mbar
Flue gas downstream of FGR/HRSG		
Mass flow	621.75 kg/s	1321.79 kg/s
Pressure	1.018 bar	1.018 bar
Temperature	84.8 °C	85.2 °C
CO_2	9.1 Vol.%	4.2 Vol.%
H_2O	10.1 Vol.%	8.7 Vol.%
N_2	76.7 Vol.%	74.3 Vol.%
O_2	3.2 Vol.%	11.9 Vol.%
Ar, NOx	0.9 Vol.%	0.9 Vol.%

5.2.1 Integration

In conformity with the SCPC case, the Greenfield case is taken into consideration also for the NGCC process. The water-steam-cycle is adapted so that the steam pressure between the IP and the LP steam turbine matches the extraction pressure required for CO_2 capture at full-load operation. The pressure drop Δp_{ext} in the steam pipe between steam turbine and reboiler is assumed to be 0.3 bar and the mean temperature difference ΔT_{reb} in the reboiler is assumed to be 10 K. The extraction pressure p_{ext} can be calculated with Equation (2).

The simplified flow sheet of the basic reboiler integration is shown in Figure 8. As for the coal case, reboiler condensate is injected into the superheated steam (spray attemperation) to reduce the temperature and prevent hot spots in the reboiler. The remaining reboiler condensate is partially returned to the water steam cycle upstream the economiser of the heat recovery steam generator to increase the temperature to 60°C and thus prevent condensation of vapour in the flue gas. The rest of the condensate is returned downstream the economiser.

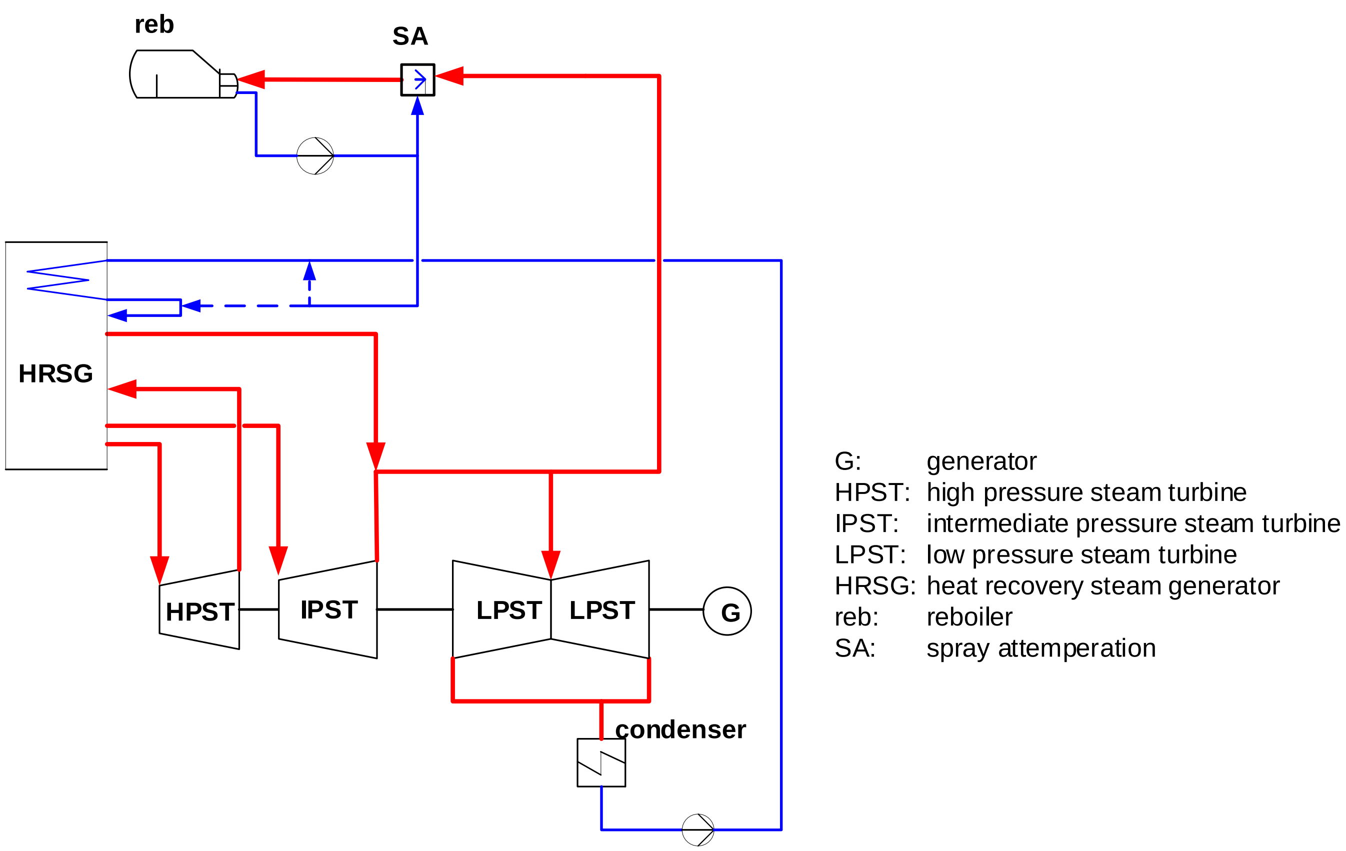

Figure 8: Basic integration for the NGCC case

A more complex waste heat integration is not applied for the NGCC case. The potential waste heat sources are similar to the sources available for the coal case, but there are no heat sinks available. Preheating the condensate, as it is done for the coal case, is possible, but does not have a positive effect on the efficiency, since an increased feed water temperature leads to an increased exhaust gas temperature. The benefit of the additional heat source is thus counterbalanced by increased exhaust gas losses. This is especially crucial for the NGCC plant with CO_2 capture, since the flue gas has to be cooled to 40 °C for good absorption. An increased exhaust gas temperature results in higher cooling duties in the capture plant.

5.3 CO_2 Compression

For CO_2 compression an integrally geared multi-stage (radial) compressor is considered (see Figure 9). After each intercooler, the condensed water is disposed with a drain valve. To further eliminate water from the CO_2 stream, an adsorptive drying unit is provided.

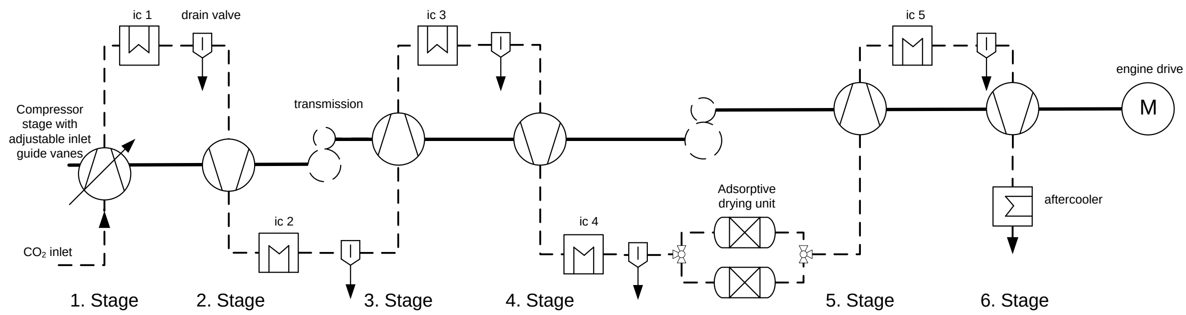

Figure 9: Schematic flow diagram of the CO_2 compressor model

A calculation method using real gas behaviour is chosen to model the non-ideal gas behaviour of CO_2 during compression and cooling. Calculating with ideal gas behaviour would lead to inaccuracies of approximately 10% related to the overall energy requirement.

In Table 5 the boundary conditions of the compressor model are listed exemplarily for a compressor with 6 stages. The pressure drop needed for the application of adsorption beds in the drying unit is assumed to be 100 mbar. Furthermore, the pressure ratio of each stage is decreased by 2% per stage, because of the inherent rotor dynamics of integrally geared compressors.

Table 5: Boundary conditions of the CO_2 compressor model

Characteristics	**Stage 1**	**Stage 2**	**Stage 3**	**Stage 4**	**Stage 5**	**Stage 6**	**Engine drive**
Pol. (/el.) efficiency	85 %	84 %	83 %	82 %	81 %	80 %	(97 %)
Pressure loss in ic/ac	20 mbar	40 mbar	60 mbar	80 mbar	100 mbar	120 mbar	-
Mechanical efficiency	99 %	99 %	99 %	99 %	99 %	99 %	99.8 %

To reach high polytropic efficiencies, high velocities are required. The inlet Mach number in front of each stage is limited to approximately 0.9 to prevent shock waves in the blade passages. The Mach number is a function of molecular weight, and therefore the polytropic efficiencies for the heavy CO_2 (~44 g/mol) are lower compared to air (~29 g/mol). All assumptions mentioned agree well with information from manufacturers [21, 22].

With regards to waste heat integration, the compression process possesses three energetic interface quantities, which have to be considered:

- the specific power duty w_{comp} (MJ_{el}/kg CO_2);
- the specific cooling duty q_{comp} (MJ_{th}/kg CO_2) accruing in the intercoolers;
- the specific waste heat $q_{wh,comp}$ (MJ_{th}/kg CO_2) at the temperature level $t_{wh,comp}$ (°C).

The part of the waste heat, which cannot be integrated in the power plant process counts to the cooling duty.

From the energetic point of view, the compressor configuration with the highest possible number of intercoolers is the most beneficial one. With consideration of waste heat integration, the best compressor configuration strongly depends on the availability of low temperature heat sinks in the power plant. Both, the quality and quantity of waste heat can be varied by the number and position of intercoolers.

The overall pressure ratio consists of the fixed outlet pressure of 110 bar and the desorber pressure of the capture process. The average pressure of each stage π_{stage} can be calculated with equation (4).

$$\pi_{\text{stage}} = \sqrt[n_{\text{stage}}]{\Pi} \qquad (4)$$

The results of equation (2) are shown in Figure 10 for 4, 6 and 8 stages. The grey shaded area shows the range of reasonable pressure ratios per stage, in this case assumed to be 1.4 – 2.1. With these three stage numbers, considered inlet pressures between 0.3 bar (8 stages) and 28.6 bar (4 stages) are covered.

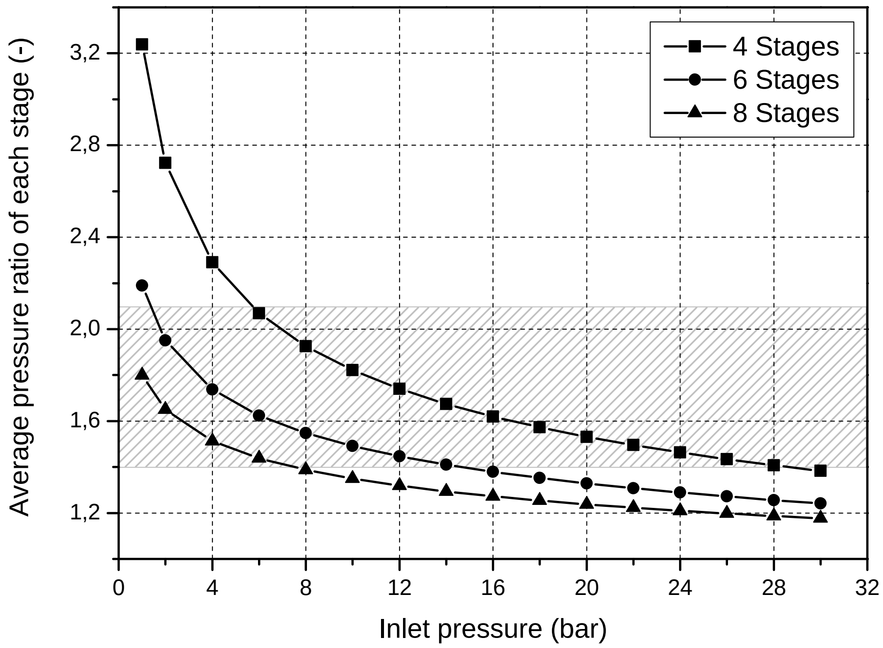

Figure 10: Average pressure ratio per stage depending on the inlet pressure for different stage numbers [18]

The interface quantities w_{comp}, q_{comp} and t_{comp} strongly depend on the number of intercoolers. Figure 11 shows the influence of the number of intercoolers on the specific power duty exemplarily for three different inlet pressures. In this case, an equivalent distribution of the intercoolers is assumed. That means that the pressure ratio of each stage upstream and downstream of an intercooler is equal. The specific power duty decreases with an increasing number of intercoolers. Furthermore, the influence of an additional intercooler decreases with an increasing number of intercoolers. For an inlet pressure of 1 bar the second intercooler leads to a decreased specific power duty of 0.1 MJ/kg CO_2. The third intercooler shows a halved effect (0.05 MJ/kJ CO_2).

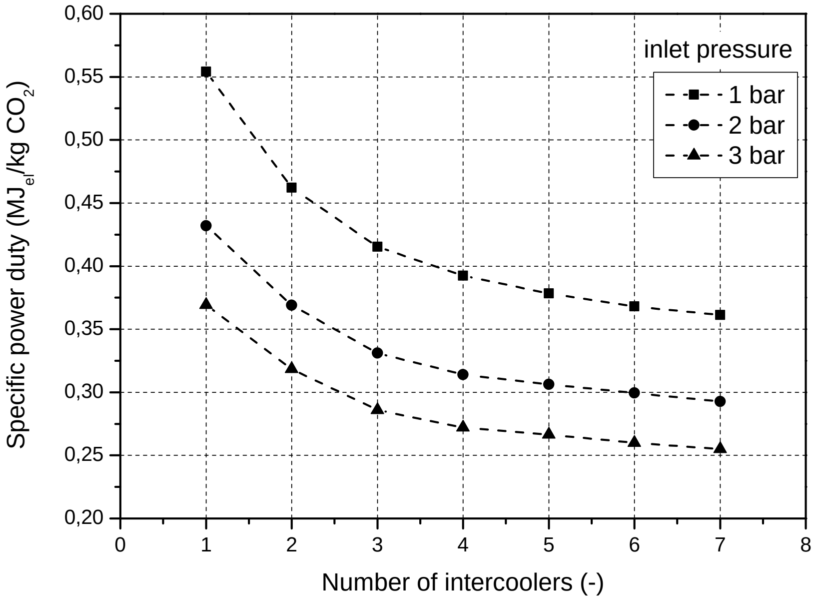

Figure 11: Specific power duty depending on the number of intercoolers [18]

Besides the specific power duty, the other interface quantities depend on the number of intercoolers as well. To systematically investigate a wide range of numbers and positions of intercoolers, three stage numbers and two different configurations for each stage number are examined. In configuration 1 the maximal possible number of intercoolers is used ($n_{IC} = n_{stage} - 1$). The overall pressure ratio is distributed equally over the stages. This leads to an almost equal temperature level in each intercooler. In configuration 2, the number of intercoolers is half of the number of stages (n_{IC} = ½ n_{stage}). The pressure ratio of each stage is the same as in configuration 1. Therefore, in configuration 2 two different temperature levels in the intercoolers exist. The temperature level depends on the presence or absence of an intercooler upstream of the former stage. Waste heat on a similar temperature level is summed up.

The distribution of waste heat over each intercooler is shown in Figure 12 exemplarily for configuration 1 with six stages. It can be observed that the waste heat for a saturated CO_2 stream first decreases and increases again with increasing intercooler number. The highest amount of waste heat accrues in the aftercooler. Two opposing effects lead to the shape of the curve:

- Steam in the CO_2 stream condenses in the intercoolers and is removed by a drain valve. The thermal energy of the steam/water phase change is discharged in the intercoolers. As the CO_2 stream is getting dryer after each intercooler, most of the thermal energy accrues in the first intercooler. Downstream of the absorptive drying unit (stage four) no thermal waste heat through water condensing occurs. The comparison with the dry CO_2 stream illustrates this effect.
- The amount of thermal energy, which is required for CO_2 cooling increases with higher pressure of the CO_2 stream. This is shown in the T,s-diagram of Figure 13. The areas below the dashed curves

represent the thermal energy transferred in the intercoolers. These areas increase when approaching the critical point. This effect explains the high waste heat in the aftercooler (Figure 12).

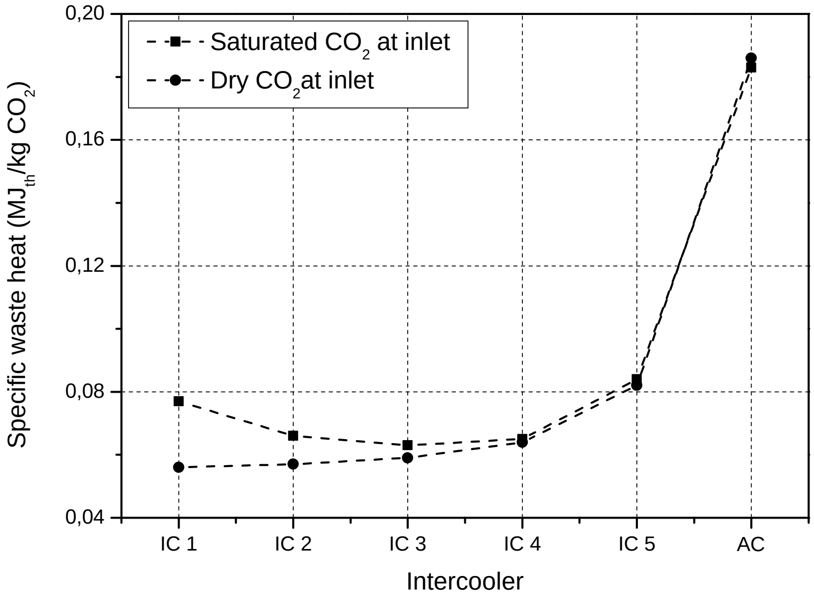

Figure 12: Distribution of waste heat over each intercooler [18]

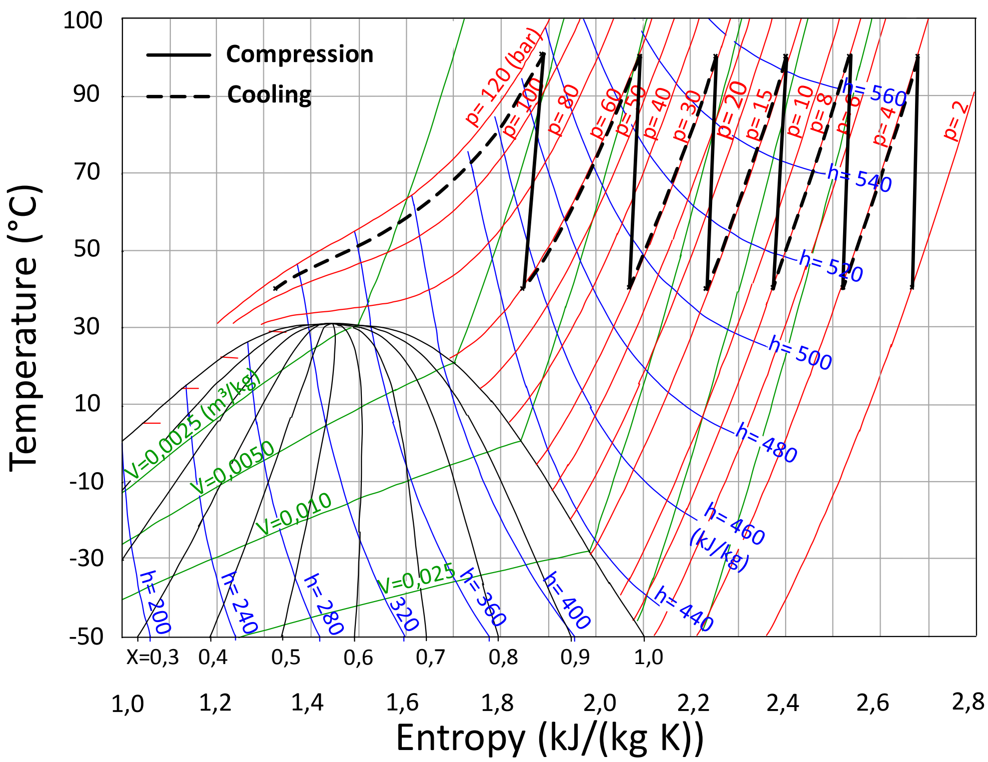

Figure 13: T,s-diagram for CO_2 compression (6 stages, 5 intercooler, 1 aftercooler) [23]

Due to the unequally distributed waste heat over the intercoolers, the intercooler positions of configuration 2 can be chosen to reach high amounts of waste heat on a high temperature level without increasing the power duty. For that purpose, the last intercooler is arranged upstream of the second to last stage. This measure raises the temperature level in the aftercooler. Compressor configuration 2 is shown in Figure 14 for each stage number. Red intercoolers represent a high temperature level, blue intercoolers represent a low temperature level. For the variant with only four stages the high temperature level only occurs in the aftercooler.

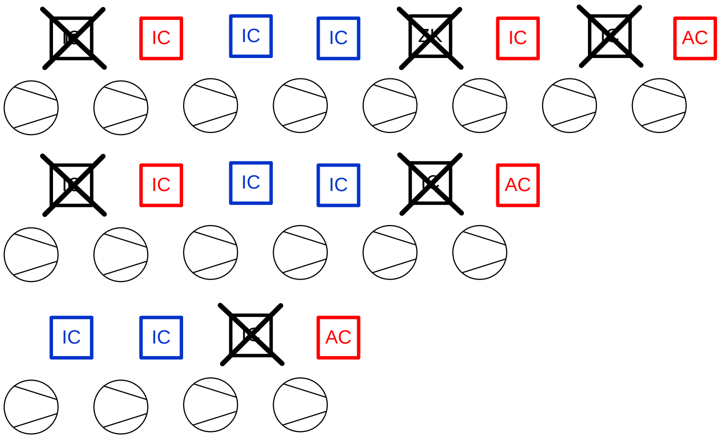

Figure 14: Compressor configuration 2, red = waste heat on high temperature level, blue = waste heat on low temperature level [18]

In Figure 15 all interface quantities are exemplarily shown for configuration 2 with six stages as a function of inlet pressure. Due to the positioning of the intercoolers, the amount of waste heat on the high temperature level is three to four times higher than the waste heat on the low temperature level. The high temperature level has temperatures between 102 °C and 223 °C. The low temperature level has temperatures between 70 °C and 121 °C.

For the overall process optimisation with heat integration all compressor configurations are taken into account. The optimal variant is determined and documented for each capture process modification.

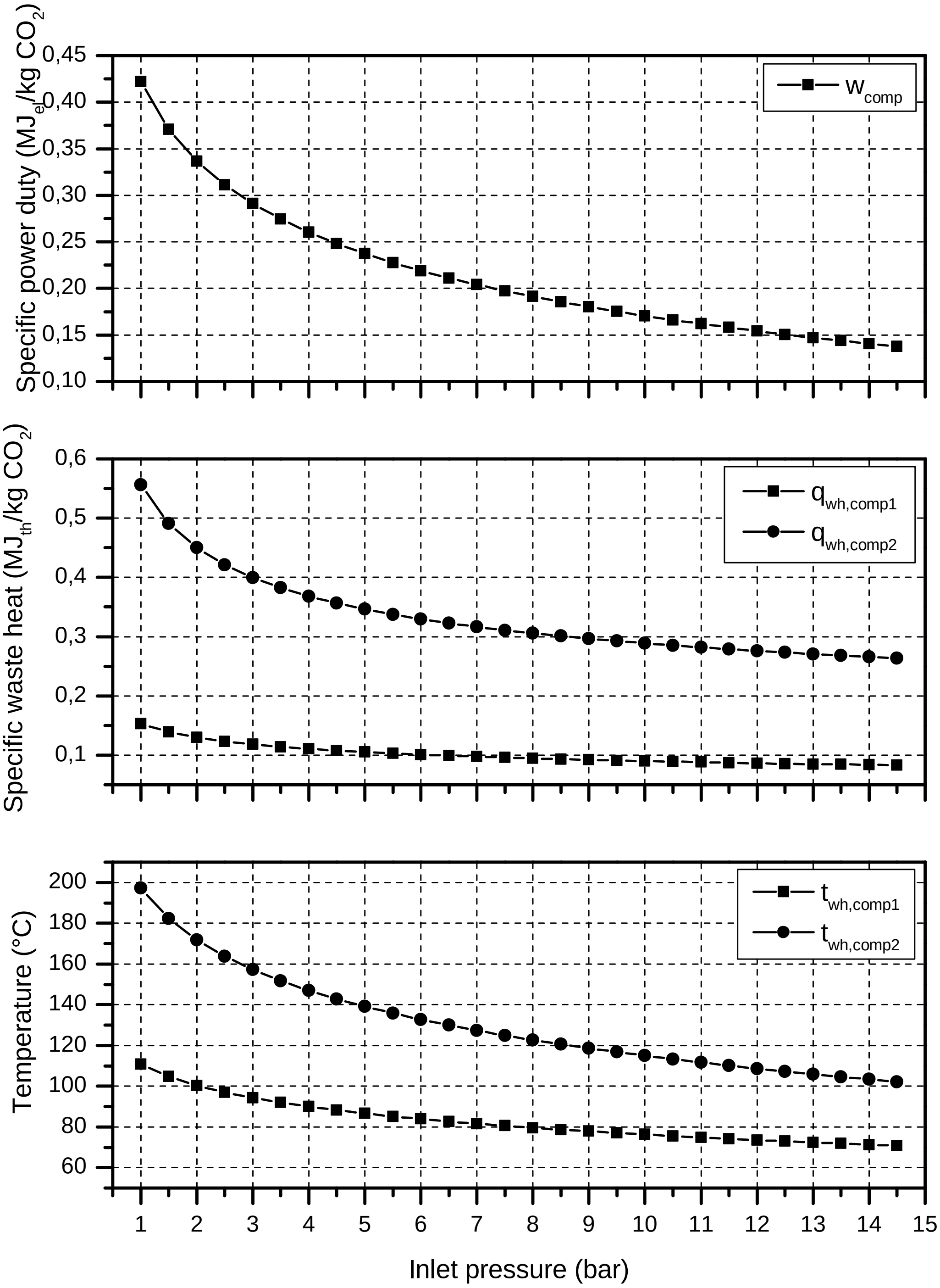

Figure 15: Interface quantities of the CO_2 compressor for configuration 2 with six stages [18]

5.4 Definition of Interface Quantities

To enable an effective procedure of overall process analysis, a clear definition of all energetic interface quantities is required. The interface quantities defined in this section will be listed for each CO_2 capture process flow sheet modification, allowing a direct comparison (see Table 6).

Table 6: Energetic interface quantities

	SCPC	NGCC
Basic integration	Heat duty q_{reb} (MJ_{th}/kg CO_2)	Heat duty q_{reb} (MJ_{th}/kg CO_2)
	Cooling duty q_{cool} (MJ_{th}/kg CO_2)	Cooling duty q_{cool} (MJ_{th}/kg CO_2)
	Power duty w_{aux} (MJ_{el}/kg CO_2)	Power duty w_{aux} (MJ_{el}/kg CO_2)
	Desorber pressure p_{des} (bar)	Desorber pressure p_{des} (bar)
	Reboiler temperature t_{reb} (°C)	Reboiler temperature t_{reb} (°C)
		Flue gas temperature upstream of the capture plant t_{flue} (°C)
Heat integration	Temperature level of waste heat t_{hi} (°C)	
	Waste heat q_{hi} (MJ_{th}/kg CO_2)	

6 CO_2 Capture Process Flow Sheet Modifications

In this chapter, different process flow sheet modifications are evaluated. In order to have a common reference for all modifications, first a capture process base case is defined and described. Afterwards, the simulation results for the capture plant are presented, followed by an energetic evaluation of the overall process. The same approach is chosen for the evaluation of all process modifications.

6.1 Base Case SCPC - A1

6.1.1 Process Characteristics

The base case for the capture plant processing the flue gas from the supercritical pulverized coal fired power plant (A1) is very similar to the basic process described in section 2.2. A two train approach is chosen, leading to an absorber diameter of 17.6 m which is below the limit of 18 m (cf. section 3.1). The process flow sheet of the base case is shown in Figure 87. In the following, the process characteristics and different means of process control are described in detail.

The solution in the capture plant is circulating in a closed loop. For steady state operation, stable mass balances are essential. The CO_2 balance is maintained by adjusting the heat duty of the reboiler. The water mass balance is maintained by adjusting the water mass flow to the washing section or a split stream downstream the OHC. Since the water balance for the cooling water cycles of the flue gas cooler and the water wash section are stable, there are only three streams left where water enters or leaves the capture plant: the flue gas stream entering the absorber, the CO_2-lean gas leaving through the stack and the separated CO_2 which is led to the compressor. These three streams have to be in balance. This is achieved by two different ways. The first control variable is the cooling water mass flow to the washing section. Increasing the water mass flow reduces the temperature of the flue gas and thus the water content of the saturated gas. Contrary, a reduced water mass flow leads to an increased temperature and more water leaving the capture plant with the flue gas. A minimum water mass flow of 50 kg/s has to be maintained, though, to prevent solvent slip. In this case, a stable water balance is achieved by directly removing water from the process. This is done behind the OHC, where an almost pure stream of water is condensed from the separated CO_2.

Intercooling in absorber affects the absorption process in different ways. The reduced solution temperature leads to a higher maximum CO_2 loading for the same partial pressure (cf. Figure 2). This is relevant for the lower part of the absorber and results in a higher CO_2 loading of the rich solvent at the absorber outlet and thus a reduced solvent mass flow for a fixed CO_2 loading of the lean solvent. At the same time, the reaction kinetics and diffusion transport mechanisms are slowed down due to the lower temperature, which leads to a lower CO_2 loading at the absorber outlet. It has to be investigated which of these opposing effects has a stronger influence. For Solvent2020, a positive effect on the energy requirement is expected, since the reaction kinetics are assumed to be fast and do not limit the absorption. This will be evaluated in the next section.

Different operating points are adjusted by changing the lean loading at the absorber inlet. An increased loading results in an increased solvent mass flow, since the capture rate is kept constant at 90% and the rich CO_2 loading is affected only slightly. The changed solvent mass flow affects the energy requirement of the capture process. There are two opposing effects: On the one hand, a reduced solvent mass flow results in a reduction of sensible heat required for the heating of the solvent. On the other hand, the temperature in the reboiler increases for lower lean CO_2 loadings and thus lower solution mass flows. The reduced lean CO_2 loading leads to a lower CO_2 partial pressure and thus, constant reboiler pressure is assumed, to a higher water partial pressure. This implies higher temperatures in the reboiler. Thus, the smallest reboiler heat duty has to be found by adjusting the lean CO_2 loading and thus the solvent mass flow. In the following figures, the specific energies are plotted against L/G, the ratio between solvent mass flow and flue gas mass flow. The flue gas mass flow is not varied for all cases, since only the design case is evaluated.

The general capture plant configuration is the same for the SCPC and the NGCC model. They differ in terms of the flue gas conditions and mass flow. The SCPC and NGCC power plant cases and the flue gas conditions can be found in Table 3 and Table 4 respectively.

6.1.2 Simulation Results

In this section, different aspects of the capture plant base case are evaluated. First, the effect of a variation of the solvent mass flow is evaluated (see Figure 16).

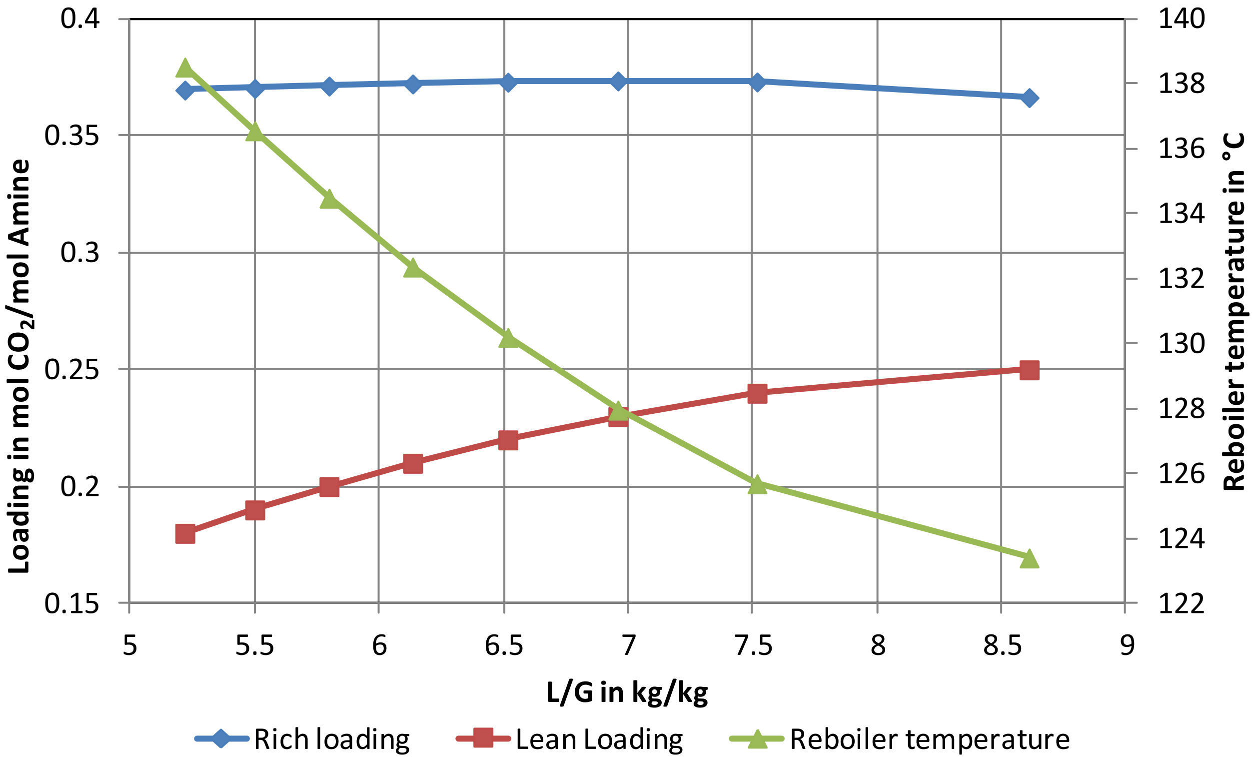

Figure 16: CO_2 Loading of the solution and reboiler temperature for different solution mass flows of a capture plant in combination with an SCPC plant (A1)

In Figure 16 the loading of the solution in mol CO_2/mol Amine downstream and upstream of the inter-cooled absorber is plotted against the ratio between lean solvent mass flow and flue gas mass flow (L/G). The reboiler temperature is shown as well. It can be seen, that the rich loading downstream the absorber does not change significantly while the lean CO_2 loading upstream the absorber increases with increasing solvent mass flow. Due to the smaller loading difference a higher solution mass flow is needed to absorb the same amount of CO_2. The reboiler temperature increases for reduced solvent mass flow and thus reduced lean loading. In order to reach low CO_2 loadings the CO_2 partial pressure has to be lower as well. Since the overall pressure in the stripper is kept constant, the water partial pressure has to be increased and thus a higher reboiler temperature is necessary.

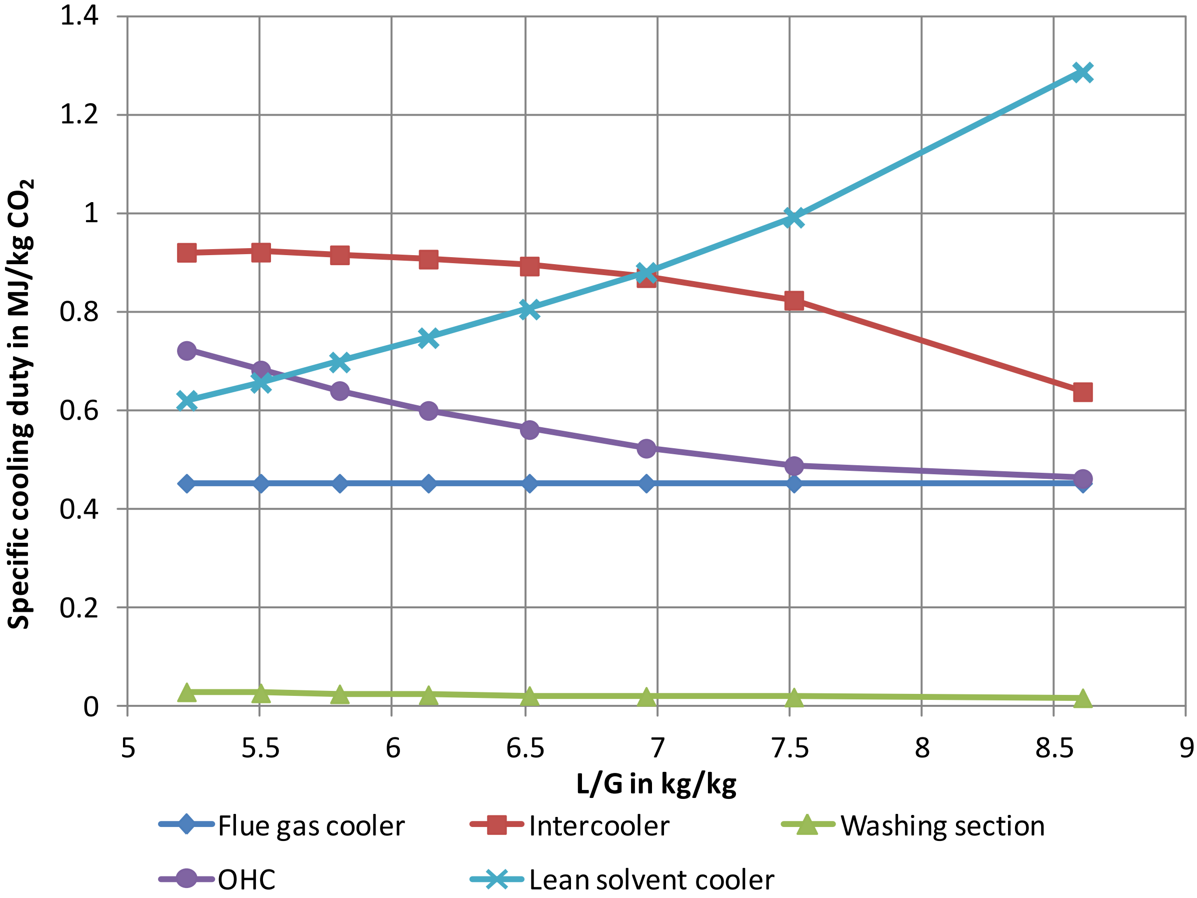

Figure 17: Specific cooling duty for different contributors to the specific cooling duty of a capture plant in combination with an SCPC plant (A1)

The cooling duty of the capture plant is the summation of five cooling duties of different components:

- flue gas cooler upstream the absorber,
- intercooler of the absorber,
- washing section downstream the absorber,
- overhead condenser downstream the desorber,
- lean solvent cooler.

The specific cooling duties of these components are shown in Figure 17 for different solvent mass flows. It can be seen that a variation of the solvent mass flow affects the different coolers in different ways. The flue gas cooler is not affected, since it is located upstream the actual capture plant. The washing section cooler requires only very low cooling duties since the temperature downstream the absorber is low and the washing section is operated at its minimum water mass flow (cf. section 6.1.1). The cooling duty of the intercooler decreases with increasing L/G. This is due to the lower temperatures in the absorber outweighing the increased mass flow. For L/G below 5.5 kg/kg this effect is inverted. The cooling duty of the OHC decreases with increasing L/G as well, since the temperature in the desorber decreases. The cooling duty of the lean solvent cooler increases with increasing L/G since the solvent mass flow increases while the temperature of the solvent entering the solvent cooler is more or less constant.

The auxiliary power is the summation of five power duties of different components:

- blower at the end of the flue gas path,
- rich solvent pump downstream the absorber,
- pump for the intercooler loop,
- pumps for the cooling water for the flue gas cooler,
- washing section.

The specific auxiliary power of these components is shown in Figure 18 for different solvent mass flows. The auxiliary power of the largest contributor, the blower, decreases with increasing L/G, since the temperature of the tail gas is reduced and thus the volume flow to the blower. The rich solvent pump is the second largest contributor to the auxiliary power, its power duty increases with increasing L/G, since more solution has to be pumped. The other three pumps are only of minor influence. The lean solvent pump is not used in the base case, since the reboiler pressure is high enough to overcome all pressure losses on the way to the absorber.

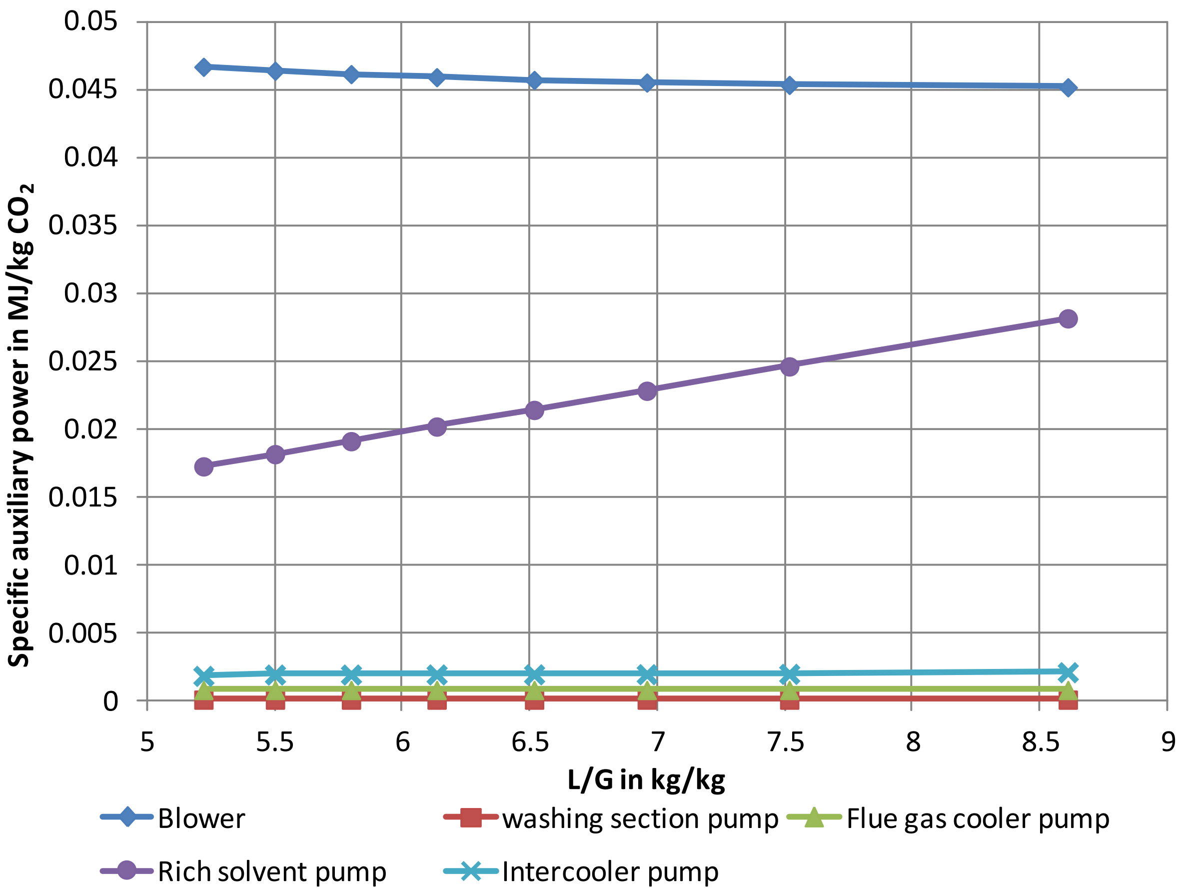

Figure 18: Specific auxiliary power for different contributors to the specific auxiliary power of a capture plant in combination with an SCPC plant (A1)

In Figure 19, the three interface quantities specific heat duty, specific cooling duty and specific auxiliary power are plotted against L/G. The two opposing effects on the specific heat duty described in section 6.1.1 can be seen in Figure 19 and result in the typical characteristic of the specific heat duty with a minimum and an increasing heat duty for increasing and decreasing L/G. The lowest specific heat duty of 2.14 MJ/kg is achieved at an L/G of 6.96 kg lean solution/kg cold flue gas. The specific cooling duty has a similar characteristic as the specific heat duty. The lowest specific cooling duty of 2.73 MJ/kg is achieved at a L/G of 6.13 kg lean solution/kg cold flue gas. The specific auxiliary power increases for increasing solvent mass flows since the increased power demand of the rich solvent pump outweighs the decreased power demand of the blower.

Compared to results from open literature for different solvents, the specific heat duty is quite low. For MEA, 3.6 MJ/kg CO_2 have been measured in pilot plants [24]. Still, there have been studies on other solvents with a significantly lower reboiler duty. For aqueous piperazine, 2.5 MJ/kg CO_2 (110.1 kJ/mol CO_2) have been reported [25]. For a mixture of MDEA and PZ, a specific reboiler duty of even below 2 MJ/kg CO_2 (86.6 kJ/mol CO_2) has been reported [26]. For Solvent2020, this low value is obtained, since a fraction of the absorption enthalpy needed for the regeneration of the solvent, is already provided in the RLHX. Due to this, approx. half of the CO_2 is already released during the heat transfer in the RLHX. This effect has been observed in the pilot plant in Heilbronn, Germany, operated by EnBW with MEA, as well, and can be problematic due to higher corrosion rates in the RLHX. Since corrosion is assumed to be of no importance for Solvent2020 in the operating range of this study, this positive effect can be used to the full extent.

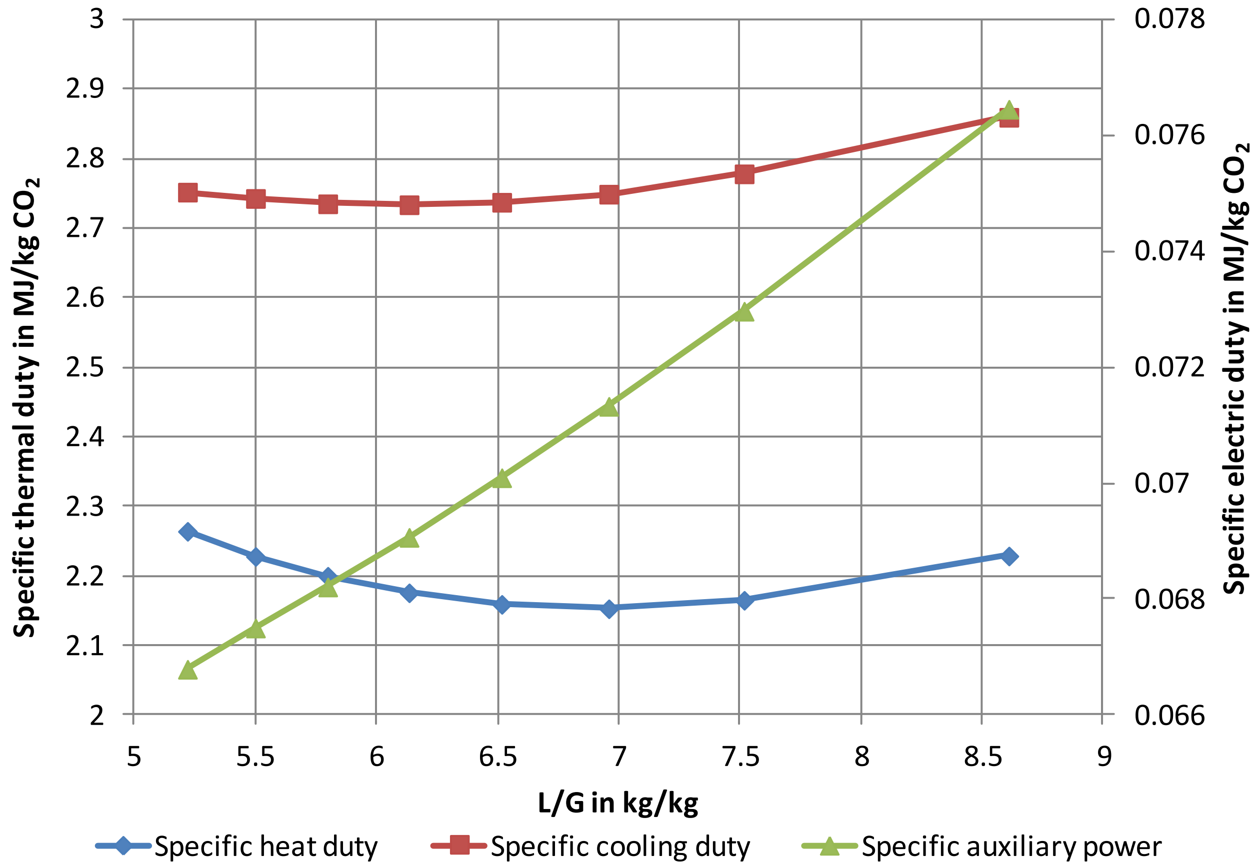

Figure 19: Specific thermal duty and specific auxiliary power of a capture plant in combination with an SCPC plant (A1)

As described in section 3.1, the absorber is equipped with an intercooler. At half height of the absorber packing, the solution is withdrawn from the absorber. At this stage, the temperature in the absorber has almost reached its maximum and intercooling is therefore most effective [27]. The withdrawn solution is cooled down to 40 °C and reintroduced into the absorber. The solvent feedback is directly downstream of the extraction. In the following, the effect of this intercooling is evaluated by comparing the base case results with the results of an identical CO_2 capture plant without intercooling.

In Figure 20, the temperature in the absorber is plotted against the relative height. The temperature profile in the absorber without intercooling is typical for an absorber. From the top of the absorber, the temperature increases due to the exothermal absorption reaction of CO_2. At the bottom of the absorber, the temperature decreases due to the cooler flue gas entering at the lower part of the absorber. Depending on the solvent mass flow, this temperature bulge can be much more distinct.

The temperature profile of the intercooled absorber shows an unsteadiness at half height where the cooled down solution is fed back, resulting in a lower temperature in the lower half of the absorber. The temperature in the upper half is reduced as well due to the lower temperature of the flue gas coming from the lower half of the absorber.

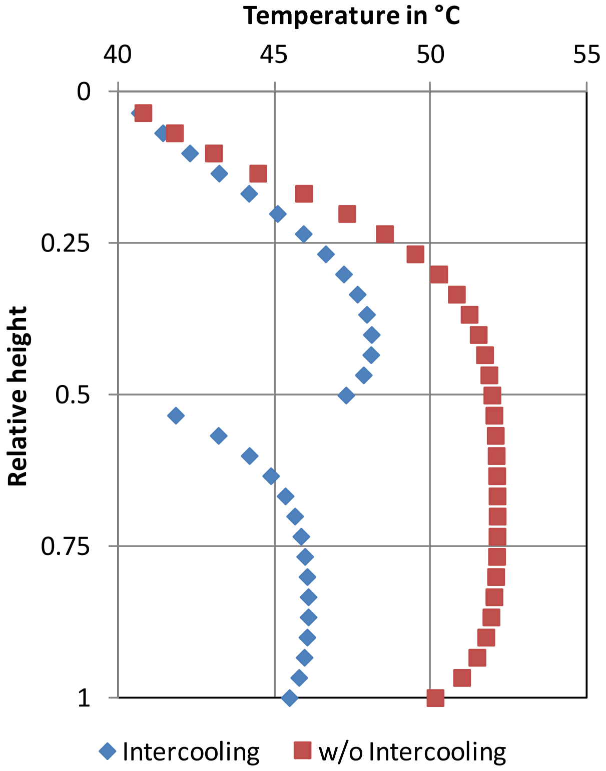

Figure 20: Temperature profile in the absorber with and without intercooling

The change in absorber temperature affects the CO_2 absorption, as can be seen in Figure 21. The loading in the absorber without intercooling increases from the top of the absorber until a steady state is nearly reached at approx. half height. Downstream, the loading increases only very slowly until it starts to increase faster near the bottom of the absorber due to the lower temperature.

Due to the lower temperature of the solution in the intercooled absorber, the CO_2 absorption capacity of the solution is increased which results in a higher rich loading. In the upper half of the absorber, the loading is lower compared to the absorber without intercooling. This results from a lower CO_2 partial pressure in the flue gas since more CO_2 has already been absorbed in the lower half.

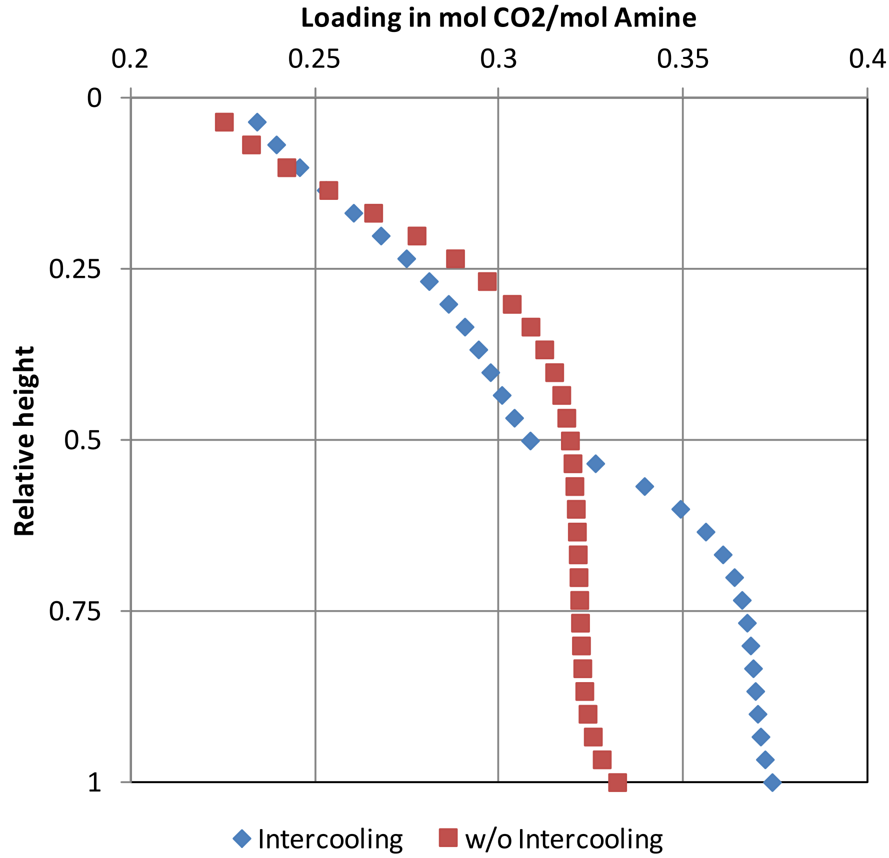

Figure 21: CO_2 loading of the solution in the absorber with and without intercooling

The effect of the absorber intercooling on the specific thermal duty and the specific auxiliary duty of the capture process can be seen in Figure 22, where the specific thermal duty and the specific auxiliary power of the capture process with and without intercooling are plotted against L/G. It can be seen that all three specific duties for the intercooled case, are reduced compared to the case without intercooling. The lowest specific heat duty is reduced by 0.27 MJ/kg CO_2, from 2.41 MJ/kg CO_2 to 2.14 MJ/kg CO_2. At the same operating point, the specific cooling duty is reduced by 0.34 MJ/kg CO_2, from 3.09 MJ/kg CO_2 to 2.75 MJ/kg CO_2, and the specific auxiliary power is reduced by 0.01 MJ/kg CO_2, from 0.079 MJ/kg CO_2 to 0.069 MJ/kg CO_2.

Despite the additional cooler and pump required for intercooling, the specific cooling duty and the specific auxiliary power do not increase when intercooling is used. This is due to the fact that the heat transferred in the intercooler has to be removed from the process by other means for the absorber without intercooling, mainly in the lean solvent cooler. The increase in auxiliary power needed for the pump is compensated by the reduced auxiliary power for other pumps, since the L/G is reduced from 9.93 kg/kg to 6.96 kg/kg. This reduction is possible due to the higher rich loading with the lean loading being nearly constant. A comparison of the interface quantities and some other process values are shown in Table 7.

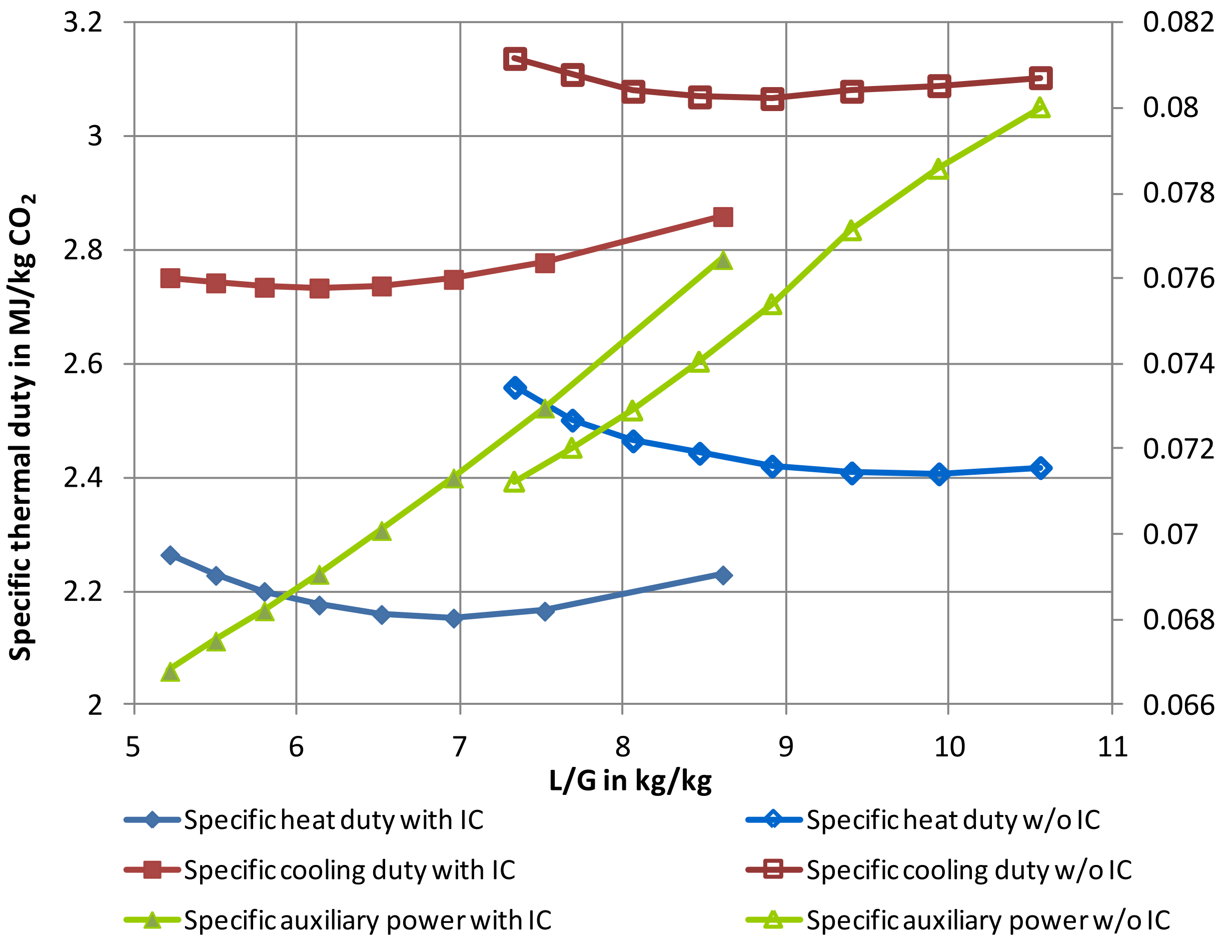

Figure 22: Specific thermal duty and specific auxiliary power of a capture plant in combination with an SCPC plant (A1) with and without intercooling

Table 7: Comparison of a capture plant in combination with an SCPC plant (A1) with and without intercooling

	Base case with intercooling	Case without intercooling
Specific heat duty in MJ/kg CO_2	2.14	2.41
Specific cooling duty in MJ/kg CO_2	2.75	3.09
Specific auxiliary power in MJ/kg CO_2	0.07	0.08
Desorber pressure in bar	5	5
Reboiler temperature in °C	128.0	125.6
Usable waste heat from OHC in MJ/kg CO_2	0.52	0.53
Temperature level of usable waste heat in °C	116.2	116.4
Lean solvent mass flow in kg/s	2934	4189
Lean loading in mol CO_2/mol Amine	0.23	0.24
Rich loading in mol CO_2/mol Amine	0.37	0.34
Rich solvent temperature in °C	45.5	50.2

The stripper pressure is an important process parameter. The CO_2 partial pressure in the stripper determines the lean loading of the solution. When the pressure in the stripper is increased and all other process values are kept constant, the CO_2 partial pressure would increase as well. In order to reach the same CO_2 partial pressure, and thus the same lean loading, for a higher stripper pressure the steam partial pressure has to be increased further. This is achieved by a higher reboiler temperature. In addition, higher stripper pressures lead to an increased power demand of the rich solution pump, while the power demand of the CO_2 compressor is reduced.

For the base case, a stripper pressure of 5 bar is chosen. This results in a reboiler temperature of 128 °C. Reducing the stripper pressure reduces the reboiler temperature, but leads to an increased specific heat duty, as can be seen in Figure 23. Higher stripper pressures are not beneficial for the overall process, since the decrease in specific heat duty is slowed down, while the reboiler temperature increases almost linearly. For this evaluation, the solvent flow rate was varied to find the operating point with the lowest specific heat duty.

The CO_2 partial pressure in the reboiler is quite high (2.5 bar), compared to a standard MEA process (0.1 bar). This behaviour is similar to the performance of the mixture of MDEA and PZ as a solvent [28]

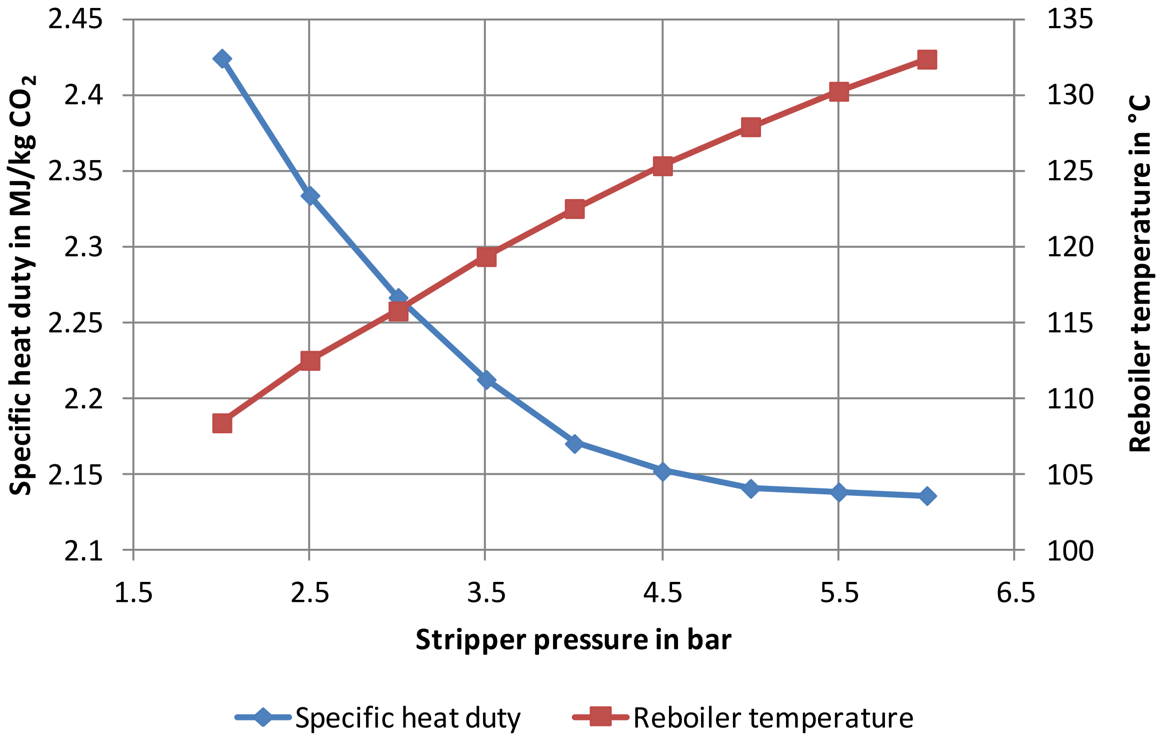

Figure 23: Specific heat duty and reboiler temperature of a capture plant in combination with an SCPC plant (A1) for different stripper pressures

6.1.3 Process Evaluation

In the previous section, the CO_2 capture plant has been evaluated without consideration of the power plant. In the following, the overall process is evaluated. As described in section 5.1, two different integration concepts can be applied. First, the basic integration is evaluated followed by the more complex waste heat integration.

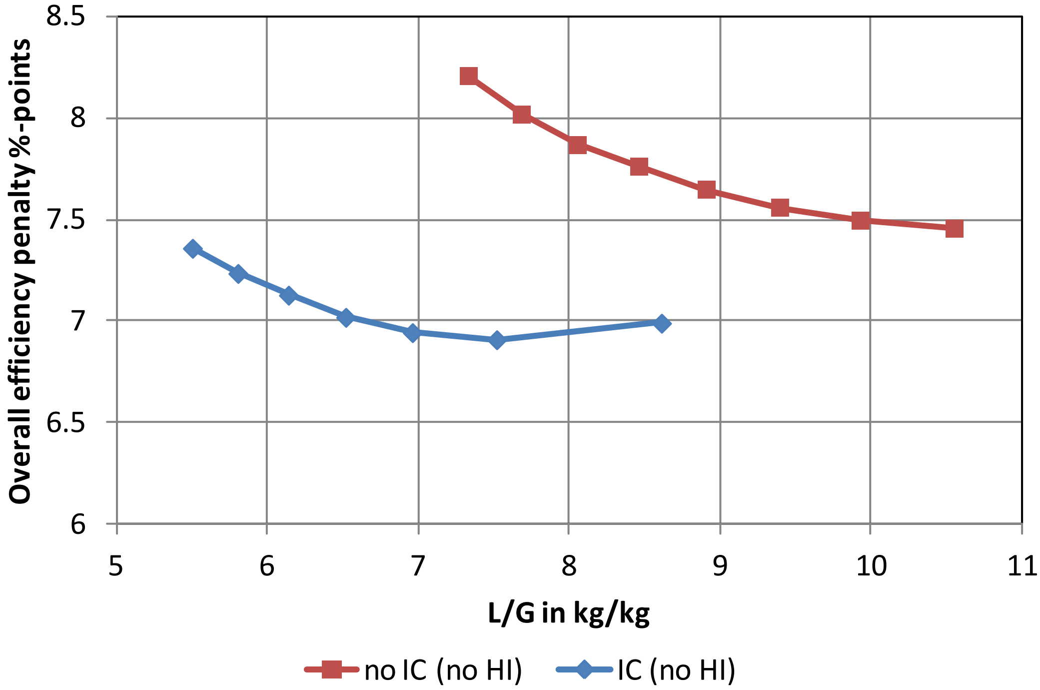

Figure 24: Overall efficiency penalty for a capture plant in combination with an SCPC plant (A1) with basic integration

In Figure 24 the overall efficiency penalty is shown for the base case and for the case without intercooling for different solvent mass flows. The overall efficiency penalty is the reduction of the net efficiency of the power plant caused by the CO_2 capture plant. The net efficiency is reduced for example by 6.9%-points from 45.2% to 38.3% when a capture plant with intercooling and an L/G of 7.5 kg/kg is used. The overall efficiency penalty includes all influences of the capture plant and is thus the value that should be compared for different process flow sheet modifications. The different contributors to the overall efficiency penalty are listed in Table 8 for the operating point with the lowest overall efficiency penalty.

Table 8: Contributors to the overall efficiency penalty for a capture plant in combination with an SCPC plant (A1) with basic integration

	Base case with intercooling	**Case without intercooling**
Steam extraction	4.16%-points	4.60%-points
Compressor duty	1.90%-points	1.90%-points
Cooling water pumps	0.23%-points	0.26%-points
Auxiliary power	0.62%-points	0.70%-points
Overall efficiency penalty	**6.91%-points**	**7.45%-points**

The largest contributor to the overall efficiency penalty is the steam extraction required for the reboiler. Due to the extracted steam, the steam mass flow to the LP turbine is reduced which results in lower power rating of the generator. The other contributors are electrical consumers and are thus directly reducing the electrical net output of the power plant. The CO_2 compressor is the largest of these consumers. The additional cooling duty of the capture plant leads to an additional power demand of the cooling water pumps. The pumps and the blower of the capture plant are combined into one value, the auxiliary power of the capture plant.

As for the specific heat duty, the overall efficiency penalty is significantly lower compared to results from previous studies with standard MEA. In a previous IEAGHG funded study, a net efficiency penalty of 12.1 %-points is obtained for an MEA case [29]. The discrepancy is due to the large difference in specific heat duty required in the reboiler caused by the different solvents. Another IEAGHG funded study shows an efficiency penalty of 8.9 %-points using Cansolv solvent, 2 bar stripper pressure, intercooling and lean vapour recompression [30]. This is already a good improvement compared to MEA with a current improved solvent.

A comparison of the specific heat duty of the capture process (cf. Figure 19) and the overall efficiency penalty in Figure 24 shows that the operating point with the lowest specific heat duty is not matching the operating point with the lowest overall efficiency penalty. While the lowest specific heat duty is obtained for an L/G of 7 kg/kg, the lowest overall efficiency penalty is obtained for an L/G of 7.5 kg/kg. This results from the lowered reboiler temperature for higher L/G (cf. Figure 16) leading to a lower required pressure of the extracted steam and thus a higher electricity production of the power plant. For higher L/G, this effect is outweighed by the increased specific heat duty of the capture plant.

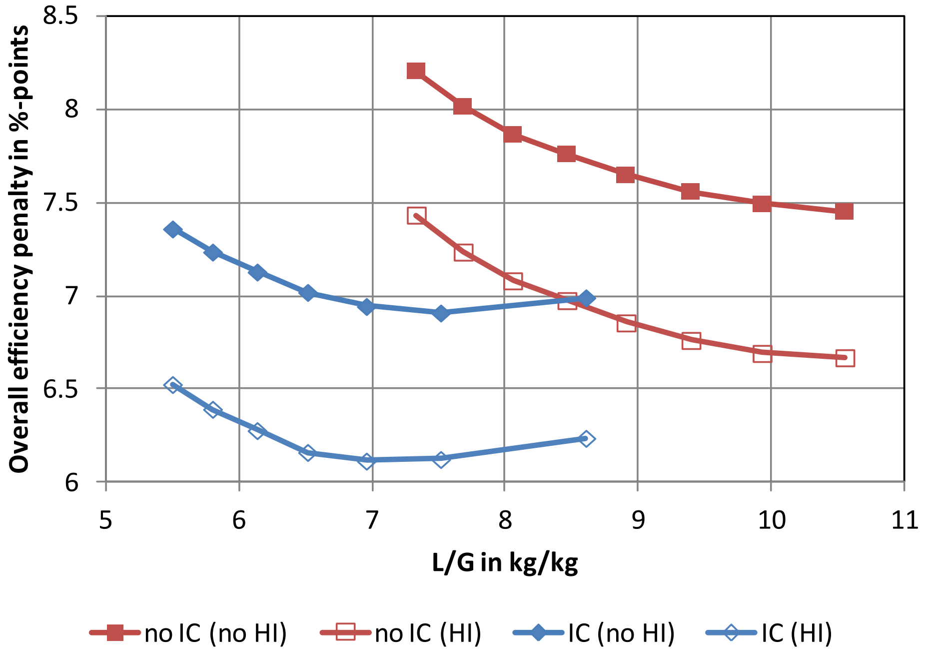

Figure 25: Overall efficiency penalty for a capture plant in combination with an SCPC plant (A1) with waste heat integration

In order to reduce the overall efficiency penalty, waste heat integration is applied. The effect can be seen in Figure 25 for both cases, the base case with intercooling and the case without intercooling. The overall efficiency penalty caused by the capture plant with intercooled absorber is reduced by 0.8%-points from 6.91%-points to 6.11%-points. The overall efficiency penalty caused by the capture plant without inter-cooled absorber is reduced by 0.78%-points from 7.45%-points to 6.67%-points. This reduction is caused by different opposing effects that can be exemplified by comparing the different contributors to the overall efficiency penalty shown in Table 8 and Table 9. In Table 9, the contributors to the overall efficiency penalty are shown for the operating point with the smallest overall efficiency penalty.

The negative values given in Table 9 for the heat integration reduce the overall efficiency penalty. They represent the saving in extraction steam for condensate preheating that is achieved by preheating the condensate of the power plant with the waste heat from the capture plant. The penalty caused by steam extraction and auxiliary power of the capture plant is not affected by the implementation of waste heat integration. The difference of these values for the base cases in Table 8 and Table 9 is due to the fact that the lowest overall efficiency penalty is achieved for different operating points. The penalty caused by the compressor duty increases for the cases with waste heat integration. This can be explained by the higher temperatures of the CO_2 in the compression train since the cooling with condensate does not allow the same low cooling temperatures as cooling with cooling water. The penalty caused by the cooling water pumps is reduced since less cooling water has to be pumped due to the cooling with condensate.

Table 9: Contributors to the overall efficiency penalty for a capture plant in combination with an SCPC plant (A1) with waste heat integration

	Base case with intercooling	**Case without Intercooling**
Steam extraction	4.21%-points	4.60%-points
Compressor duty	2.06%-points	2.06%-points
Cooling water pumps	0.21%-points	0.23%-points
Auxiliary power	0.60%-points	0.70%-points
Heat integration	-0.97%-points	-0.92%-points
Overall efficiency penalty	**6.11%-points**	**6.67%-points**

The effect of a reduced stripper pressure can be seen in Figure 26. The lowest overall efficiency penalty for the basic integration case as well as for the waste heat integration case is achieved for a stripper pressure of 5 bar. Higher stripper pressures lead to an increased reboiler temperature and thus an increased penalty due to steam extraction. Lower stripper pressures lead to an increased specific auxiliary power for CO_2 compression as well as an increased specific heat duty of the reboiler (cf. Figure 23).

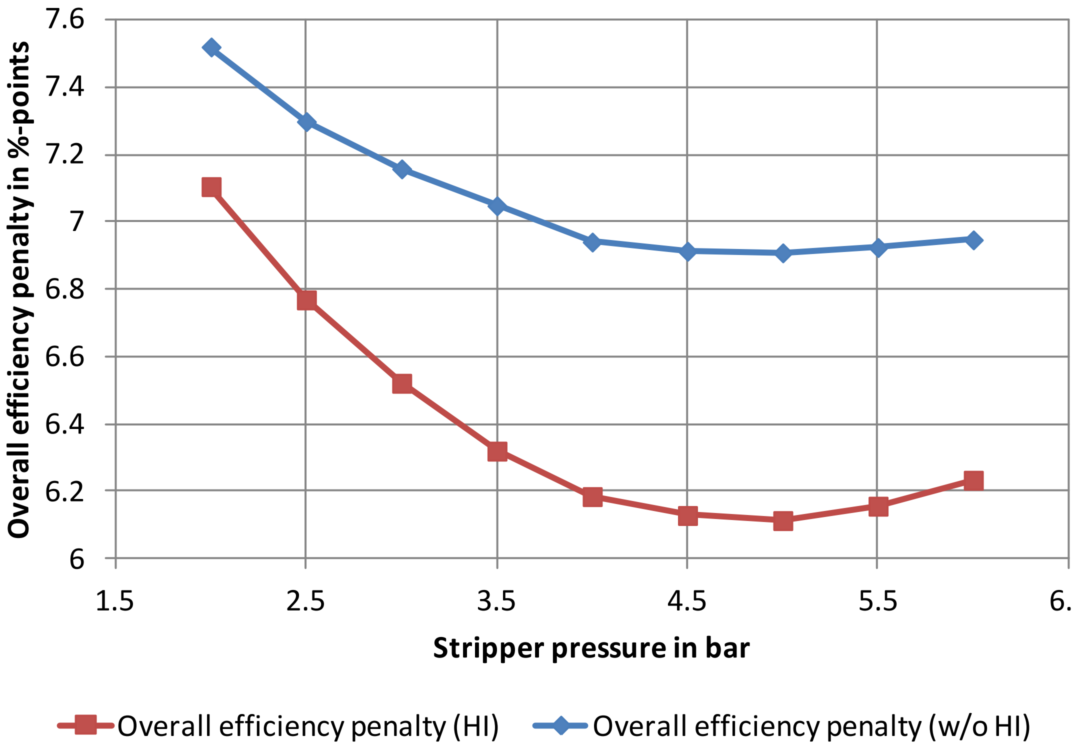

Figure 26: Overall efficiency penalty for a capture plant in combination with an SCPC plant (A1) for different stripper pressures

The different contributors to the overall efficiency penalty for different stripper pressures are shown in Table 10 exemplarily for the Base Case as well as for the case with a stripper pressure of 2 bar. It can be seen that despite the higher specific heat duty, the penalty due to steam extraction is reduced due to the lower reboiler temperature. Still, the overall efficiency penalty is increased since the compressor duty is increased significantly.

Table 10: Contributors to the overall efficiency penalty for a capture plant in combination with an SCPC plant (A1) with basic integration for different stripper pressures

	Base case (5 bar stripper pressure)	**Case with 2 bar stripper pressure**
Steam extraction	4.16%-points	4.08%-points
Compressor duty	1.90%-points	2.54%-points
Cooling water pumps	0.23%-points	0.26%-points
Auxiliary power	0.62%-points	0.64%-points
Overall efficiency penalty	**6.91%-points**	**7.52%-points**

For comparison, the results for a standard MEA (30 wt%) capture process with 2 bar reboiler pressure are listed in Table 11. The reference power plant is identical to the one chosen for this study. There are no process modifications like intercooling or advanced waste heat integration incorporated in the capture plant. The boundary conditions are: specific reboiler duty 3.47 MJ/kg CO_2, specific cooling duty 3.83 MJ/kg CO_2, specific auxiliary power 0.054 MJ/kg CO_2 and reboiler temperature 120.8 °C. It can be seen that the penalties due to steam extraction and cooling water pumps are increased since the specific reboiler duty as well as the specific cooling duty are higher for the MEA case. The penalty due to the compression of CO_2 is identical to the Solvent2020 case with 2 bar stripper pressure. The penalty due to auxiliary power is lower for the MEA case since the solution mass flow is significantly reduced.

Table 11: Contributors to the overall efficiency penalty for a capture plant in combination with an SCPC plant

	MEA 30 wt% case
Steam extraction	6.47%-points
Compressor duty	2.54%-points
Cooling water pumps	0.32%-points
Auxiliary power	0.47%-points
Overall efficiency penalty	**9.80%-points**

6.2 Base Case NGCC - B1

6.2.1 Process Characteristics

For the base case of the CO_2 capture plant processing the flue gas from a natural gas combined cycle plant (B1), two different approaches are evaluated. First, the flue gas from the power plant without flue gas recirculation is processed in a three train capture plant. The resulting absorber diameter for the operating point with the lowest heat duty is 16.2 m. In addition, a two train capture plant is simulated for the flue gas from the power plant with flue gas recirculation. The resulting absorber diameter is 14.5 m. The process flow sheet of the base case is shown in the annex (Figure 96).

For the CO_2 capture plant in combination with the SCPC plant it was shown that an intercooled absorber results in a significantly lower specific heat duty. Thus, an intercooled absorber is used for the NGCC case, as well.

The means of process control are the same as described for the SCPC case with one exception: The extraction of steam for the reboiler results in a lower condensate mass flow to the economiser since a fraction of the condensate coming from the reboiler is reintroduced downstream the economiser. This leads to an increased flue gas temperature upstream the capture plant since less heat can be transferred in the economiser. The flue gas temperature is thus an interface quantity for the capture plant and an iterative approach has to be applied.

6.2.2 Simulation Results

The specific duties of the CO_2 capture plant for both NGCC cases are shown in Figure 27. As for the coal case, the typical behaviour of the specific thermal duty and the specific auxiliary duty can be seen. The cooling duty and the heat duty show minima, while the auxiliary power increases for increasing solution mass flow. The lowest specific heat duty for the capture plant without FGR of 2.84 MJ/kg CO_2 is obtained for an L/G of 2.4 kg/kg. For the same operating point, the specific cooling duty adds up to 3.96 MJ/kg CO_2, the specific auxiliary power adds up to 0.202 MJ/kg CO_2. Incorporation of the FGR leads to a reduction in the specific heat duty by 0.47 MJ/kg CO_2 (2.37 MJ/kg CO_2). For the same operating point, the specific cooling duty is reduced by 0.48 MJ/kg CO_2 (3.48 MJ/kg CO_2), the specific auxiliary power is reduced by 0.099 MJ/kg CO_2 (0.103 MJ/kg CO_2).

For the NGCC case without flue gas recirculation (FGR), the CO_2 concentration in the flue gas is significantly lower. In order to achieve a capture rate of 90%, the lean loading of the solution has to be much lower for the case without FGR and amounts to 0.16 for the operating point with the lowest specific heat duty. For the case with FGR, a higher lean loading of 0.21 mol CO_2/mol amine is obtained. The regeneration of the solution to lower lean loadings needs more energy since the partial pressure of CO_2 in the vapour phase in the stripper has to be smaller and more water has to be evaporated. This leads to a higher reboiler temperature and a higher specific heat duty. In addition, the rich loading is increased as well due to the higher CO_2 content in the flue gas for the case with FGR which results in a reduced reboiler heat duty, too.

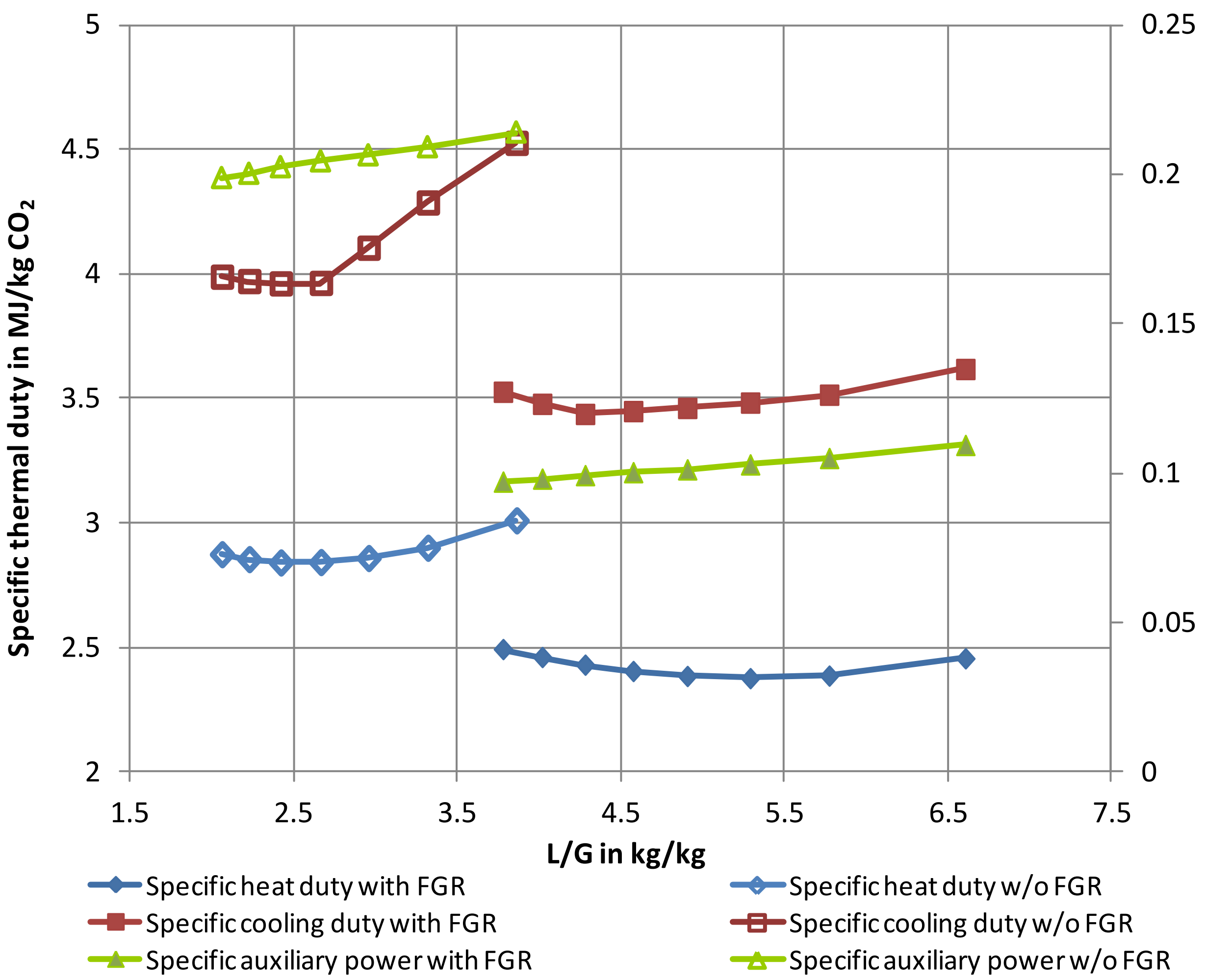

Figure 27: Specific thermal duty and specific auxiliary power of a capture plant in combination with an NGCC plant (B1) with and without flue gas recirculation

The cooling duty for the case without FGR is higher since more heat has to be removed from the process. This is due to the higher specific heat duty, as well as the higher flue gas temperature. As explained in section 6.2.1, the flue gas temperature of the power plant is increased when a CO_2 capture plant is equipped. Since the reboiler temperature for the case without FGR is higher (cf. Table 12), less reboiler condensate can be reintroduced upstream the economiser for a mixing temperature of 60 °C. Thus, less heat can be exchanged with the flue gas.

The specific auxiliary power is reduced significantly for the case with FGR. This is due to the reduced flue gas mass flow and thus a lower power demand for the flue gas blower downstream the absorber. The power demand of the pumps is not changed despite the higher L/G for the case with FGR. The solution mass flow in one train is increased by around 53.5% but, as said before, there are only two trains necessary for the case with FGR, while the plant without FGR has to consist of three trains.

Table 12: Interface quantities of a capture plant in combination with an NGCC plant (B1) with and without flue gas recirculation (FGR)

	Case with FGR	Case without FGR
Specific heat duty in MJ/kg CO_2	2.37	2.84
Specific cooling duty in MJ/kg CO_2	3.48	3.96
Specific auxiliary power in MJ/kg CO_2	0.103	0.202
Desorber pressure in bar	5	5
Reboiler temperature in °C	132.4	142.3
Lean solvent mass flow in kg/s	1612	1050
Lean loading in mol CO_2/mol Amine	0.21	0.16
Rich loading in mol CO_2/mol Amine	0.34	0.29
Flue gas temperature upstream the capture plant in °C	109.8	124.5

As for the coal case (A1), a stripper pressure of 5 bar is chosen. This results in a reboiler temperature of 132.4 °C. Reducing the stripper pressure reduces the reboiler temperature, but leads to an increased specific heat duty, as can be seen in Figure 28. Similar to the coal case, the reduction of specific heat duty is slowed down for higher stripper pressures, while the reboiler temperature increases linearly.

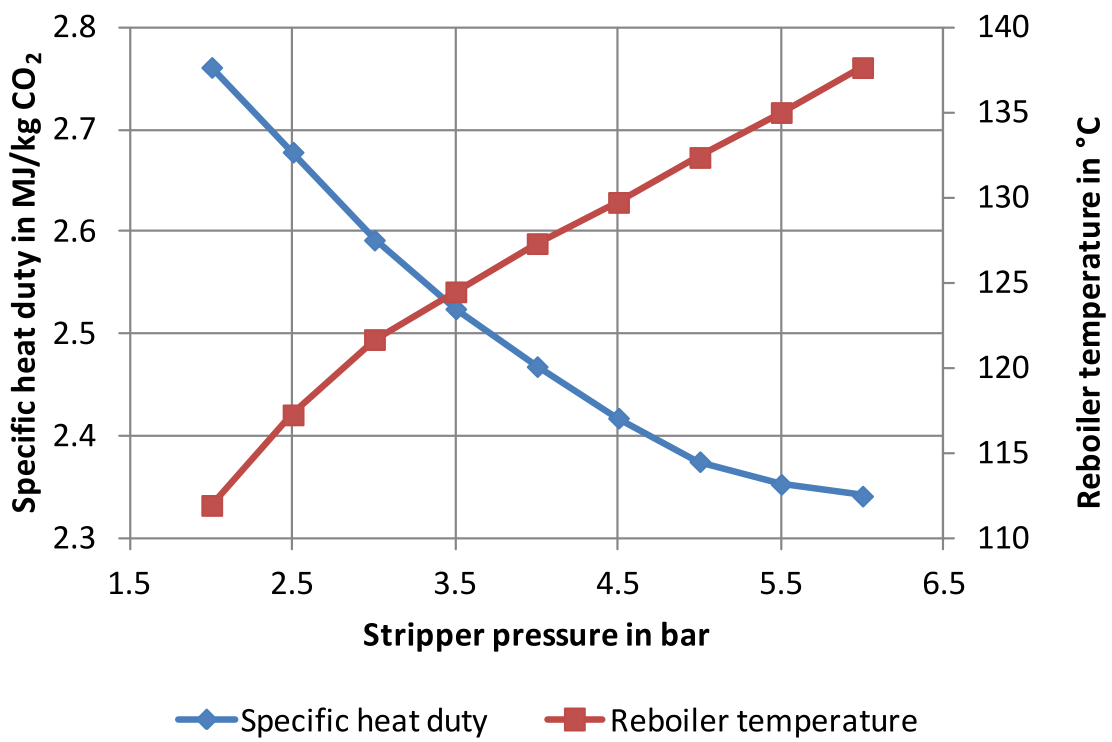

Figure 28: Specific heat duty and reboiler temperature of a capture plant in combination with an NGCC plant (B1) for different stripper pressures

6.2.3 Process Evaluation

The overall efficiency penalty of the CO_2 capture plant for the NGCC case is shown in Figure 29. Both cases, with and without flue gas recirculation, are evaluated for different L/G ratio.

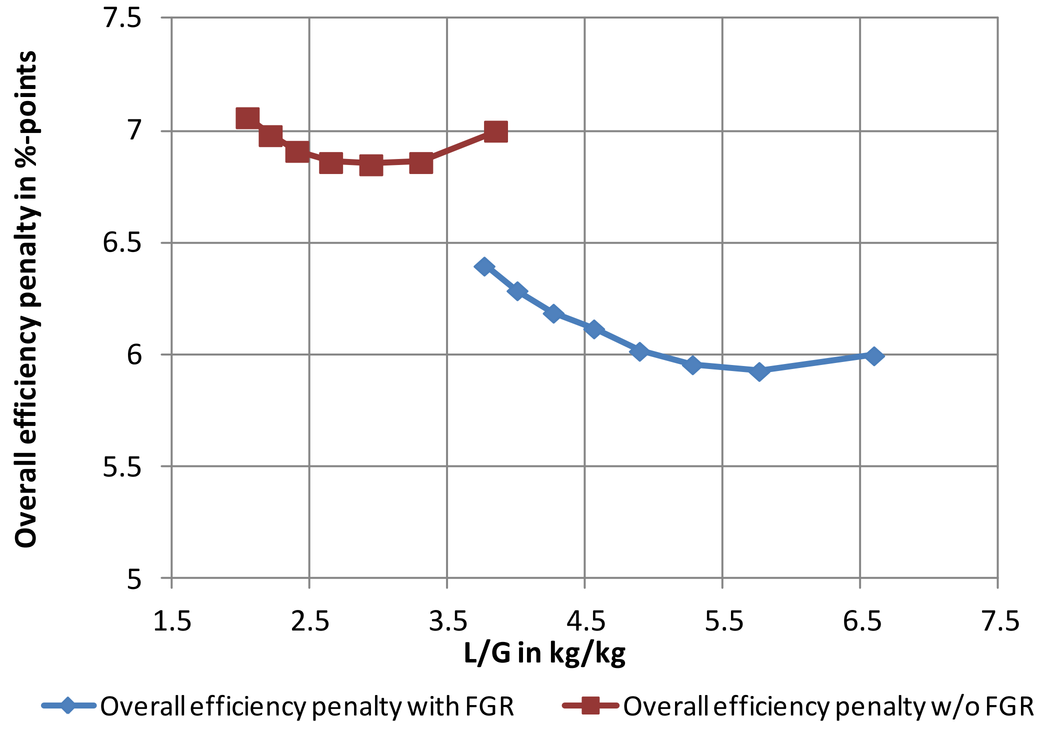

Figure 29: Overall efficiency penalty for a capture plant in combination with an NGCC plant (B1) with and without flue gas recirculation

For the case without FGR, the lowest overall efficiency penalty achieved is 6.84%-points. Again, the largest contributor is the loss due to steam extraction, which causes almost two thirds of the overall efficiency penalty. Incorporating the FGR reduces the efficiency penalty by 0.91%-points to 5.93%-points. This reduction is mainly caused by the reduced loss due to steam extraction, which is reduced by 0.97%-points, and the reduced auxiliary power of the capture plant, which is reduced by 0.54%-points. Due to the higher CO_2 partial pressure in the flue gas, the rich loading downstream the absorber is higher as well, leading to a reduced solvent mass flow and reduced pumping duty. Since less solution has to be heated up, the reboiler duty is reduced as well. Furthermore, the flue gas mass flow is significantly reduced, which leads to a decreased blower duty. As described in section 5.2, the net efficiency of the reference power plant with FGR is reduced compared to the reference power plant without FGR by 0.62%-points. Since the FGR is only applied to enhance the performance of the capture plant, this loss is added to the efficiency penalty of the capture plant with FGR.

Table 13: Contributors to the overall efficiency penalty for a capture plant in combination with an NGCC plant (B1)

	Base Case with FGR	Case without FGR
Steam extraction	3.45%-points	4.42%-points
Compressor duty	1.20%-points	1.20%-points
Cooling water pumps	0.12%-points	0.15%-points
Auxiliary power	0.53%-points	1.07%-points
Flue gas recirculation	0.62%-points	
Overall efficiency penalty	**5.93%-points**	**6.84%-points**

For the process modifications, only the case with FGR is evaluated since the overall efficiency penalty as well as all specific energy demands are significantly lower compared to the case without FGR.

Comparing the results for the coal case and the natural gas case shows that the overall efficiency penalty for the natural gas case is found to be slightly lower. This is due to the lower carbon content of the fuel and thus the flue gas leading to a higher specific energy demand but a lower overall energy demand.

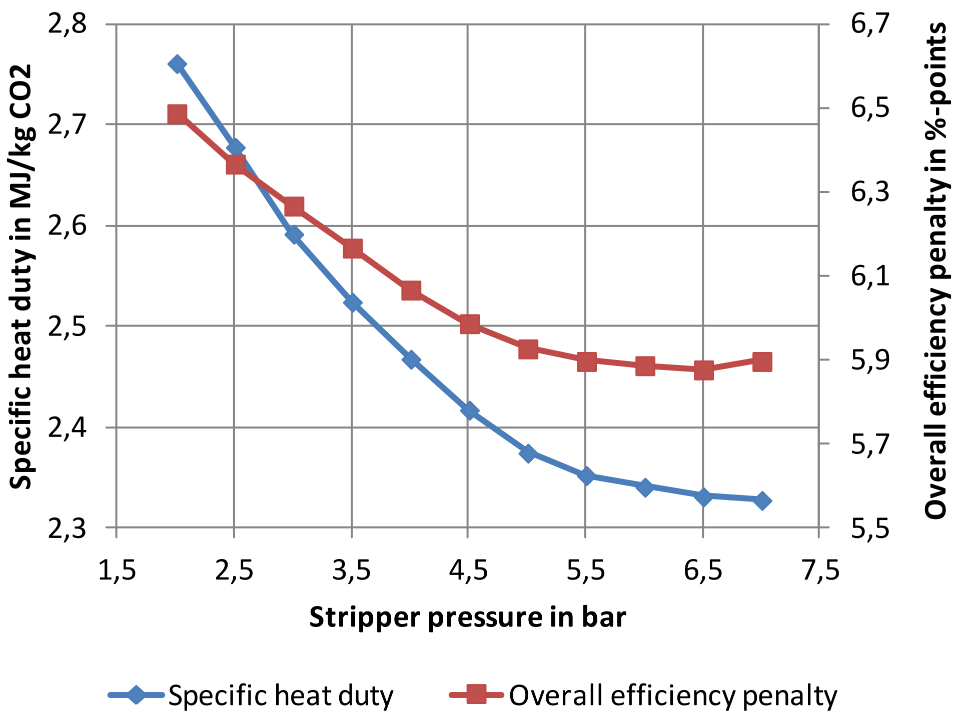

Figure 30: Overall efficiency penalty for a capture plant in combination with an NGCC plant (B1) for different stripper pressures

The effect of a reduced stripper pressure on the overall efficiency penalty can be seen in Figure 30. The lowest overall efficiency penalty is achieved for a stripper pressure of 6.5 bar. Higher stripper pressures lead to an increased reboiler temperature and thus an increased penalty due to steam extraction. Lower stripper pressures lead to an increased specific auxiliary power for CO_2 compression as well as an increased specific heat duty of the reboiler (cf. Figure 28).

The different contributors to the overall efficiency penalty for different stripper pressures are shown in Table 14 exemplarily for the NGCC Base Case as well as for the case with a stripper pressure of 2 bar. All cases are with absorber intercooling. It can be seen that despite the higher specific heat duty, the penalty due to steam extraction is slightly reduced due to the lower reboiler temperature. Still, the overall efficiency penalty is increased since the compressor duty is increased significantly.

Table 14: Contributors to the overall efficiency penalty for a capture plant in combination with an NGCC plant (B1) for different stripper pressures with absorber intercooling

	Base case (5 bar stripper pressure)	**Case with 2 bar stripper pressure**
Steam extraction	3.45%-points	3.43%-points
Compressor duty	1.20%-points	1.72%-points
Cooling water pumps	0.12%-points	0.14%-points
Auxiliary power	0.53%-points	0.57%-points
Flue gas recirculation	0.62%-points	0.62%-points
Overall efficiency penalty	**5.93%-points**	**6.48%-points**

The results for a standard MEA (30 wt%) capture process with 2 bar reboiler pressure are listed in Table 15. The reference power plant is identical to the one chosen for this study. There are no process modifications like absorber intercooling or advanced waste heat integration incorporated in the capture plant. The boundary conditions are: specific reboiler duty 3.68 MJ/kg CO_2, specific cooling duty 4.49 MJ/kg CO_2, specific auxiliary power 0.084 MJ/kg CO_2 and reboiler temperature 120.7 °C. It can be seen that the penalties due to steam extraction and cooling water pumps are increased since the specific reboiler duty as well as the specific cooling duty are higher for the MEA case. The penalty due to the compression of CO_2 is identical to the Solvent2020 case with 2 bar stripper pressure. The penalty due to auxiliary power is lower for the MEA case since the solution mass flow is significantly reduced.

Table 15: Contributors to the overall efficiency penalty for a capture plant without absorber intercooling in combination with an SCPC plant

	MEA 30 wt% case
Steam extraction	4.90%-points
Compressor duty	1.72%-points
Cooling water pumps	0.18%-points
Auxiliary power	0.44%-points
Flue gas recirculation	0.62%-points
Overall efficiency penalty	**7.86%-points**

6.3 Multi-Component Column

In state-of-the-art European power plants flue gas cleaning measures, such as a denitrification unit, an electrostatic precipitator, and also a desulphurisation unit (FGD unit) are applied. For the FGD, a spray column using limestone solution has been well-established in the last decades. More than 95% of the FGD units in power stations and industrial facilities are reliably operated on the basis of this process technology [31]. Due to its good performance in terms of SO_2 capture and high availability no other technologies were considered in the power plants recently.

In current research activities, further optimisation of the FGD performance is targeted. Andritz Energy & Environment developed the REAPLUS concept, which has been installed in the RWE power plant Niederaußem, Germany as a pilot plant (see Figure 31). The difference to standard desulphurisation lies in the staggered sequence of the scrubbing process and in improved contact between lime slurry and flue-gas SO_2. Downstream the washing section an additional wet electrostatic precipitator is installed. First pilot runs have shown promising potential for a techno-economic improvement [31, 32].

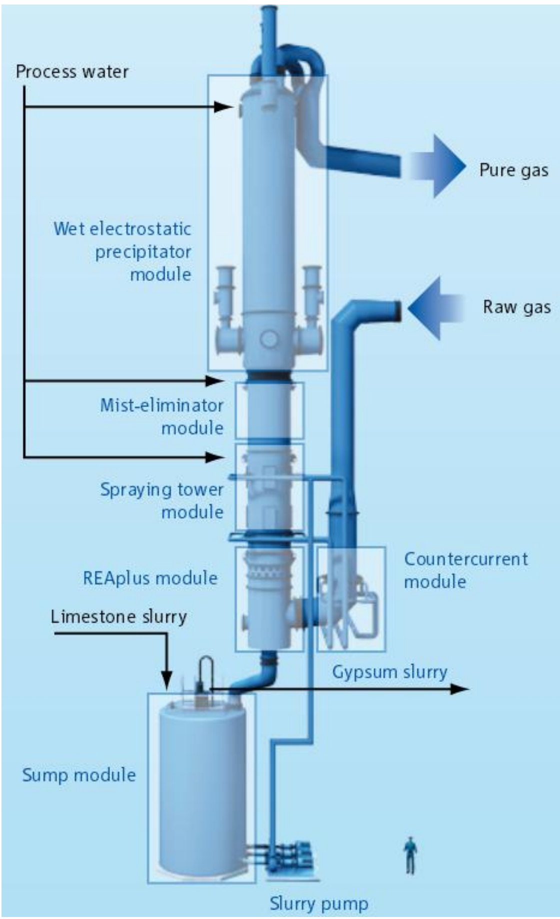

Figure 31: Sketch of the RWE REAPLUS concept [31]

For the Australian Case (no installed FGD unit), a combined capture of SO_2 and CO_2 is investigated in several research activities. Cansolv has shown first approaches for a common column for both SO_2 and CO_2 capture. However, as shown in Figure 32, the column is internally split by a water wash section to separate the SO_2 capture from the CO_2 capture [33].

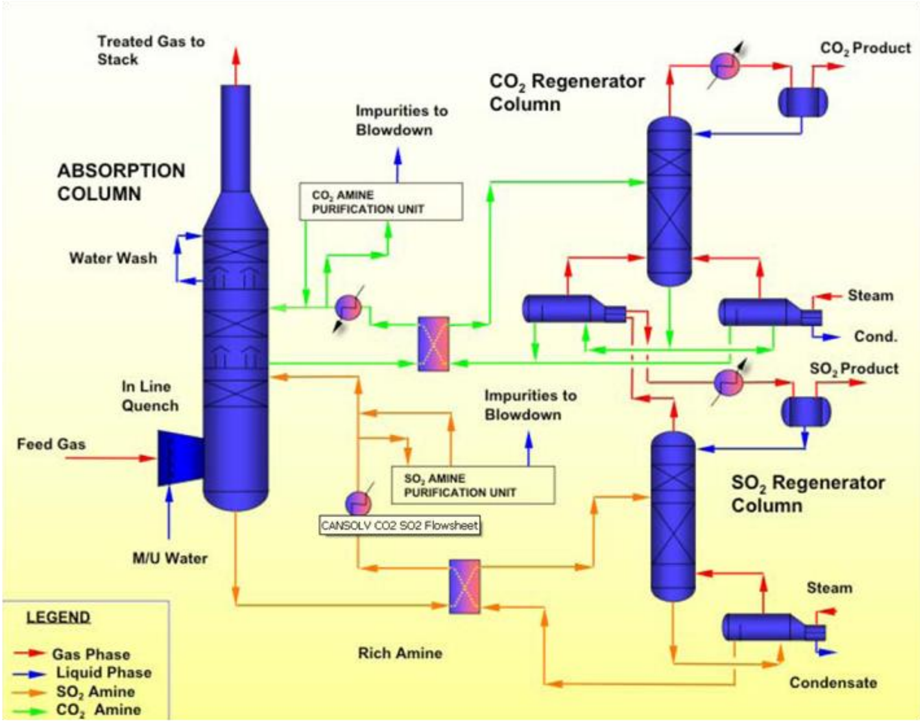

Figure 32: Flow sheet of combined SO_2-CO_2 capture by Cansolv [33]

A simultaneous capture of SO_2 and CO_2 has been investigated by CSIRO and TNO analysing different solvents and process concepts. The most promising process concept is called CASPER (see Figure 33) using 'potassium beta-alanate' as a solvent for SO_2 and CO_2. Overall process analysis have been performed showing that the energetic potential of combined SO_2-CO_2 capture is comparable to state-of-the-art CO_2 capture technologies (e. g. based on MEA) in combination with a standard FGD unit [34].

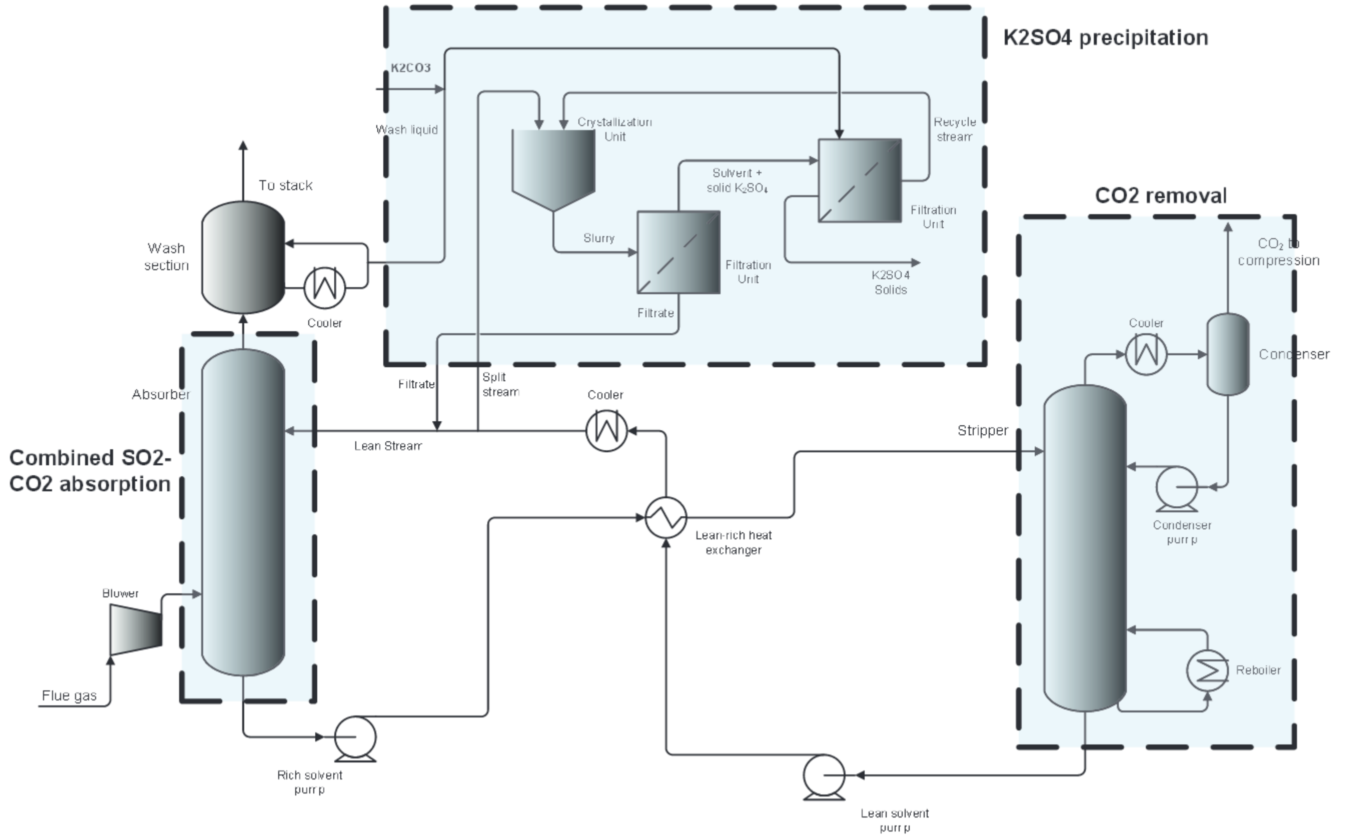

Figure 33: Flow sheet of the CASPER process by TNO [34]

In this study it was decided to exclude combined SO_2-CO_2 capture for the following reasons:

- In general, the FGD unit has a minor effect on the overall costs of electricity. Thus, the potential of improvement is very limited.
- The complexity of a combined SO_2 and CO_2 capture process leads to a lower expected availability of the combined capture process. As the power plant is not allowed to operate without SO_2 capture the increased process complexity will (in contrast to a separated CO_2 capture plant) directly lead to lower power plant availability.
- Due to the increasing grid feed-in of renewable energy sources, fossil-fuelled power plants are forced to operate in part-load mode more frequently in the near future. A CO_2 capture unit could in this case serve as a regulator for electricity generation. During high electricity demand the steam extraction for solvent regeneration could be reduced to directly increase the net output while decreasing the CO_2 capture rate and vice versa. In a combined SO_2-CO_2 capture process this benefit is inapplicable as SO_2 capture is mandatory.
- Combined SO_2-CO_2 capture requires special solvent characteristics which do not agree with the chosen Solvent2020 characteristics. The investigation of promising solvents and the development of the corresponding property model for combined SO_2 and CO_2 capture are beyond the scope of this work.

6.4 Vapour recompression

6.4.1 Process Characteristics

Vapour recompression is a process modification that reduces the reboiler heat duty by replacing it with auxiliary power in a compressor. Vapour is extracted from the process, compressed and reintroduced into the stripper. There are different process configurations possible, in which the vapour is extracted from different positions in the stripper. In some cases the vapour is taken directly from the stripper at different heights, or a liquid solvent stream is flashed to a lower pressure in order to release vapour which is then recompressed. The way of using the compressed vapour can be different as well. It can be fed back directly into the stripper, or the heat is transferred in a heat exchanger before the reintroduction [35, 36, 4, 37].

The effect of vapour recompression is strongly depending on the used solvent as well as the stripper pressure and temperature. A simple vapour recompression was tested for example for different solvents at the Esbjerg Pilot plant by DONG [24]. The result of these tests showed that the effect of vapour recompression was strongest for the solvents with high specific energy consumption. This was explained by the fact that the energy required for water evaporation is high for these solvents. The potential reduction which can be achieved by vapour recompression is therefore high as well. The reboiler heat duty for monoethanolamine (MEA) was reduced by 20%, while the reduction for CESAR I, a blend of aminomethylpropanol and piperazine, was reduced by about 13%.

In order to achieve high efficiency with vapour recompression, the vapour should consist mainly of steam. The compression of the steam changes the amount of heat of evaporation and is thus providing more heat to the stripper then is supplied by the compressor. A high CO_2 content of the vapour would not have a positive effect on the process, though. The CO_2 is throttled to a lower pressure and afterwards compressed to stripper pressure without any energetic advantage for the capture process. Therefore, the CO_2 content in the vapour should be low. Simulations of a simple recompression configuration for MEA have shown a steam content of the vapour of about 96 Vol.%.

In this study, two process configurations are evaluated for this concept. First a modification considering only one flash/compressor is analysed, which is among the process modifications that have only little influence on the complexity of the capture process. The lean solvent leaving the stripper is throttled to a lower pressure thus evaporating a part of the solvent. The vapour is flashed, compressed to the pressure in the stripper and led back to the reboiler, thus reducing the heat duty of the reboiler while the auxiliary power is increased. A schematic flow diagram of the stripper is shown in Figure 34.

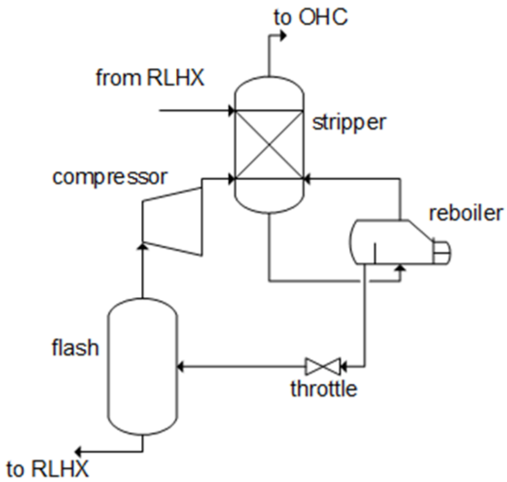

Figure 34: Schematic flow diagram of a simple one flash/compressor configuration

For the modelling of this process modification, the following additional assumptions were made: The compressor in the simulation is modelled with an isentropic efficiency of 0.83, the mechanical efficiency is assumed to be 0.99. The vapour pressure has to be higher than the stripper pressure to make the reintroduction possible. An overpressure of 10% is therefore specified between the recompressed vapour and the stripper which takes into account the losses due to friction in the pipes as well.

6.4.2 SCPC power plant results - A2

The simulations for the simple one flash/compressor configuration were executed for different flash pressures. For each pressure the L/G was varied to find the operating point with the lowest specific heat duty. As an example, the complete flow sheet for a flash pressure of 4 bar for further information on this process flow sheet modification can be found in the annex (Figure 88).

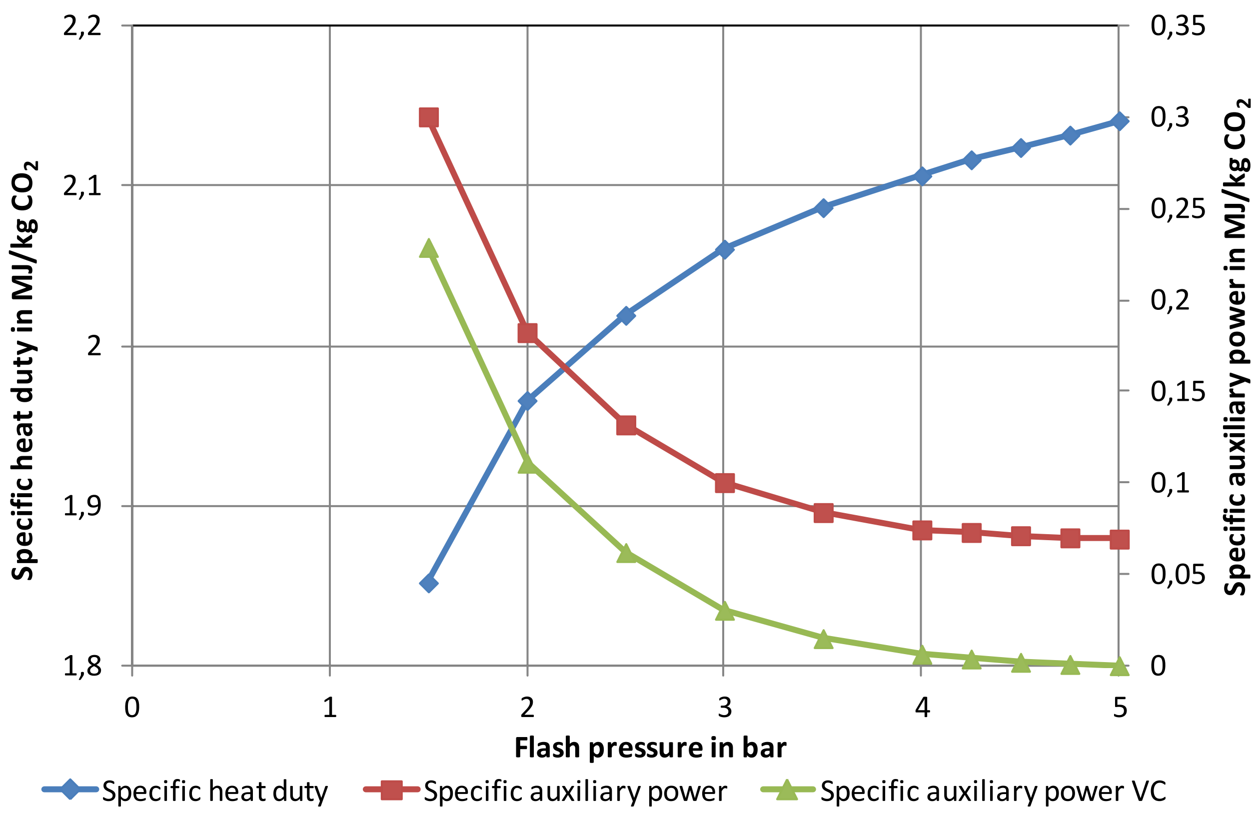

Figure 35: Specific heat duty and specific auxiliary power of a capture plant with vapour recompression in combination with an SCPC plant (A2) and specific auxiliary power of the vapour compressor for different flash pressures

The specific heat duty and the specific auxiliary power of the capture plant and the specific auxiliary power of the vapour compressor are shown in Figure 35 for different flash pressures. The operating points are the ones with the lowest specific heat duty for each flash pressure. The values shown for a flash pressure of 5 bar are the values for the base case. It can be seen that the specific heat duty is reduced as expected from 2.14 MJ/kg CO_2 for the base case to 1.85 MJ/kg CO_2 for a flash pressure of 1.5 bar. On the other hand, the specific auxiliary power is increased from 0.07 MJ/kg CO_2 to 0.3 MJ/kg CO_2. The increase in auxiliary power results from the additional compressor. The auxiliary power for the other consumers is also reduced, which can be seen in Figure 35 showing the difference between the specific auxiliary power of the capture plant and the compressor. For a flash pressure of 1.5 bar, the compressor has a specific auxiliary power of 0.23 MJ/kg CO_2, while the other electrical consumers in the capture plant have a specific auxiliary power of 0.07 MJ/kg CO_2.

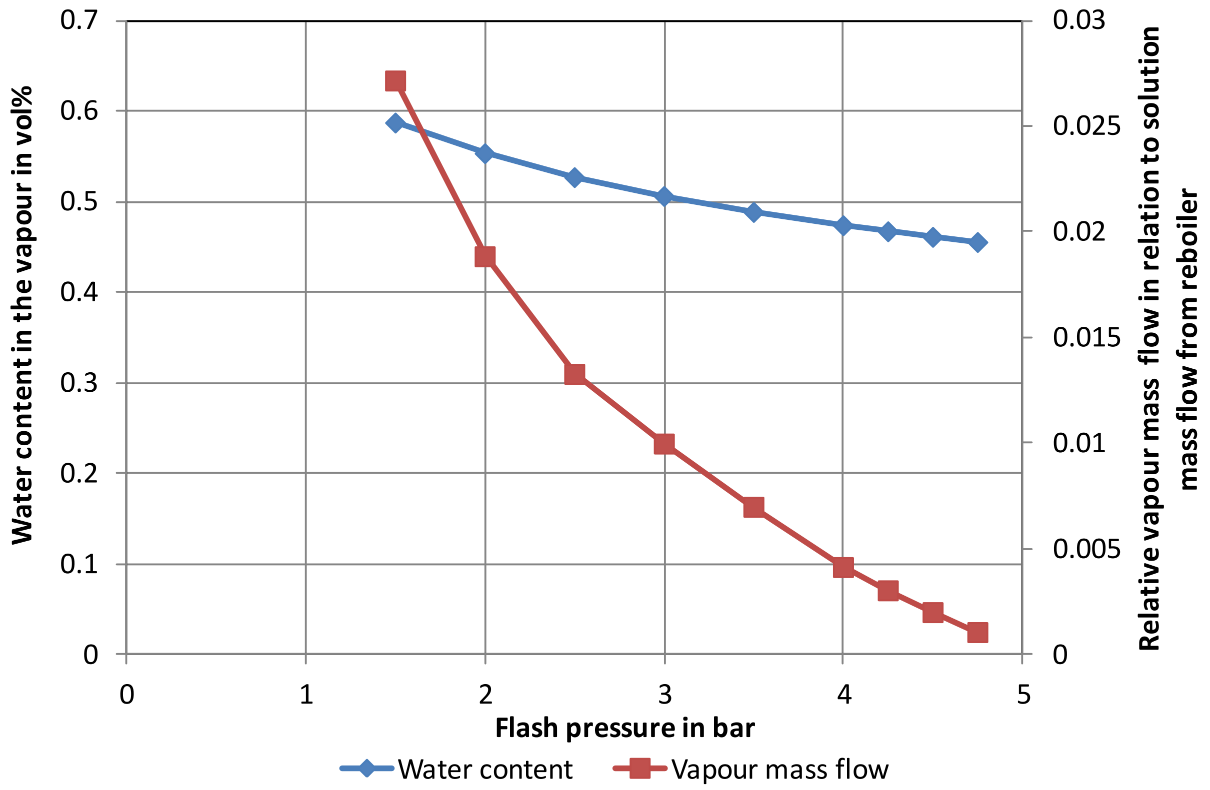

Figure 36: Relative vapour mass flow and water content in the vapour for different flash pressures in a capture plant with vapour recompression in combination with an SCPC plant (A2)

The vapour stream should consist mostly of water for vapour recompression to have the most positive effect. In Figure 36, the water content in the vapour, as well as the ratio between recompressed vapour mass flow and lean solution mass flow from the reboiler are shown for different flash pressures. The water content increases for decreasing flash pressures from around 45% to almost 60% and is thus much lower compared to the water content for a capture plant operated with MEA (>95 vol%). This is due to the low temperatures required in the reboiler since the water partial pressure in the flash is equivalent to the water vapour pressure for the respective temperature. Low temperatures at high pressures result in a low water partial pressure and thus a low water content. The mass flow of recompressed vapour increases with decreasing flash pressure since more water is evaporated, but even for low flash pressures, the mass flow is only a small fraction of the lean solvent mass flow.

In Table 16, the interface quantities are shown for the base case and for the case with vapour recompression and a flash pressure of 1.5 bar. For both cases, the operating point with the lowest specific heat duty is chosen. It can be seen that the operating point with vapour recompression has a lower lean loading and thus a lower solvent mass flow. Without any other changes in the process, this would lead to a significantly higher reboiler temperature (cf. Figure 16). Due to the effect of vapour recompression, the reboiler temperature is increased only slightly compared to the base case, though. The usable waste heat and its temperature level are decreased leading to a lower potential for waste heat integration.

Table 16: Interface quantities of a capture plant in combination with an SCPC plant for base case (A1) and case with vapour recompression (A2)

	SCPC base case	Vapour recompression with 1.5 bar flash pressure
Specific heat duty in MJ/kg CO_2	2.14	1.85
Specific cooling duty in MJ/kg CO_2	2.75	2.50
Specific auxiliary power in MJ/kg CO_2	0.07	0.29
Desorber pressure in bar	5	5
Reboiler temperature in °C	128.0	129.6
Usable waste heat from OHC in MJ/kg CO_2	0.52	0.29
Temperature level of usable waste heat in °C	116.2	102.6
Lean solution mass flow in kg/s	2934	1993

For the simple one flash/compressor configuration, the overall efficiency penalty is shown in Figure 37, for both heat integration cases, the basic integration and the waste heat integration. All data is valid for the operating point with the lowest overall efficiency penalty.

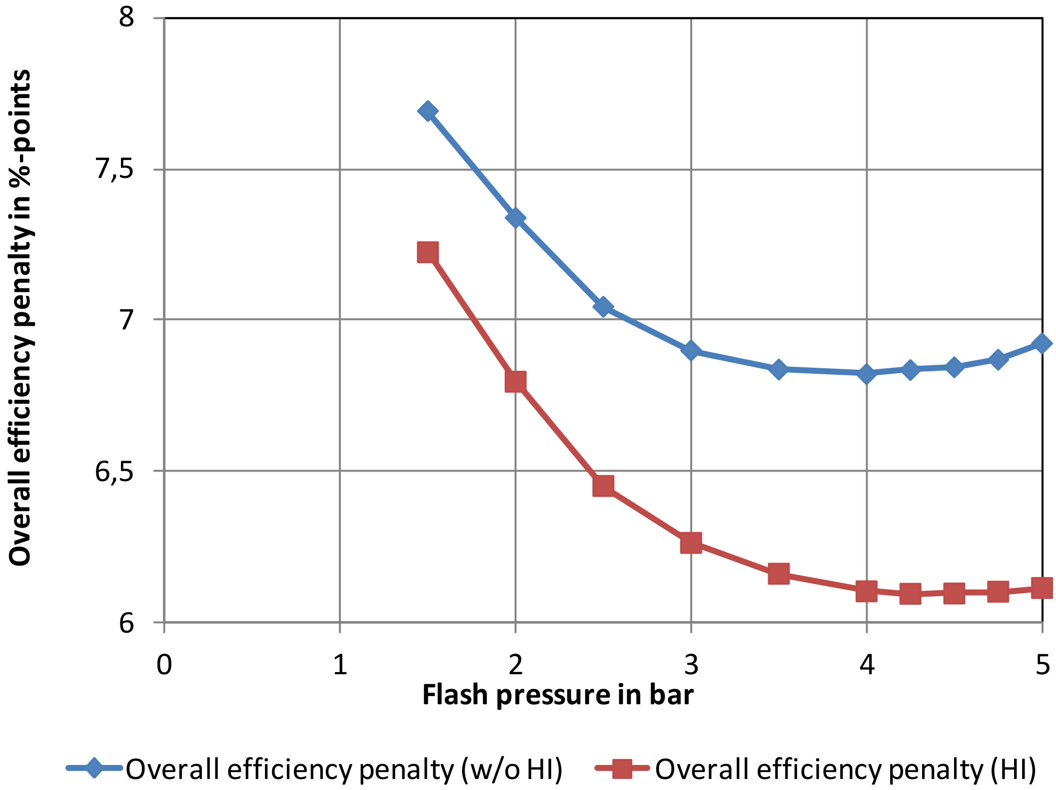

Figure 37: Overall efficiency penalty for a capture plant with vapour recompression in combination with an SCPC plant (A2)

Results from Figure 35 show that the positive effect on the overall efficiency penalty is very small. The lowest overall efficiency penalty of 6.82% for the basic integration case is achieved at a flash pressure of 4 bar. Compared to the base case this is a reduction by 0.09%-points. The effect on the case with waste heat integration is even smaller. For a flash pressure of 4.25 bar the overall efficiency penalty is 6.09%, which is a reduction by 0.02%-points. This is due to the fact that the significant increase of specific auxiliary power nearly completely compensates the positive effect of the reduced specific heat duty. This can be seen in Table 17, where the contributors to the overall efficiency penalty are shown for the base case, as well as the vapour recompression cases with and without heat integration. Compared to the base case, the penalty due to steam extraction is reduced by 0.14%-points, while the penalty due to auxiliary power of the capture plant is increased by 0.05%-points. Comparing the cases with waste heat integration shows that the penalty due to steam extraction is reduced by 0.13%-points for the vapour recompression case, while the penalty due to auxiliary power of the capture plant is increased by 0.03%-points. The positive effect of heat integration is reduced by 0.08%-points since the temperature level of usable waste heat as well as the amount of heat is reduced (cf. Table 20).

Table 17: Contributors to the overall efficiency penalty of a capture plant in combination with an SCPC plant for base case (A1) and case with vapour recompression (VR) (A2)

	SCPC base case w/o HI	**VR w/o HI**	**SCPC base case with HI**	**VR with HI**
Steam extraction	4.16%-points	4.02%-points	4.21%-points	4.08%-points
Compressor duty	1.90%-points	1.90%-points	2.06%-points	2.06%-points
Cooling water pumps	0.23%-points	0.23%-points	0.21%-points	0.21%-points
Auxiliary power	0.62%-points	0.67%-points	0.60%-points	0.63%-points
Heat integration			-0.97%-points	-0.89%-points
Overall efficiency penalty	**6.91%-points**	**6.82%-points**	**6.11%-points**	**6.09%-points**

6.4.3 NGCC power plant results - B2

The specific heat duty and the specific auxiliary power of the capture plant and the specific auxiliary power of the vapour compressor for NGCC case are shown in Figure 38 for different flash pressures. The operating points are the ones with the lowest specific heat duty for each flash pressure showed in Figure 36.

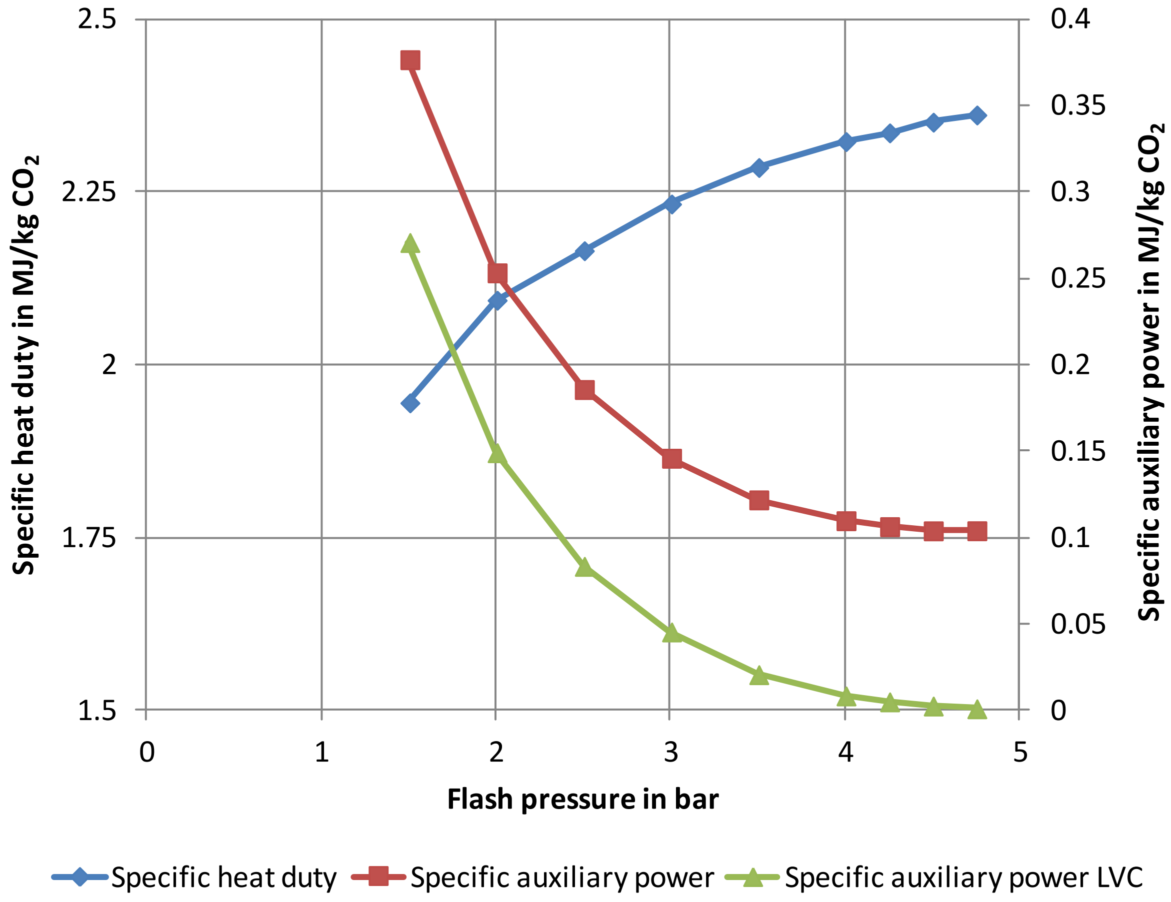

Figure 38: Specific heat duty and specific auxiliary power of a capture plant with vapour recompression in combination with an NGCC plant (B2) and specific auxiliary power of the vapour compressor for different flash pressures

It can be seen that the specific heat duty is reduced even more than for the coal case by 0.42 MJ/kg CO_2 from 2.37 MJ/kg CO_2 for the base case to 1.95 MJ/kg CO_2 for a flash pressure of 1.5 bar. On the other hand, the specific auxiliary power is increased by 0.28 MJ/kg CO_2 from 0.10 MJ/kg CO_2 to 0.38 MJ/kg CO_2. The increase in specific auxiliary power is smaller compared to the coal case and is again resulting from the additional compressor. The reboiler temperature is reduced compared to the base case. This is due to the effect of vapour recompression outweighing the reduced solution mass flow which would result in a higher reboiler temperature without vapour recompression. The flue gas temperature upstream the capture plant is reduced as well. Since less heat is required in the reboiler, less steam has to be extracted from the IP/LP crossover. Thus, more power is produced in the LP steam turbine and more condensate is available downstream the condenser - therefore more heat can be removed from the flue gas for condensate pre-heating.

Table 18: Interface quantities of a capture plant in combination with an NGCC plant for base case (B1) and case with vapour recompression (B2)

	NGCC base case	Vapour recompression with 1.5 bar flash pressure
Specific heat duty in MJ/kg CO_2	2.37	1.95
Specific cooling duty in MJ/kg CO_2	3.48	3.26
Specific auxiliary power in MJ/kg CO_2	0.10	0.38
Desorber pressure in bar	5	5
Reboiler temperature in °C	132.4	129.6
Flue gas temperature upstream of the capture plant in °C	109.8	103.9

In Figure 39, the overall efficiency penalty is shown for a capture plant at an NGCC plant with vapour recompression for different flash pressures. All data are valid for the operating point with the lowest overall efficiency penalty.

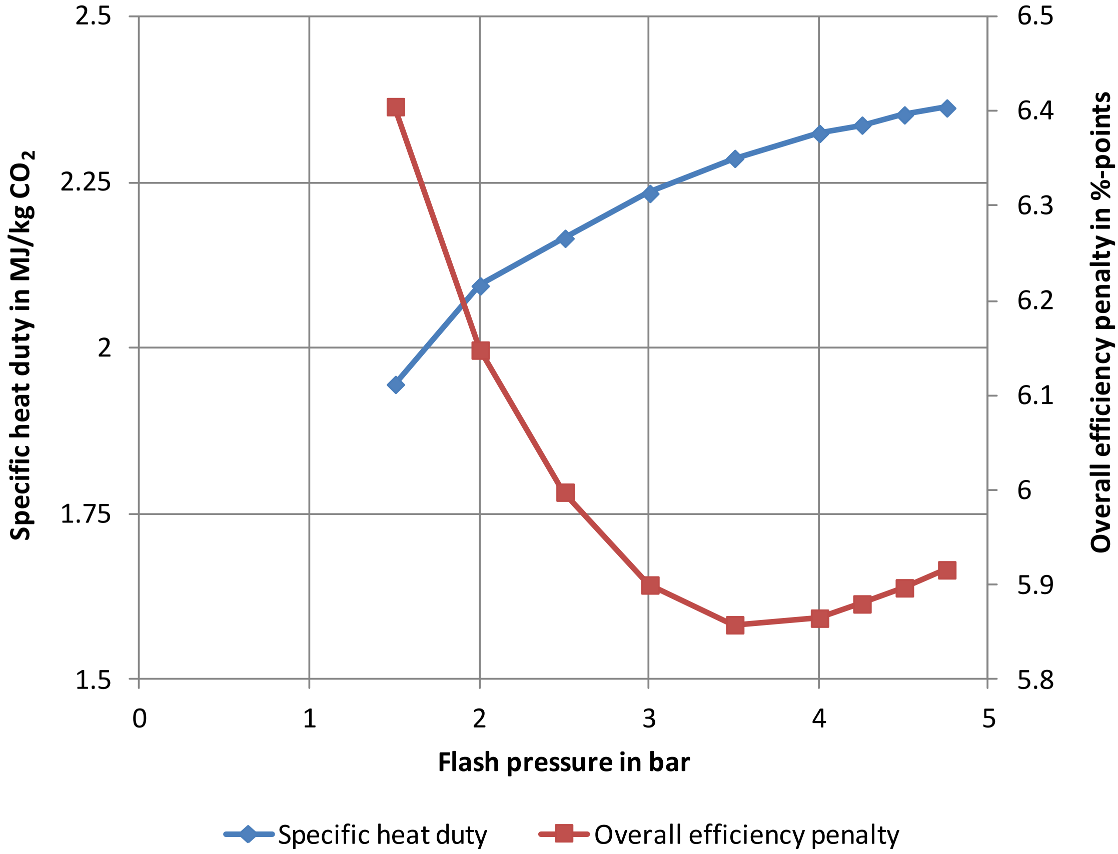

Figure 39: Overall efficiency penalty and specific heat duty for a capture plant with vapour recompression in combination with an NGCC plant (B2)

The lowest overall efficiency penalty is reached for a flash pressure of 3.5 bar. Compared to the base case, it is reduced by 0.07%-points, from 5.93%-points to 5.86%-points. The reduction is in the same order of magnitude as for the coal case, but is achieved at a lower flash pressure. The reduced specific heat duty as well as the reduced reboiler temperature resulted in a reduction of the penalty due to steam extraction by 0.18%-points, which can be seen in Table 19. This is partially outweighed by the increase of the penalty due to auxiliary power of the capture plant by 0.11%-points. For comparison, the overall efficiency penalty for a flash pressure of 1.5 bar is shown as well in Table 15. It can be seen that the penalty due to steam extraction is reduced by 0.72%-points, while the penalty due to auxiliary power of the capture plant is increased by 1.19%-points. The increase of the penalty due to auxiliary power is caused by the additional vapour compressor. Since the other penalties are nearly constant, the overall efficiency penalty is increased by 0.48%-points.

Table 19: Contributors to the overall efficiency penalty of a capture plant in combination with an NGCC plant for base case (B1) and case with vapour recompression (B2)

	NGCC base case	**VR, 3.5 bar flash pressure**	**VR, 1.5 bar flash pressure**
Steam extraction	3.45%-points	3.27%-points	2.73%-points
Compressor duty	1.20%-points	1.20%-points	1.20%-points
Cooling water pumps	0.12%-points	0.13%-points	0.14%-points
Auxiliary power	0.53%-points	0.64%-points	1.72%-points
Flue gas recirculation	0.62%-points	0.62%-points	0.62%-points
Overall efficiency penalty	**5.93%-points**	**5.86%-points**	**6.41%-points**

6.5 Multi-pressure Stripper

6.5.1 Process Characteristics

In the second vapour recompression case, an advanced process modification is analysed in which the desorber is divided into different pressure sections. The entire rich solution stream is passing through the sections following the drop of pressure. The pressure sections are fed with stripping steam using compressed vapour from the next lower pressure stage. This process modification is also referred to as multi-pressure stripper (MPS). In literature, reductions in reboiler duty for MEA have been reported between 20 and 30% [4, 27]. A schematic flow diagram of the modified stripper is shown in Figure 40.

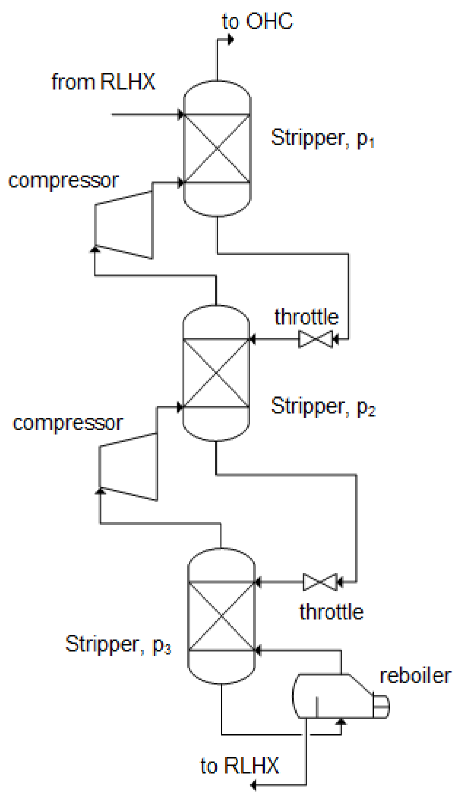

Figure 40: Schematic flow diagram of a multi-pressure stripper

For the multi-pressure stripper two pressure levels have to be varied. The pressure in the first stripper section is kept constant at 5 bar to ensure the most promising process with low energy requirements. The pressure levels in the other sections are varied. The complete flow sheet for pressures of 4 bar in the second section and 3.2 bar in the third section can exemplarily be found in the annex (Figure 89).

For the graphic account of the results, the pressure ratios between the stripper sections are used: the pressure ratio between the first and the second section p_1/p_2 as well as between the second and the third section p_2/p_3 with the pressure in the top section p_1, the pressure in the middle section p_2 and the pressure in the bottom section p_3.

The pressure ratios between the three sections can be chosen independently from each other. In order to compare different operating conditions, a characteristic number ξ is defined: the ratio between the pressure ratios is $\xi = (p_2/p_3)/(p_1/p_2)$. For $\xi > 1$ the pressure ratio between the second and the third section is larger than the pressure ratio between the first and the second section. For $\xi < 1$ the pressure ratio between the second and the third section is smaller than the pressure ratio between the first and the second section.

6.5.2 SCPC power plant results - A3

In Figure 41, the specific heat duty and the specific auxiliary power are shown for different pressure ratios p_1/p_2 and different ξ. It can be seen that the specific heat duty decreases for all ξ with increasing pressure ratio. This is consistent with the results for the simple vapour recompression, since a higher pressure ratio is equivalent to a lower pressure in the low pressure sections. For $\xi = 1$ a pressure ratio of 1.11 is equivalent to $p_2 = 4.5$ bar and $p_3 = 4.05$ bar, while a pressure ratio of 1.67 is equivalent to $p_2 = 3$ bar and $p_3 = 1.8$ bar. The results for the specific auxiliary power are consistent as well: a higher pressure ratio and thus a lower pressure in the low pressure sections lead to a higher specific auxiliary power, since the vapour coming from the low pressure sections has to be compressed to a higher pressure.

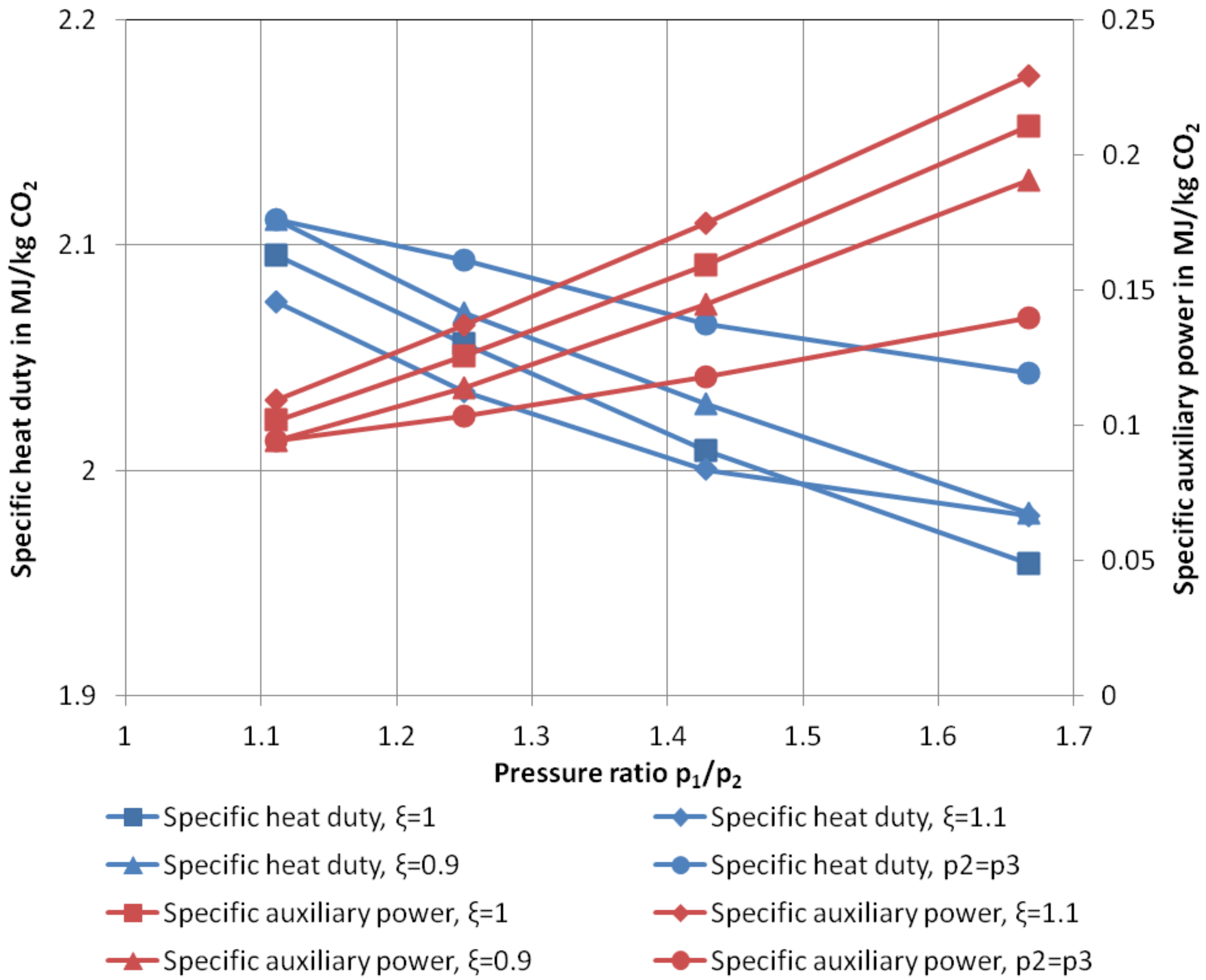

Figure 41: Specific heat duty and specific auxiliary power of a capture plant with multi-pressure stripper in combination with an SCPC plant (A3) for different operating conditions

A further reduction of the pressure in the third stripper section, equivalent to a higher value of ξ leads to an increased effect on both energy demands: the specific heat duty decreases while the specific auxiliary power increases. The opposite effect can be seen for a lower value of ξ. For comparison, the energy demands for operating points with $p_2 = p_3$ are shown as well. This is equivalent to a simpler process configuration with only two pressure sections. As expected, the effect on both energy demands is the smallest for all pressure ratios.

In Table 20, the interface quantities for the base case and for a case with a multi-pressure stripper with $p_1/p_2 = 1.67$ and $\xi = 1$ are shown. For both cases the operating point with the lowest specific heat duty is chosen. It can be seen that the reboiler temperature is significantly reduced for the multi-pressure case. This is due to the low pressure in the third stripper section resulting in a low CO_2 partial pressure and thus a low required steam partial pressure.

Table 20: Interface quantities and process values of a capture plant in combination with an SCPC plant for base case (A1) and case with multi-pressure stripper (A3)

	SCPC base case	MPS with $p_1/p_2 = 1.67$ and $\xi = 1$
Specific heat duty in MJ/kg CO_2	2.14	1.96
Specific cooling duty in MJ/kg CO_2	2.75	2.56
Specific auxiliary power in MJ/kg CO_2	0.07	0.21
Desorber pressure section 1 in bar	5	5
Desorber pressure section 2 in bar	-	3
Desorber pressure section 3 in bar	-	1.8
Reboiler temperature in °C	128.0	112.0
Usable waste heat from OHC in MJ/kg CO_2	0.52	0.31
Temperature level of usable waste heat in °C	116.2	104.4

The results for the overall efficiency penalty when a multi-pressure stripper is used are shown in Figure 42 for the basic integration and in Figure 43 for the advanced waste heat integration. It can be seen that the overall efficiency penalty is higher for all cases than for the base case. This is due to the fact, that the negative effect of the additional auxiliary power for the compression outweighs the positive effect of the reduced specific heat duty.

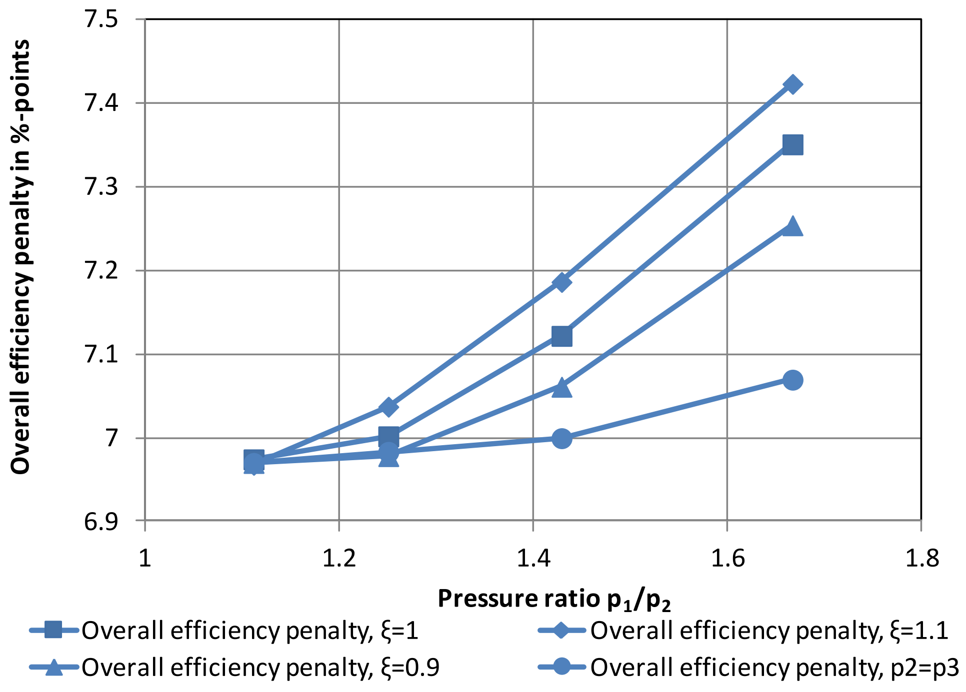

Figure 42: Overall efficiency penalty for a capture plant with multi-pressure stripper and basic integration in combination with an SCPC plant (A3)

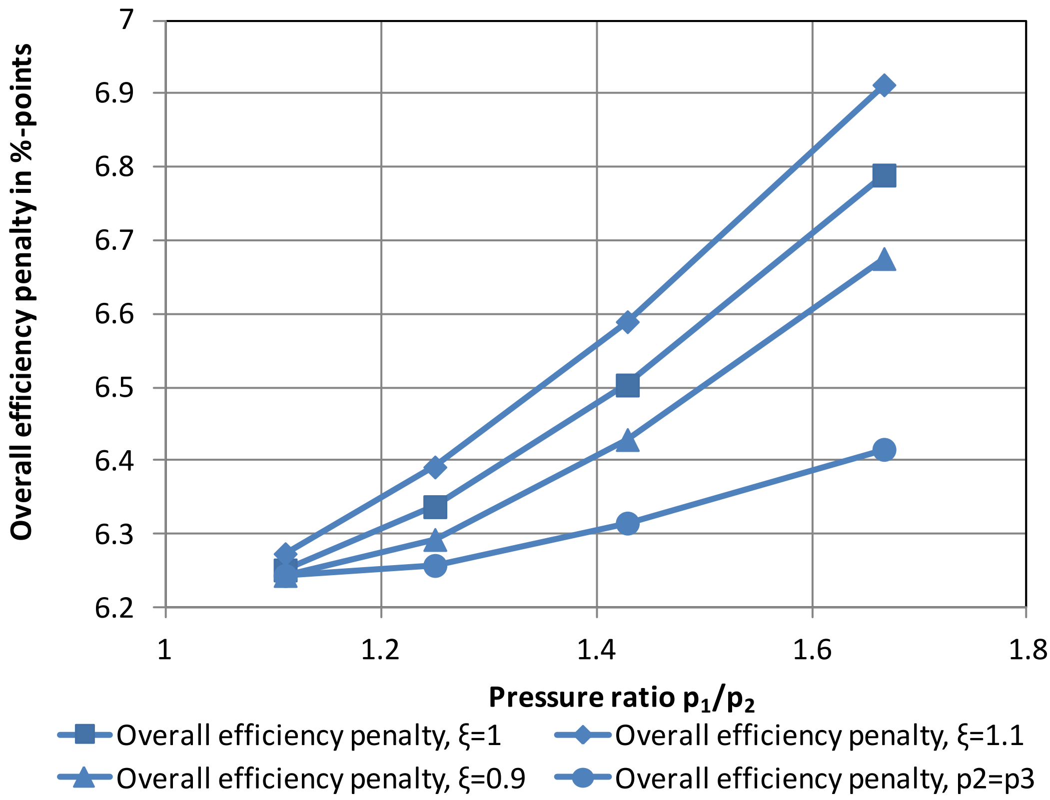

Figure 43: Overall efficiency penalty for a capture plant with multi-pressure stripper and waste heat integration in combination with an SCPC plant (A3)

The contributors to the overall efficiency penalty are shown Table 21 for the base case as well as for the multi-pressure stripper cases with and without heat integration. Compared to the base case, the penalty due to steam extraction is reduced by 0.22%-points, while the penalty due to auxiliary power of the capture plant is increased by 0.28%-points. Comparing the cases with waste heat integration shows that the penalty due to steam extraction is reduced by 0.24%-points for the multi-pressure stripper case, while the penalty due to auxiliary power of the capture plant is increased by 0.28%-points. The positive effect of heat integration is reduced by 0.10%-points since the temperature level of usable waste heat as well as the amount of heat is reduced (cf. Table 20).

In summary it can be stated, that the overall efficiency penalty for this modification is always higher than the overall efficiency penalty for the base case, which achieves a lowest overall efficiency penalty of 6.11%-points.

Table 21: Contributors to the overall efficiency penalty for a capture plant in combination with an SCPC plant for base case (A1) and case with multi-pressure stripper (A3)

	SCPC base case w/o HI	**MPS w/o HI, $p_1/p_2 = 1.1$ and $\xi = 1$**	**SCPC base case with HI**	**MPS with HI, $p_1/p_2 = 1.1$ and $\xi = 1$**
Steam extraction	4.16%-points	3.94%-points	4.21%-points	3.97%-points
Compressor duty	1.90%-points	1.90%-points	2.06%-points	2.06%-points
Cooling water pumps	0.23%-points	0.23%-points	0.21%-points	0.21%-points
Auxiliary power	0.62%-points	0.90%-points	0.60%-points	0.88%-points
Heat integration			-0.97%-points	-0.87%-points
Overall efficiency penalty	**6.91%-points**	**6.97**%-points	**6.11%-points**	**6.25%-points**

6.5.3 NGCC power plant results - B3

In Figure 44, the specific heat duty and the specific auxiliary power for a capture plant with multi-pressure stripper are shown for different pressure ratios p_1/p_2 and different ξ. It can be seen that the results are similar to the coal case. The specific heat duty decreases for all ξ with increasing pressure ratio, the specific auxiliary power increases with increasing pressure ratio. A further reduction of the pressure in the third stripper section, equivalent to a higher value of ξ leads to an increased effect on both energy demands: the specific heat duty decreases while the specific auxiliary power increases. The opposite effect can be seen for a lower value of ξ. For identical pressures in the second and third stripper $p_2 = p_3$ the effect on both energy demands is smallest for all pressure ratios.

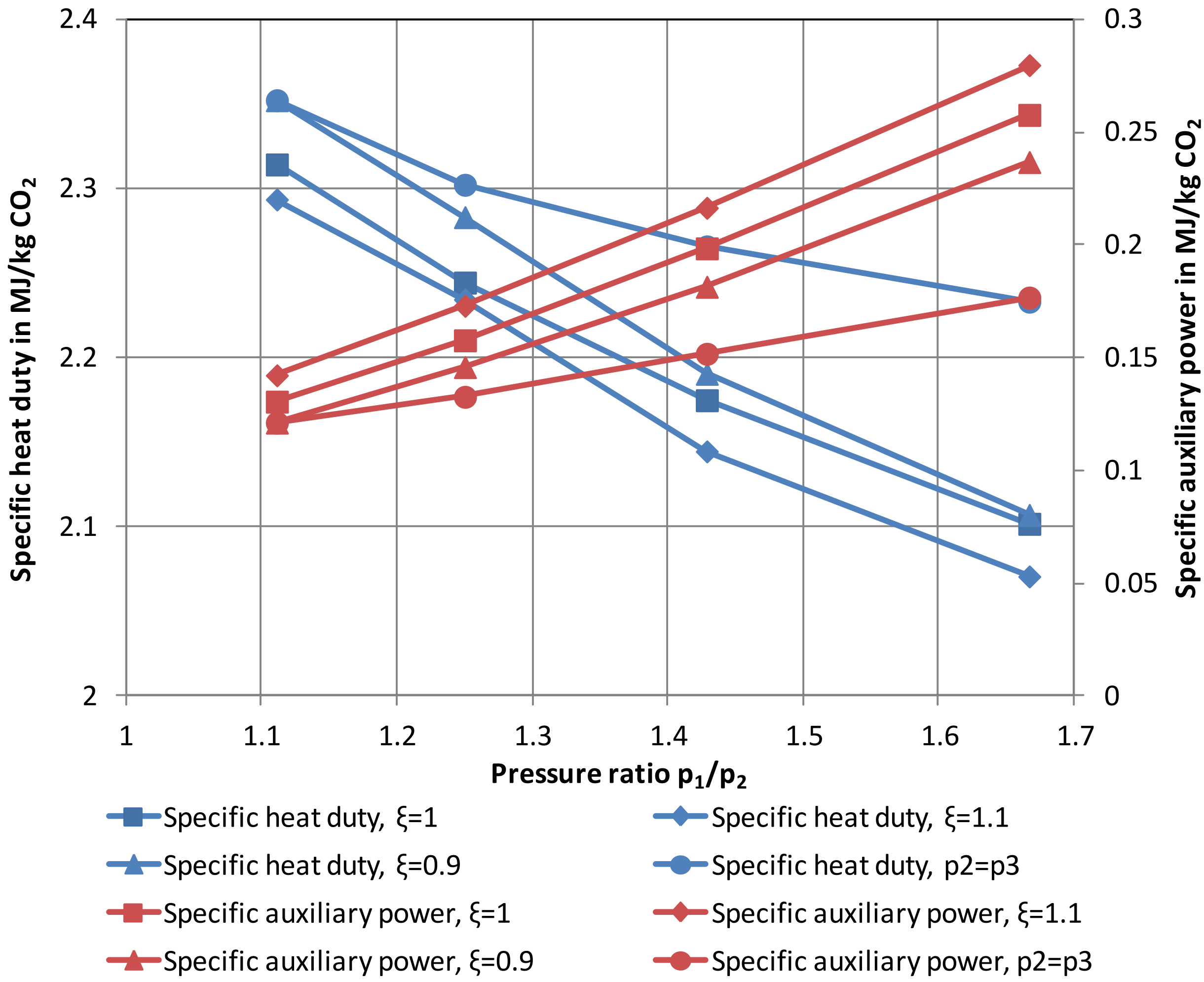

Figure 44: Specific heat duty and specific auxiliary power of a capture plant with multi-pressure stripper in combination with an NGCC plant (B3) for different operating conditions

In Table 20, the interface quantities for the NGCC base case and for a case with a multi-pressure stripper with $p_1/p_2 = 1.67$ and $\xi = 1.1$ are shown. For both cases the operating point with the lowest specific heat duty is chosen. As for the coal case, it can be seen that the reboiler temperature is significantly reduced for the multi-pressure case. The flue gas temperature is reduced as well, since less steam is needed for regeneration and more condensate is available in the economiser.

Table 22: Interface quantities and process values of a capture plant in combination with an NGCC plant for base case (B1) and case with multi-pressure stripper (B3)

	NGCC base case	MPS with $p_1/p_2 = 1.67$ and $\xi = 1.1$
Specific heat duty in MJ/kg CO_2	2.37	2.07
Specific cooling duty in MJ/kg CO_2	3.48	3.15
Specific auxiliary power in MJ/kg CO_2	0.10	0.28
Desorber pressure section 1 in bar	5	5
Desorber pressure section 1 in bar	-	3
Desorber pressure section 1 in bar	-	1.64
Reboiler temperature in °C	132.4	110.9
Flue gas temperature upstream of the capture plant in °C	109.8	95.5

The results for the overall efficiency penalty when a multi-pressure stripper is used are shown in Figure 45. for the NGCC case. It can be seen that the overall efficiency penalty for small pressure ratios is lower compared to the base case, but increases for higher pressure ratios. The overall efficiency penalty for all configurations is lowest for a pressure ratio $p_1/p_2 = 1.25$.

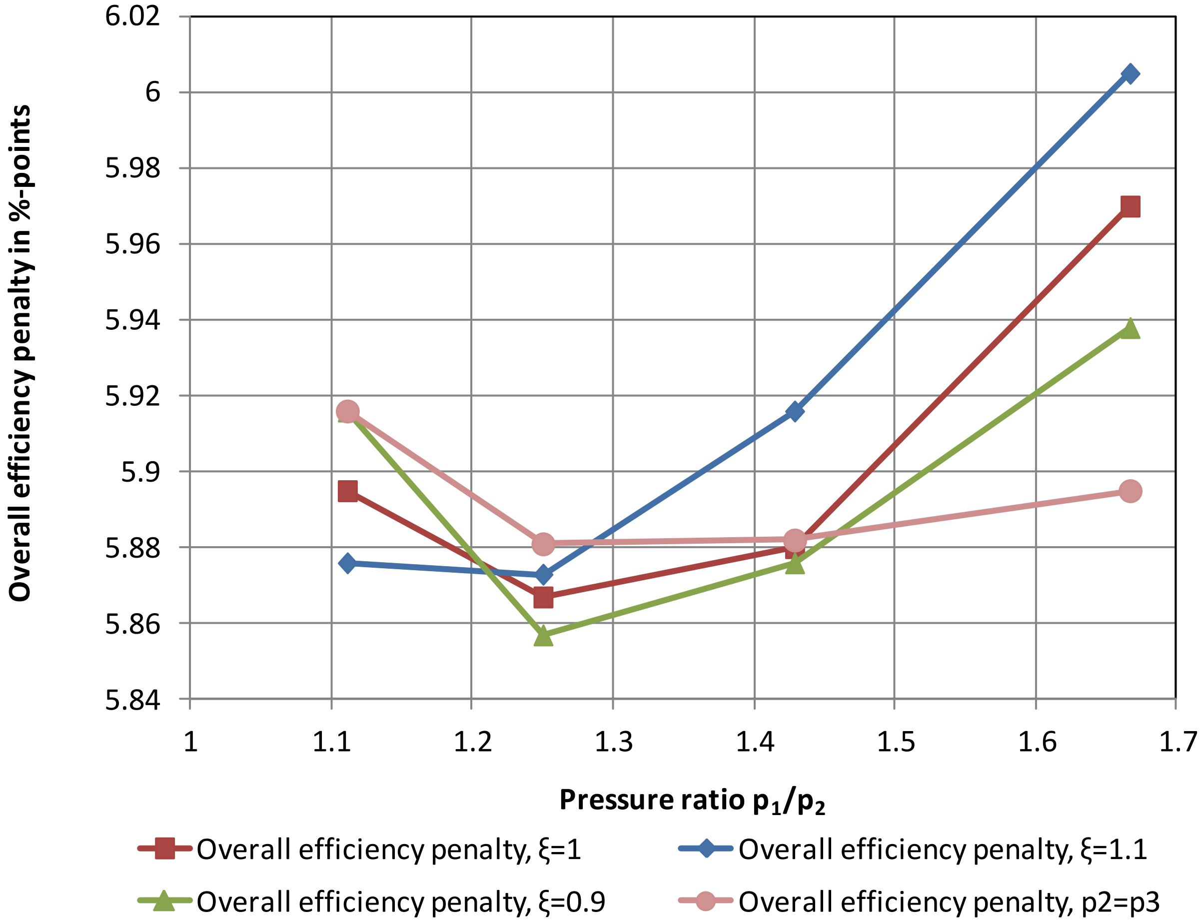

Figure 45: Overall efficiency penalty for a capture plant with multi-pressure stripper in combination with an NGCC plant (B3)

In Table 23, the contributors to the overall efficiency penalty are shown for the NGCC base case, the case with the lowest overall efficiency penalty and the case with the lowest specific heat duty. It can be seen that the overall efficiency penalty is reduced by 0.07%-points for the case with the lowest overall efficiency penalty ($p_1/p_2 = 1.25$, $\xi = 0.9$). This is due to the lower penalty by steam extraction which is reduced by 0.30%-points. The lower specific heat duty as well as the lower reboiler temperature are both contributing to this reduction. The increase of the penalty due to auxiliary power of the capture plant by 0.23%-points does not outweigh this benefit. For the case with the lowest specific heat duty the effect is reversed. The reduction of the penalty due to steam extraction is smaller (0.83%-points) compared to the increase of the penalty due to auxiliary power of the capture plant (0.91%-points).

Table 23: Contributors to the overall efficiency penalty for a capture plant in combination with an NGCC plant for base case (B1) and case with multi-pressure stripper (B3)

	NGCC base case	**MPS with $\mathbf{p_1/p_2 = 1.25}$ and $\boldsymbol{\xi = 0.9}$**	**MPS with $\mathbf{p_1/p_2 = 1.67}$ and $\boldsymbol{\xi = 1.1}$**
Steam extraction	3.45%-points	3.15%-points	2.62%-points
Compressor duty	1.20%-points	1.20%-points	1.20%-points
Cooling water pumps	0.12%-points	0.13%-points	0.12%-points
Auxiliary power	0.53%-points	0.76%-points	1.44%-points
Flue gas recirculation	0.62%-points	0.62%-points	0.62%-points
Overall efficiency penalty	**5.93%-points**	**5.86%-points**	**6.01%-points**

6.6 Heat-integrated stripping column

6.6.1 Process Characteristics

In a standard capture process, the heat exchange between rich and lean solution takes place in the rich-lean heat exchanger and is achieved before the rich solution enters the stripping column. In a heat-integrated stripping column, the heat exchanger is integrated into the stripper. The rich solution from the absorber is introduced directly into the stripper, where it is heated up by lean solution which is conducted in counter current flow. In literature, the equivalent work is claimed to be reduced by around 20% for different solvents [5].

In this study, a simplified process configuration which is shown in Figure 44 is evaluated. The stripper is divided into two sections. The rich solution from the RLHX is partially regenerated in the upper section and is then cross heat exchanged in an interheater with hot lean solution from the reboiler. Afterwards, it is further regenerated in the lower section. The lean solution from the interheater is led to the RLHX. This configuration is less complex compared to a heat-integrated stripping column, but has similar advantages. The temperature at the stripper head is reduced, since the lean solution is already cooled in the interheater which results in a lower temperature in the RLHX. This leads to less water in the overhead vapour and thus less heat that has to be transferred in the overhead condenser (OHC). In addition, the low temperature at the stripper head leads to an increased temperature gradient in the column, when the reboiler temperature is kept constant or is increased. Thus, the conditions especially in the upper part of the stripper are closer to equilibrium.

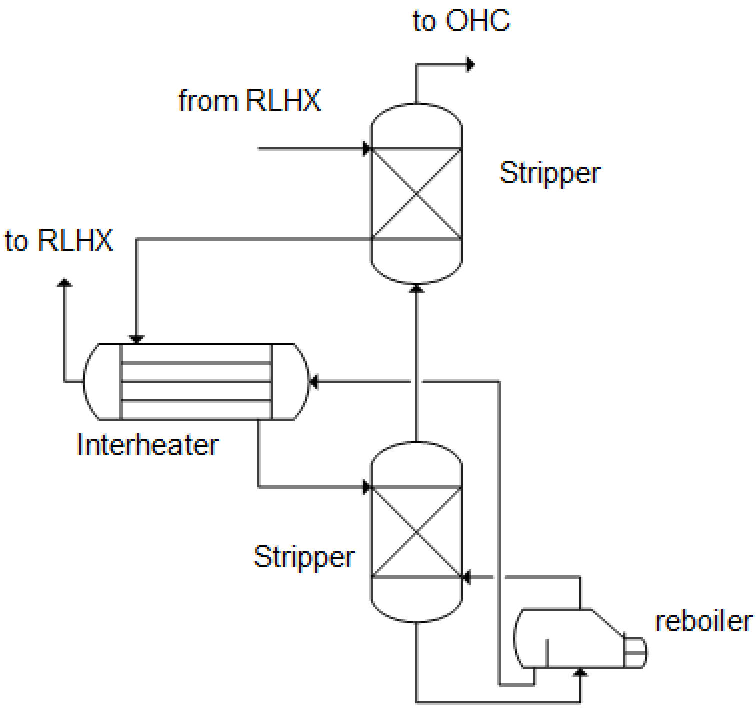

Figure 46: Schematic flow diagram of an interheated stripping column

6.6.2 SCPC power plant case - A4

For the evaluation of the capture plant with an interheated stripper, the heat which is transferred in the interheater was varied. For each heat duty, L/G was varied to find the operating point with the lowest specific reboiler heat duty. The results for the capture process with an interheated stripper are shown in Figure 47, where the specific heat duty and the specific auxiliary power are plotted against the relative interheater duty (RID). The RID is the heat duty of the interheater as a fraction of the reboiler heat duty. Other heat duties, like the RLHX duty, could be used as a basis as well, but doing so would not change the shape of the curves. The reboiler duty is chosen since it changes less than the RLHX duty.

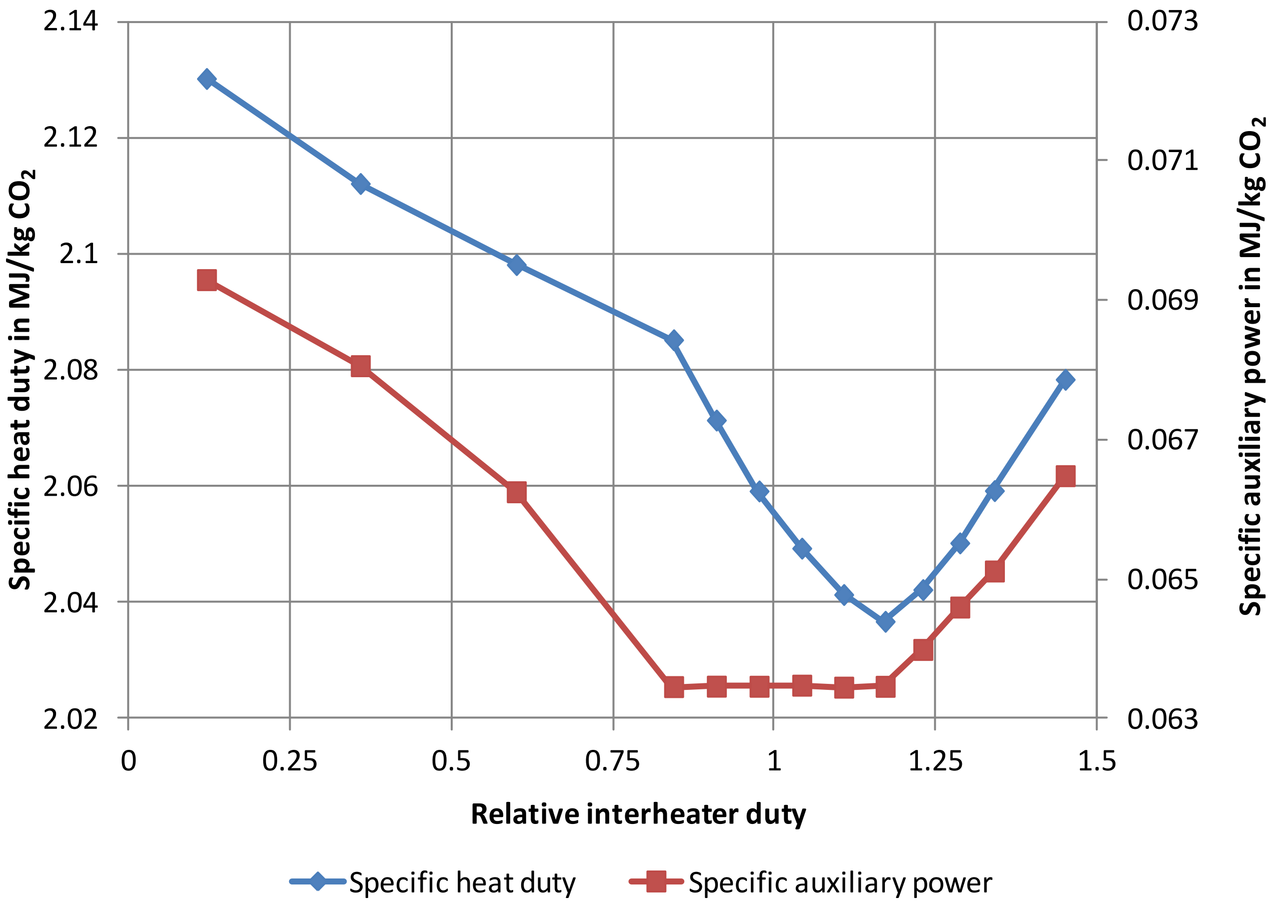

Figure 47: Specific heat duty and specific auxiliary power of a capture plant with interheated stripper in combination with an SCPC plant (A4) for different relative interheater duties

It can be seen that the specific heat duty decreases when more heat is transferred in the interheater. Since a fraction of the energy contained in the lean solution stream is recycled to the stripper, less steam is required in the reboiler. The specific auxiliary power decreases as well. This is due to the lower auxiliary power of the rich solution pump since the optimal L/G decreases as well. In the detailed simulations it can be seen that the lowest possible L/G is reached for an RID of 0.8. A further reduction of the solvent mass flow would lead to reboiler temperatures of more than 150 °C. Thus, the specific auxiliary power does not change for higher interheater duties up to an RID of 1.2. When the interheater duty is increased further, the L/G has to be increased as well. This is due to the temperatures in the interheater. Since the solution mass is not changed between an RID of 0.8 and 1.2, the LMTD has to be reduced to allow for higher heat flow rates. Following the assumptions for the RLHX, a minimum LMTD of 5 K is assumed for the interheater as well. For an RID of 1.2, this limit is reached and the solution mass flow has to be increased to allow for higher heat duties. This leads to a higher energy demand for the lean solution pump and thus a higher specific auxiliary power. The specific heat duty of the reboiler increases as well.

Table 24: Interface quantities of a capture plant in combination with an SCPC plant for base case (A1) and case with interheated stripper (A4)

	SCPC base case	Interheated stripper with RID 1.2
Specific heat duty in MJ/kg CO_2	2.14	2.04
Specific cooling duty in MJ/kg CO_2	2.75	2.54
Specific auxiliary power in MJ/kg CO_2	0.07	0.06
Desorber pressure in bar	5	5
Reboiler temperature in °C	128.0	150.7
Usable waste heat from OHC in MJ/kg CO_2	0.52	0.34
Temperature level of usable waste heat in °C	116.2	106.5

The interface quantities for the base case and for the interheated stripper case with the lowest specific heat duty are shown in Table 24. The specific heat duty is decreased by 0.1 MJ/kg CO_2 from 2.14 MJ/kg CO_2 to 2.04 MJ/kg CO_2. The specific cooling duty and the specific auxiliary duty are decreased by 0.21 MJ/kg CO_2 and 0.1 MJ/kg CO_2, respectively. Since the solution mass flow is reduced, the reboiler temperature has to be increased to allow for a higher water partial pressure in the stripper and thus lower lean loadings. The temperature level of usable waste heat is decreased since a fraction of the heat available in the lean solution is transferred in the interheater. Thus, the temperatures in the RLHX and in the stripper head are reduced.

The effect of an interheated stripper on the overall efficiency penalty is shown in Figure 48. It can be seen that the use of an interheater results in a marginal reduction for the case without waste heat integration. For the case with waste heat integration the overall efficiency penalty is increased for all operating points. This is due to the high sensitivity of the reboiler temperature due to changed lean loadings for Solvent2020 compared to other solvents. A relatively small change in lean loading results in a steep change of the reboiler temperature and thus in the losses due to steam extraction.

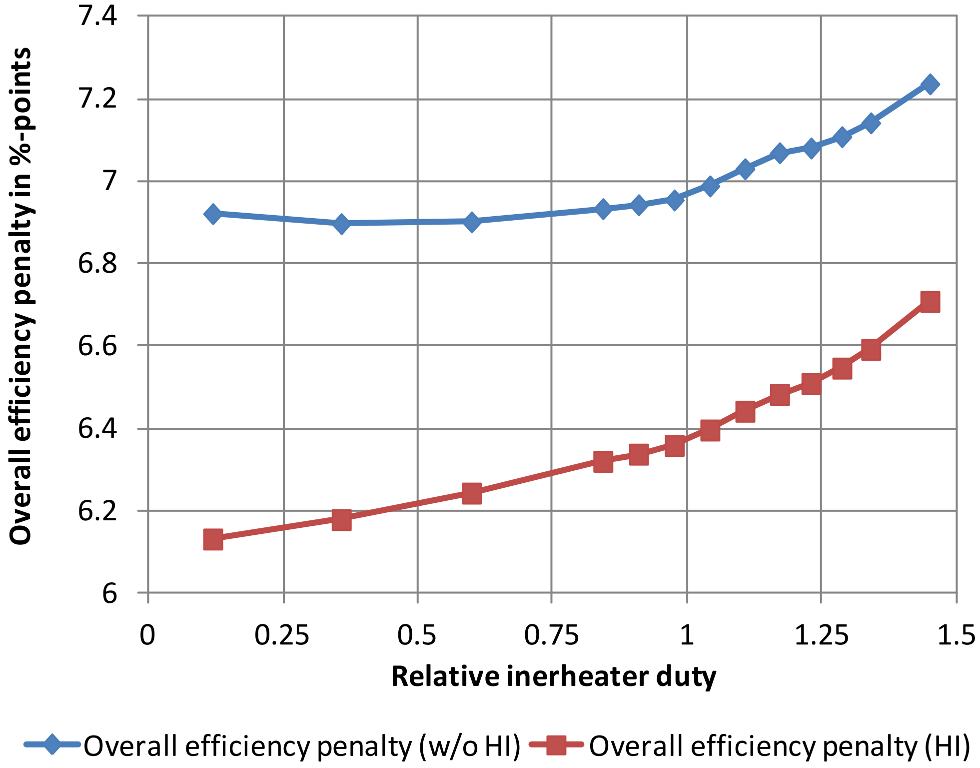

Figure 48: Overall efficiency penalty of a capture plant with interheated stripper in combination with an SCPC plant (A4) with and without heat integration

The positive effect on the overall efficiency penalty is very small (cf. Figure 48). The lowest overall efficiency penalty of 6.89%-points for the basic integration case without heat integration is achieved at an RID of 0.35. Compared to the base case this is a reduction by 0.02%-points. This can be seen in Table 25, were the contributors to the overall efficiency penalty are shown for the base case as well as two cases with interheated stripper. In addition to the operating point with the lowest overall efficiency penalty, the operating point with the lowest specific heat duty is shown, which is obtained for an RID of 1.2. Compared to the base case, only the penalty due to auxiliary power of the capture plant is reduced for an RID of 0.35. This is due to the lower solution mass flow and thus a lower energy demand of the lean solution pump. For an RID of 1.2 the penalty due to auxiliary power of the capture plant is decreased even further, but the penalty due to steam extraction is increased due to the higher reboiler temperature which outweighs the lower auxiliary power.

Table 25: Contributors to the overall efficiency penalty of a capture plant in combination with an SCPC plant for base case (A1) and case with interheated stripper (A4) without waste heat integration

	SCPC base case without HI	**Interheated stripper with RID 0.35, without HI**	**Interheated stripper with RID 1.2, without HI**
Steam extraction	4.16%-points	4.16%-points	4.38%-points
Compressor duty	1.90%-points	1.90%-points	1.90%-points
Cooling water pumps	0.23%-points	0.23%-points	0.22%-points
Auxiliary power	0.62%-points	0.60%-points	0.57%-points
Overall efficiency penalty	**6.91%-points**	**6.89%-points**	**7.07**%-points

For the cases with waste heat integration no reduction of the overall efficiency penalty is achieved. Compared to the base case for an RID of 0.35, the penalty due to steam extraction and auxiliary power of the capture plant are reduced by 0.04%-points and 0.02%-points, respectively, but the positive effect of heat integration is reduced. This is due to the reduced temperature in the OHC (cf. Table 24). For an RID of 1.2, the penalty due to steam extraction is increased due to the high reboiler temperature, while the positive effect of heat integration is reduced even further.

In summary it can be stated, that the overall efficiency penalty for this modification with advanced waste heat integration is always higher than the overall efficiency penalty for the base case which achieves a lowest overall efficiency penalty of 6.11%-points.

Table 26: Contributors to the overall efficiency penalty of a capture plant in combination with an SCPC plant for base case (A1) and case with interheated stripper (A4) with waste heat integration

	SCPC base case with HI	**Interheated stripper with RID 0.35, with HI**	**Interheated stripper with RID 1.2, with HI**
Steam extraction	4.21%-points	4.17%-points	4.38%-points
Compressor duty	2.06%-points	2.06%-points	2.06%-points
Cooling water pumps	0.21%-points	0.21%-points	0.20%-points
Auxiliary power	0.60%-points	0.58%-points	0.57%-points
Heat integration	-0.97%-points	-0.86%-points	-0.72%-points
Overall efficiency penalty	**6.11%-points**	**6.18%-points**	**6.48**%-points

6.6.3 NGCC power plant case - B4

As for the coal case, the heat which is transferred in the interheater was varied. For each interheater duty L/G was varied to find the operating point with the lowest specific reboiler heat duty. The results for the capture process of an NGCC power plant with an interheated stripper are shown in Figure 49, where the specific heat duty and the specific auxiliary power are plotted against the RID.

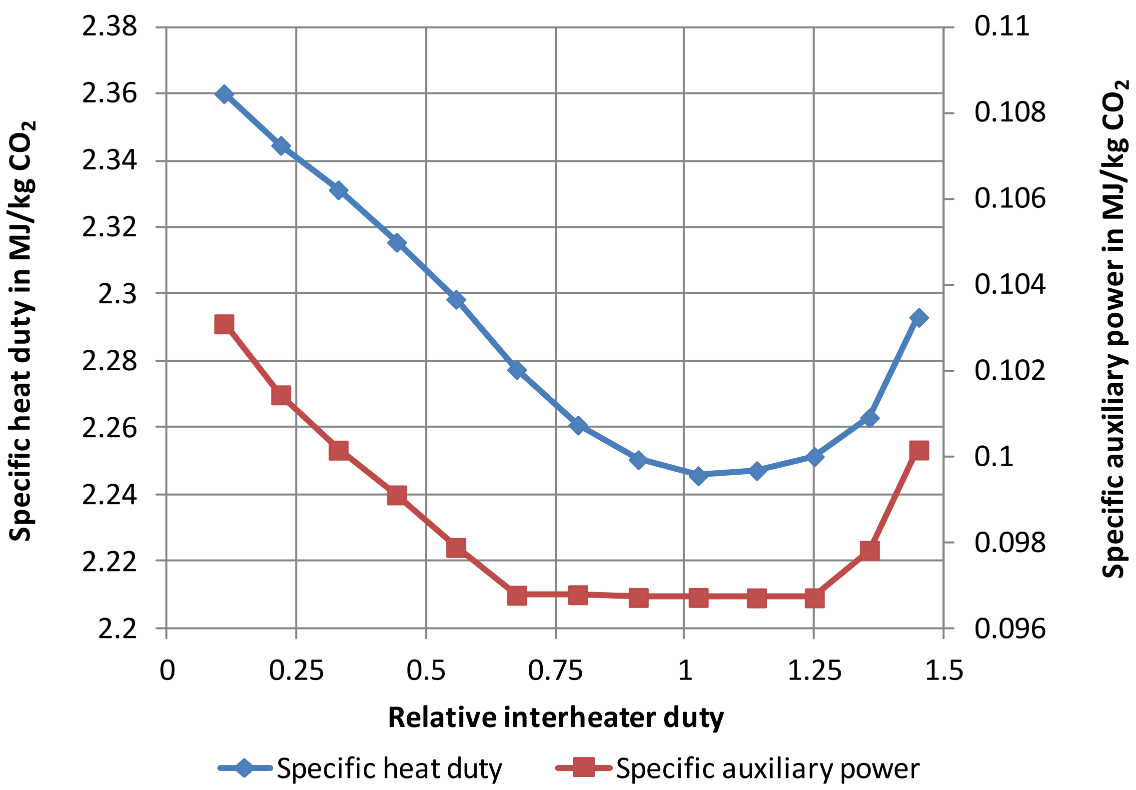

Figure 49: Specific heat duty and specific auxiliary power of a capture plant with interheated stripper in combination with an NGCC plant (B4) for different interheater heat duties

As for the coal case, the specific heat duty decreases when more heat is transferred in the interheater, since less heat has to be transferred in the reboiler. The specific auxiliary power decreases marginally due to the lower power demand of the rich solution pump since the optimal L/G decreases. For an RID of 0.67, the lowest possible L/G is reached. A further reduction of the solvent mass flow would lead to reboiler temperatures of more than 150 °C. Thus, the specific auxiliary power does not change for higher interheater duties up to an RID of 1.25. When the interheater duty is increased further, the L/G has to be increased as well to ensure a minimum LMTD of 5 K in the interheater. This leads to a higher energy demand for the lean solution pump and thus a higher specific auxiliary power.

Table 27: Interface quantities of a capture plant in combination with an NGCC plant for base case (B1) and case with interheated stripper (B4)

	NGCC base case	Interheated stripper with RID 1.03
Specific heat duty in MJ/kg CO_2	2.37	2.25
Specific cooling duty in MJ/kg CO_2	3.48	3.22
Specific auxiliary power in MJ/kg CO_2	0.10	0.10
Desorber pressure in bar	5	5
Reboiler temperature in °C	132.4	149.2
Flue gas temperature upstream of the capture plant in °C	109.8	114.1

The interface quantities for the base case and for the interheated stripper with the lowest specific heat duty are shown in Table 27. The specific heat duty is decreased by 0.12MJ/kg CO_2 from 2.37 MJ/kg CO_2 to 2.25 MJ/kg CO_2. The specific cooling duty is decreased by 0.26 MJ/kg CO_2, the decrease in auxiliary power cannot be seen in Table 27 due to rounding. Since the solution mass flow is reduced, the reboiler temperature has to be increased to allow for higher water partial pressure in the stripper and thus lower lean loadings.

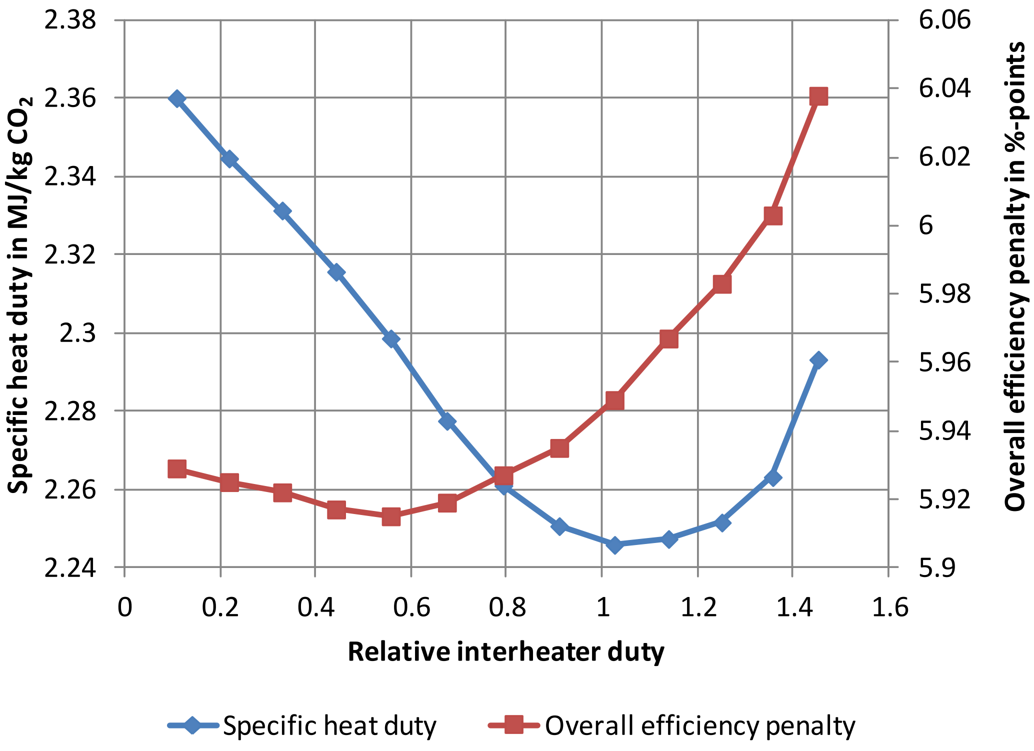

Figure 50: Specific heat duty and overall efficiency penalty of a capture plant with interheated stripper in combination with an NGCC plant (B4)

As for the coal case, the positive effect on the overall efficiency penalty is very small (cf. Figure 50). The lowest overall efficiency penalty of 5.92%-points is achieved at an RID of 0.56. Compared to the base case, this is a reduction by 0.01%-points. For higher RID the overall efficiency penalty increases, but not as fast as for the coal case. The contributors to the overall efficiency penalty are shown in Table 28 for the base case as well as for two cases with interheated stripper. The operating point with an RID of 0.56 is the operating point with the lowest overall efficiency penalty. For an RID of 1.03 the lowest specific heat duty is achieved. Compared to the base case, only the penalty due to steam extraction is reduced for an RID of 0.56. This is due to the reduced specific heat duty. For an RID of 1.03 the penalty due to steam extraction is increased since the high reboiler temperature outweighs the reduced specific heat duty.

Table 28: Contributors to the overall efficiency penalty of a capture plant in combination with an NGCC plant for base case (B1) and case with interheated stripper (B4)

	NGCC base case	**Interheated stripper with RID 0.56**	**Interheated stripper with RID 1.03**
Steam extraction	3.45%-points	3.44%-points	3.48%-points
Compressor duty	1.20%-points	1.20%-points	1.20%-points
Cooling water pumps	0.12%-points	0.12%-points	0.12%-points
Auxiliary power	0.53%-points	0.53%-points	0.52%-points
Flue gas recirculation	0.62%-points	0.62%-points	0.62%-points
Overall efficiency penalty	**5.93%-points**	**5.92%-points**	**5.96%-points**

6.7 Improved split flow process

6.7.1 Process Characteristics

A split flow process is defined by splitting a solvent stream and using it for different means. In the literature, there are several process concepts which can be defined as split flow processes. In the following, two of these concepts are evaluated.

A concept first described in 1934 by Shoeld [38] is shown in Figure 51. A fraction of the solution is withdrawn at half height of the absorber and the stripper. These split streams are partially loaded or partially regenerated, respectively. The partial loaded solution withdrawn from the absorber is heated in a heat exchanger by the partial regenerated solution from the stripper. Afterwards, it is fed to the stripper at half height. The cooled partial regenerated solution is fed to the absorber at half height. As for the base case, the lean solution from the stripper sump and the rich solution from the absorber sump are cross heat exchanged and led to the head of the opposing column. This modification is intended to reduce the reboiler duty, since only a fraction of the solution has to be regenerated to the lowest loading. On the other hand, an increased solution mass flow might be necessary since the working range of a fraction of the solution is reduced.

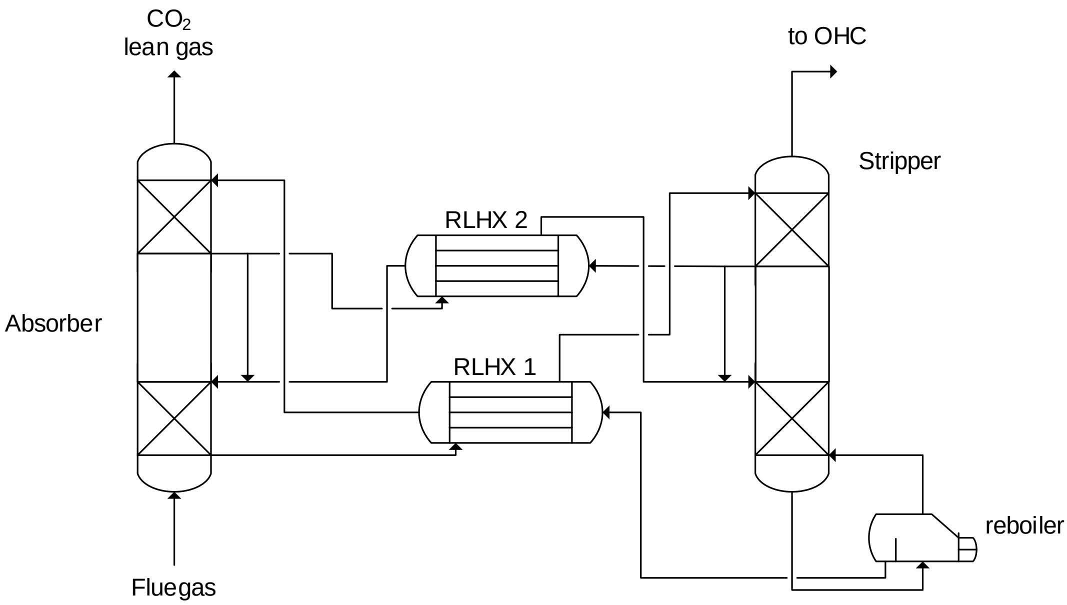

Figure 51: Schematic flow diagram of the split flow process by Shoeld [38]

A different split flow concept was suggested by Eisenberg and Johnson [39] and is shown in Figure 52. A portion of the rich solution coming from the absorber is branched off. This stream bypasses the RLHX and is led directly to the top of the desorber. The bulk of the rich solution is led to the RLHX and enters the stripper below the stripper top section.

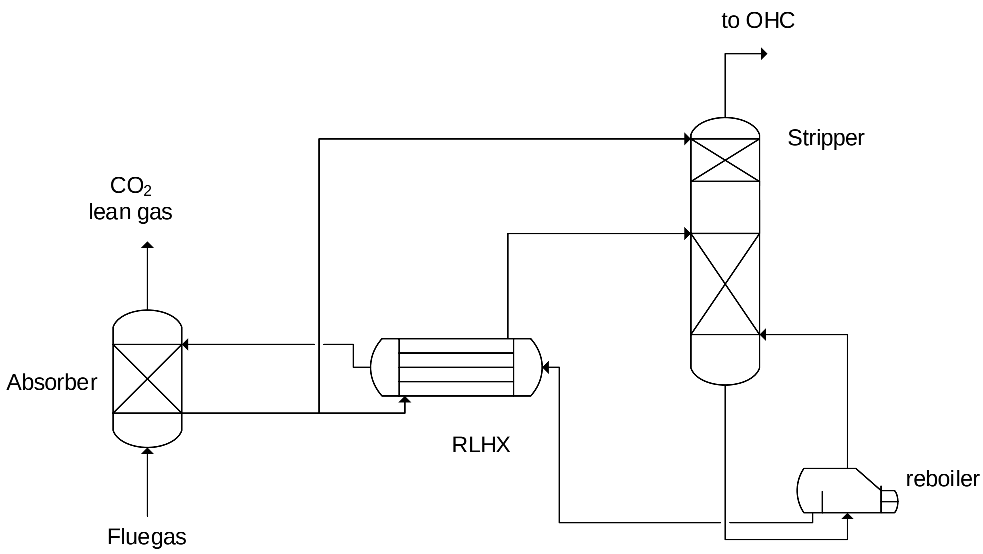

Figure 52: Schematic flow diagram of the split flow process by Eisenberg and Johnson [39]

Reducing the rich solution mass flow to the RLHX has a positive effect on the heat transfer. The mass flow of rich solution from the absorber is generally larger than the mass flow of lean solution from the stripper. This is due to the absorbed CO_2 in the rich solution. Thus, the heat capacity stream of the rich solution is higher as well. For a LMTD of 5 K in the RLHX, this results in a temperature difference of more than 5 K on the hot side of the RLHX, while the temperature difference at the cold side is less than 5 K. This can be seen in Figure 53, where the simplified temperature profile in the RLHX is shown for the base case. The temperature difference at the cold side of the RLHX is 47.5 °C-45.6 °C=1.9 K, the temperature difference at the hot side is 128 °C-117.6 °C=10.4 K. Reducing the rich solution mass flow to the RLHX reduces this imbalance and leads to higher temperatures of the rich solution downstream the RLHX.

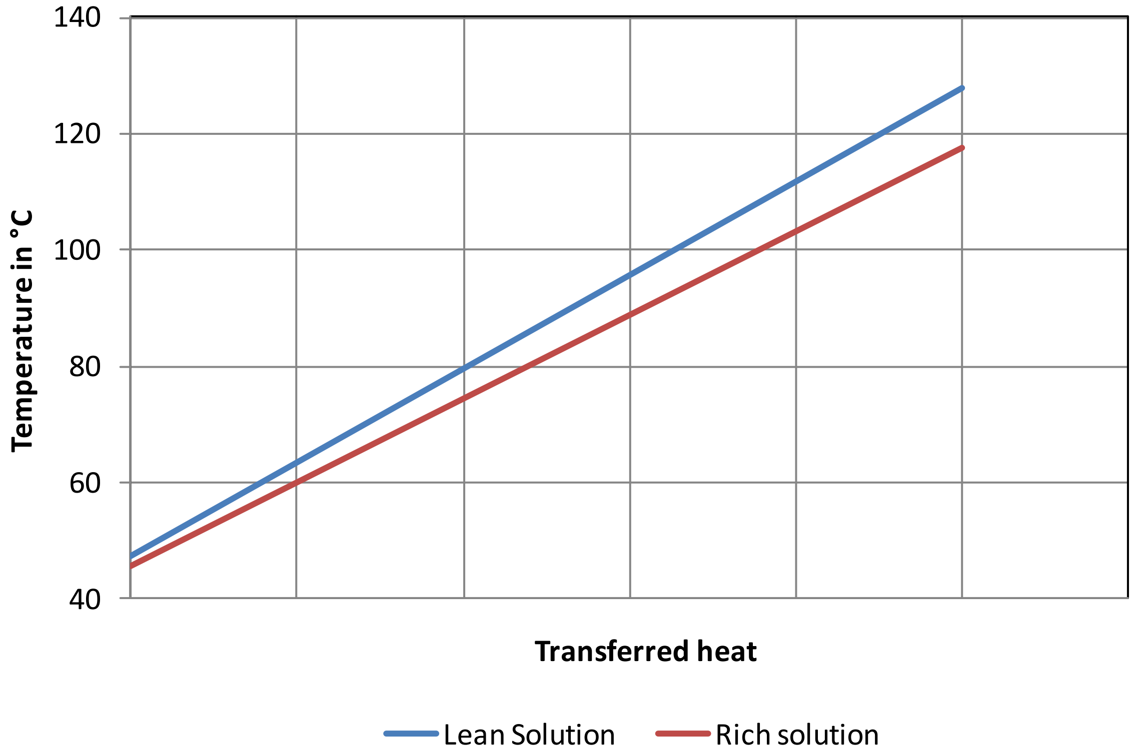

Figure 53: Simplified temperature profile in the RLHX for the base case

At the head of the stripper, the temperature is reduced significantly due to the cold rich solution that is fed to the top section. Thus, more steam is condensed in the stripper and the energy of vaporisation is kept in the stripper. In combination with the increased temperature of the rich solution downstream the RLHX, this results in a reduced reboiler duty. In addition, the cooling duty of the overhead condenser is reduced significantly. On the other hand, the cooling duty of the lean solution cooler is increased, since the temperature of the lean solution at the cold side of the RLHX is increased. This increase is expected to be smaller than the decrease in cooling duty at the OHC, since only latent heat is needed at the lean solution cooler.

6.7.2 SCPC power plant results - A5

For the split flow process described by Shoeld (cf. Figure 51) different split ratios were investigated. The split ratio is defined for the absorber and the stripper in the same way. It is the ratio between the split stream withdrawn from the absorber or the stripper, respectively, and the total solution mass flow at half height of the respective column. For a first evaluation, the split ratios for the absorber and the stripper are kept equal. The split ratios are varied between 0.1 (only a small split stream of semi lean solution is exchanged between the columns) and 1 (the solution is removed completely and led to the other column). The L/G is varied for every split ratio to reach the operating point with the lowest specific reboiler duty. The results are shown in Figure 54.

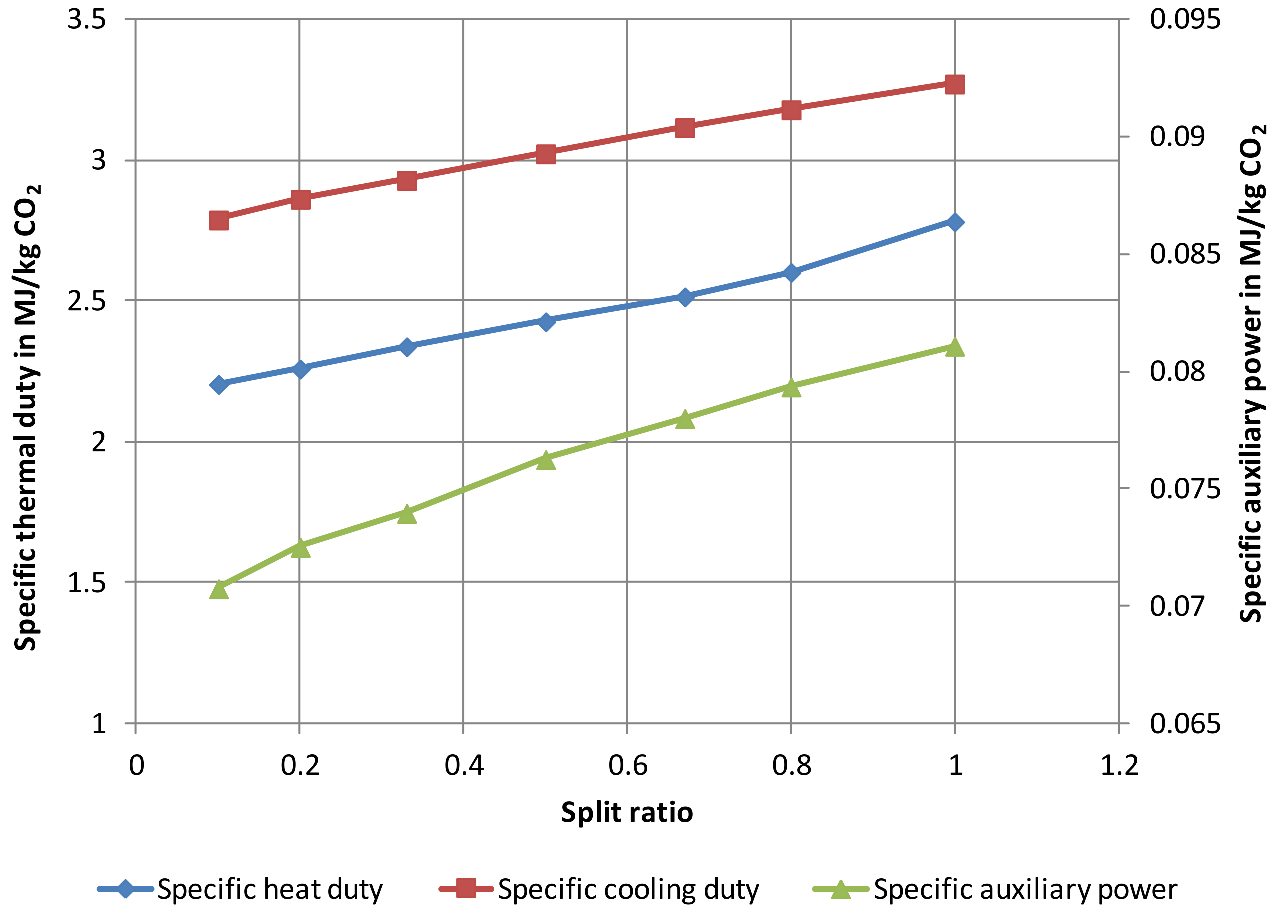

Figure 54: Specific thermal duty and specific auxiliary power of a capture plant with a split flow configuration suggested by Shoeld in combination with an SCPC plant

It can be seen from Figure 54 that the specific heat duty as well as the specific cooling duty and the specific auxiliary power increase with increasing split ratio. The specific heat duty increases from 2.21 MJ/kg to 2.78 MJ/kg. A comparison of these results with the performance of the base case (2.14 MJ/kg) shows an increase in specific heat duty for all operating points. For the specific cooling duty (base case: 2.73 MJ/kg) and the specific auxiliary power (base case: 0.069 MJ/kg) the same conclusion can be made.

The main reason for the increased specific heat duty can be found in the stripper. Since Solvent2020 is assumed to be very fast, the desorption of CO_2 takes place in large part at the bottom of the stripper where the stripping steam from the reboiler is introduced. The solution that is withdrawn at half the stripper height has to be heated to approx. reboiler temperature, while it is regenerated only partially. It has to be noted that the loading downstream the upper section and upstream the lower section are not equal, since the semi rich split stream from half the absorber height is added upstream the lower section. An overall process evaluation is not performed since it can be clearly seen from the specific thermal duty and specific auxiliary power of the capture plant that the effect on the overall process will be negative. The reboiler temperature is nearly constant for different split ratios and does not have a positive effect on the overall process either.

For the split flow process by Eisenberg and Johnson [39], a part of the rich solution is removed upstream the RLHX. The split stream is led to the top of the stripper, while the bulk of the rich solution is fed to the stripper 2 m below the top. The split ratio, the ratio between bypass mass flow and total mass flow of rich solution, is varied between 0.01 and 0.2. For each split ratio, L/G is varied to reach the operating point with the lowest specific heat duty. The results are shown in Figure 55. Note that the L/G variation for each split ratio is carried out for discrete values for L/G. This leads to sharp bends in the shape of the specific auxiliary power.

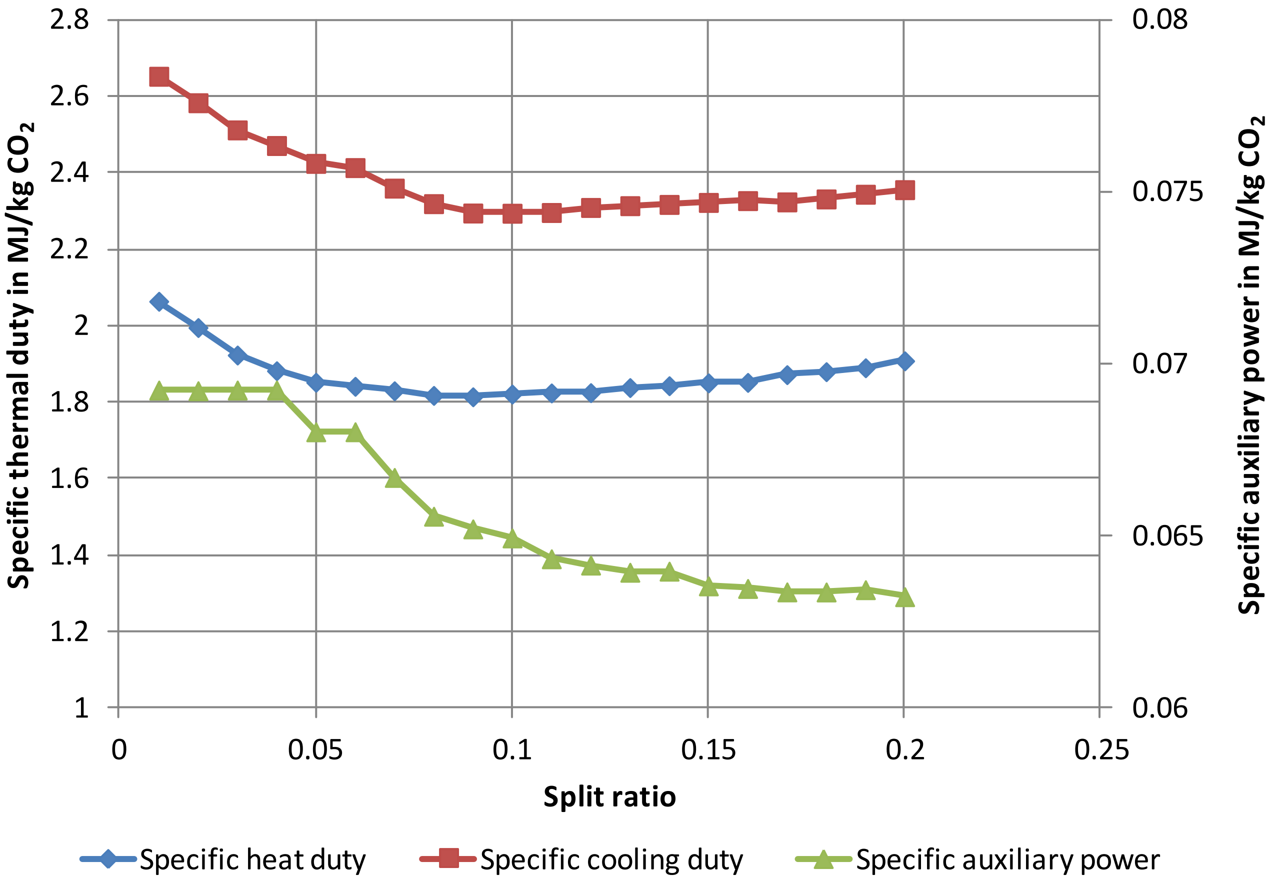

Figure 55: Specific thermal duty and specific auxiliary power of a capture plant with a split flow configuration suggested by Eisenberg and Johnson in combination with an SCPC plant (A5)

With increasing split ratio, the specific heat duty decreases until a minimum of 1.81 MJ/kg CO_2 is reached at a split ratio of 0.09. Compared to the base case, this is a reduction by 0.33 MJ/kg CO_2. For higher split ratios, the specific heat duty increases. The specific cooling duty has a similar trend as the specific heat duty. It is reduced by 0.45 MJ/kg CO_2 to 2.3 MJ/kg CO_2 for a split ratio of 0.09 and increases for higher split ratios. For the same operating point, the specific auxiliary power is reduced by 0.004 MJ/kg CO_2 to 0.065 MJ/kg CO_2. This is due to the fact that the operating point with the lowest heat duty for higher split ratios is reached with lower L/G. Thus, the power demand of the rich solution pump is reduced.

Table 29: Interface quantities of a capture plant in combination with an SCPC plant for base case (A1) and case with split flow configuration suggested by Eisenberg and Johnson (A5)

	SCPC base case	**Split flow with a split ratio of 0.09**
Specific heat duty in MJ/kg CO_2	2.14	1.81
Specific cooling duty in MJ/kg CO_2	2.75	2.30
Specific auxiliary power in MJ/kg CO_2	0.069	0.065
Desorber pressure in bar	5	5
Reboiler temperature in °C	128.0	138.4
Usable waste heat from OHC in MJ/kg CO_2	0.524	0.163
Temperature level of usable waste heat in °C	116.2	88.5

The interface quantities for the operating point with the lowest specific heat duty are given in Table 29. It can be seen that the reboiler temperature is increased from 128 °C for the base case to 138.4 °C. This is due to the reduction of solution mass flow from 2934 kg/s for the base case to 2196 kg/s. For higher split ratios, and thus lower solution mass flows, the reboiler temperature is increased even more. This can be seen in Figure 56, where the reboiler temperature is shown for different split ratios. Again, the discrete values for the evaluation of L/G can be seen. The first four split ratios, for example, are shown at the same L/G, followed by a lower L/G for split ratios 0.05 and 0.06.

As explained in section 6.7.1, the temperature at the stripper head is reduced significantly. This can be seen in Table 29 at the temperature level of usable waste heat, which is reduced by almost 30 °C. The lower temperature affects the usable waste heat as well, which is reduced by 68.9%.

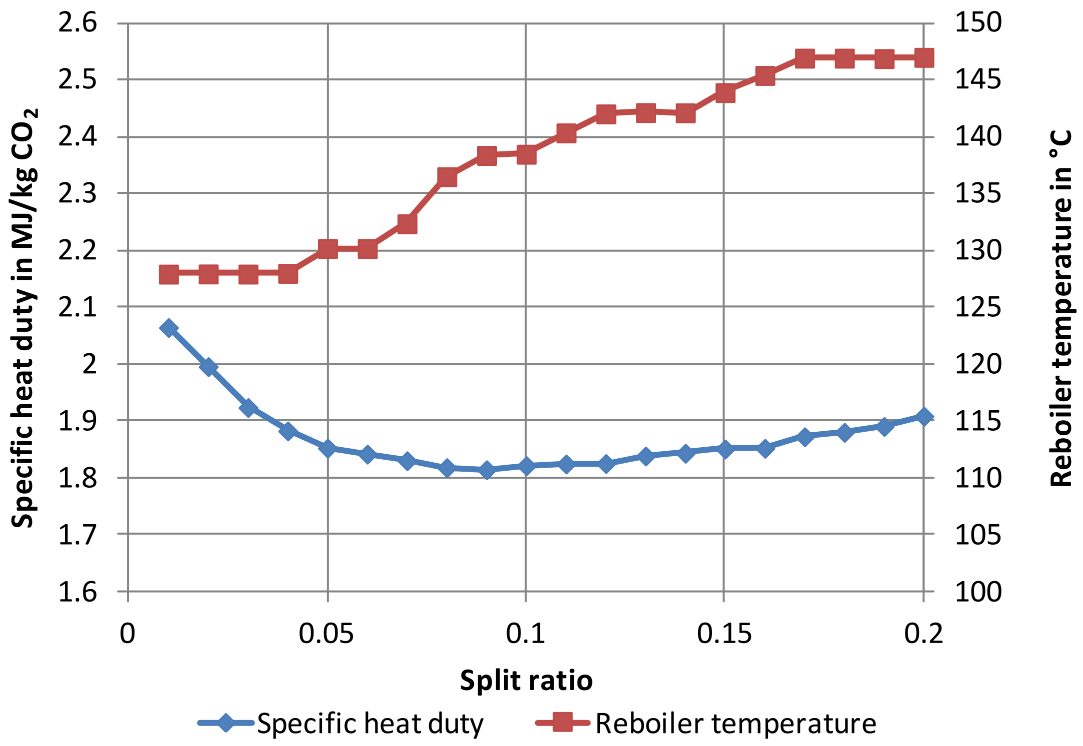

Figure 56: Specific heat duty and reboiler temperature of a capture plant with a split flow configuration suggested by Eisenberg and Johnson in combination with an SCPC plant (A5) for different split ratios

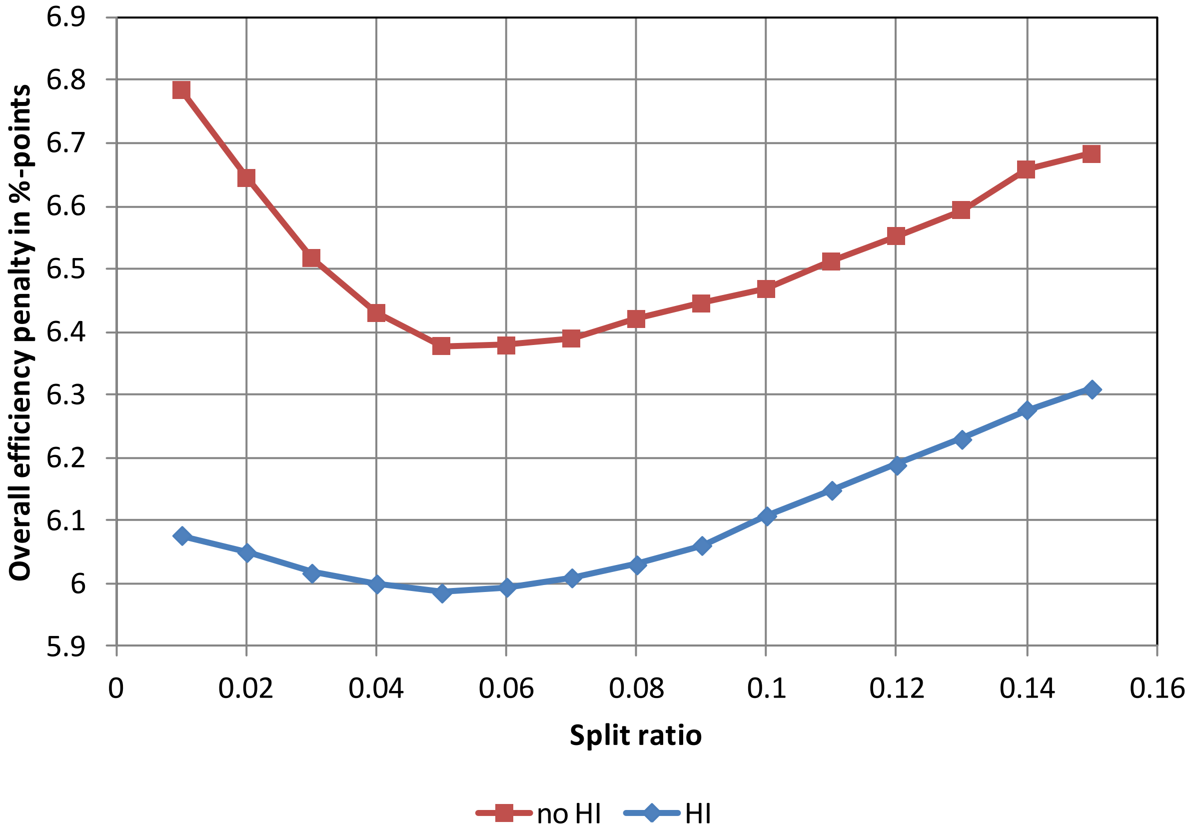

Figure 57: Overall efficiency penalty for a capture plant with a split flow configuration suggested by Eisenberg and Johnson [39] in combination with an SCPC plant (A5)

The overall efficiency penalty for different split ratios is shown in Figure 57. The values are given for the operating points with the lowest overall efficiency penalty. It can be seen that the lowest energy penalty of 6.38%-points for the case without heat integration is reached for a split ratio of 0.05. Compared to the base case, this is a reduction of 0.53%-points. This reduction is mainly caused by the reduced specific heat duty which decreases the penalty due to steam extraction by 0.49%-points. The penalty due to auxiliary power of the capture plant and additional cooling water pumps is decreased as well, as can be seen in Table 30.

Table 30: Contributors to the overall efficiency penalty of a capture plant in combination with an SCPC plant for base case (A1) and split flow configuration suggested by Eisenberg and Johnson (A5) without advanced waste heat integration

	Base case without HI	**Split flow with a split ratio of 0.05, w/o HI**	**Split flow with a split ratio of 0.09, w/o HI**
Steam extraction	4.16%-points	3.67%-points	3.77%-points
Compressor duty	1.90%-points	1.90%-points	1.90%-points
Cooling water pumps	0.23%-points	0.21%-points	0.20%-points
Auxiliary power	0.62%-points	0.60%-points	0.57%-points
Overall efficiency penalty	**6.91%-points**	**6.38**%-points	**6.44**%-points

Comparing Figure 57 with Figure 55, where the specific thermal duty and the specific auxiliary power of the capture plant are displayed, shows that all three energy duties are further reduced up to a split ratio of 0.09. Still, the overall efficiency penalty is higher for a split ratio of 0.09. This is due to the higher reboiler temperature (cf. Figure 56) which leads to a higher penalty due to steam extraction despite the reduced specific reboiler heat duty. The reduced penalties due to auxiliary power of the capture plant and the additional cooling water pumps do not compensate this increase.

Table 31: Contributors to the overall efficiency penalty of a capture plant in combination with an SCPC plant for base case (A1) and split flow configuration suggested by Eisenberg and Johnson (A5) with advanced waste heat integration

	Base case with HI	**Split flow with a split ratio of 0.05, with HI**	**Split flow with a split ratio of 0.09, with HI**
Steam extraction	4.21%-points	3.71%-points	3.77%-points
Compressor duty	2.06%-points	2.06%-points	2.06%-points
Cooling water pumps	0.21%-points	0.18%-points	0.18%-points
Auxiliary power	0.60%-points	0.59%-points	0.58%-points
Heat integration	-0.97%-points	-0.55%-points	-0.52%-points
Overall efficiency penalty	**6.11%-points**	**5.99%-points**	**6.38%-points**

For the case with advanced waste heat integration, the lowest overall efficiency penalty of 5.99%-points is reached for a split ratio of 0.05, too. Compared with the base case this is a reduction of 0.12%-points. The positive effect of the split flow modification is much smaller with waste heat integration. This is due to the reduced temperature at the stripper head and thus a lower temperature level of the integrated waste heat which reduces the positive effect of waste heat integration by 0.42%-points from 0.97%-points to 0.55%-points.

6.7.3 NGCC power plant results - B5

For the CO_2 capture plant equipped to an NGCC power plant solely the split flow process by Eisenberg and Johnson [39] is evaluated. Again, different bypass mass flows were tested by varying the split ratio of the splitter upstream the RLHX. For each split ratio L/G is varied to reach the operating point with the lowest specific heat duty. In Figure 58 the specific thermal duty and the specific auxiliary power of the capture plant are shown for different split ratios. The process flow sheet of a split flow case with a split ratio of 0.1 is shown in Figure 100.

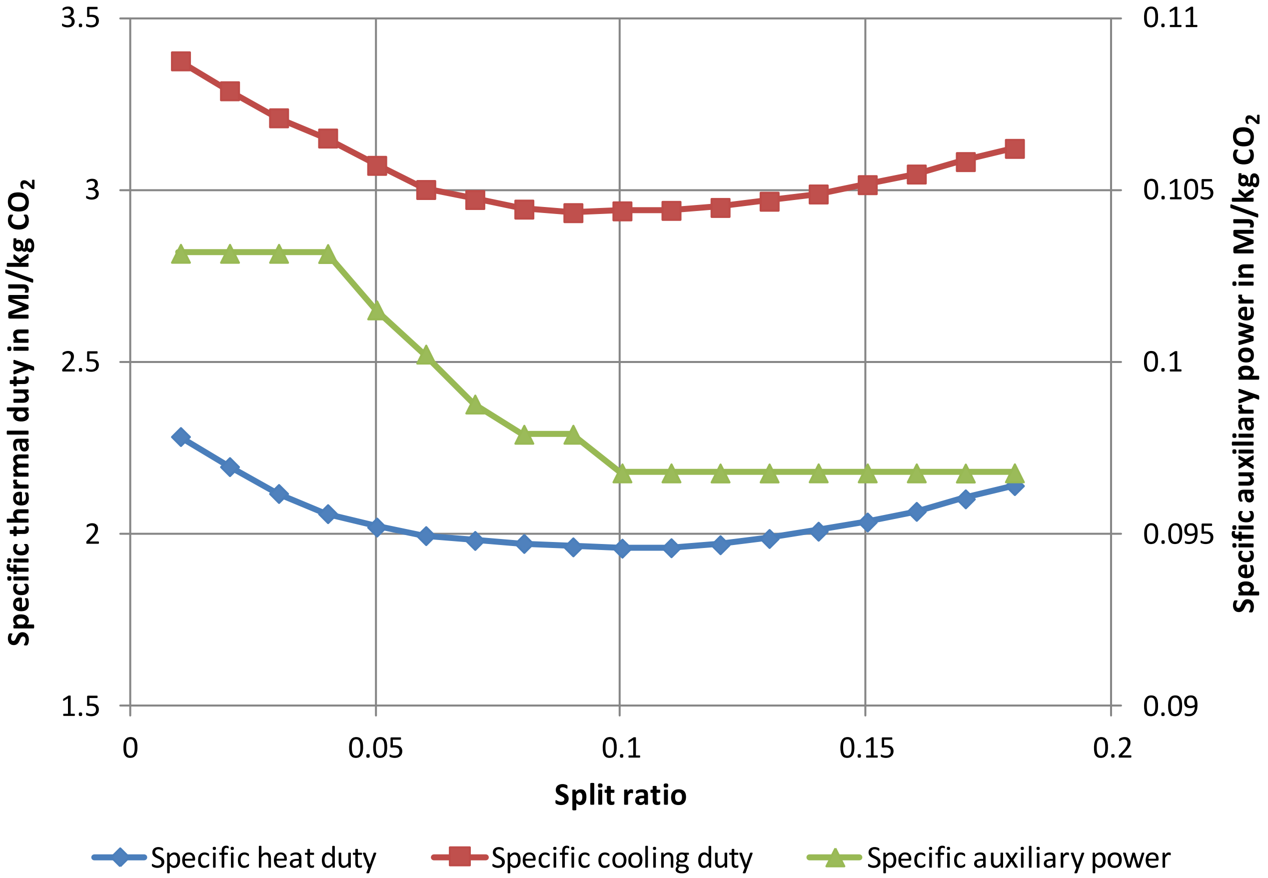

Figure 58: Specific thermal duty and specific auxiliary power of a capture plant with a split flow configuration suggested by Eisenberg and Johnson in combination with an NGCC plant (B5)

For increasing split ratio, the specific heat duty decreases until a minimum of 1.96 MJ/kg CO_2 is reached at a split ratio of 0.1. Compared to the base case, this is a reduction by 0.41 MJ/kg CO_2. For the same operating point, the specific cooling duty is reduced by 0.54 MJ/kg CO_2 from 3.48 to 2.94 MJ/kg CO_2. The specific auxiliary power is reduced as well, since the lowest heat duty is reached for lower L/G with increasing split ratios. At a split ratio of 0.1 the lowest possible L/G is reached, a further decrease would lead to reboiler temperature of more than 150 °C. The interface quantities for the NGCC base case as well as a split flow case with a split ratio of 0.1 are shown in Table 32.

Table 32: Interface quantities of a capture plant in combination with an NGCC plant for base case (B1) and case with split flow configuration suggested by Eisenberg and Johnson (B5)

	NGCC base case	Split flow with a split ratio of 0.1
Specific heat duty in MJ/kg CO_2	2.37	1.96
Specific cooling duty in MJ/kg CO_2	3.48	2.94
Specific auxiliary power in MJ/kg CO_2	0.10	0.10
Desorber pressure in bar	5	5
Reboiler temperature in °C	132.4	149.2
Flue gas temperature upstream of the capture plant in °C	109.8	110.8

The overall efficiency penalty for different split ratios is shown in Figure 59. The values are given for the operating points with the lowest overall efficiency penalty. It can be seen that the lowest overall efficiency penalty of 5.46%-points is reached for a split ratio of 0.06. Compared to the base case, this is a reduction by 0.47%-points. For higher split ratios, the overall efficiency penalty increases despite the decreasing specific heat duty, since the reboiler temperature increases.

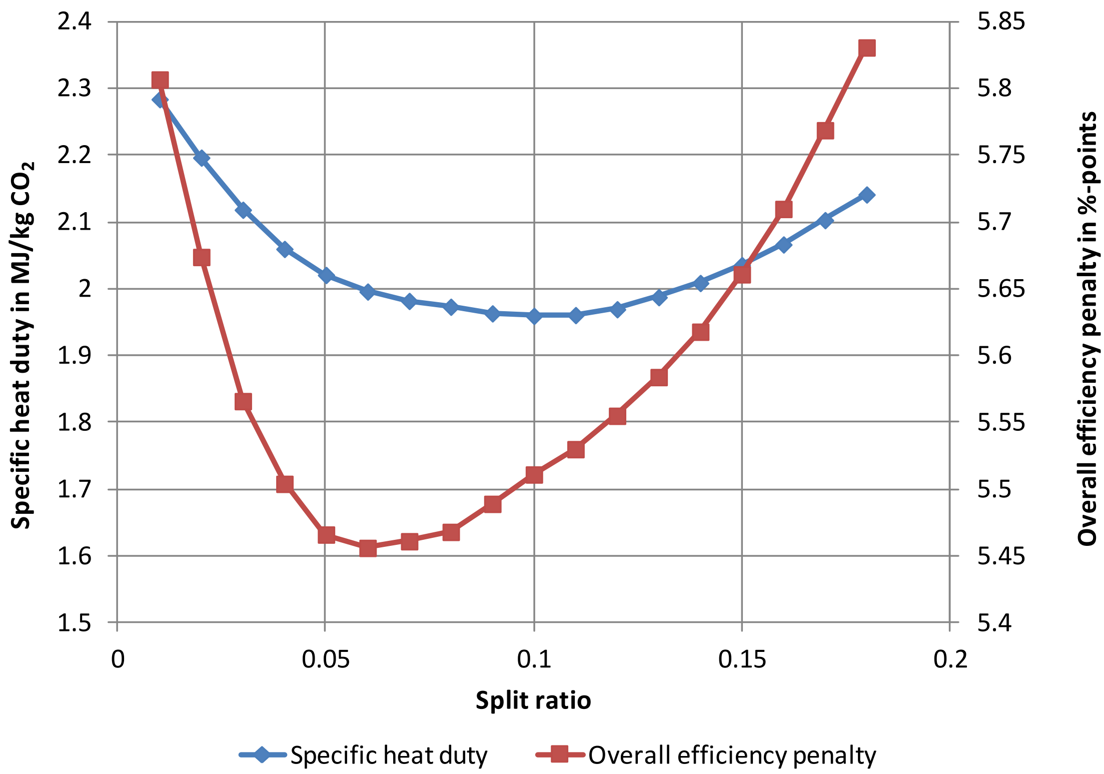

Figure 59: Specific heat duty and overall efficiency penalty for a capture plant with split flow configuration suggested by Eisenberg and Johnson in combination with an NGCC plant (B5)

Table 33: Contributors to the overall efficiency penalty of a capture plant in combination with an NGCC plant for base case (B1) and split flow configuration suggested by Eisenberg and Johnson (B5)

	NGCC base case	**Split flow with a split ratio of 0.06**	**Split flow with a split ratio of 0.1**
Steam extraction	3.45%-points	2.99%-points	3.06%-points
Compressor duty	1.20%-points	1.20%-points	1.20%-points
Cooling water pumps	0.12%-points	0.12%-points	0.12%-points
Auxiliary power	0.53%-points	0.52%-points	0.51%-points
Flue gas recirculation	0.62%-points	0.62%-points	0.62%-points
Overall efficiency penalty	**5.93%-points**	**5.46%-points**	**5.51%-points**

The reduction of the overall efficiency penalty is caused by the reduced penalty due to steam extraction, as can be seen in Table 33. All of the other contributors are changed only marginally, while the penalty due to steam extraction is reduced by 0.46%-points. For the case with the lowest specific heat duty, the case with a split ratio of 0.1, the penalty due to steam extraction is increased due to the increased reboiler temperature. The L/G is reduced leading to a further reduction of the auxiliary power of the capture plant.

6.8 Matrix stripping

6.8.1 Process Characteristics

Matrix stripping is one of the most complex flow sheet modifications investigated in this study. The stripper is divided into several pressure stages, which are fed by rich solvent. In various possible configurations, partial regenerated solution is extracted from different parts of the stripper columns and fed to columns at lower pressure. In this study, the configuration shown in Figure 60 is investigated. The solution is regenerated at three different pressure levels in the high pressure (HP) stripper, the intermediate pressure (IP) stripper and the low pressure (LP) stripper. The rich solution stream from the RLHX is split and fed to the heads of the stripper columns. Different distributions are evaluated. For the evaluation of the results the split ratio is defined as the ratio between a split stream to one of the strippers and the solution stream from the RLHX.

The partial regenerated solution at the bottom of the HP- and IP stripper is fed to the IP- and LP-stripper, respectively, at half height. The reboiler duties for the HP- and IP-stripper are adjusted to reach a CO_2 loading at the bottom equal to the loading in the lower pressure stripper where the solution is fed in. This leads to a minimum of required heat as well as similar temperatures in the reboilers. Thus, one common extraction for the steam from the power plant can be used. The lean solution from the LP-stripper bottom is led to the RLHX.

The effect of matrix stripping is similar to the effect of a multi pressure stripper, but without the auxiliary power of additional compressors in between the pressure sections. In the high pressure section, CO_2 is regenerated with a low specific heat duty, but the reboiler temperature is lower compared to a single high pressure stripper since the CO_2 loading at the sump does not need to be as low. In the low pressure section a low CO_2 loading of the lean solution is reached with lower reboiler temperatures compared to a single high pressure stripper. Altogether an increased specific heat duty and a decreased reboiler temperature are expected leading to a reduced overall efficiency penalty.

Another advantage of matrix stripping claimed in literature is a reduced power demand of the compression train. The CO_2 from each stripper column is send to a separate stage of the compressor minimising the compressor work since some of the CO_2 streams start the compression at higher pressures. In this study, this effect is assumed not to be usable. The pressure ratio between the different stripper sections is small compared to the assumed pressure ratio over one stage of the compressor (cf. section 5.3). Thus, an additional compressor would be necessary to overcome the pressure difference between the stripper sections. To reduce the complexity of the flow sheet, the vapour from all strippers is throttled to the pressure of the LP-stripper and merged. A positive side effect of this configuration is that only one overhead condenser is needed which reduces complexity even more.

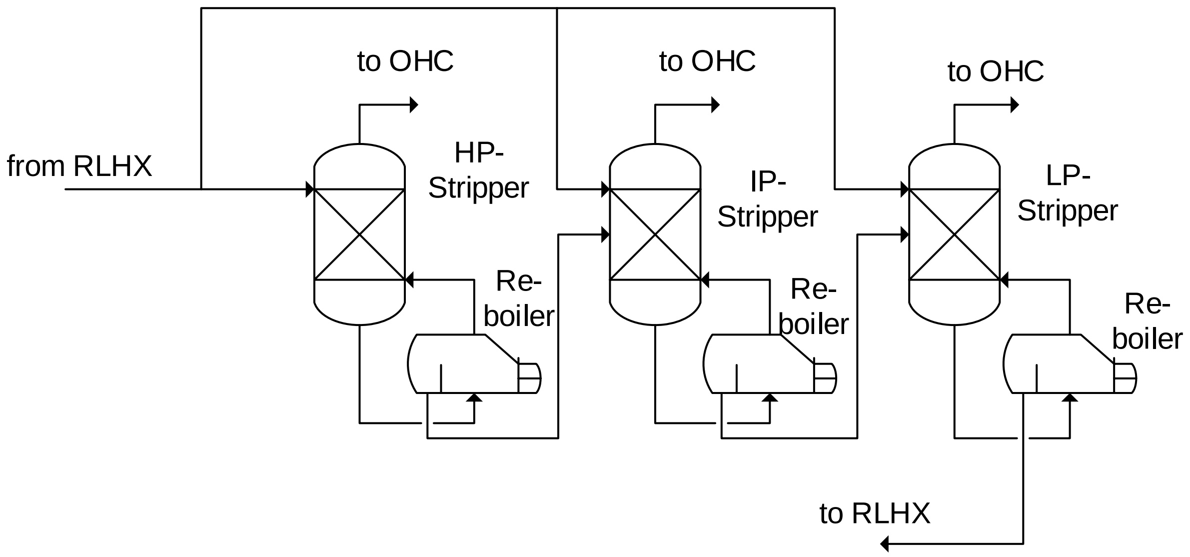

Figure 60: Schematic flow diagram of the stripper configuration for matrix stripping

6.8.2 SCPC power plant results - A6

For the simulations of the matrix stripping process, the pressure in the HP-stripper was fixed at 5 bar. This is done since an increased HP-stripper pressure would probably result in lower specific heat duties, but this reduction would not be due to the complex flowsheet modification but rather the increased stripper pressure. Using a higher stripper pressure for matrix stripping should thus be referenced to a case with the same high stripper pressure and is not considered in this study.

At first, the pressure in the IP-stripper was set to 4 bar, equivalent to a pressure ratio of 1.25 between HP- and IP-stripper. Between IP- and LP-stripper, the same pressure ratio is applied resulting in 3.2 bar LP pressure. The split ratios for the IP- and LP-stripper are in a first step chosen to be identical and varied between 0.05 and 0.4. A split ratio of 0.05 means that 5% of the rich solution mass flow are led to the IP- and LP-stripper, respectively, while 90% are led to the HP-stripper. A split ratio of 0.4 means that 40% of the rich solution are led to the IP- and LP-stripper, respectively, while 20% are led to the HP-stripper. For each split ratio L/G is varied to find the operating point with the lowest specific heat duty. The results are shown in Figure 61.

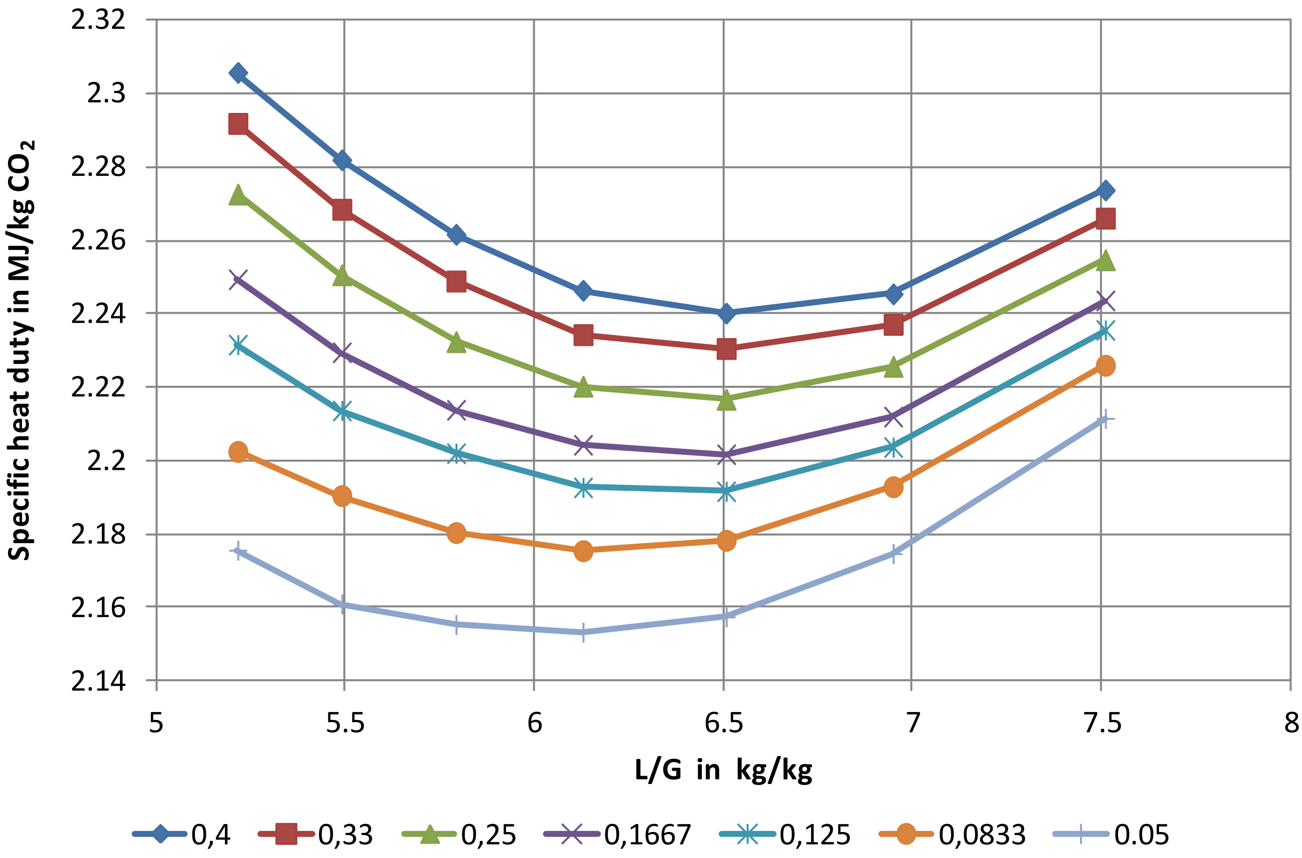

Figure 61: Specific heat duty of a capture plant with matrix stripping in combination with an SCPC plant (A6) for different split ratios to IP- and LP-stripper

It can be seen that the lowest specific heat duty is reached for a split ratio of 0.05. Still, this lowest specific heat duty of 2.15 MJ/kg CO_2 is higher compared to the specific heat duty of 2.14 MJ/kg CO_2 for the base case. Apparently, matrix stripping does not have a positive effect on the specific heat duty for Solvent2020. The more solution is fed to the lower pressure stripper, the higher the specific heat duty. This is due to the higher specific heat duty required for regeneration at low pressures (cf. section 6.1.2). A positive effect of matrix stripping is a reduced reboiler temperature. The temperature in the reboilers of the strippers with lower pressure is lower, as stated before (cf. section 6.1.2). The reboiler temperature of the HP-stripper is reduced as well, since the CO_2 loading of the solution does not need to be as low as for the base case. For the operating points with the lowest specific heat duty, the reboiler temperature is reduced from 128 °C for the base case to around 120 °C. Whether or not this results in a benefit for the overall process will be evaluated in the following section.

For the operating points with the lowest specific heat duty, the specific thermal duty and the specific auxiliary power are shown in Figure 62 for different split ratios. The specific cooling duty has a similar course as the specific heat duty. The auxiliary power is reduced for small split ratios since the lowest specific heat duty for small split ratios is reached for lower L/G (cf. Figure 61) and thus lower auxiliary power of the rich solution pump.

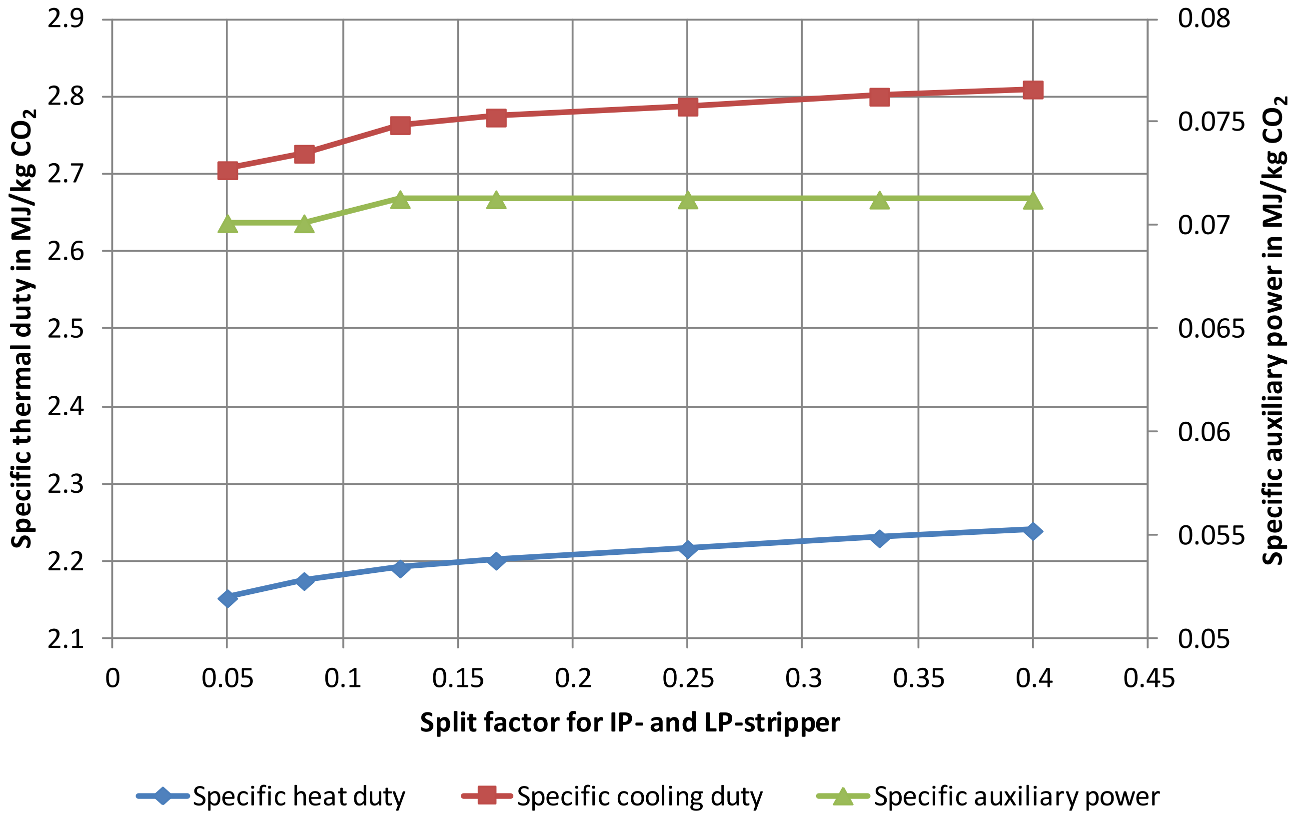

Figure 62: Specific thermal duty and specific auxiliary power of a capture plant with matrix stripping in combination with an SCPC plant (A6) for identical split factors for IP- and LP-stripper

In a next step, the split ratios for the IP- and LP-stripper are varied independently. One split ratio is fixed at 0.125, while the other is varied between 0.05 and 0.4. The results for the operating points with the lowest specific heat duty are shown in Figure 63. For comparison, the specific heat duties of the cases with identical split ratios for both strippers are shown as well. The effects that can be seen in the diagram are the same already seen in Figure 62. The less solution is led to the HP-stripper, the higher the specific heat duty. In addition, it can be seen that the influence of the solution led to the LP-stripper is higher compared to the solution led to the IP-stripper. A variation of the mass flow led to the IP-stripper (red curve) leads to a small change in specific heat duty, while a variation of the mass flow to the LP-stripper (green curve) has a higher impact.

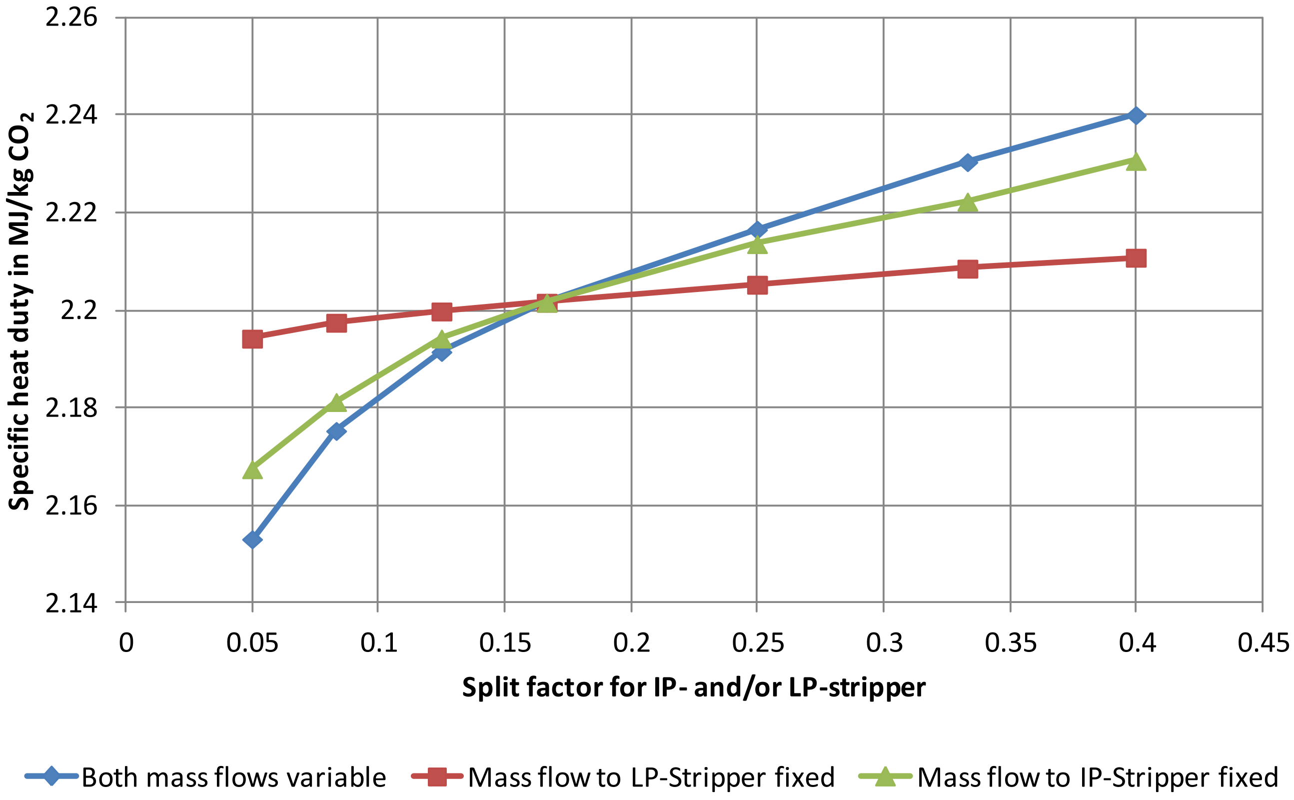

Figure 63: Specific heat duty of a capture plant with matrix stripping in combination with an SCPC plant (A6) for different split factors for IP- and LP-stripper

In a next step, the pressures in the IP- and LP-strippers are varied. The split ratio for all operating points is set to 0.125 for both, IP- and LP-stripper. Again, the pressure ratio between HP- and IP-stripper is chosen to be identical with the pressure ratio between IP- and LP-stripper. In addition to the pressure ratio of 1.25 already evaluated, pressure ratios of 1.11 (IP 4.5 bar, LP 4.05 bar) and 1.43 (IP 3.5 bar, LP 2.45 bar) are chosen. In Figure 64, the specific heat duty for different solution mass flows is shown. The lowest specific heat duty of 2.16 MJ/kg CO_2 is reached for a pressure ratio of 1.11. Still, the specific heat duty is higher compared to the base case. The reboiler temperature for the operating point with the lowest specific heat duty is reduced to 123 °C and is thus higher compared to the case with a pressure ratio of 1.25.

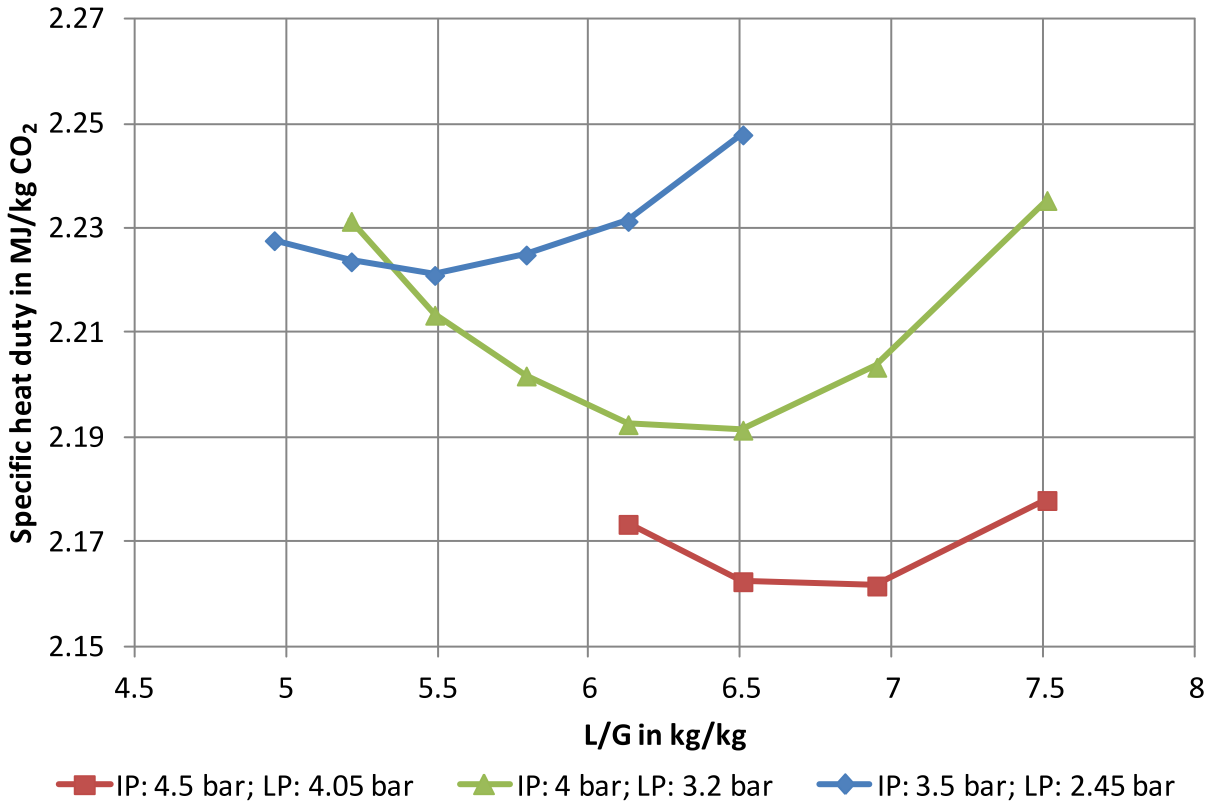

Figure 64: Specific heat duty of a capture plant with matrix stripping in combination with an SCPC plant (A6) for different pressure levels

The interface quantities for the base case as well as for the matrix stripping case with split ratios of 0.125 for IP- and LP-stripper and stripper pressures of 4.5 bar and 4.15 bar for the IP- respectively LP-stripper are shown in Table 34. It can be seen that the specific heat duty is slightly increased from 2.14 to 2.16 MJ/kg CO_2, while the specific cooling duty and the specific auxiliary power are not changed. The reboiler temperature is decreased from 128.0 to 122.9 °C. As mentioned before, the three strippers are equipped with separate reboilers with similar temperatures. In this operating point, the reboiler for the LP-stripper is operated at the highest temperature level of 122.9 °C. The steam required for the heating of the HP- and IP-stripper is throttled to a slightly lower pressure since the required temperature level for these reboilers is 121.7 °C respectively 120.4 °C. Since the temperature level in the strippers is reduced, the temperature and the amount of available waste heat in the overhead condenser are reduced as well. The overhead vapour from the three strippers is merged resulting in a combined temperature of 111.7 °C.

Table 34: Interface quantities of a capture plant in combination with an SCPC plant for base case (A1) and case with matrix stripping (A6)

	SCPC base case	Matrix stripping, IP: 4.5 bar, LP: 4.05 bar, split ratio IP and LP: 0.125
Specific heat duty in MJ/kg CO_2	2.14	2.16
Specific cooling duty in MJ/kg CO_2	2.75	2.75
Specific auxiliary power in MJ/kg CO_2	0.069	0.069
HP-stripper pressure in bar	5	5
IP-stripper pressure in bar	-	4.5
LP-stripper pressure in bar	-	4.05
Reboiler temperature in °C	128.0	122.9
Usable waste heat from OHC in MJ/kg CO_2	0.524	0.49
Temperature level of usable waste heat in °C	116.2	111.7

The overall efficiency penalty for different split ratios is shown in Figure 65. The values are given for the operating points with the lowest overall efficiency penalty and a pressure ratio of 1.11 between the strippers. It can be seen that the overall efficiency penalty is higher for all split ratios compared to the base case. With increasing split ratio, the overall efficiency penalty increases.

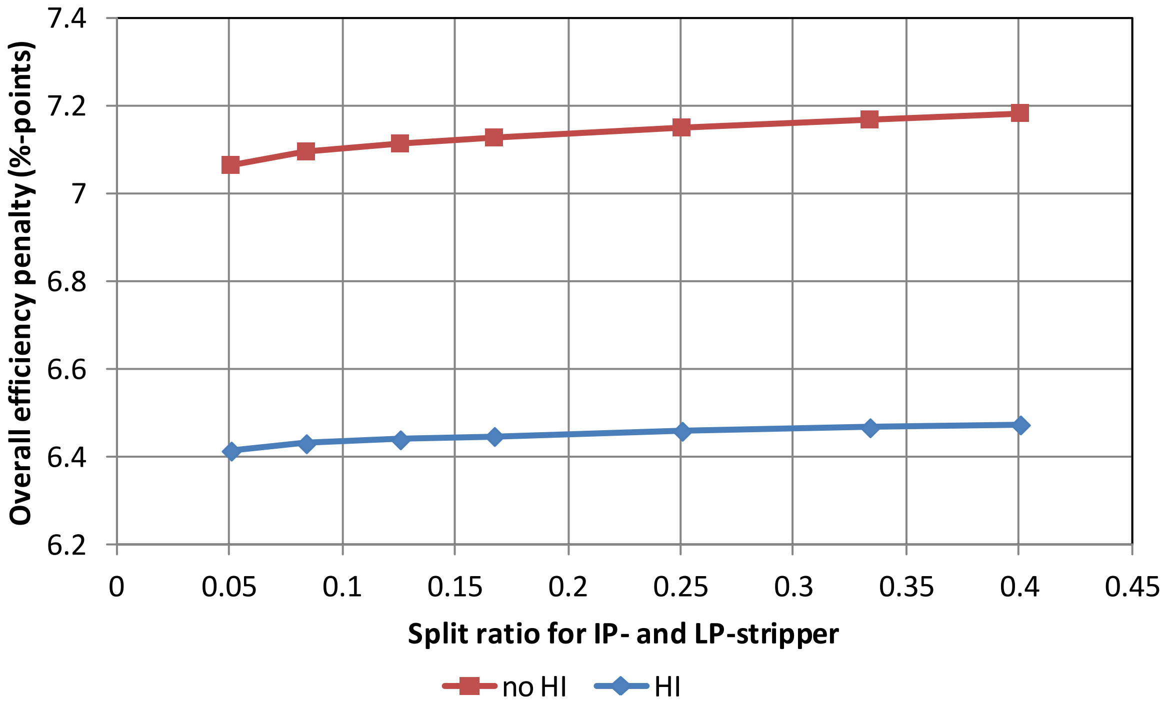

Figure 65: Overall efficiency penalty for a capture plant with matrix stripping in combination with an SCPC plant (A6) for identical split factors for IP- and LP-stripper

The different contributors to the overall efficiency penalty are shown in Table 35. It can be seen that the penalty due to steam extraction is reduced by 0.21%-points for the case without waste heat integration, since the reduced reboiler temperature overcompensates the increased specific heat duty. Still, the effect on the overall process is negative since the CO_2 compressor duty is increased by 0.36%-points. For the case with advanced waste heat integration, the penalty due to steam extraction is reduced by 0.26%-points, while the penalty due to the CO_2 compressor duty is increased by 0.41%-points. In addition, the positive effect of advanced waste heat integration is reduced, since the temperature level, as well as the amount of available waste heat in the overhead condenser is reduced.

The overall efficiency penalty of the matrix stripping process could be reduced by using three different CO_2 compressors for the CO_2 from the different columns. For the operating point with advanced heat integration shown in Table 35, this would reduce the penalty due to the compressor duty from 2.47 to 2.20%-points and the overall efficiency penalty from 6.41 to 6.14%-points. So, the complexity is increased even more, but the overall efficiency penalty is still higher compared to the base case.

In summary it can be stated, that the overall efficiency penalty for this modification is always higher than the overall efficiency penalty for the base case which achieves a lowest overall efficiency penalty of 6.11%-points.

Table 35: Contributors to the overall efficiency penalty of a capture plant in combination with an SCPC plant for base case (A1) and case with matrix stripping (A6) with and without advanced waste heat integration

	SCPC base case without HI	**Matrix stripping, split ratio: 0.05, w/o HI**	**SCPC base case with HI**	**Matrix stripping, split ratio: 0.05, with HI**
Steam extraction	4.16%-points	3.95%-points	4.21%-points	3.95%-points
Compressor duty	1.90%-points	2.26%-points	2.06%-points	2.47%-points
Cooling water pumps	0.23%-points	0.23%-points	0.21%-points	0.21%-points
Auxiliary power	0.62%-points	0.62%-points	0.60%-points	0.60%-points
Heat integration			-0.97%-points	-0.83%-points
Overall efficiency penalty	**6.91%-points**	**7.07%-points**	**6.11%-points**	**6.41%-points**

6.8.3 NGCC power plant results - B6

For the CO_2 capture plant with matrix stripping equipped to an NGCC power plant similar configurations as for the coal case are evaluated. First, the split ratio for IP- and LP-stripper is identical, followed by a variation of only one split ratio. Afterwards, the pressure levels in the strippers are varied. In Figure 66, the specific heat duties of a capture plant in combination with an NGCC plant are shown for different split ratios. The IP-stripper pressure is 4 bar and the LP-stripper pressure is 3.2 bar. As for the coal case, the specific heat duty is higher for all operating points compared to the base case. Again, the increase due to the split ratio for the LP-stripper is higher than the increase due to the IP-stripper split ratio.

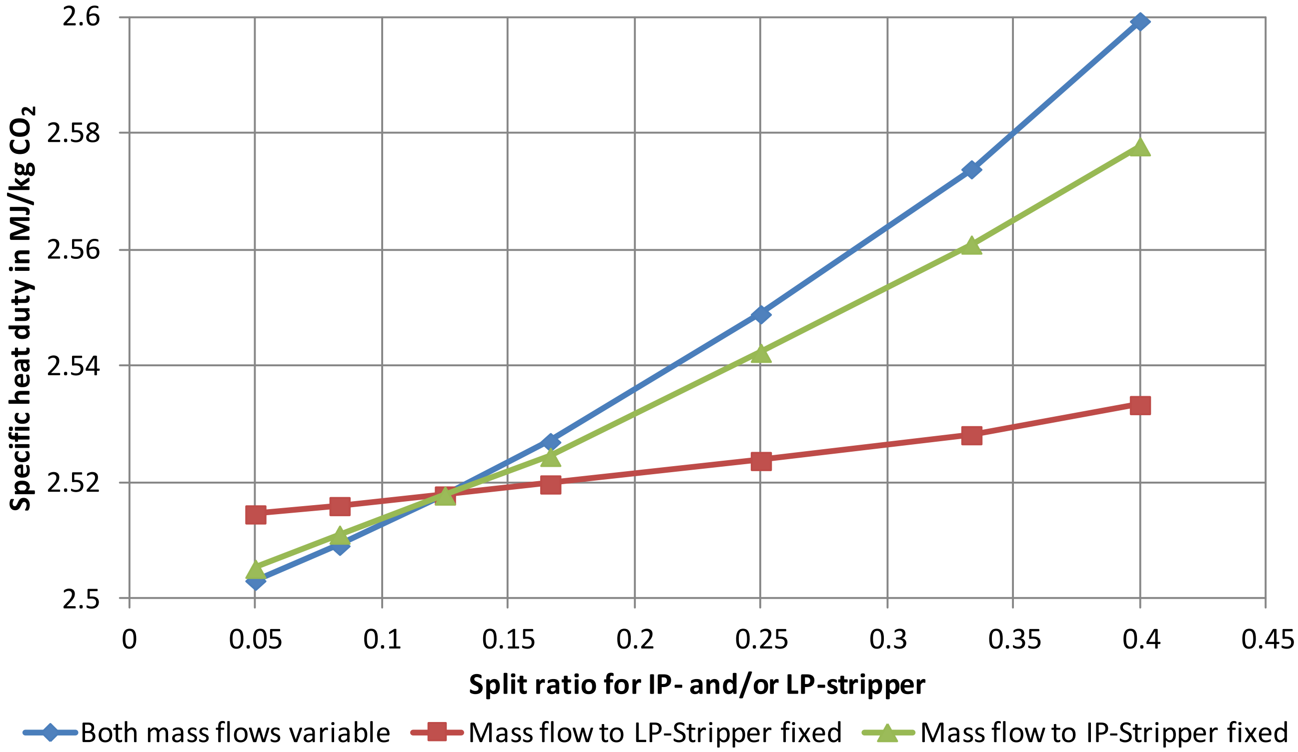

Figure 66: Specific heat duty of a capture plant with matrix stripping in combination with an NGCC plant (B6) for different split factors for IP- and LP-stripper

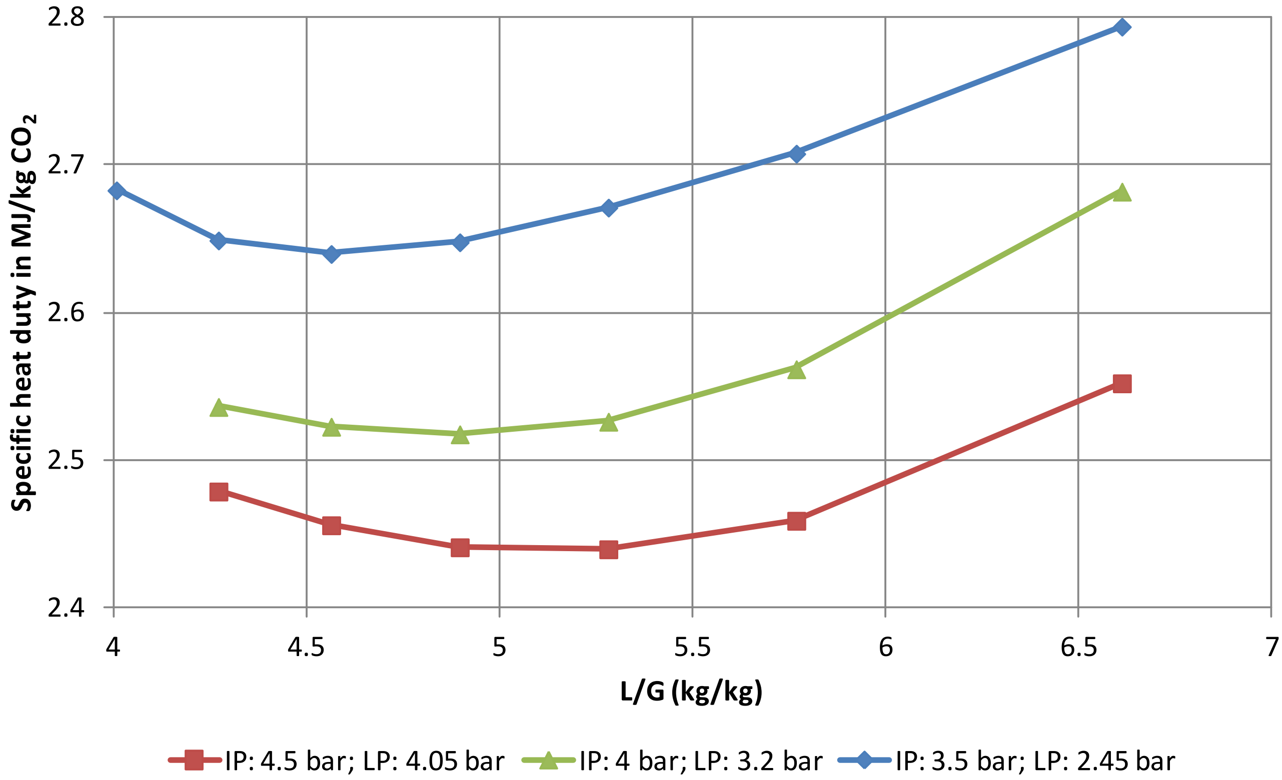

Figure 67: Specific heat duty of a capture plant with matrix stripping in combination with an NGCC plant (B6) for different pressure levels

The results of the variation of the stripper pressure are shown in Figure 67. It can be seen that the specific heat duty decreases with increasing pressure level, but is still higher compared to the base case. For lower stripper pressures the lowest specific heat duty is reached for lower solution mass flows, as for the coal case. For the operating point with the lowest specific heat duty, the interface quantities are shown in Table 36. The specific heat duty is increased by 0.07 MJ/kg CO_2, while the reboiler temperature is reduced by 3.3 °C. The reduced flue gas temperature is due to the reduced reboiler temperature. The temperature of the condensate coming from the reboiler is lower, and thus more heat can be removed from the flue gas.

Table 36: Interface quantities of a capture plant in combination with an NGCC plant for base case (B1) and case with matrix stripping (B6)

	NGCC base case	Matrix stripping, IP: 4.5 bar, LP: 4.05 bar, split ratio IP and LP: 0.125
Specific heat duty in MJ/kg CO_2	2.37	2.44
Specific cooling duty in MJ/kg CO_2	3.48	3.52
Specific auxiliary power in MJ/kg CO_2	0.10	0.10
HP-stripper pressure in bar	5	5
IP-stripper pressure in bar	-	4.5
LP-stripper pressure in bar	-	4.05
Reboiler temperature in °C	132.4	129.1
Flue gas temperature upstream of the capture plant in °C	109.8	108.7

The resulting overall efficiency penalty for the NGCC case is shown in Figure 68 for different stripper pressures and different split ratios. It can be seen that lower pressure ratios in the stripper lead to significantly lower overall efficiency penalties. As for the coal case, lower split ratios lead to lower overall efficiency penalties, as well. But still the efficiency penalty for all operating points is higher compared to the base case.

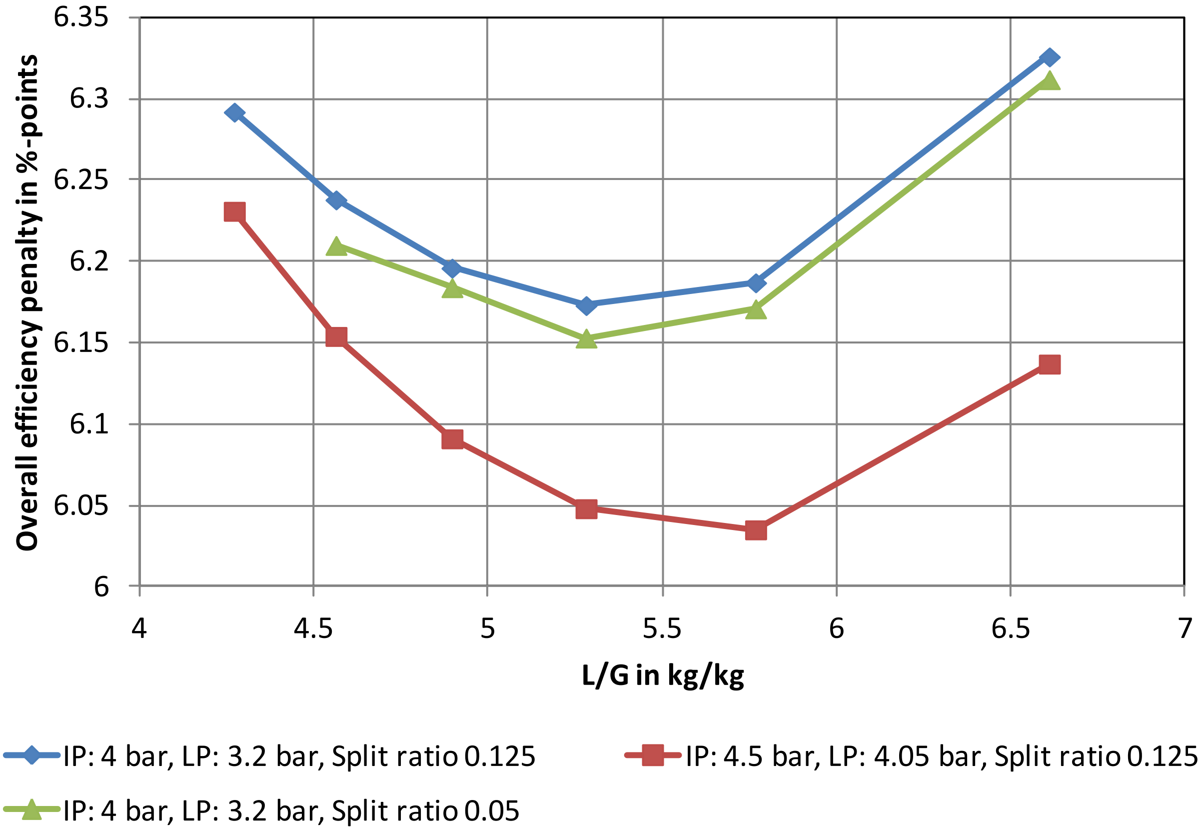

Figure 68: Overall efficiency penalty for a capture plant with matrix stripping in combination with an NGCC plant (B6) for different operating conditions

The different contributors to the overall efficiency penalty are shown in Table 37. It can be seen that the penalty due to steam extraction is reduced by 0.03%-points which is less reduction compared to the coal case. This is due to the lower reduction in reboiler temperature for the NGCC case (cf. Table 34 and Table 36). The increase in penalty due to the compressor duty is smaller, too, since less CO_2 has to be compressed compared to the coal case. All in all, matrix stripping does not have a positive effect on the overall process.

In summary it can be stated, that the overall efficiency penalty for this modification is always higher than the overall efficiency penalty for the base case which achieves a lowest overall efficiency penalty of 6.11%-points.

Table 37: Contributors to the overall efficiency penalty of a capture plant in combination with an NGCC plant for base case (B1) and case with matrix stripping (B6)

	NGCC base case	**Matrix stripping, IP: 4.5 bar, LP: 4.05 bar, split ratio IP and LP: 0.125**
Steam extraction	3.45%-points	3.42%-points
Compressor duty	1.20%-points	1.31%-points
Cooling water pumps	0.12%-points	0.13%-points
Auxiliary power	0.53%-points	0.55%-points
Flue gas recirculation	0.62%-points	0.62%-points
Overall efficiency penalty	**5.93%-points**	**6.04%-points**

6.9 Various heat integration options - overhead condenser

6.9.1 Process Characteristics

There are many different process modifications proposed in literature where heat is integrated in different parts of the process. One possible modification is to use heat from the flue gas upstream the capture plant to increase the temperature of a semi lean solution stream extracted from the stripper with a claimed reduction of reboiler duty by 6.7% [40]. In another modification, heat from the overhead condenser is used to heat up the rich solution upstream the stripper [41]. A reduction of the reboiler duty by 30% is claimed for this modification. Other possible heat sources are reboiler condensate or hot flue gas downstream of the economiser.

In this study, two different heat integration concepts are evaluated. First, heat from the overhead condenser is used to heat up the rich solution. In a second modification, the reboiler condensate is used to heat up semi lean solution which is extracted from the stripper.

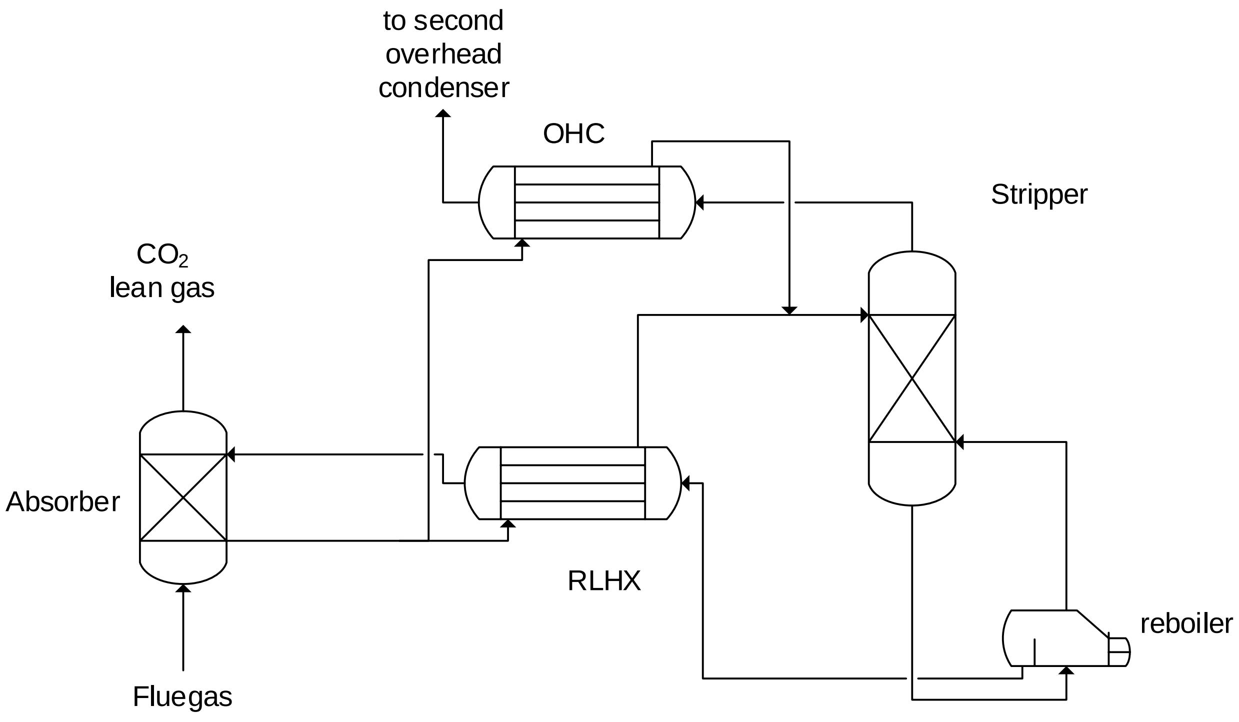

Figure 69: Schematic flow diagram of the overhead condenser heat integration

The integration of heat from the overhead condenser is shown in Figure 69. A fraction of the rich solution from the absorber bypasses the rich-lean heat exchanger and is lead to the overhead condenser. There, it is heated up with the overhead vapour from the stripper. The heated rich solution from the OHC is merged with the hot rich solution stream from the RLHX. Therefore, the sensible heat as well as the latent heat from the overhead vapour is used to heat up the stripper feed stream and less heat is needed in the reboiler. For the basic coal case and for the NGCC case, this heat would otherwise be lost to the cooling water. Thus, a reduced specific heat duty is expected. An LMTD of 5 K is assumed for the OHC. Downstream the OHC, there is a second OHC where the overhead vapour is cooled down to the same temperature as in the base case. This is done to ensure that the temperature of the CO_2 upstream the compressor is the same for both cases.

6.9.2 SCPC power plant results - A7

The results for the integration of heat from the OHC are shown in Figure 70. It can be seen that the specific heat duty decreases with decreasing solution mass flow. The lowest specific heat duty of 1.71 MJ/kg CO_2 is reached at an L/G of 5.2 which is significantly lower compared to the base case, where the lowest specific heat duty of 2.14 MJ/kg CO_2 is reached for an L/G of 6.9. Due to the additional heat from the OHC, the stripper inlet temperature is increased and less heat is needed in the reboiler. The specific cooling duty is reduced by 0.56 MJ/kg CO_2. The reduction of specific cooling duty is larger compared to the reduction of specific heat duty, since the specific cooling duty is reduced by two effects. On the one hand, the heat which is brought into the process is reduced due to the lower reboiler duty. On the other hand, the cooling of the OHC is reduced significantly since a large fraction of the heat is transferred to the rich solution. Only a small heat flow has to be transferred to the cooling water to ensure a temperature of 40 °C downstream the OHC.

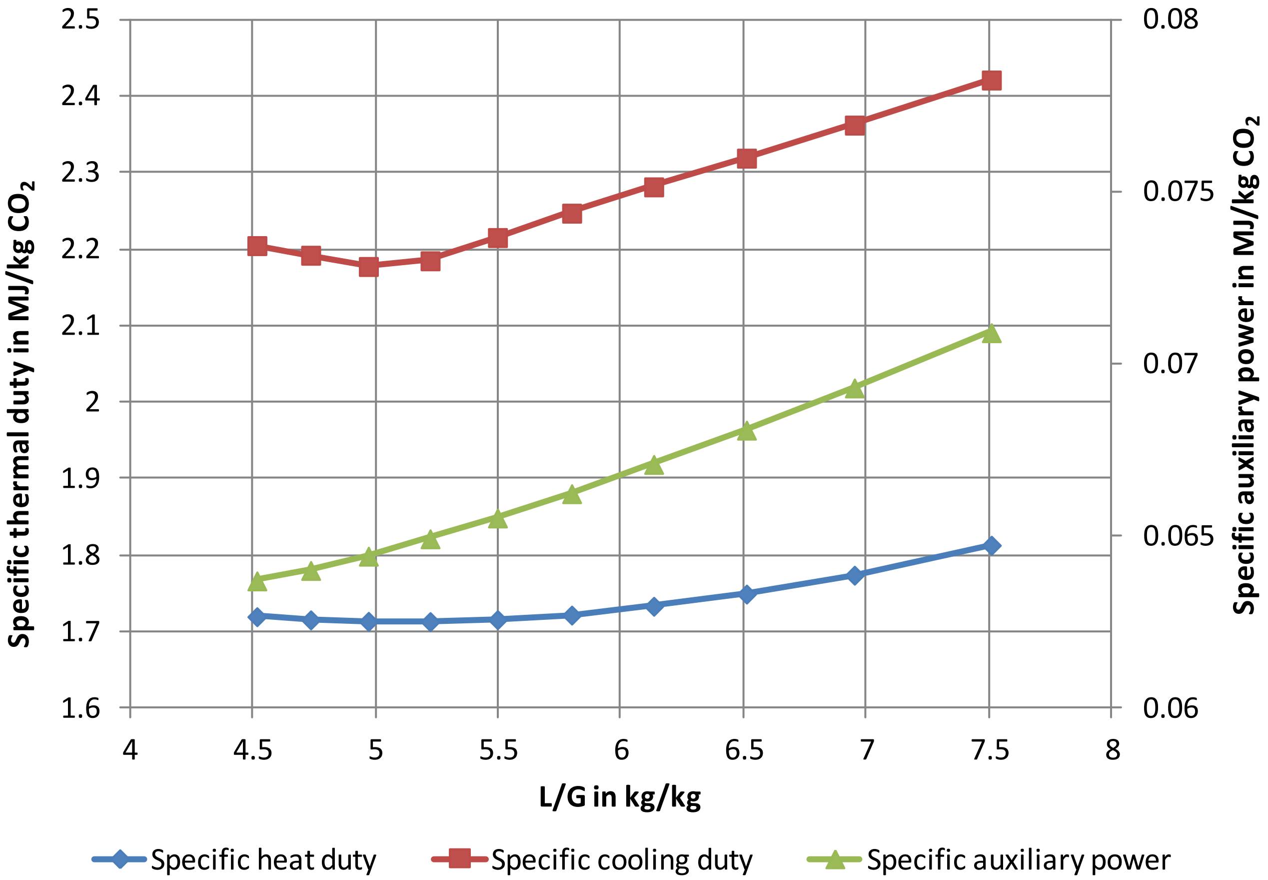

Figure 70: Specific thermal duty and specific auxiliary power of a capture plant with OHC heat integration in combination with an SCPC plant (A7)

The interface quantities for the OHC heat integration case with the lowest specific heat duty as well as for the base case are shown in Table 38. In addition to the reduction of specific heat and cooling duty, the specific auxiliary power is reduced as well, since the duty of the solution pump is reduced. Due to the lower L/G a lower lean loading is needed, which results in an increase of the reboiler temperature by 10.6 °C. Since a large fraction of the heat duty of the OHC is used for heating up the rich solution, only a small amount of waste heat would be available for integration into the power plant. Since the heat is furthermore available at a very low temperature, the integration is not practical any more. Thus, the advanced waste heat integration for this modification includes only the heat from the CO_2 compressor. The residual heat from the OHC is transferred to the cooling water.

Table 38: Interface quantities of a capture plant in combination with an SCPC plant for base case (A1) and case with overhead condenser heat integration (A7)

	SCPC base case	OHC heat integration
Specific heat duty in MJ/kg CO_2	2.14	1.71
Specific cooling duty in MJ/kg CO_2	2.75	2.19
Specific auxiliary power in MJ/kg CO_2	0.069	0.065
Desorber pressure in bar	5	5
Reboiler temperature in °C	128.0	138.6
Usable waste heat from OHC in MJ/kg CO_2	0.524	0.034
Temperature level of usable waste heat in °C	116.2	50.0

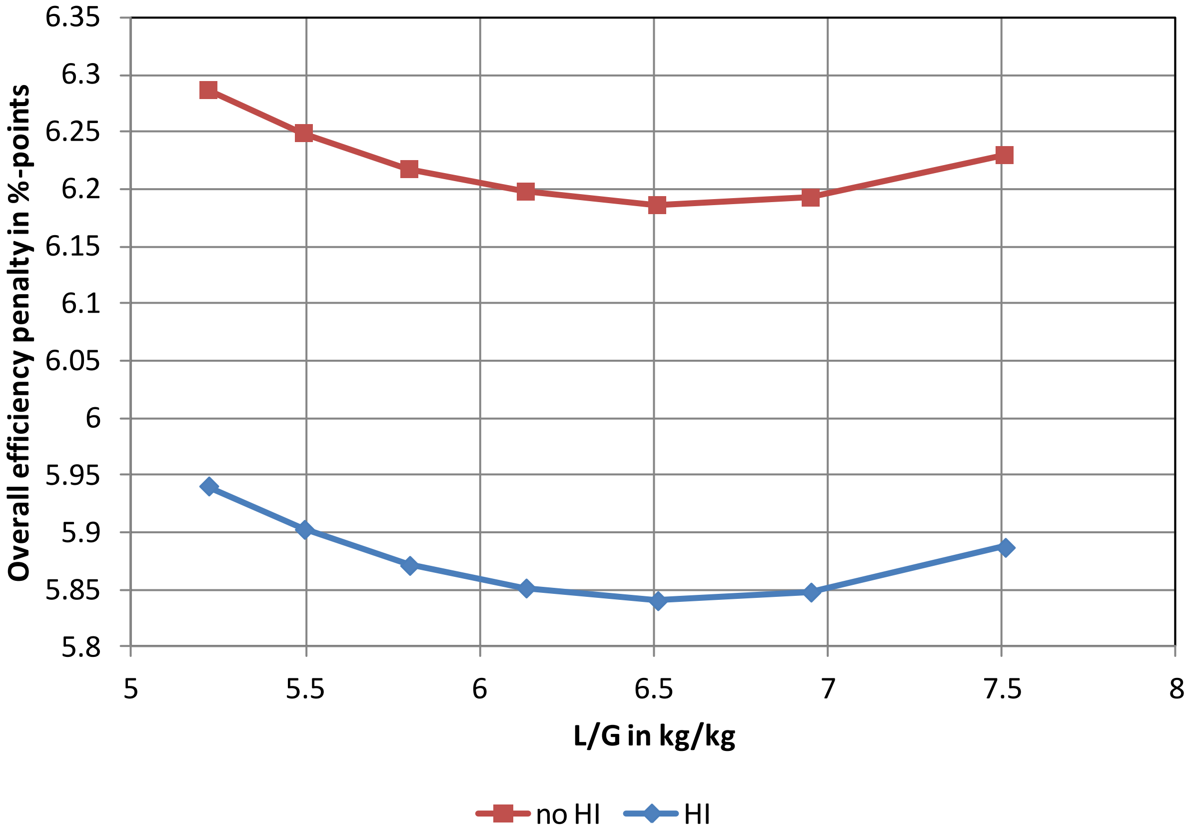

Figure 71: Overall efficiency penalty for a capture plant with overhead condenser heat integration in combination with an SCPC plant (A7)

The effect on the overall process can be seen in Figure 71 where the overall efficiency penalty is shown for different solution mass flows. The lowest overall efficiency penalty is reached with an L/G of 6.5 for the case without heat integration as well as for the case with heat integration. Compared to the base case, the overall efficiency penalty is reduced by 0.72%-points for the case without heat integration. This is due to the lower specific heat duty resulting in a reduced penalty due to steam extraction. The penalties for auxiliary power and cooling duty are reduced, too.

For the case with heat integration into the power plant, the overall efficiency penalty is reduced by 0.27%-points. The penalty due to steam extraction is reduced even further compared to the case without heat integration, but the positive effect of heat integration is reduced since the heat from the OHC is not available anymore. Altogether, the reduction of the overall efficiency penalty is smaller, but there is still a positive effect of this modification.

Table 39: Contributors to the overall efficiency penalty of a capture plant in combination with an SCPC plant for base case (A1) and case with overhead condenser heat integration (A7)

	SCPC base case without HI	**OHC heat integration, w/o HI**	**SCPC base case with HI**	**OHC heat integration, with HI**
Steam extraction	4.16%-points	3.49%-points	4.21%-points	3.49%-points
Compressor duty	1.90%-points	1.90%-points	2.06%-points	2.06%-points
Cooling water pumps	0.23%-points	0.20%-points	0.21%-points	0.17%-points
Auxiliary power	0.62%-points	0.59%-points	0.60%-points	0.59%-points
Heat integration			-0.97%-points	-0.48%-points
Overall efficiency penalty	**6.91%-points**	**6.19%-points**	**6.11%-points**	**5.84%-points**

6.9.3 NGCC power plant results - B7a

The results for the integration of heat from the OHC into the capture plant for the NGCC case are shown in Figure 72. As for the coal case, the L/G of the operating point with the lowest specific heat duty is much smaller compared to the base case. The lowest specific heat duty of 1.83 MJ/kg CO_2 is reached with an L/G of 3.4, while the base case has an L/G of 5.3. This is a reduction by 0.54 MJ/kg CO_2. The specific cooling duty for the same operating point is reduced by 0.64 MJ/kg CO_2. The lowest specific cooling duty is reached at a higher L/G. This is due to the increased cooling duty for the water wash which is required for operating points with L/G below 4.3. For these operating points, a higher water mass flow in the water wash is required to ensure a stable water balance. Again, the reduced solution mass flow at low L/G leads to a decreased specific auxiliary power.

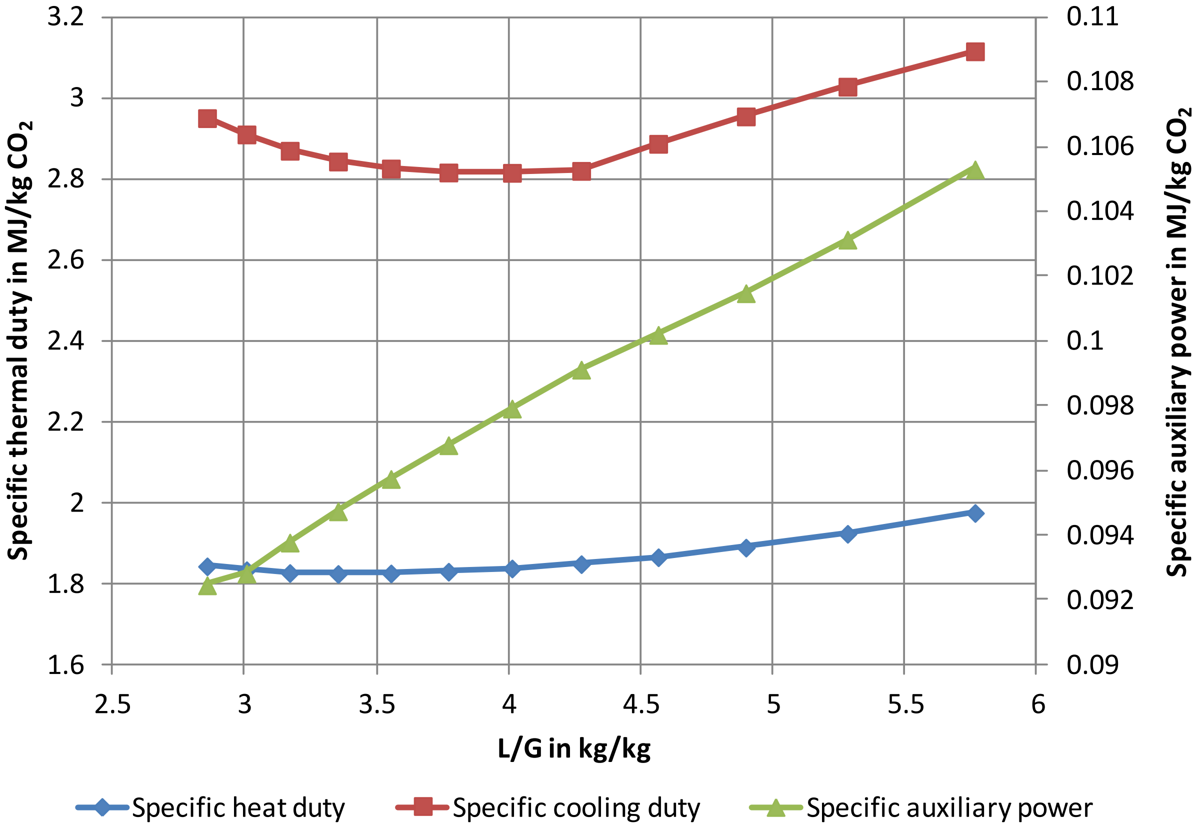

Figure 72: Specific thermal duty and specific auxiliary power of a capture plant with overhead condenser heat integration in combination with an NGCC plant (B7a)

The interface quantities for the NGCC case with OHC heat integration with the lowest specific heat duty as well as for the base case are shown in Table 40. Due to the lower L/G a lower lean loading is needed, which results in an increase of the reboiler temperature by 13.2 °C. Due to the higher temperature of the reboiler condensate, less heat can be removed from the flue gas and the flue gas temperature is slightly increased.

Table 40: Interface quantities of a capture plant in combination with an NGCC plant for base case (B1) and case with overhead condenser heat integration (B7a)

	NGCC base case	OHC heat integration
Specific heat duty in MJ/kg CO_2	2.37	1.83
Specific cooling duty in MJ/kg CO_2	3.48	2.84
Specific auxiliary power in MJ/kg CO_2	0.10	0.09
Desorber pressure in bar	5	5
Reboiler temperature in °C	132.4	145.6
Flue gas temperature upstream of the capture plant in °C	109.8	111.1

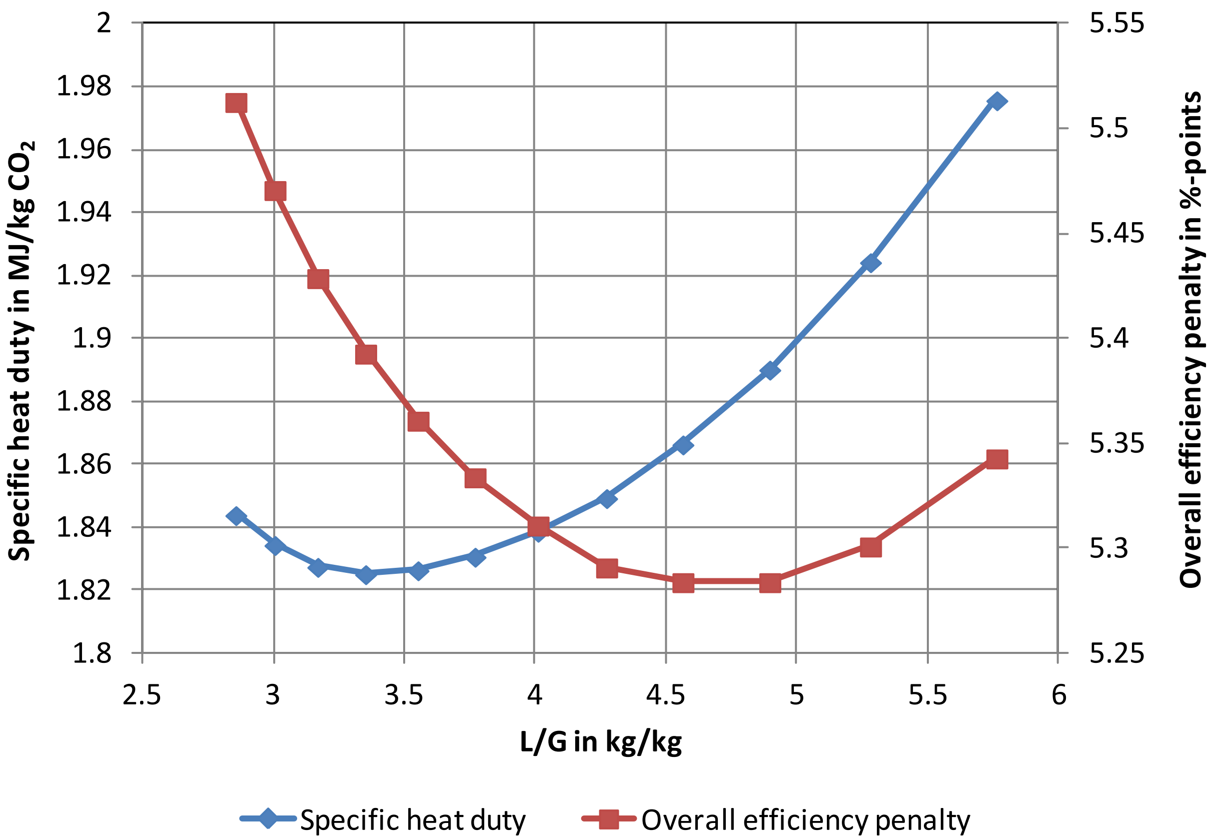

Figure 73: Overall efficiency penalty for a capture plant with overhead condenser heat integration in combination with an NGCC plant (B7a)

The effect on the overall process can be seen in Figure 73 where the overall efficiency penalty and the specific heat duty are shown for different solution mass flows. The lowest overall efficiency penalty of 5.28%-points is reached at an L/G of 4.6. Compared to the base case, this is a reduction by 0.65%-points, as can be seen in Table 41. The reduction is due to the reduced penalty caused by steam extraction, which results from the lower specific heat duty. For lower L/G, the increased reboiler temperature outweighs the reduced specific heat duty and the penalty due to steam extraction increases. For the operating point with the lowest specific heat duty, the overall efficiency penalty is thus higher.

Table 41: Contributors to the overall efficiency penalty of a capture plant in combination with an NGCC plant for base case (B1) and case with overhead condenser heat integration (B7a)

	NGCC base case	OHC heat integration, lowest overall efficiency penalty	OHC heat integration, lowest specific heat duty
Steam extraction	3.45%-points	2.83%-points	2.97%-points
Compressor duty	1.20%-points	1.20%-points	1.20%-points
Cooling water pumps	0.12%-points	0.11%-points	0.11%-points
Electrical duty	0.53%-points	0.52%-points	0.49%-points
Flue gas recirculation	0.62%-points	0.62%-points	0.62%-points
Overall efficiency penalty	**5.93%-points**	**5.28%-points**	**5.39%-points**

6.10 Various heat integration options - reboiler condensate

6.10.1 Process Characteristics

The second heat integration option, the integration of heat from reboiler condensate, is shown in Figure 74. The reboiler condensate, leaving the reboiler is heat exchanged with a semi lean solution stream extracted at half height of the stripper. The feedback of the solution is directly downstream of the extraction. An LMTD of 5 K is assumed for the heat exchanger. For the coal case, the reboiler condensate is already integrated into the preheating train. Thus, this modification is evaluated only for the NGCC case. Here, the reboiler condensate would otherwise be led to the economiser with a higher temperature. Since there is too much heat available in the flue gas in the base case (cf. section 5.2.1) a negative effect on the overall process is not expected. Still, the positive effect on the overall process is expected to be smaller compared to the first heat integration modification, since only sensible and no latent heat is available for integration.

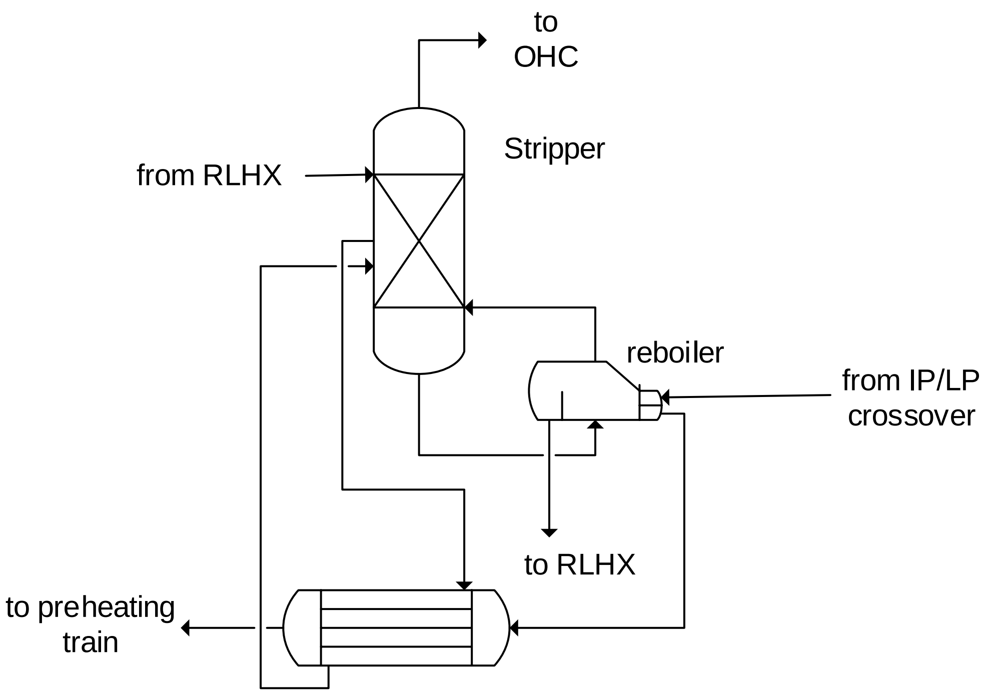

Figure 74: Schematic flow diagram of the reboiler condensate heat integration

6.10.2 NGCC power plant results - B7b

The results for the integration of heat from the reboiler condensate into the capture plant for the NGCC case are shown in Figure 75. It can be seen that the effect on the specific heat duty is smaller compared to the OHC heat integration. This is due to the small amount of available heat in the reboiler condensate. The temperature of the semi-lean solution extracted from the desorber is 122.6 °C. The temperature of the reboiler condensate results from the temperature in the desorber bottom and the temperature approach in the reboiler and adds up to 144.5 °C. Thus, the reboiler condensate can be cooled down by only around 20 °C. The specific heat duty is thus reduced by only 0.08 MJ/kg CO_2 from 2.37 to 2.29 MJ/kg CO_2. The specific cooling duty, the specific auxiliary power and the reboiler temperature for the same operating point are not changed compared to the base case. The heat source for the desorber is changed, while the rest of the process is not affected by the modification. The flue gas temperature is decreased, since the temperature of the reboiler condensate is reduced and more heat can be transferred from the flue gas. The interface quantities are shown in Table 42.

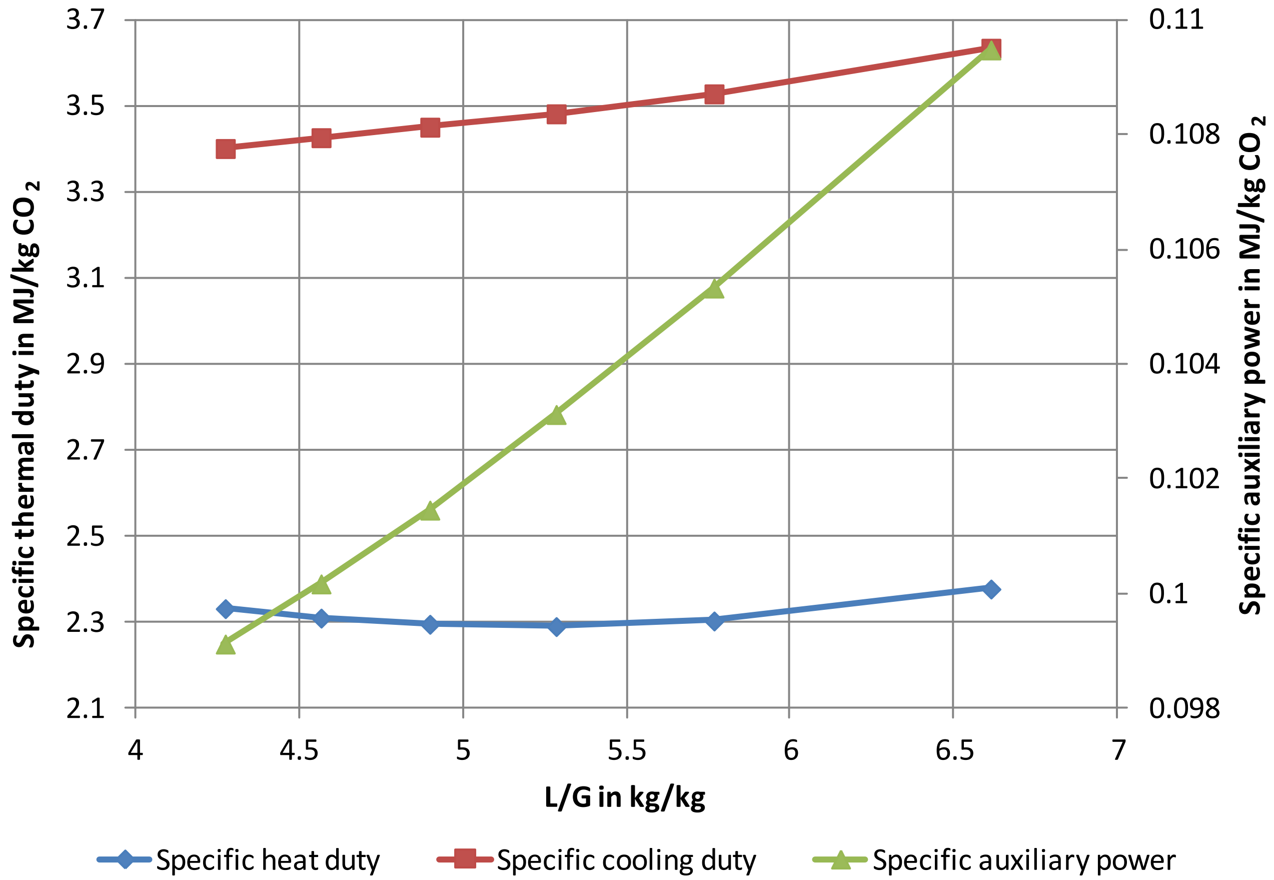

Figure 75: Specific thermal duty and specific auxiliary power of a capture plant with reboiler condensate heat integration in combination with an NGCC plant (B7b)

Table 42: Interface quantities of a capture plant in combination with an NGCC plant for base case (B1) and case with reboiler condensate (RC) heat integration (B7b)

	NGCC base case	RC heat integration
Specific heat duty in MJ/kg CO_2	2.37	2.29
Specific cooling duty in MJ/kg CO_2	3.48	3.48
Specific auxiliary power in MJ/kg CO_2	0.10	0.10
Desorber pressure in bar	5	5
Reboiler temperature in °C	132.4	132.4
Flue gas temperature upstream of the capture plant in °C	109.8	104.4

The effect on the overall process can be seen in Figure 76 where the overall efficiency penalty and the specific heat duty are shown for different solution mass flows. The overall efficiency penalty is reduced by 0.1%-points from 5.93 to 5.83%-points. The detailed list of contributors to the overall efficiency penalty is shown in Table 43. It can be seen that the reduction of the overall efficiency penalty is due to the lower penalty caused by steam extraction. As for the OHC heat integration, the lowest specific heat duty does not result in the lowest overall efficiency penalty, since the increased reboiler temperature outweighs the reduced specific heat duty.

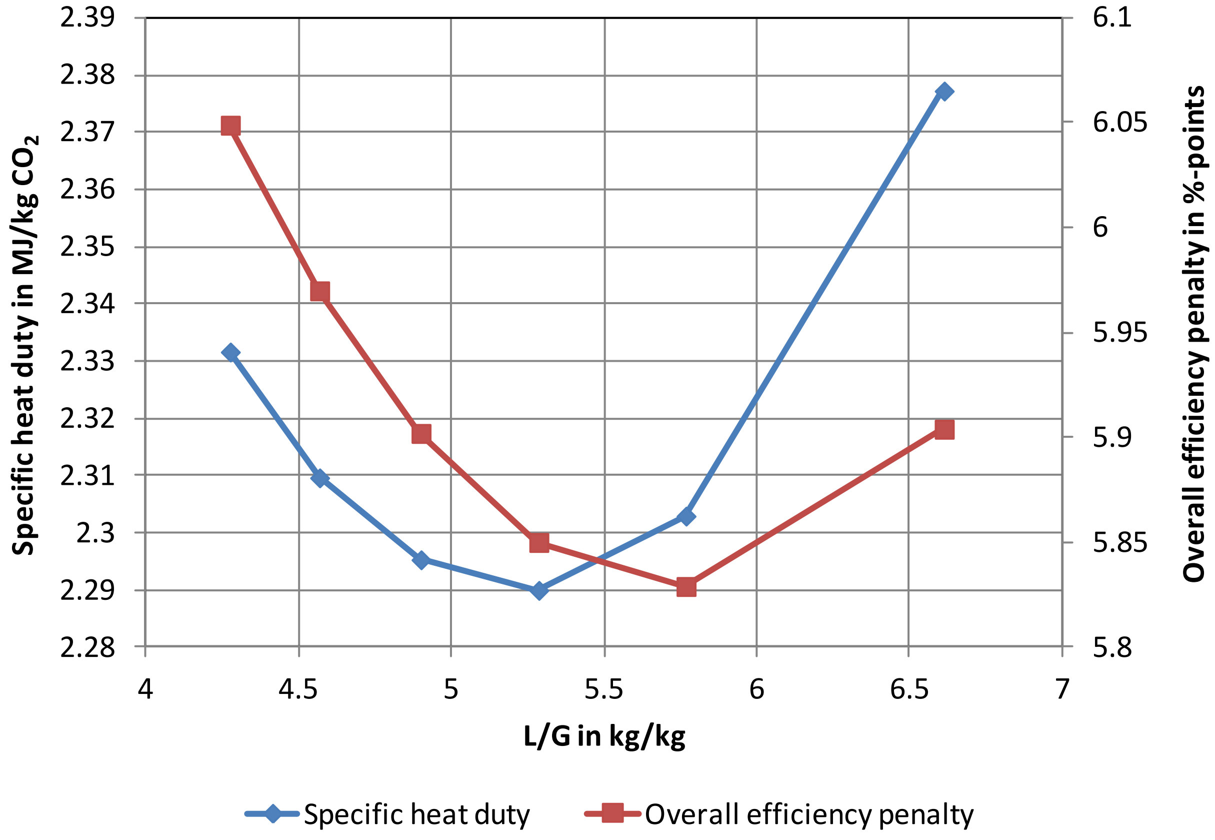

Figure 76: Overall efficiency penalty for a capture plant with reboiler condensate heat integration in combination with an NGCC plant (B7b)

Table 43: Contributors to the overall efficiency penalty of a capture plant in combination with an NGCC plant for base case (B1) and case with reboiler condensate (RC) heat integration (B7b)

	NGCC base case	**RC heat integration, lowest overall efficiency penalty**	**RC heat integration, lowest specific heat duty**
Steam extraction	3.45%-points	3.35%-points	3.38%-points
Compressor duty	1.20%-points	1.20%-points	1.20%-points
Cooling water pumps	0.12%-points	0.12%-points	0.12%-points
Electrical duty	0.53%-points	0.53%-points	0.53%-points
Flue gas recirculation	0.62%-points	0.62%-points	0.62%-points
Overall efficiency penalty	**5.93%-points**	**5.83%-points**	**5.85%-points**

6.11 Improved process flow sheet modification - Vapour recompression and split flow

6.11.1 Process Characteristics

In each of the following two sections, two of the flow sheet modifications described in the previous sections are combined in a single flow sheet. In this section, vapour recompression (cf. section 6.4) and the split flow process (cf. section 6.7) are combined as shown in Figure 75. For the split flow process, a lower overall efficiency penalty is achieved, although the reboiler temperature is increased. This could be beneficial for the vapour recompression, since more water is expected to be evaporated during flashing due to the higher reboiler temperature. The high CO_2 content in the vapour was assumed to be one of the main reasons for the bad performance of the vapour recompression case.

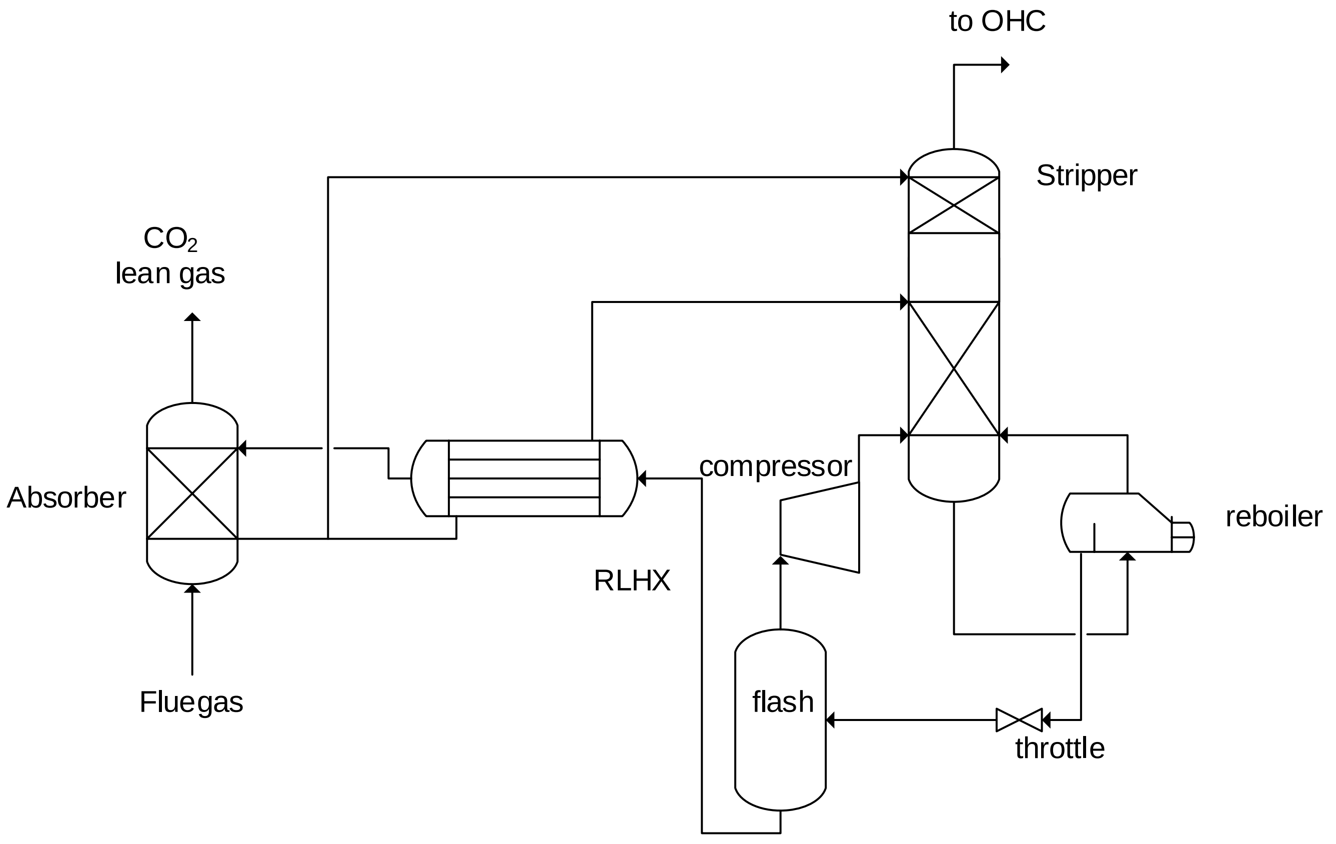

Figure 77: Schematic flow diagram of the combination of vapour recompression and split flow process

6.11.2 SCPC power plant results - A8

For the combination of vapour recompression and split flow process in combination with an SCPC plant, the specific heat duty and the specific auxiliary power are shown in Figure 78 for different flash pressures. For each flash pressure, the L/G as well as the split ratio are varied to find the operating points with the lowest specific heat duty. It can be seen that the specific heat duty increases for decreasing flash pressure up to a maximum at 3.5 bar flash pressure and decreases when the flash pressure is decreased further. This is due to the fact that vapour recompression leads to a higher temperature gradient in the desorber. The flashed vapour is reintroduced at a high temperature and more CO_2 is stripped in the bottom of the desorber. Simultaneously, the lean solution is cooled down during throttling, which reduces the temperature level in the RLHX and thus the temperature of the rich solution at the desorber head. For that reason, the positive effect of the split flow, which resulted from the high temperature at the desorber head, is reduced. For lower flash pressures, the split ratio is thus decreased from 0.09 at a flash pressure of 4.75 bar to 0.04 for a flash pressure of 1.5 bar. Still, the specific heat duty is smaller compared to the vapour recompression case. The specific auxiliary power increases for decreasing flash pressure. The effect is similar to the vapour recompression case and is not affected by the split flow.

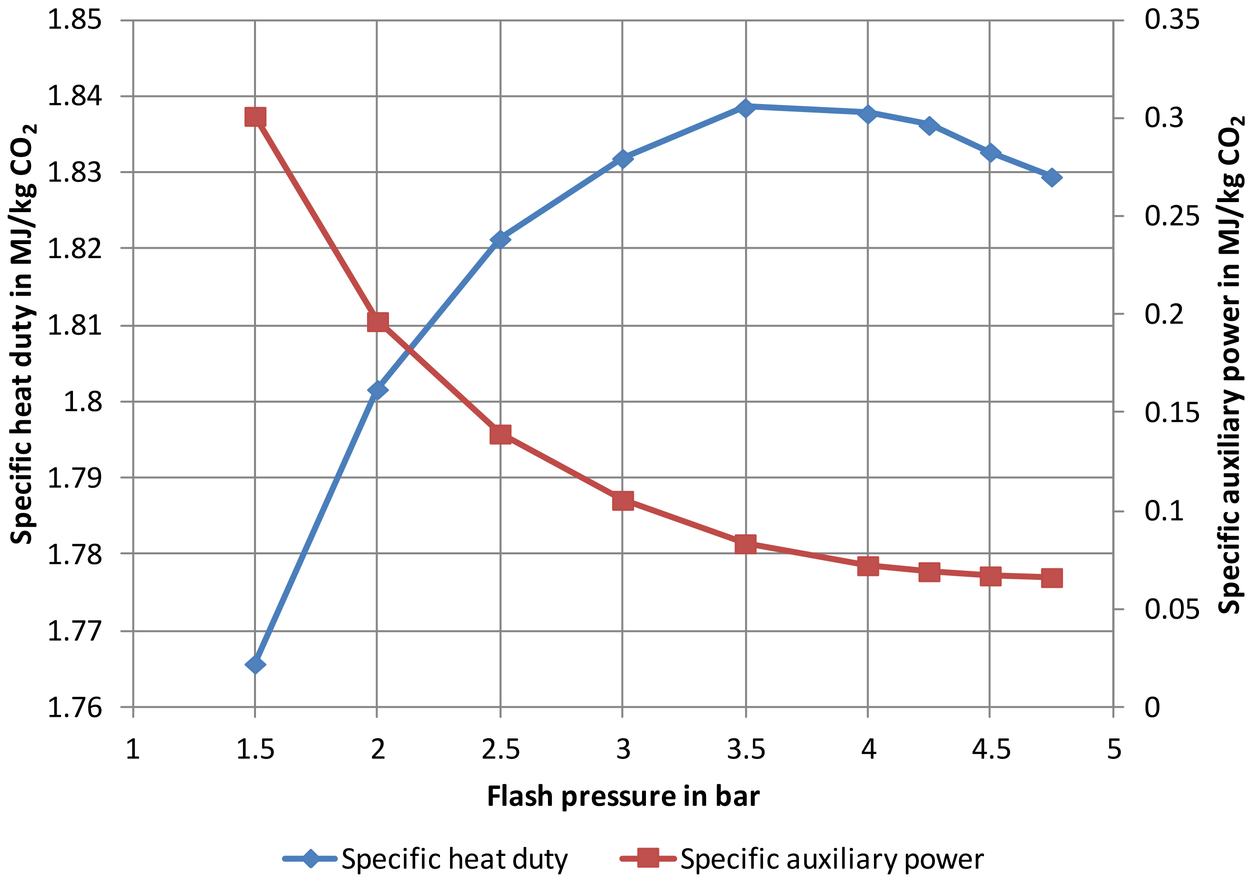

Figure 78: Specific heat duty and specific auxiliary power of a capture plant with vapour recompression and split flow in combination with an SCPC plant (A8) for different flash pressures

The interface quantities for the operating point with the lowest specific heat duty are shown in Table 44. For comparison, the interface quantities for the base case as well as for the vapour recompression case with the same flash pressure are shown. It can be seen that the specific heat duty for the combination of vapour recompression and split flow is 0.08 MJ/kg CO_2 lower than for the vapour recompression case. For the base case, the positive effect of the split flow was much higher (0.33 MJ/kg CO_2, cf. section 6.7).

The reboiler temperature, which was expected to be increased and thus to have a positive effect on the vapour recompression, is the same for both modifications. The amount of usable waste heat is further reduced since the split stream is led to the desorber head reducing the temperature of the overhead vapour.

Table 44: Interface quantities of a capture plant in combination with an SCPC plant for base case (A1), case with vapour recompression (A2) and case with vapour recompression and split flow (A8)

	SCPC base case	**Split flow and vapour recompression, flash pressure 1.5 bar**	**Vapour recompression, flash pressure 1.5 bar**
Specific heat duty in MJ/kg CO_2	2.14	1.77	1.85
Specific cooling duty in MJ/kg CO_2	2.75	2.41	2.50
Specific auxiliary power in MJ/kg CO_2	0.069	0.30	0.29
Desorber pressure in bar	5	5	5
Reboiler temperature in °C	128.0	129.6	129.6
Usable waste heat from OHC in MJ_{th}/kg CO_2	0.524	0.124	0.29
Temperature level of usable waste heat in °C	116.2	83.3	102.6

In Figure 79, overall efficiency penalties are shown for cases without and with advanced heat integration for the combination of vapour recompression and split flow as well as for the vapour recompression case varying the flash pressure. It can be seen that the lowest overall efficiency penalties for the combination are reached for the highest flash pressure. While the penalty showed a small increase for the vapour recompression case towards higher flash pressures, this increase cannot be seen for the combination. This is due to the decrease in specific heat duty towards higher flash pressures for the combination. Comparison with the vapour recompression case shows that the overall efficiency penalty is smaller for the combination for all operating points. In Table 45, the contributors to the overall efficiency penalty are shown for the case with the lowest overall efficiency penalty. Since the flash pressure is only slightly below the desorber pressure of 5 bar, the results are similar to the results obtained for the split flow process (cf. section 6.7.2).

In summary it can be stated, that the overall efficiency penalty for this combination is always higher than or equal to the overall efficiency penalty for the split flow process alone which achieves a lowest overall efficiency penalty of 5.99%-points.

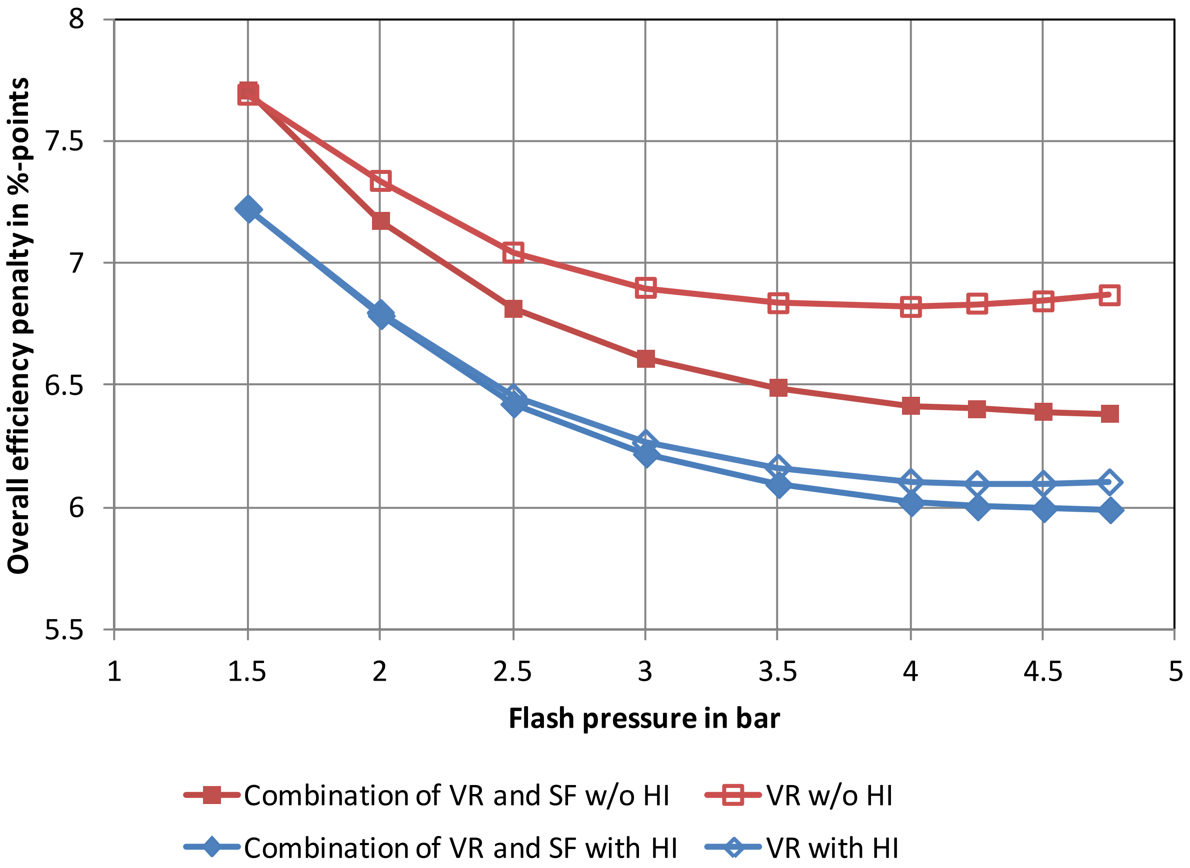

Figure 79: Overall efficiency penalty for a capture plant with vapour recompression combined with split flow (A8) as well as vapour recompression (A2) in combination with an SCPC plant with and without heat integration for different flash pressures

Table 45: Contributors to the overall efficiency penalty of a capture plant in combination with an SCPC plant for base case (A1) and case with vapour recompression and split flow (A8) with and without heat integration

	SCPC base case without HI	**SF and VR, flash pressure 4.75 bar, w/o HI**	**SCPC base case with HI**	**SF and VR, flash pressure 4.75 bar, with HI**
Steam extraction	4.16%-points	3.66%-points	4.21%-points	3.69%-points
Compressor duty	1.90%-points	1.90%-points	2.06%-points	2.06%-points
Cooling water pumps	0.23%-points	0.21%-points	0.21%-points	0.19%-points
Electrical duty	0.62%-points	0.61%-points	0.60%-points	0.61%-points
Heat integration			-0.97%-points	-0.55%-points
Overall efficiency penalty	**6.91%-points**	**6.38%-points**	**6.11%-points**	**5.99%-points**

6.11.3 NGCC power plant results - B8

The results for the NGCC case are similar to the results for the SCPC case. In Figure 80, the specific heat duty and the specific auxiliary power for the NGCC case are shown for different flash pressures. It can be seen that the specific heat duty is again increasing for decreasing flash pressures up to a maximum at 3 bar. For lower flash pressures, the specific heat duty decreases. The lowest specific heat duty of 1.93 MJ/kg CO_2 is reached for the lowest flash pressure evaluated. As for the vapour recompression case, the specific auxiliary power increases significantly for lower flash pressures.

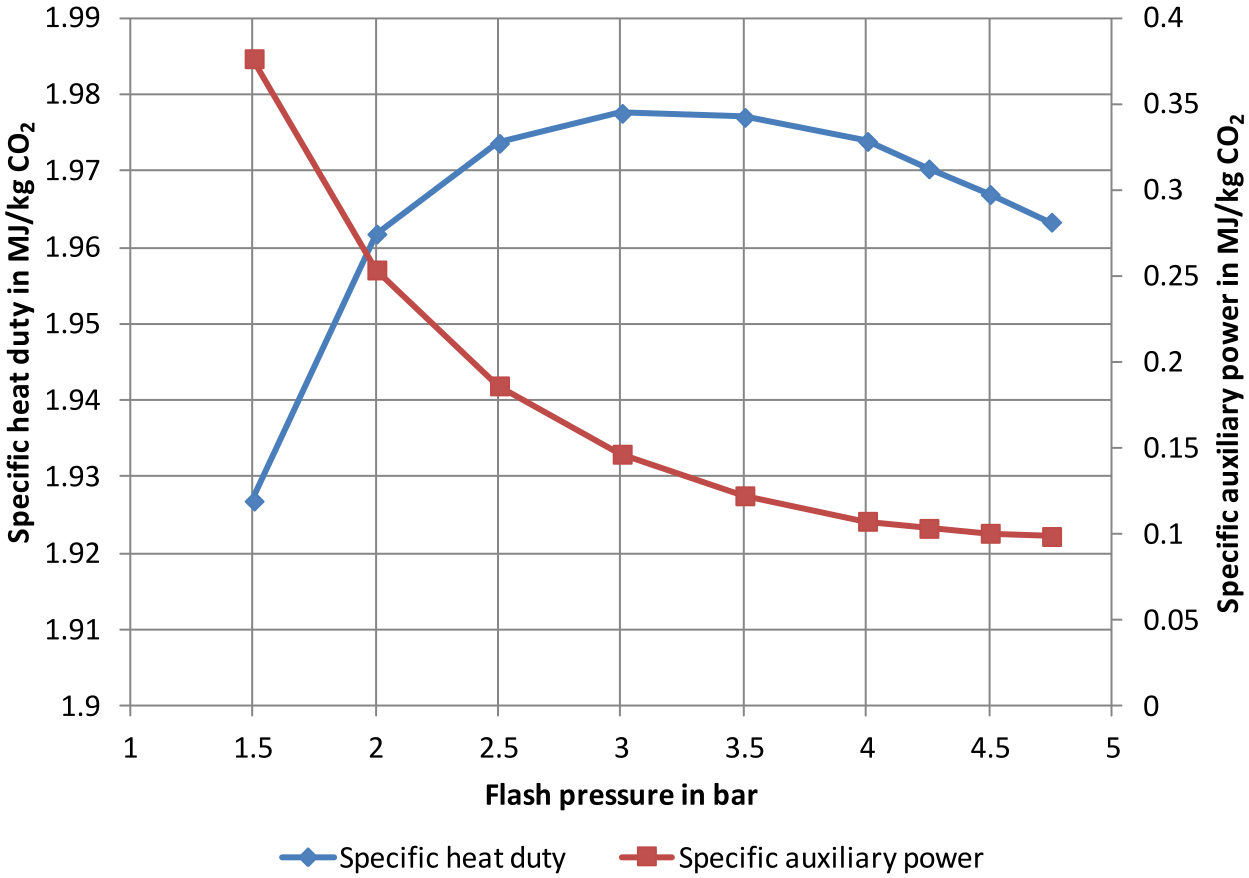

Figure 80: Specific heat duty and specific auxiliary power of a capture plant with vapour recompression and split flow in combination with an NGCC plant (B8) for different flash pressures

The interface quantities for the NGCC case are shown in Table 46. For comparison, the interface quantities for the base case and the vapour recompression case are shown as well. It can be seen that the difference in specific heat duty between vapour recompression case and the combination of vapour recompression and split flow is very small for this operating point. This is due to the reduced effect of the split flow for low flash pressures, as described in the previous section. The split ratio for this operating point is reduced even further as for the coal case to 0.02. The specific cooling duty and the flue gas temperature upstream the flue gas cooler are slightly decreased compared to the vapour recompression case, since less heat is transferred to the capture plant.

Table 46: Interface quantities of a capture plant in combination with an NGCC plant for base case (B1), case with vapour recompression (B2) and case with vapour recompression and split flow (B8)

	NGCC base case	SF and VR, flash pressure 1.5 bar	VR, flash pressure 1.5 bar
Specific heat duty in MJ/kg CO_2	2.37	1.93	1.95
Specific cooling duty in MJ/kg CO_2	3.48	3.19	3.26
Specific auxiliary power in MJ/kg CO_2	0.10	0.38	0.38
Desorber pressure in bar	5	5	5
Reboiler temperature in °C	132.4	129.6	129.6
Flue gas temperature upstream of the capture plant in °C	109.8	103.6	103.9

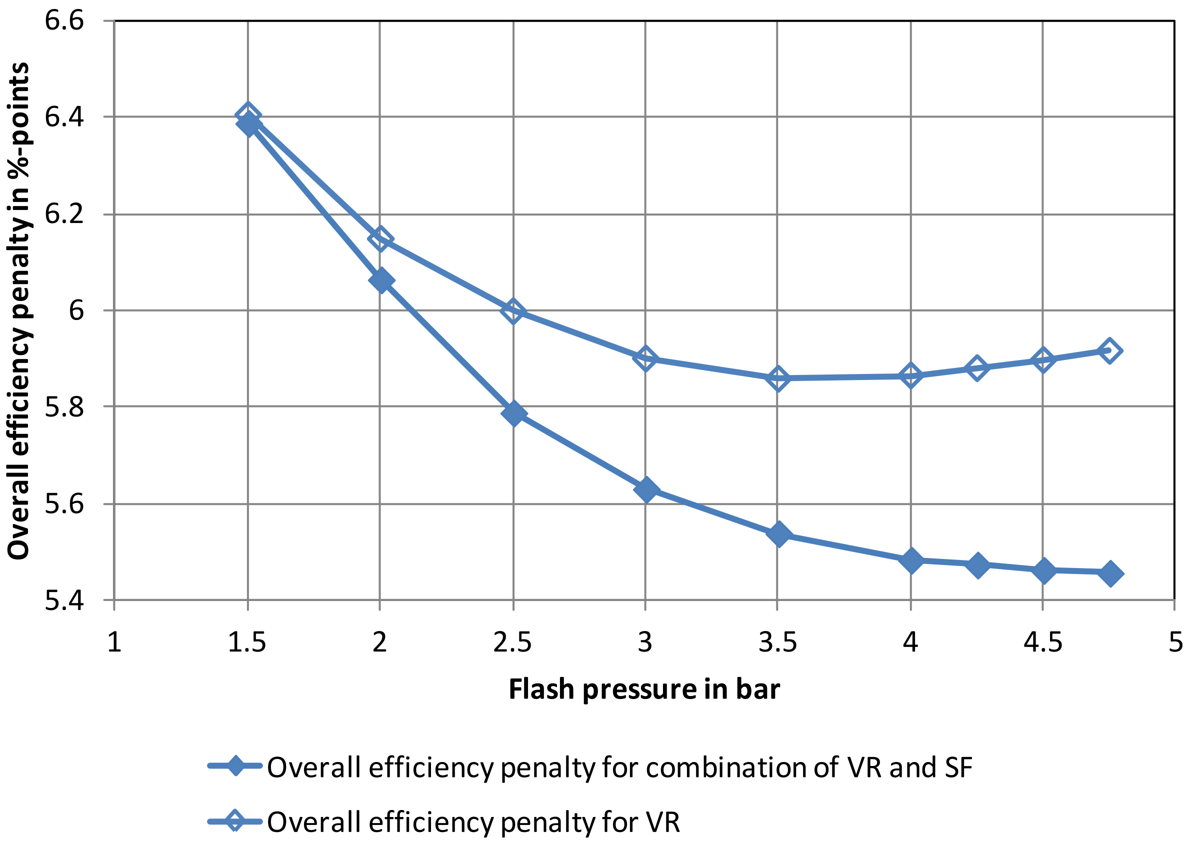

Figure 81: Overall efficiency penalty for a capture plant with vapour recompression combined with split flow (B8) as well as vapour recompression (B2) in combination with an NGCC plant for different flash pressures

The overall efficiency penalty for the NGCC case is shown in Figure 81. For comparison, the overall efficiency penalty for the vapour recompression case is shown as well. It can be seen that the overall efficiency penalty increases for decreasing flash pressures without the minimum obtained for vapour recompression only. The lowest overall efficiency penalty is thus achieved for the highest flash pressure evaluated. Since the effect of vapour recompression on the overall process is very small for this operating point, the obtained results are very similar to the split flow case. This can be seen in Table 47, where the contributors to the overall efficiency penalty are shown. In addition to the base case and the combination of vapour recompression and split flow, the contributors for the split flow case and for the vapour recompression case with the lowest overall efficiency penalty are shown as well. It can be seen that the overall efficiency penalty for the combination is reduced compared to the vapour recompression case, since low specific heat duties and thus low penalties due to steam extraction are reached for higher flash pressures.

In summary it can be stated, that the overall efficiency penalty for this combination is always higher than or equal to the overall efficiency penalty for the split flow process alone which achieves a lowest overall efficiency penalty of 5.46%-points.

Table 47: Contributors to the overall efficiency penalty of a capture plant in combination with an NGCC plant for base case (B1), case with vapour recompression and split flow (B8), case with split flow only (B5), and case with vapour recompression only (B2)

	NGCC base case	**SF and VR, flash pressure 4.75 bar**	**SF**	**VR, flash pressure 3.5 bar**
Steam extraction	3.45%-points	2.99%-points	2.99%-points	3.27%-points
Compressor duty	1.20%-points	1.20%-points	1.20%-points	1.20%-points
Cooling water pumps	0.12%-points	0.12%-points	0.12%-points	0.13%-points
Electrical duty	0.53%-points	0.52%-points	0.52%-points	0.64%-points
Flue gas recirculation	0.62%-points	0.62%-points	0.62%-points	0.62%-points
Overall efficiency penalty	**5.93%-points**	**5.46%-points**	**5.46%-points**	**5.86%-points**

6.12 Improved process flow sheet modification - Heat-integrated stripper and overhead condenser heat integration

6.12.1 Process Characteristics

The second combination of process flow sheet modifications is the combination of the heat-integrated stripper (HIS) (cf. section 6.6) and the overhead condenser (OHC) heat integration (cf. section 6.9). The HIS is not beneficial for the overall process since the reduced specific heat duty is outweighed by the increased reboiler temperature. The integration of heat from the OHC results in a significant decrease of specific heat duty and overall efficiency penalty. When both modifications are combined, the increased reboiler temperature due to the HIS could increase the amount of heat available in the OHC and thus enlarge the positive effect of the heat integration. The schematic flow diagram for this combination is shown in Figure 82.

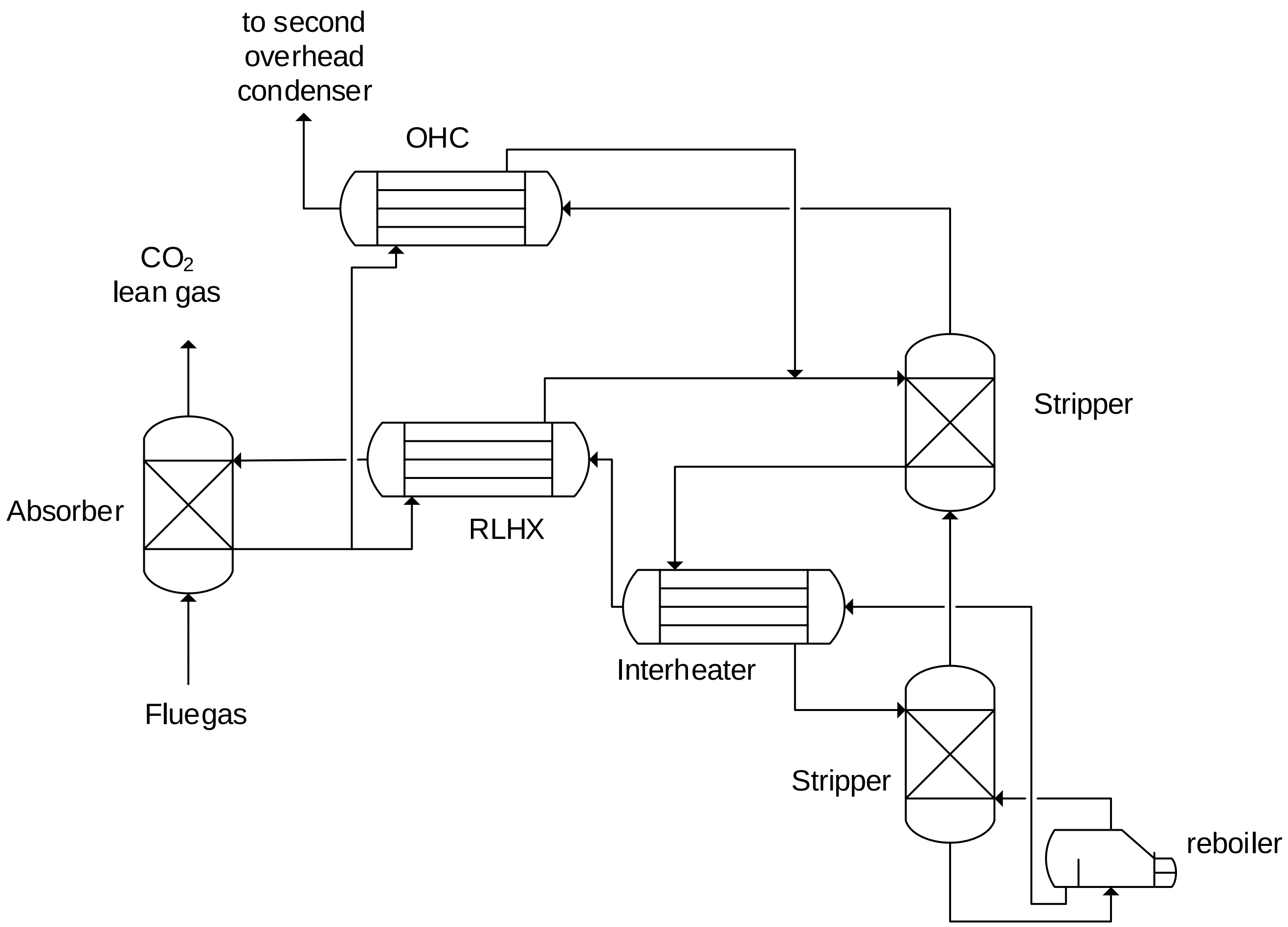

Figure 82: Schematic flow diagram of the combination of heat-integrated stripper and overhead condenser heat integration

6.12.2 SCPC power plant results - A9

As for the interheated stripper, the relative interheater duty (RID) is defined as the ratio between the heat duty in the interheater and the reboiler heat duty. For every RID, the L/G is varied to find the operating point with the lowest specific heat duty. The specific thermal duties and the specific auxiliary power for these operating points are shown in Figure 83. It can be seen that the specific heat duty increases for increasing RID. This can be explained by the fact that the use of an interheater for a fixed lean loading, and thus a fixed reboiler temperature, reduces the temperature of the rich solution entering the RLHX. A reduction of the specific heat duty is thus only possible, when the reboiler temperature is increased as well. The operating point with the lowest specific heat duty for the OHC heat integration alone has a high reboiler temperature, already. The potential for a further increase of the reboiler temperature is thus very small and is outweighed by the negative effect of the reduced desorber inlet temperature.

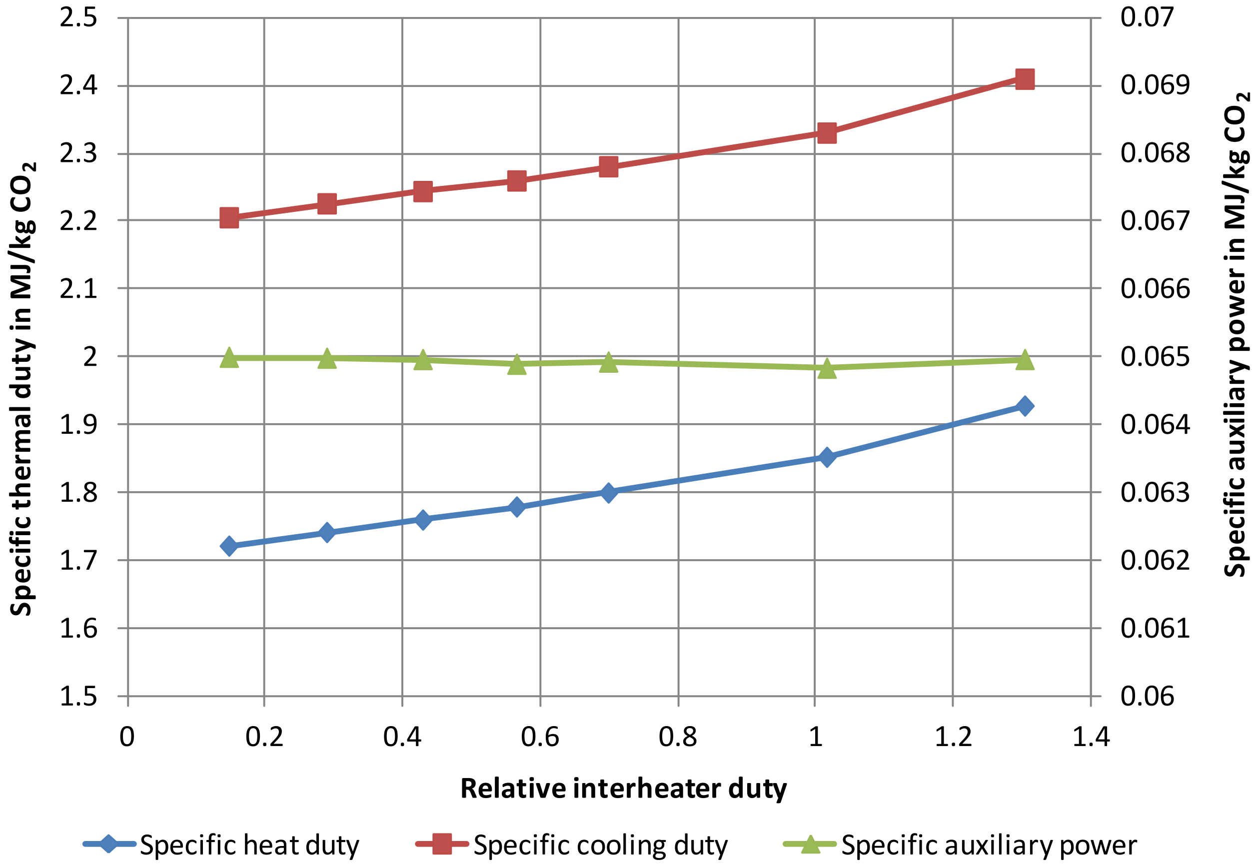

Figure 83: Specific thermal duties and specific auxiliary power of a capture plant with heat-integrated stripper and overhead condenser heat integration in combination with an SCPC plant (A9) for different relative interheater duties

The interface quantities for the base case and for the combination of a HIS and OHC heat integration are shown in Table 48. For comparison, the interface quantities for the OHC heat integration case are shown as well. The operating point for the combination is the one with the lowest evaluated heat duty in the interheater. Since only a small amount of heat is transferred, the interface quantities for the combination are similar to the interface quantities of the OHC heat integration case. Still, the specific heat duty as well as the specific cooling duty is increased slightly compared to the OHC heat integration case. The amount of usable waste heat is even further reduced than for the OHC heat integration case.

Table 48: Interface quantities of a capture plant in combination with an SCPC plant for base case (A1), case with heat-integrated stripper and overhead condenser heat integration (A9), and case with overhead condenser heat integration (A7)

	SCPC base case	HIS and OHC heat integration	OHC heat integration
Specific heat duty in MJ/kg CO_2	2.14	1.72	1.71
Specific cooling duty in MJ/kg CO_2	2.75	2.20	2.19
Specific auxiliary power in MJ/kg CO_2	0.069	0.065	0.065
Desorber pressure in bar	5	5	5
Reboiler temperature in °C	128.0	138.5	138.6
Usable waste heat from OHC in MJ/kg CO_2	0.524	0.033	0.034
Temperature level of usable waste heat in °C	116.2	50.0	50.0

The overall efficiency penalties for the combination of HIS and OHC heat integration are shown for the cases with and without advanced heat integration in Figure 84. For every RID the operating point with the lowest overall efficiency penalty is shown. For comparison the overall efficiency penalty for the case with HIS is shown as well. It can be seen that the overall efficiency penalty is reduced compared to the HIS case due to the positive effect of the OHC heat integration. In Table 49, the contributors to the overall efficiency penalty are shown for the case with the lowest overall efficiency penalty. Since only a small amount of heat is transferred in the interheater, the results are similar to the results obtained for the OHC heat integration case (cf. section 6.9.2).

In summary it can be stated, that the overall efficiency penalty for this combination is always higher than the overall efficiency penalty for the OHC heat integration alone which achieves a lowest overall efficiency penalty of 6.19%-points for the case without advanced waste heat integration and 5.84%-points for the case with advanced waste heat integration.

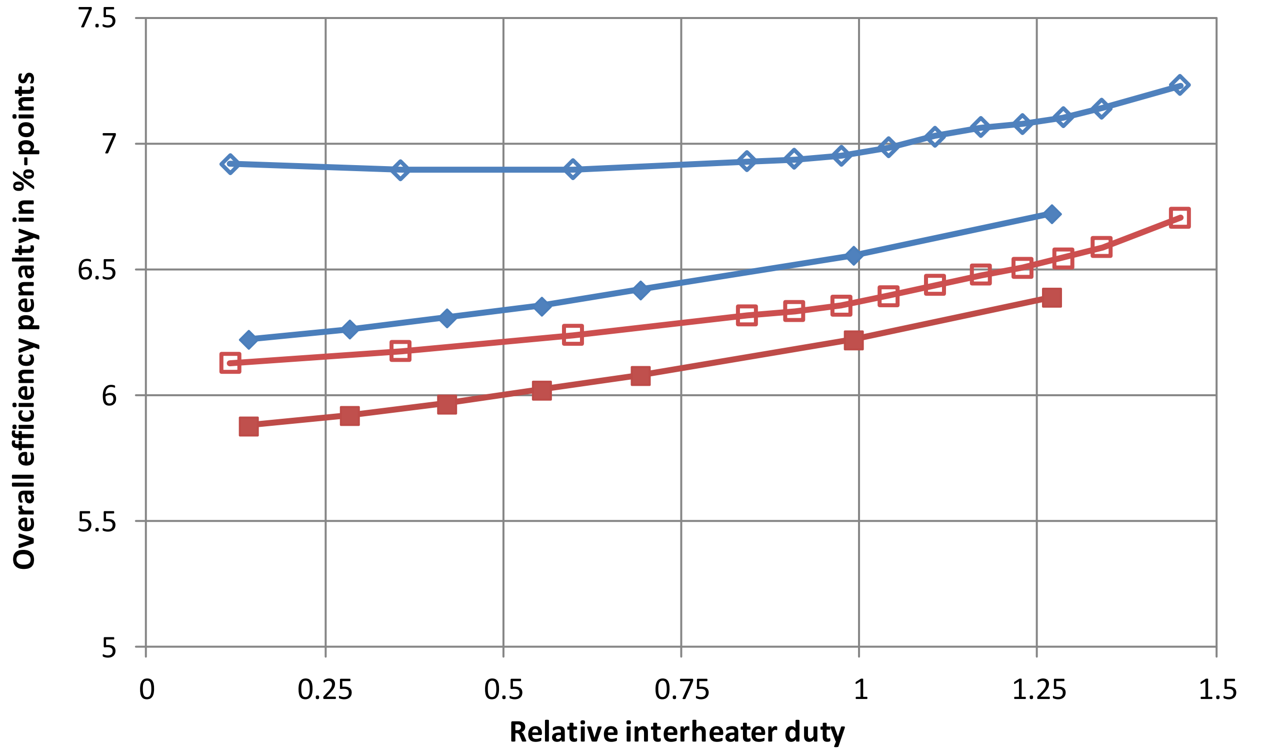

Figure 84: Overall efficiency penalty for a capture plant with heat-integrated stripper and overhead condenser heat integration (A9) and a capture plant with heat-integrated stripper (A4) in combination with an SCPC plant with and without heat integration for different relative interheater duties

Table 49: Contributors to the overall efficiency penalty of a capture plant in combination with an SCPC plant for base case (A1) and case with heat-integrated stripper and overhead condenser heat integration (A9) with and without advanced waste heat integration

	SCPC base case without HI	**HIS and OHC heat integration, w/o HI**	**SCPC base case with HI**	**HIS and OHC heat integration, with HI**
Steam extraction	4.16%-points	3.55%-points	4.21%-points	3.55%-points
Compressor duty	1.90%-points	1.90%-points	2.06%-points	2.06%-points
Cooling water pumps	0.23%-points	0.20%-points	0.21%-points	0.17%-points
Electrical duty	0.62%-points	0.57%-points	0.60%-points	0.57%-points
Heat integration			-0.97%-points	-0.48%-points
Overall efficiency penalty	**6.91%-points**	**6.23%-points**	**6.11%-points**	**5.88%-points**

6.12.3 NGCC power plant results - B9

The results for the NGCC case are similar to the results for the SCPC case. In Figure 85, the specific thermal duties and the specific auxiliary power for the NGCC case are shown for different RID. The specific heat duty, as well as the specific cooling duty are increasing for higher RID. The lowest specific heat duty is reached when the interheater has nearly no influence. The specific auxiliary power is not changed significantly since the operating points with the lowest specific heat duty for different RID are obtained for the same lean loading and thus the same solution mass flow.

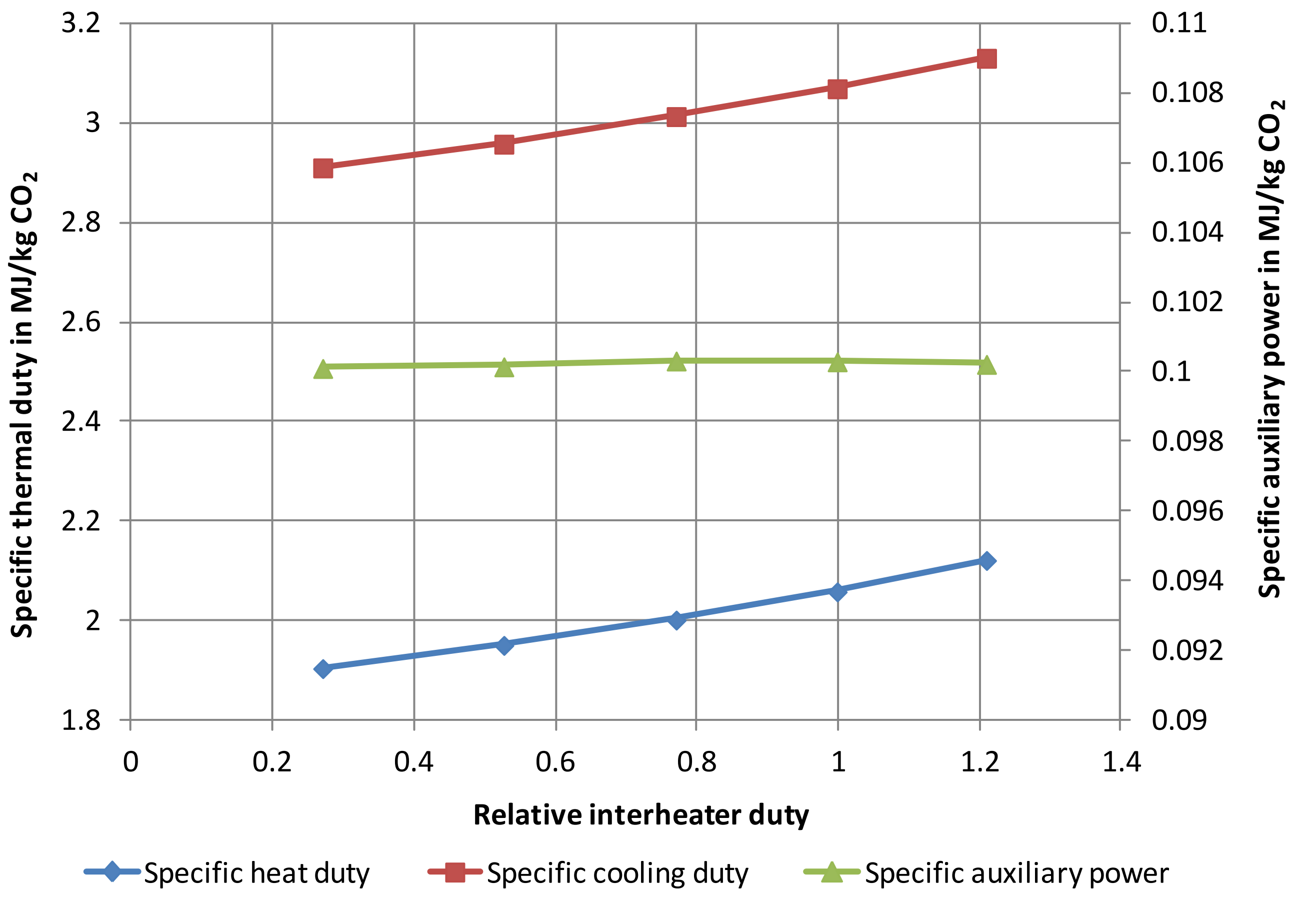

Figure 85: Specific thermal duties and specific auxiliary power of a capture plant with heat-integrated stripper and overhead condenser heat integration in combination with an NGCC plant (B9) for different relative interheater duties

The interface quantities for the base case and for the combination of a HIS and OHC heat integration are shown in Table 50. For comparison, the interface quantities for the OHC heat integration case are shown as well. As for the coal case, the interface quantities for the combination are very similar to the OHC heat integration case, since only a small amount of heat is transferred in the interheater. The flue gas temperature upstream of the capture plant is slightly increased since more heat is needed in the reboiler.

Table 50: Interface quantities of a capture plant in combination with an NGCC plant for base case (B1), case with heat-integrated stripper and overhead condenser heat integration (B9), and case with overhead condenser heat integration (B7a)

	NGCC base case	HIS and OHC heat integration	OHC heat integration
Specific heat duty in MJ/kg CO_2	2.37	1.90	1.83
Specific cooling duty in MJ/kg CO_2	3.48	2.91	2.84
Specific auxiliary power in MJ/kg CO_2	0.10	0.10	0.09
Desorber pressure in bar	5	5	5
Reboiler temperature in °C	132.4	145.6	145.6
Flue gas temperature upstream of the capture plant in °C	109.8	111.3	111.1

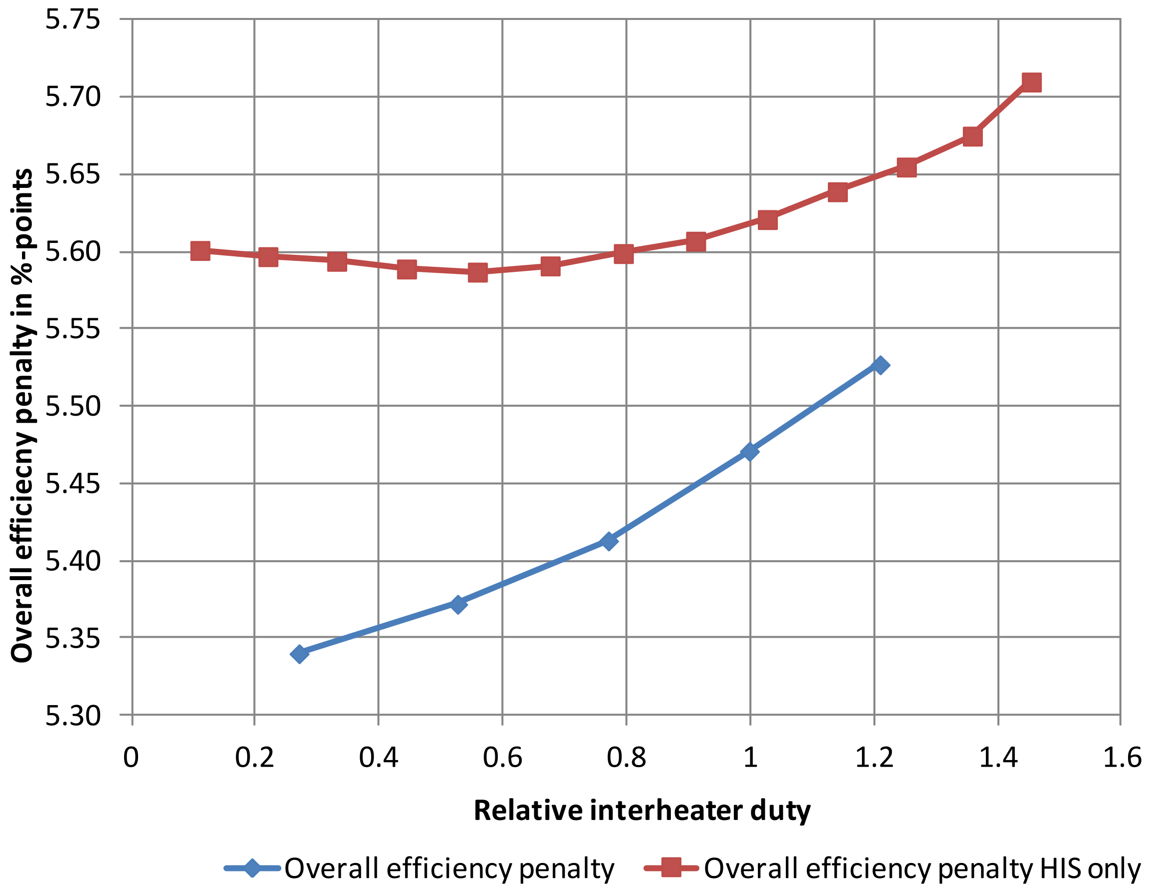

Figure 86: Overall efficiency penalty for a capture plant with heat-integrated stripper and overhead condenser heat integration (B9) and a capture plant with heat-integrated stripper (B4) in combination with an NGCC plant for different relative interheater duties

The overall efficiency penalties for the combination of HIS and OHC heat integration are shown in Figure 86. For every RID, the operating point with the lowest overall efficiency penalty is shown. For comparison, the overall efficiency penalty for the case with HIS alone is shown as well. As for the coal case, it can be seen that the overall efficiency penalty is reduced compared to the HIS case due to the positive effect of the OHC heat integration. In Table 51 the contributors to the overall efficiency penalty are shown for the case with the lowest overall efficiency penalty. Again, the results are similar to the results obtained for the OHC heat integration case (cf. section 6.9.3).

In summary it can be stated, that the overall efficiency penalty for this combination is always higher than the overall efficiency penalty for the OHC heat integration alone which achieves a lowest overall efficiency penalty of 5.28%-points.

Table 51: Contributors to the overall efficiency penalty of a capture plant in combination with an NGCC plant for base case (B1), case with heat-integrated stripper and overhead condenser heat integration (B9), and case with overhead condenser heat integration (B7a)

	NGCC base case	**HIS and OHC heat integration**	**OHC heat integration**
Steam extraction	3.45%-points	2.89%-points	2.83%-points
Compressor duty	1.20%-points	1.20%-points	1.20%-points
Cooling water pumps	0.12%-points	0.11%-points	0.11%-points
Electrical duty	0.53%-points	0.52%-points	0.52%-points
Flue gas recirculation	0.62%-points	0.62%-points	0.62%-points
Overall efficiency penalty	**5.93%-points**	**5.34%-points**	**5.28%-points**

7 Qualitative Analysis

In the qualitative analysis, the CO_2 capture process flow sheet modifications are investigated under aspects that differ from the energetic evaluation but are also important for overall analysis. The main aspect is the behaviour of the capture unit and the overall process in the whole operation range and under varying conditions. For the base, case a number of different aspects is analysed and for the modifications the main points are elaborated. The analysed aspects are:

- the impact of an increased CO_2 capture rate,
- the impact of the power plant size on the equipment requirement,
- the limitations from solvent properties on the process flow sheet modifications performance,
- the suitability of commercially available improved solvents on the performance of different process modifications,
- the impact of change in impurity concentration in the flue gas on solvent degradation, solvent make-up, corrosion, waste generation etc.,
- the operational flexibility requirement for part load operation of the power plant,
- the process control requirement in normal power plant operating conditions, issues related to retrofitting of an existing plant by looking at available utilities, space, power plant efficiency etc.
- site specific limitations like water availability, environment conditions etc.

The aspects on the limitation from the solvent, the operational flexibility in part load, the process control requirement and issues regarding the retrofitting are discussed for the different flow sheet modifications in detail and an overview is given in Table 52. The other aspects are discussed for the modifications in general in the description of the base cases.

7.1 Effect of increased CO_2 capture rate:

The behaviour of the process at higher capture rates than in the reference case are relevant because capture rates of more than 90% could be temporarily necessary to reach an average capture rate of 90% during the year. The reference capture rate is 90%; reducing the capture rate leads to lower reboiler heat duties, while higher capture rates increase the reboiler heat duty significantly. This is due to the higher or lower lean loading required for lower respectively higher capture rates.

For the solvent 7 m MEA the specific heat duty increases by 3% for a capture rate of 95% [42]. A reduction to a capture rate of 70% reduces the heat duty by 3%. The consequences for an SCPC overall process are a higher power loss for higher heat duties or a generation of additional electric energy for lower capture rates. For 7 m MEA, a higher capture rate of 95% leads to an additional power loss of approximately 3%. With a reduced capture rate of 70% there is the possibility to generate around 5% additional power. These values, especially the values for the overall process, are very site specific [42].

For the base case of the SCPC power plant with Solvent2020 used in this study, the additional heat duty is around 4% to reach a capture rate of 95%. The additional losses in the overall process is expected to be in the same order of magnitude. The capture plants with the different process modifications are expected to behave in the same way. All modifications cover improvements at the desorber but not at the absorber. This means that for a higher capture rate the solution mass flow and the lean loading have to be manipulated, because the rich loading is coupled with the absorber. Processes which integrate the heat more efficiently will benefit from higher solution mass flows and processes with a flat response of the specific heat duty on the L/G or the lean loading will benefit from a further reduction of the lean loading.

7.2 Size of power plant

The power plant size is a boundary condition and therefore very variable. To examine the impact of the power plant size on the process equipment requirement, a possible variation in the power output is shown in the following. Due to the possibility to build multiple parallel trains, there is no limitation in power plant size by the capture plant. The determining factor for the number of parallel trains is the absorber, in the base case for an SCPC plant the absorber diameter is around 17.6 m with a limit of 18 m, (cf. chapter 4). For an SCPC with power output of more than 900 MW_{el}, this will result in more than two trains. The base case of an NGCC plant results in an absorber diameter of 14.5 m. For the modifications there are no further limitations. The components within the process of the different modifications can be built in parallel trains.

7.3 Impact of solvent properties

In some cases the solvent properties can limit the performance of the process flow sheet modification. The characteristic solvent properties are described in chapter 3. For the overall process analysis, the most important solvent property is the interaction between the specific interface quantities and the process parameters desorber pressure, lean loading and reboiler temperature. The reboiler temperature is limited by the degradation potential and the desorber pressure. In the base cases, no limitation of the solvent properties are significant. The impact on the different modifications is shown in Table 52. The vapour recompression could be more efficient if the solvent has a better CO_2 regeneration performance and less CO_2 will be in the vapour downstream the flash. This point is discussed in section 6.4. This behaviour is also negative for the multi-pressure stripper. In the heat-integrated stripping column, the reboiler temperature is a real limit because in some cases the temperature could exceed 150 °C. The improved process flow sheet modifications which include either a vapour recompression or a heat-integrated stripping column are therefore also limited by the solvent properties.

The issue of the suitability of commercially available improved solvents on the performance of different process modifications is of interest. Most available solvents are suitable for the most of the process modifications evaluated in this study. The benefit between the performance of the modification and the reference case could be larger, especially the vapour recompression could be more promising, see section 6.4, but with the actual solvents the reference case would not be that efficient. For a reliable conclusion, the solvents have to be modelled and examined in detail with the capture plant and an overall process analysis is necessary.

The solvent characteristics in degradation, solvent make-up and corrosion depend on the impurities of the flue gas and the temperature level in the reboiler. In this study a mixture of tertiary amine and polyamine is used. The degradation potential is lower than for primary amines. A lower degradation potential is beneficial for the solvent make-up rate, the fouling of the system, the corrosion rate and the reclaimer waste. The corrosivity is also lower for these solvents [43]. For a better behaviour a pre-treatment column for lower SO_x and NO_x concentration in the flue gas could be necessary. The impact of the impurities is similar for all process modifications. For capture plants with multi-pressure stripping and matrix stripping the reboiler temperature is higher and therefore the solvent degradation potential is higher.

7.4 Effect of power plant operation flexibility at part load conditions

Another important issue is the operational flexibility requirement for part load operation of the power plant. In part load, the boundary conditions for the capture plant deviate from those for full load. The flue gas composition and mass flow are different for varying loads. For an SCPC the CO_2 content decreases due to a higher air excess and the mass flow decreases. This leads to a lower specific reboiler heat duty in part load because of a closer approach to equilibrium in the absorber and a lower LMTD in the RLHX, both caused by overdesigned equipment. But for the overall process the efficiency penalty increases because of higher losses in the steam conditioning process in part load. In part load the IP/LP crossover pressure decreases according to Stodola's law. Therefore a pressure maintaining valve is necessary to guarantee a certain steam pressure level for the reboiler [15]. Also the specific auxiliary power of the CO_2-compressor depends on the load. In part load, the specific auxiliary power is higher due to lower efficiencies of the compressor. A further efficiency reduction occurs due to a bypass operation of the compressor, which could extend the operation range [42]. For an NGCC plant similar results regarding the steam extraction can be expected, since the steam turbines and the steam conditioning behave like in an SCPC plant. The impact on the different flow sheet modifications is shown in Table 52. It can be expected that the vapour recompression and the multi-pressure stripping will have higher losses in part load, because the efficiency of fans decreases in part load operation. Processes with heat exchangers can operate more efficiently at part load, because the heat exchangers are overdesigned and the temperature approach is smaller in part load operation. Modifications with heat integration benefit from smaller temperature differences and reduced losses. The matrix stripping has an advantage in part load. The pressure of the first desorber, which influences the necessary steam pressure, can be reduced easily without influencing the compressor much and therefore the reboiler temperature decreases as well as the losses for the steam conditioning.

7.5 Process control requirement

The process control is necessary to reach a value for the control variable by setting the actuating variable. At normal power plant operating conditions the most important control variable is the capture rate. The capture rate should be 90% and can be reached by manipulating the solvent flow and the reboiler heat duty. The control has to respect the overall efficiency of the power plant and should operate the capture plant in an operation regime with the lowest efficiency penalty for a capture rate of 90%. Other control variables are in subsidiary controls, like certain levels of temperature in heat exchanger. The requirement in control of the capture plant rises with more complexity in the flow sheet modifications and the choice of free variables. As shown in Table 52, the most complex modification is the matrix stripping. Here, the degree of freedom is the largest and the split factor and the pressure level need a control loop. Most of the modifications have a slight increase in the complexity compared to the base case.

7.6 Retrofitting to an existing power plant

All process analysis in chapter 6 were done for the case of a Greenfield power plant. When retrofitting an existing power plant, other issues like space and available utilities have to be considered and the IP/LP crossover pressure is of major importance. The design crossover pressure of the power plant influences the choice of the optimal process flow sheet modification. The temperature of the reboiler gets a higher sensitivity; at lower crossover pressures a lower reboiler temperature is significantly beneficial due to lower losses in steam conditioning and the other way round. This impact is shown in Table 52. The multi-pressure stripper has a very low temperature level in the reboiler compared to the base case and is therefore adequate for lower IP/LP crossover pressure. The matrix stripping has a reboiler temperature between the base case and the multi-pressure stripper. The heat integrated stripping column has the highest reboiler temperature and is therefore suitable for higher IP/LP crossover pressures. The other modifications show slight increases in the reboiler temperature.

The available space for a retrofit is very site specific. The different flow sheet modifications are similar in the required space compared to the base case. A general conclusion on this point cannot be drawn.

The retrofit of a capture plant into an existing NGCC plant is more complicated than into an SCPC plant, because a flue gas recirculation has to be installed to enrich the CO_2 content. This will lead to an adaptation of the whole gas turbine which may not be applicable for a retrofit. For the water-steam-cycle of an NGCC plant similar behaviour like in an SCPC plant is expected.

Further issues are site specific limitations like water availability, environmental conditions, etc. The process has a neutral water balance, therefore the capture process itself does not need water in normal operation condition. However, the water availability is important for the cooling section and therefore lower cooling duties in the process flow sheet modifications are beneficial in this point. Environmental conditions influence the efficiency of the power plant significantly, especially the gas turbine efficiency, but this influence is found to be equal for all process modifications.

Table 52: Impacts of the key parameters on the different flow sheet modifications

Process flow sheet modification	Limitations from solvent properties	Operational flexibility in part load	Process control requirement	Retrofitting to an existing power plant
Case A1 (SCPC BC)	○	○	○	○
Case B1 (NGCC BC)	○	○	○	○
Case A2 (SCPC VR)	--	-	○	○
Case B2 (NGCC VR)	--	-	○	○
Case A3 (SCPC MPS)	-	--	-	++
Case B3 (NGCC MPS)	-	--	-	++
Case A4 (SCPC HIS)	-	○	○	--
Case B4 (NGCC HIS)	-	○	○	--
Case A5 (SCPC SF)	○	○	-	-
Case B5 (NGCC SF)	○	○	-	-
Case A6 (SCPC MS)	○	+	--	+
Case B6 (NGCC MS)	○	+	--	+
Case A7 (SCPC OHC HI)	○	+	-	-
Case B7 (NGCC OHC/RC HI)	○	+	-	-
Case A8 (SCPC VR + SF)	-	-	-	-
Case B8 (NGCC VR + SF)	-	-	-	-
Case A9 (SCPC HIS + OHC HI)	-	+	-	-
Case B9 (NGCC HIS + OHC HI)	-	+	-	-

Notes:

++: very positive +: positive ○ : neutral -: negative --: very negative

SCPC cases were evaluated with advanced heat integration

BC: base case, VR: vapour recompression, MPS: multi-pressure stripper, HIS: heat integrated stripper, SF: split flow, MS: matrix stripping, OHC HI: overhead condenser heat integration, RC HI: reboiler condensate heat integration

8 Economic Evaluation

After the technical evaluation of different PCC process flow sheet modifications, it is necessary to investigate these process modifications from an economic point of view. This will enable taking account not only of the efficiency increase chances but also of the costs connected to them. The economic evaluation has been conducted regarding the additional capital costs of the CO_2 capture plant due to the major equipment items as well as the connected costs for instrumentation and controls, piping, electrical equipment, etc. The capital costs have been broken down into equipment, installation and further direct costs, indirect costs such as engineering and supervision, construction expenses etc. as well as financial costs like profit, contingency and interest costs.

Furthermore, annual operating costs of the PCC process have been taken into account and broken down into the main items as shown in Section 8.1.

For each process flow sheet modification two economic indicators have been calculated: the Cost of Electricity CoE in €/MWh and the cost of CO_2 avoidance in €/t_{CO2}. These figures allow for a direct comparison with the reference coal and natural gas power plants without CO_2 capture.

8.1 Evaluation Procedure

In this section the economic evaluation procedure is shown as an example for the base case of a CO_2 capture plant in combination with an SCPC power plant. The procedure is kept the same for all process flow sheet modifications provided thereafter. Data for each flow sheet modification with relevant details and differences are presented in Section 8.2.

8.1.1 Capital costs (CAPEX)

The first step for the evaluation of capital costs consists in drawing up a list of equipment for the CO_2 capture plant, which is shown in Table 53.

Table 53: List of Equipment for the base case of a capture plant in combination with an SCPC power plant (for one train)

List of Equipment	Ref. 2010								
Number of absorber trains	2								
Component	Type	Material (S: steal; SS stainless steal; CI: cast iron)	Number per train	Total number	Reference value	PEC per train (€)	PEC (k€)	Installation Costs per train (k€)	Installation Costs (k€)
Absorber shell	incl. collectors and distributors	S	1	2	547989 kg	1,272,409	2,545	954	1,909
Absorber packing	Mellapak Plus 252 Y	-	1	2	2433 m3	3,034,986	6,070	1,214	2,428
Absorber extras	Platforms and ladders	-	1	2	17.6 m	173,071	346	69	138
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	4	8	0.928 m3/s	520,707	1,041	221	443
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	4	8	606 kW	181,946	364	73	146
RL heat exchanger	Plates, sealed	SS	29	58	1075 m2	1,172,997	2,346	528	1,056
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	4	8	971 m2	149,310	299	67	134
Desorber overhead condenser	Pipe bundle	SS	1	2	648 m2	157,460	315	71	142
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.685 m	19,045	38	8	15
Desorber shell	incl. collectors and distributors	S	1	2	288387 kg	755,917	1,512	567	1,134
Desorber packing	Mellapak Plus 252 Y	-	1	2	650 m3	811,361	1,623	325	649
Desorber extras	Platforms and ladders	-	1	2	9.1 m	85,820	172	34	69
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	7	14	2859 m2	2,903,183	5,806	1,306	2,613
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	1001 m2	408,539	817	184	368
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.073 m3/s	23,671	47	10	20
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	70 kW	8,953	18	4	7
Activated-C filter	Inlet filter	SS	2	4	77 m2	237,009	474	172	344
Mechanical filter	Vertical plates	SS	4	8	77 m2	29,925	60	22	43
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	31 m3	65,833	132	26	53
Surge tank	Small field errected tank, incl. stairs etc.	S	5	10	1077 m3	796,907	1,594	319	638
ID fan	Axial fan with guide vane	S	1	2	328 m3/s	1,847,421	3,695	831	1,663
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	3627 kW	96,580	193	39	77
Heater to stack	Gasketed plate & frame	S	2	4	631 m2	26,577	53	12	24
DCC	incl. collectors and distributors	S	1	2	277411 kg	732,939	1,466	550	1,099
DCC surfaces	Plates, sealed	S	3	6	790 m2	47,571	95	21	43
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.558 m3/s	32,838	66	14	28
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	66 kW	17,124	34	7	14
Washing section (cooler)	Plates, sealed	S	1	2	93 m2	2,940	6	1	3
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.050 m3/s	15,026	30	6	13
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9 kW	3,389	7	1	3
Intercooler	Plates, sealed	S	3	6	752 m2	45,784	92	21	41
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	4	8	0.928 m3/s	77,475	155	33	66
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	4	8	60 kW	31,527	63	13	25
CO2 compressor	Integrally geared, 6 stages, intercooled incl. Driving engine	-	4	8	19.91 kg/s	15,429,312	30,859	6,943	13,886
Overall PEC (2010)							**62,431**		**29,331**
					Year of cost analysis		2010		
					CEPCI (2012)		1.04		
Overall PEC (2012)							**64,787**		**30,438**

In the List of Equipment all items are listed along with a short description, the number of components per parallel train of CO_2 capture units and the total number of components needed. The table shows a reference value for each item, which derives from the thermodynamical dimensioning of the components of the CO_2 capture plant on the basis of the simulations.

For each item the Purchased Equipment Costs (PEC) and Installation Costs are calculated on the basis of the reference value using cost correlations to be found in the literature for different components:

- absorber and desorber columns with extras, DCC [44]
- absorber and desorber packing [45]
- all remaining components (pumps, fans, electrical motors, heat exchangers, etc.) [46]

Due to the fact that most of the price information is only available in US$, a conversion factor for US$/€ has been taken into account. Since the correlations used are valid only for a reference year, different conversion factors were used depending on the reference year. Afterwards, the calculated costs are corrected using the cost index *Chemical Engineering Chemical Plant Index* (CECPI) which is published monthly on the journal *Chemical Engineering*.

As a result, the total Purchased Equipment Costs and Installation Costs for the CO_2 capture plant have been calculated.

Capital costs include additional direct and indirect costs, which are calculated scaling the PEC with appropriate factors as shown in Table 54. As a result of this calculation the Total Plant Costs (TPC) are determined. In addition to that, interest costs as well as start-up expense, owners' costs and spare parts costs are taken into account using scaling factors on the base of TPC. The final result is the Total Capital Requirement (TCR), which is equivalent to CAPEX.

Table 54: Capital costs calculation for the base case of a capture plant in combination with an SCPC power plant

	CAPEX			
		Base	*Factor*	*Result (Mio €)*
	Purchased Equipment Cost (PEC)			**64.79**
	Purchased-Equipment Delivery	PEC	0.10	6.48
Direct Costs	**Purchased Equipment Delivered Costs (PEDC)**			**71.27**
	Installation	PEDC		30.44
	Instrumentation and Controls	PEDC	0.43	30.79
	Piping	PEDC	0.68	48.46
	Electrical Equipment and Materials	PEDC	0.20	14.11
	Buildings	PEDC	0.12	8.20
	Yard Improvements	PEDC	0.10	7.13
	Service Facilities	PEDC	0.30	21.38
Indirect Costs	**Total Direct Costs (TDC)**			**231.77**
	Engineering and Supervision	PEDC	0.33	23.52
	Construction Expenses	PEDC	0.41	29.22
	Contractor's Fee	PEDC	0.04	2.85
	Total Indirect Costs (TIC)			**55.59**
	Profit	TDC+TIC	0.05	14.37
	Contingency	TDC+TIC	0.10	28.74
	Total Plant Cost (TPC)			**330.46**
	Interest During Construction	Year 1	Year 2	Year 3
	Expenditure Schedule	20%	45%	35%
	Project Costs	66.09	148.71	115.66
	Interest During Construction	6.03	20.13	32.43
	Funding Requirement	72.12	168.83	148.09
	Sum			**389.05**
	Startup Expense	FCI	0.06	19.83
	Owners Costs	FCI	0.07	24.52
	Spare Parts	TPC	0.005	1.65
	Total Capital Requirement (TCR, CAPEX)			**435.05**

8.1.2 Annual operating costs (OPEX)

Annual operating costs include the cost of the consumables (such as cooling water make-up, solvent make-up, etc.), maintenance and repairs, operating labour, taxes, insurance and administrative costs. All these contributions are estimated through scaling factors. The single items and the Total Operating Expenses (OPEX) are presented in Table 55.

Table 55: Annual operating costs calculation for the base case of a capture plant in combination with an SCPC power plant

OPEX					
		Base	*Factor*	*Consumable Amount*	*Result (Mio €/year)*
Consumables	Cooling water make up	1 m³/GJ_th; 0.2 € / m³	0.2 € / GJ_th	11666122.6 GJ_th/yr	2.33
	Solvent make up	1.5 kg solvent / t CO2; 1.5 € / kg solvent	2.25 € / t_CO2	4269595.97 t_CO2/yr	9.61
	Inhibitor make up	solvent make-up cost	0.2		1.92
	NaOH make up	0.13 kg NaOH / t CO2; 350 € / t NaOH	0.0455 € / t_CO2	4269595.97 t_CO2/yr	0.19
	Activated C consumption	0.075 kg C / t CO2; 4230 € / t C	0.31725 € / t_CO2	4269595.97 t_CO2/yr	1.35
Maintenance and Repairs (M)		TPC	0.015		4.96
Operating Labor (OL)		18 Technicians			1.08
Direct Supervisory and Clercial Labor (SL)		OL	0.3		0.32
Operating Supplies		M	0		0.00
Laboratory Charges		OL	0		0.00
Variable (Direct) Operating Costs					**21.77**
Plant Overhead Costs		**M + OL + SL**	**0.25**		**1.59**
Taxes		TPC	0.005		1.65
Insurance		TPC	0.005		1.65
Fixed Charges					**3.30**
Administrative Costs		M	0.12		0.59
Distribution and Marketing Costs		OPEX	0		0.00
R&D Costs		OPEX	0		0.00
General Expenses					**0.59**
Total Operating Expenses (OPEX)					**27.26**

8.1.3 Cost of Electricity

The Cost of Electricity expresses the cost of the production of one MWh (€/MWh). The Cost of Electricity for a power plant with CO_2 capture consists of 5 contributions:

$$CoE = CoE_{ref} + \Delta CoE_{output} + \Delta CoE_{CAPEX} + \Delta CoE_{OPEX} + \Delta CoE_{T\&S}$$

With:

- ΔCoE_{ref} Cost of Electricity of the reference power plant without CO_2 capture;
- ΔCoE_{output} increase of the Cost of Electricity due to the decrease of net power output;
- ΔCoE_{CAPEX} increase of the Cost of Electricity due to additional capital costs;
- ΔCoE_{OPEX} increase of the Cost of Electricity due to additional operating costs;
- $\Delta CoE_{T\&S}$ increase of the Cost of Electricity due to transport and storage costs of the captured CO_2 (10€/tonne CO_2 stored).

The last 4 terms of the equation are calculated as follows:

$$\Delta CoE_{output} = CoE_{ref}\left(\frac{P_{ref}}{P} - 1\right)$$

$$\Delta CoE_{CAPEX} = \frac{\alpha \cdot CAPEX}{P \cdot t}$$

$$\alpha = \frac{i(1+i)^{t_{PL}}}{(1+i)^{t_{PL}} - 1}$$

$$\Delta CoE_{OPEX} = \frac{OPEX}{P \cdot t}$$

$$\Delta CoE_{T\&S} = CoT\&S \cdot (e_{CO_2} - e_{CO_2,ref})$$

With:

- P, P_{ref}: net power output of the power plant with CO_2 capture and of the reference power plant w/o CO_2 capture;
- α: annuity factor, which describes a linear amortisation over the Project Lifetime t_{PL} with an interest rate *i*;
- t: power plant operating hours per year;
- CAPEX, OPEX: total capital costs and annual operating costs as described in Sections 8.1.1 and 8.1.2;
- CoT&S: Cost of Transport&Storage of the captured CO_2 in €/t;
- e_{CO2}, $e_{CO2,ref}$: specific CO_2 emissions of the power plant with CO_2 capture and of the reference power plant w/o CO_2 capture in t/MWh.

8.1.4 Cost of CO_2 avoidance

The cost of CO_2 avoidance expresses the financial effort necessary to avoid a ton of CO_2. It is calculated as follows:

$$c_{CO_2,avoided} = \frac{CoE - CoE_{ref}}{e_{CO_2,ref} - e_{CO_2}}$$

8.2 Economic Evaluation of Process Flow Sheet Modifications

In this section each process flow sheet modification will be evaluated from the economic point of view, highlighting which additional equipment items are required along with their costs and which repercussions the modified process will produce on the costs of the items of the base case capture plant. These changes affect the Purchased Equipment Costs. Due to the fact that CAPEX are directly proportional to PEC, the contribution factor ΔCoE_{CAPEX} is also proportional to PEC. Moreover, differences with regard to OPEX will be presented, affecting the contribution term ΔCoE_{OPEX}, as well as with regard to the power plant net efficiency, which affects the contribution terms ΔCoE_{output} and $\Delta CoE_{T\&S}$ and the specific CO_2 emissions e_{CO2}.

In this way the influence of each flow sheet modification on the economic indicators CoE and costs of CO_2 avoidance will be explained and justified.

8.2.1 SCPC power plant

Relevant economic data for the calculation of the different flow sheet modification economic indicators are shown in Table 56.

Table 56: Economic data for the SCPC power plant

Project Life Time	**t_{PL}**	25	yr
Interest Rate	**i**	8	%
Specific Capital Investment		1,700	€ / kW_{el} (net)
Operating hours per year	**t**	7,446	h / yr
Fuel Price		2.4	€/GJ
Man power		80	-
Labour cost		60,000	€ / (man yr)
Cost of Electricity (w/o capture)	**CoE_{ref}**	42.22	€/MWh

8.2.1.1 Base Case

In the base case for the capture plant in combination with an SCPC power plant the major equipment costs are represented by the following components, accounting for about 75% of the Purchased Equipment Costs (cf. Table 53):

- CO_2 compressor (49% of PEC)
- absorber packing (9.7% of PEC)
- reboiler (9.3% of PEC)
- ID fan (5.9% of PEC)

The other components account for less than 5% of PEC each. OPEX amount to 27.26 M€/yr. This results in Cost of Electricity for the base case of 68.29 €/MWh and cost of CO_2 avoidance of 38.32 €/t CO_2.

An alternative Base Case capture plant with higher desorber pressure has also been investigated. Higher costs for desorber shell, rich/lean heat exchanger and rich solution and intercooler pump motors lead to higher CAPEX (+2.8%). The net efficiency penalty of the process is also higher than for the Base Case, resulting in higher CoE (68.93 €/MWh) and costs of CO_2 avoidance (39.31 €/t_{CO2}).

8.2.1.2 Vapour recompression

This modification requires an additional tank for the production of flash vapour and the separation from the liquid fraction as well as an additional centrifugal compressor in order to re-inject the vapour into the desorber.

The flash tank contributes in a very limited manner to the additional costs, representing only 0.1% of capture plant PEC. The additional compressor is causing an increase of the capture plant PEC by 2.3%. The dimensioning of the other equipment items in the modified process is almost unchanged in comparison to the base case and yields no cost reduction potential.

Higher CAPEX along with almost unchanged OPEX (27.45 M€/yr) and an only slightly better net efficiency (39.14% instead of 39.12%) lead to higher CoE (68.43 €/MWh) and costs of CO_2 avoidance (38.54 €/t_{CO2}) in comparison to the base case.

8.2.1.3 Multi-pressure stripper

The additional equipment items for this modification consist of two desorber columns (with packing and extras) and two centrifugal compressors raising the pressure of the CO_2 vapour from lower- to higher-pressure desorber in two stages.

Centrifugal compressors are the most expensive additional items, accounting for 7.4% of the capture plant PEC, while desorber columns (w/o reboiler and reclaimer) account for 7.8% of PEC instead of 5.3% for the base case.

The other equipment items require a smaller dimensioning, thus producing lower PEC. However, the high costs of the additional equipment outweigh these savings. Moreover, OPEX rise to 28.35 M€/yr and the net efficiency decreases, so that this modification turns out to be the most expensive among the modifications considered in this study.

8.2.1.4 Heat-integrated stripping column

This modification only requires an additional heat exchanger, the stripper interheater. The shifting of the heat exchange from the rich/lean heat exchanger (RLHX) to the stripper interheater, however, leads to a smaller dimensioning of the RLHX. Equipment costs for the RLHX (base case) equal thus the sum of the costs for RLHX and interheater in the modified process. The rest of the equipment requires the same dimensioning, so that in the end PEC for this modification are unvaried in comparison to the base case. Also OPEX are unvaried.

The slightly higher values for CoE (68.39 €/MWh) and costs of CO_2 avoidance (38.48 €/t_{CO2}) in comparison to the base case (as shown in Table 58) are merely due to the inferior net efficiency of the modified process.

8.2.1.5 Improved split flow process

This modification requires no additional equipment. Moreover, the following components require a smaller dimensioning, leading to lower capture plant PEC:

- rich solution pump and motor
- rich/lean heat exchanger
- desorber overhead condenser, condensate return tank
- reboiler and reclaimer
- reboiler condensate pump and motor
- filters

For this reason and due to lower OPEX (26.81 M€/yr) and a better net efficiency, the modified process yields a cost reduction potential of 0.9%-points in comparison to the base case.

8.2.1.6 Matrix stripping

In this case, three desorber columns are needed instead of a single one. The three columns are connected in parallel, so that additional equipment items consist not only of two desorber columns, but also two additional reboilers and reclaimers as well as desorber overhead condensers and condensate return tanks. The sum of this group of items accounts for 19.2% of the capture plant PEC, in comparison to the 16.5% for the base case.

The rest of the equipment requires the same dimensioning as for the base case, yielding no cost reduction potential.

Due to higher PEC and higher OPEX (27.60 M€/yr) as well as a lower net efficiency, the CoE and the costs of CO_2 avoidance (respectively 68.95 €/MWh and 39.33 €/t_{CO2}) are higher for this modification than for the base case. This modification represents the second most expensive (in terms of CoE increase) modification among the modifications presented in this study for the SCPC power plant, after the multi-pressure stripper modification.

8.2.1.7 OHC heat integration

This modification only requires an additional overhead condenser/rich solution heat exchanger switched in parallel to the RLHX. It accounts for only 0.9% of the capture plant PEC.

Furthermore, the following equipment items require a smaller dimensioning than for the base case, leading to lower prices:

- rich/lean heat exchanger
- desorber overhead condenser
- reboiler and reclaimer

The capture plant PEC are lower than for the base case. Together with lower OPEX (26.74 M€/yr) and a higher net efficiency the modified process leads to lower CoE and costs of CO_2 avoidance (respectively 67.65 €/MWh and 37.35 €/t_{CO2}) in comparison to the base case.

8.2.1.8 Vapour recompression + split flow

This modification is a combination of vapour recompression and improved split flow process modification presented previously. Compared to the base case, higher equipment costs for the additional flash tank and flash vapour compressor are outweighed by lower costs for the equipment items listed in section 8.2.1.5. The OPEX for the combination compared to the split flow process are slightly increased, but in the end the modified process shows lower CoE (67.78 €/MWh) and costs of CO_2 avoidance (37.57 €/t_{CO2}) in comparison to the split flow process alone.

8.2.1.9 Heat-integrated stripper + OHC heat integration

The combination of heat integration into the stripper and the integration of the overhead condenser requires two additional heat exchangers: the stripper interheater and the overhead condenser/rich solution HX. Compared to the base case, higher equipment costs due to additional items are outweighed by lower costs for the equipment items listed in Section 8.2.1.7. Compared to the process with OHC heat integration alone, the capture plant PEC as well as the OPEX are almost the same, but a lower net efficiency leads to higher CoE and costs of CO_2 avoidance (respectively 67.71 €/MWh and 37.45 €/t_{CO2}).

8.2.1.10 SCPC power plant flow sheet modifications overview

Table 57 gives an overview of the additional equipment and modified main equipment for each flow sheet modification along with the variations of the PEC in comparison to the Base Case. For a complete listing please refer to Table 65 to Table 73 in the Appendix.

Table 57: Additional equipment and modified main equipment with relative PEC variations for SCPC power plant flow sheet modifications

Modification	Additional equipment	PEC (k€)	Modified main equipment	PEC difference to Base Case (k€)
Vapour recompression	Flash tank	71	Rest	-343
	Lean vapour CO2-Compressor	1452		
	Subtotal	1523		-343
	Total	**1180**		
Multi-pressure stripper	Desorber2 shell	1174	Desorber1 shell	-451
	Desorber2 packing	547	Desorber1 packing	-986
	Desorber2 extras	133	Desorber1 extras	-35
	Desorber3 shell	1041	Reboiler	-79
	Desorber3 packing	547	Reclaimer	-11
	Desorber3 extras	133	Rest	-215
	Multi-stripper compressor2	2469		
	Multi-stripper compressor3	2687		
	Subtotal	8731		-1778
	Total	**6953**		
Heat-integrated stripping column	Stripper interheater	64	Rest	-100
	Subtotal	64		-100
	Total	**-36**		
Improved split flow process	No additional equipment	0	Solvent pump (rich)	-21
			Solvent pump motor (rich)	-10
			RL heat exchanger	-89
			Desorber overhead condenser	-133
			Condensate return tank	-24
			Reboiler	-794
			Reclaimer	-75
			Condensate pump	-2
			Condensate pump motor	-2
			Activated-C filter	-21
			Mechanical filter	-3
			Rest	-79
	Subtotal	0		-1252
	Total	**-1252**		
Matrix stripping	Desorber overhead condenser2	97	Desorber overhead condenser1	-137
	Condensate return tank2	8	Condensate return tank1	-20
	Desorber2 shell	559	Desorber1 shell	-737
	Desorber2 packing	593	Desorber1 packing	-1143
	Desorber2 extras	111	Desorber1 extras	-61
	Reboiler2	543	Reboiler1	-5130
	Reclaimer2	144	Reclaimer1	-638
	Overhead condenser3	197	Rest	-160
	Reflux drum3	20		
	Desorber3 shell	854		
	Desorber3 packing	1102		
	Desorber3 extras	145		
	Reboiler3	4863		
	Reclaimer3	720		
	Subtotal	9955		-8026
	Total	**1929**		

Modification	Additional equipment	PEC (k€)	Modified main equipment	PEC difference to Base Case (k€)
OHC heat integration	OH rich split heat exchanger	571	RL heat exchanger	-124
			Desorber overhead condenser	-225
			Reboiler	-987
			Reclaimer	-104
			Rest	-188
	Subtotal	571		-1628
	Total	**-1057**		
Vapour recompression + split flow	Flash tank	71	Absorber shell	-92
	Lean vapour CO2-Compressor	201	Absorber packing	-340
			Surge tank	-315
			Solvent pump (rich)	-74
			Solvent pump motor (rich)	-33
			RL heat exchanger	-326
			Desorber overhead condenser	-142
			Condensate return tank	-25
			Reboiler	-838
			Reclaimer	-82
			Condensate pump	-2
			Condensate pump motor	-2
			Activated-C filter	-72
			Mechanical filter	-9
			Rest	25
	Subtotal	273		-2328
	Total	**-2055**		
Heat-integrated stripping column + OHC heat integration	Stripper interheater	64	Absorber shell	-110
	OH rich split heat exchanger	737	Absorber packing	-407
			Surge tank	-351
			RL heat exchanger	-427
			Desorber overhead condenser	-220
			Reboiler	-1043
			Reclaimer	-112
			Rest	-275
	Subtotal	801		-2945
	Total	**-2144**		

Table 58 gives a summary of the Cost of Electricity and of CO_2 avoidance costs for the SCPC modifications obtained from the evaluation process. CoE_{ref} is the original value for the SCPC power plant without CO_2 separation and stated for better comparability.

As can be seen, CO_2 separation increases the CoE relatively by 60.2 to 64.7%. The base case shows an increase of 61.7%, which can be converted to CO_2 avoidance costs of 38.32 €/t_{CO2}. The process modifications vapour recompression, multi-pressure stripper, heat-integrated stripping column and matrix stripping show even higher increases of the CoE and of the CO_2 avoidance costs, respectively, with multi-pressure stripper being by far the most expensive one (40.17 €/t_{CO2}).

The process modifications improved split flow process, OHC heat integration, vapour recompression + split flow and heat-integrated stripper + OHC heat integration yield cost reduction potential compared to the base case. The first combination of process modifications benefits from both single modifications, showing a bigger cost reduction potential than the single modifications alone. The second combination of process modifications benefits from the cost reduction for the OHC heat integration, but results however more expensive than the OHC heat integration alone. The modification OHC heat integration shows the lowest CoE (67.71 €/MWh) and CO_2 avoidance costs (37.35 €/t_{CO2}).

Table 58: Economic indicators for SCPC power plant flow sheet modifications

	CoE_{ref}	CoE	relative change of CoE	$c_{CO2,avoided}$
	€/MWh	€/MWh	%	€/t_{CO2}
Base case	42.22	68.29	61.7%	38.32
Vapour recompression	42.22	68.43	62.1%	38.54
Multi-pressure stripper	42.22	69.53	64.7%	40.17
Heat-integrated stripping column	42.22	68.39	62.0%	38.48
Improved split flow process	42.22	67.87	60.7%	37.69
Matrix stripping	42.22	68.95	63.3%	39.33
OHC heat integration	42.22	67.65	60.2%	37.35
Vapour recompression + split flow	42.22	67.78	60.5%	37.57
Heat-integrated stripper + OHC heat integration	42.22	67.71	60.4%	37.45

Moreover, a detailed overview of the net efficiency penalty, CAPEX, OPEX and the different contributions to the total CoE for SCPC power plant flow sheet modifications is offered in Table 63.

8.2.2 NGCC power plant

Relevant economic data for the calculation of the different flow sheet modification economic indicators are shown in Table 59.

Table 59: Economic data for the NGCC power plant

Project Life Time	t_{PL}	25	yr
Interest Rate	**i**	8	%
Specific Capital Investment		750	€ / kW_{el} (net)
Operating hours per year	**t**	7,446	h / yr
Fuel Price		7.5	€/GJ
Man power		80	-
Labour cost		60,000	€ / (man yr)
Cost of Electricity (w/o capture)	CoE_{ref}	59.50	€/MWh

8.2.2.1 Base Case

In the base case for the capture plant in combination with an NGCC power plant the major equipment costs are represented by the following components, accounting for about 75% of the capture plant PEC:

- CO_2 compressor (46% of PEC)
- absorber packing (10.1% of PEC)
- reboiler (7.9% of PEC)
- ID fan (7.8% of PEC)

The other components account for less than 5% of PEC each. Note that the cost-setting equipment is the same as for the SCPC power plant, although the numbers are slightly different. OPEX amounts to 16.21 M€/yr. This results in Cost of Electricity for the base case of 76.82 €/MWh and cost of CO_2 avoidance of 54.76 €/t CO_2.

An alternative Base Case capture plant with higher desorber pressure has also been investigated. As in the combination with the SCPC power plant, higher costs for main equipment and a lower net efficiency yield no cost reduction potential in comparison with the Base Case capture plant with lower desorber pressure.

8.2.2.2 Vapour recompression

This process modification requires an additional tank for the production of flash vapour and the separation from the liquid fraction as well as an additional centrifugal compressor in order to re-inject the vapour into the desorber.

The flash tank contributes in a very limited manner to the additional costs, representing only 0.1% of PEC. The additional compressor causes an increase of the capture plant PEC by 3.9%. Since the lean solution downstream the desorber is throttled to a lower pressure, an additional solvent pump (lean) is necessary. The dimensioning of the other equipment items in the modified process is almost unchanged in comparison to the base case, with exclusion of the rich-lean heat exchanger and of the heater to stack, which require a smaller dimensioning. However these benefits do not compensate the costs of additional equipment.

Higher PEC (and thus higher CAPEX) along with higher OPEX (16.51 M€/yr) are not outweighed by the increase of net efficiency, leading to higher CoE and costs of CO_2 avoidance (respectively 76.99 €/MWh and 55.27 €/t_{CO2}) in comparison to the base case.

8.2.2.3 Multi-pressure stripper

The additional equipment items for this modification consist of two desorber columns (with packing and extras) and two centrifugal compressors raising the pressure of the CO_2 vapour from the lower-pressure to the higher-pressure desorber in two stages.

Centrifugal compressors are the most expensive additional items, accounting for 9.8% of the capture plant PEC, while desorber columns (w/o reboiler and reclaimer) account for 6.1% of PEC instead of 4.7% for the base case.

Again, an additional lean solution pump is necessary, which compensates any cost benefits. The other equipment items require a smaller dimensioning, thus producing lower PEC. The high costs of the additional items outweigh however these savings. Furthermore, OPEX raise to 17.05 M€/yr. Under these conditions the small increase of net efficiency achieved with this modification does not outweigh the higher costs, so that this modifications comes out as the most expensive one considered in this study for the NGCC power plant.

8.2.2.4 Heat-integrated stripping column

This modification requires only an additional heat exchanger, the stripper interheater. The shifting of the heat transfer from the rich/lean heat exchanger (RLHX) to the stripper interheater leads to a smaller dimensioning of the RLHX. In total the sum of the costs for RLHX and interheater in the modified process are lower than the single RLHX for the base case. Moreover, the rest of the equipment requires a smaller dimensioning, leading to lower PEC for this modification. OPEX also decrease.

The slightly lower values for CoE (76.51 €/MWh) and costs of CO_2 avoidance (53.77 €/t_{CO2}) in comparison to the base case (as shown in Table 61) are merely due to lower CAPEX and OPEX, since the net efficiency of the modified process amounts to 51.55% - an increase of only 0.01%-points in comparison to the base case.

8.2.2.5 Improved split flow process

This modification requires no additional equipment. Moreover, following components require a smaller dimensioning, leading to lower capture plant PEC:

- rich solution pump and motor
- rich/lean heat exchanger
- desorber overhead condenser, condensate return tank
- reboiler and reclaimer
- reboiler condensate pump and motor
- filters
- heater to stack

For this reason and due to a better efficiency, the modified process shows lower CAPEX and OPEX (15.79 M€/yr) than the base case, thus yielding a cost reduction potential of 1.5%-points.

8.2.2.6 Matrix stripping

In this case, three desorber columns are needed instead of a single one. The three columns are connected in parallel, so that additional equipment items consist not only in two desorber columns, but also two additional reboilers and reclaimers as well desorber overhead condensers and condensate return tanks. The sum of this group of items accounts for 17.6% of the capture plant PEC, in comparison to the 14.5% for the base case. The rest of the equipment requires a bigger dimensioning as for the base case, yielding no cost reduction potential.

As the resulting CAPEX as well as OPEX (16.77 M€/yr) are higher than those of the base case and as the net efficiency is lower, the CoE and the costs of CO_2 avoidance (respectively 77.39 €/MWh and 56.57 €/t_{CO2}) for this modification are not beneficial.

8.2.2.7 OHC heat integration

This modification requires only an additional overhead condenser/rich solution heat exchanger switched in parallel to the RLHX. It accounts for only 1.0% of the final PEC.

Furthermore, following equipment items require a smaller dimensioning than for the base case, leading to lower prices:

- rich/lean heat exchanger
- desorber overhead condenser
- reboiler and reclaimer

The capture plant PEC are lower than for the base case, which results in lower CAPEX. The OPEX are also decreased (15.80 M€/yr). Together with a higher net efficiency of the modified process, this leads to lower CoE and costs of CO_2 avoidance (respectively 75.73 €/MWh and 51.21 €/t_{CO2}) in comparison to the base case.

8.2.2.8 Reboiler condensate integration

For this modification, a new heat exchanger is required, whose additional cost can be neglected. The reboiler can be designed smaller, but almost all other equipment costs are increased. Although the CAPEX and the OPEX (16.29 M€/yr) are slightly higher than those of the base case, the CoE is lower (76.73 €/MWh) due to a better net efficiency. This improvement is however very small, amounting to -0.1% of relative CoE change in comparison to the base case.

8.2.2.9 Vapour recompression + split flow

This modification is a combination of vapour recompression and improved split flow process modification presented previously. Compared to the base case, higher equipment costs for the additional flash tank and flash vapour compressor are outweighed by lower costs for the equipment items listed in Section 8.2.2.5.

The CAPEX and the OPEX (15.83 M€/yr) are lower than those of the base case. For this reason and due to higher net efficiency, the modified process shows lower CoE and costs of CO_2 avoidance (respectively 75.95 €/MWh and 51.94 €/t_{CO2}) in comparison to the base case. Still, the CoE and costs of CO_2 avoidance in comparison to the split flow process alone are not reduced.

8.2.2.10 Heat-integrated stripper + OHC heat integration

The combination of heat integration into the stripper and the integration of the overhead condenser requires two additional heat exchangers: the stripper interheater and the overhead condenser/rich solution HX. Higher equipment costs due to additional items are outweighed by lower costs for the equipment items listed in Section 8.2.2.7 as well as the heater to stack and surge tank. In total the capture plant PEC are lower than for the base case but not lower than for OHC heat integration alone.

Both, CAPEX and OPEX (15.80 M€/yr) are lower, the net efficiency is higher than for the base case, so that the overall CoE (75.80 €/MWh) and costs of CO_2 avoidance (51.46 €/t_{CO2}) are lower. But as for the SCPC case, the CoE and costs of CO_2 avoidance in comparison to the OHC heat integration alone are not reduced.

8.2.2.11 NGCC power plant flow sheet modifications overview

Table 60 gives an overview of the additional equipment and modified main equipment for each flow sheet modification along with the variations of the PEC in comparison to the Base Case. For a complete listing please refer to Table 74 to Table 83 in the Appendix.

Table 60: Additional equipment and modified main equipment with relative PEC variations for NGCC power plant flow sheet modifications

Modification	Additional equipment	PEC (k€)	Modified main equipment	PEC difference to Base Case (k€)
Vapour recompression	Flash tank	71	RL heat exchanger	-165
	Lean vapour CO2-Compressor	2110	Heater to stack	-699
	Solvent pump (lean)	751	Rest	-181
	Solvent pump motor (lean)	40		
	Subtotal	2972		-1044
	Total	**1928**		
Multi-pressure stripper	Desorber2 shell	526	Desorber1 shell	-381
	Desorber2 packing	270	Desorber1 packing	-592
	Desorber2 extras	95	Desorber1 extras	-35
	Desorber3 shell	646	Reboiler	-74
	Desorber3 packing	270	Reclaimer	-12
	Desorber3 extras	95	Rest	-875
	Multi-stripper compressor2	1627		
	Multi-stripper compressor3	2900		
	Solvent pump (lean)	751		
	Solvent pump motor (lean)	89		
	Subtotal	7268		-1970
	Total	**5298**		
Heat-integrated stripping column	Stripper interheater	153	RL heat exchanger	-386
			Rest	-1803
	Subtotal	153		-2188
	Total	**-2036**		
Improved split flow process	No additional equipment	0	Solvent pump (rich)	-18
			Solvent pump motor (rich)	-8
			RL heat exchanger	-77
			Desorber overhead condenser	-112
			Condensate return tank	-18
			Reboiler	-348
			Reclaimer	-59
			Condensate pump	-2
			Condensate pump motor	-1
			Activated-C filter	-13
			Mechanical filter	-2
			Heater to stack	-740
			Rest	-140
	Subtotal	0		-1537
	Total	**-1537**		

Modification	Additional equipment	PEC (k€)	Modified main equipment	PEC difference to Base Case (k€)
Matrix stripping	Desorber overhead condenser2	79	Desorber overhead condenser1	-151
	Condensate return tank2	6	Condensate return tank1	-20
	Desorber2 shell	380	Desorber1 shell	-473
	Desorber2 packing	433	Desorber1 packing	-680
	Desorber2 extras	97	Desorber1 extras	-54
	Reboiler2	285	Reboiler1	-2923
	Reclaimer2	76	Reclaimer1	-466
	Overhead condenser3	183	Rest	996
	Reflux drum3	19		
	Desorber3 shell	493		
	Desorber3 packing	753		
	Desorber3 extras	123		
	Reboiler3	3.147		
	Reclaimer3	531		
	Subtotal	6605		-3772
	Total	**2832**		
OHC heat integration	OH rich split heat exchanger	399	RL heat exchanger	-148
			Desorber overhead condenser	-170
			Reboiler	-725
			Reclaimer	-81
			Rest	-318
	Subtotal	399		-1442
	Total	**-1043**		
Reboiler condensate integration	Reboiler condensate heat exchanger	20	Reboiler	-66
			Reclaimer	-11
			Rest	397
	Subtotal	20		320
	Total	**340**		
Vapour recompression + split flow	Flash tank	71	Absorber shell	-17
	Lean vapour CO2-Compressor	123	Absorber packing	-57
			Surge tank	-33
			Solvent pump (rich)	-18
			Solvent pump motor (rich)	-8
			RL heat exchanger	-96
			Desorber overhead condenser	-116
			Condensate return tank	-19
			Reboiler	-339
			Reclaimer	-57
			Condensate pump	-2
			Condensate pump motor	-1
			Activated-C filter	-13
			Mechanical filter	-2
			Heater to stack	-713
			Rest	-29
	Subtotal	195		-1519
	Total	**-1324**		
Heat-integrated stripping column + OHC heat integration	Stripper interheater	64	Surge tank	-65
	OH rich split heat exchanger	349	RL heat exchanger	-199
			Desorber overhead condenser	-172
			Reboiler	-690
			Reclaimer	-75
			Heater to stack	-740
			Rest	414
	Subtotal	413		-1526
	Total	**-1112**		

Table 61 gives a summary of the Cost of Electricity for the NGCC modifications obtained from the evaluation process. CoE_{ref} is the original value for the NGCC power plant without CO_2 separation and stated for better comparability.

As can be seen, CO_2 separation increases the CoE relatively by 27.3 to 30.2%. The base case shows an increase of 29.1%, which can be converted to CO_2 avoidance costs of 54.76 €/t_{CO2}. The process modifications vapour recompression, multi-pressure stripper and matrix stripping show even higher increases of the CoE and of the CO_2 avoidance costs, respectively, with multi-pressure stripper being the most expensive one by far (56.76 €/t_{CO2}).

The process modifications heat-integrated stripping column, improved split flow process, OHC heat integration, vapour recompression + split flow, heat-integrated stripper + OHC heat integration and reboiler condensate integration yield cost reduction potential compared to the base case. The modification OHC heat integration shows the lowest avoidance costs (51.21 €/t_{CO2}).

Table 61: Economic indicators for NGCC power plant flow sheet modifications

	CoE_{ref}	CoE	relative change of CoE	$c_{CO2,avoided}$
	€/MWh	€/MWh	%	€/t_{CO2}
Base case	59.50	76.82	29.1%	54.76
Vapour recompression	59.50	76.99	29.4%	55.27
Multi-pressure stripper	59.50	77.46	30.2%	56.76
Heat-integrated stripping column	59.50	76.51	28.6%	53.77
Improved split flow process	59.50	75.92	27.6%	51.85
Matrix stripping	59.50	77.39	30.1%	56.57
OHC heat integration	59.50	75.73	27.3%	51.21
Reboiler condensate integration	59.50	76.73	29.0%	54.46
Vapour recompression + split flow	59.50	75.95	27.6%	51.94
Heat-integrated stripper + OHC heat integration	59.50	75.80	27.4%	51.46

Moreover, a detailed overview of the net efficiency penalty, CAPEX, OPEX and the different contributions to the total CoE for NGCC power plant flow sheet modifications is offered in Table 64.

9 Identification of Gaps and Future Recommendations

The process modifications analysed in this study are suitable for the application of post combustion capture in power plants. The reliability has to be investigated for all process modifications to ensure that no negative implications on the power plant process occur.

One of the most challenging point is the development of a solvent with a very good performance. This solvent has to be tested in pilot plants and it is necessary to develop an exact property model of the solvent which describes the solvent with the effects of all process modifications. Also a very good behaviour in degradation and corrosion is necessary. A solvent with a very good energetic performance is not applicable when the tendency for degradation is very high.

The interaction between different solvents and process modifications is the crucial point. While some solvents with a certain modification can show an improvement in efficiency other solvents might not reach this improvement.

The different process modifications have to be realised in pilot plants and the reliability of the process modification has to be high to ensure an application in power plants. In certain campaigns long-time tests in pilot plants with flue gas of power plants have to be done to estimate the behaviour in operation.

To evaluate the process modifications it is very important to do an overall process analysis. A number of process modifications leads to a reduced specific heat duty, while the reboiler temperature is increased, resulting in no positive effect on the overall process.

The sensitivity of the logarithmic temperature difference of the RLHX is very high [7]. For all modifications a temperature difference of 5 K is set. This could lead to very large heat exchangers but this is technically feasible.

In the cases with vapour recompression the compressor which reintroduces the vapour into the column is a large electrical consumer. The efficiency and the operation regime of the component are relevant for the best operating point and the overall efficiency. This applies also to the multi-pressure stripping.

Limitations of the solvent can influence the process strongly and inhibit improvements. In the cases with the heat-integrated stripping column this is an important factor that affects the process. During the evaluation of solvents these limits have to be considered and solvents have to be improved from this point of view.

There is a possibility of various numbers of improved split flow processes. In this study the most promising modification is analysed. For other solvents, different split flow processes might by more efficient.

In the matrix stripping case losses occur due to the fact that the CO_2 compressor inlet pressure is adapted to the lowest pressure. The CO_2 compression could be more efficient using the higher CO_2 outlet pressure of the desorber columns without throttling. The highest temperature of different reboilers specifies the steam pressure of the steam extraction and the steam to the reboiler with lower temperature is throttled down. The difference between the reboiler temperatures should thus be as small as possible.

In the heat integration cases the temperature differences in the heat exchangers define the usable heat and therefore it is an important key process parameter. A reduction of the temperature difference could improve the process. This is very important for processes with solvents which have significant higher specific energy demands.

The flue gas recirculation for the NGCC process is not state-of-the-art and therefore disputable but the use of the recirculation is necessary for the post combustion capture to increase the CO_2 content in the flue gas and improve the CO_2 absorption process. Without flue gas recirculation the CO_2 capture processes are not efficient and are leading to higher efficiency penalties. Therefore the development and improvement of an NGCC plant with flue gas recirculation is necessary.

In this study, the power plants are considered as Greenfield and the IP/LP crossover pressure is optimised for the full load nominal point. It could be necessary to evaluate the behaviour in part load and optimise the IP/LP crossover pressure for this operation regime. This could lead to a higher IP/LP crossover pressure in full load, but the efficiency in part load would be better.

10 Summary and Outlook

For this study, different process flow sheet modifications of post combustion CO_2 capture unit in combination with SCPC and NGCC power plant have been evaluated. A generic optimised solvent has been chosen including a solvent property model for the simulation of the process in ASPEN Plus® software. Reference plants for the SCPC and the NGCC plant were defined and simulated. For each SCPC and NGCC power plant, a CO_2 capture plant base case was simulated to have a common basis for all process modifications.

An energetic evaluation and optimisation has been performed for the following process flow sheet modifications:

- vapour recompression
- multi-pressure stripper
- interheated stripper
- split flow process
- matrix stripping
- overhead condenser heat integration
- reboiler condensate heat integration
- combination of vapour recompression and split flow process
- combination of interheated stripper and overhead condenser heat integration

The most important interface quantities specific heat duty, specific cooling duty, specific auxiliary power, reboiler temperature, and desorber pressure were obtained from the process energetic evaluation. These were used to conduct an overall process evaluation for every process flow sheet modification in order to quantify the influence of the modified CO_2 capture plant on the overall process performance. The overall efficiency penalty was used as a characteristic value to rate the effect on the overall process performance. This is defined as the difference between the net efficiency of the reference power plant and the net efficiency of a power plant equipped with a CO_2 capture plant incorporating the respective process flow sheet modification. The overall efficiency penalty for different process modifications is shown in Table 62.

Table 62: Overall efficiency penalty for the evaluated process flow sheet modifications

	SCPC case in %-points	NGCC case in %-points
Base case	6.11	5.93
Vapour recompression	6.09	5.86
Multi-pressure Stripper	6.25	5.86
Heat-integrated stripping column	6.18	5.92
Improved split flow process	5.99	5.46
Matrix stripping	6.41	6.04
Overhead condenser heat integration	5.84	5.28
Reboiler condensate heat integration	-	5.83
Combination of vapour recompression and split flow process	5.99	5.46
Combination of interheated stripper and overhead condenser heat integration	5.88	5.34

The process with the lowest overall efficiency penalty is the overhead condenser heat integration. Compared to the base case, a reduction of the overall efficiency penalty by 0.37%-points for the SCPC case and 0.65%-points for the NGCC case compared to the base case was obtained. The results for the improved split flow process show a considerable reduction of the overall efficiency penalty, especially for the NGCC case. The other modifications do not improve the overall process, for some modifications the overall efficiency penalty is even higher compared to the base case. This was noticed in almost all process flow sheet modification cases due to the higher reboiler temperatures, making it necessary to use steam of a higher quality to heat the reboiler. This effect overcompensates the positive influence of the reduced specific heat duty, which was observed for almost all process flow sheet modifications. This illustrates the importance of an overall process evaluation.

A comparison of the results for SCPC and NGCC cases shows that the NGCC case generally benefits more from the process flow sheet modifications. This is mainly due to the fact that the SCPC base case is designed with a waste heat integration using heat from the overhead condenser and the CO_2 compressor for the preheating of the feed water. Modifications that reduce the temperature in the desorber head are thus less effective for the SCPC case since the amount of available waste heat is reduced.

It has to be noted that these results strongly depend on the properties of the selected solvent and as well as on the boundary conditions selected for the processes. Therefore a general conclusion regarding to the benefit of one of the process flow sheet modifications cannot be drawn. For a new solvent, a similar evaluation has to be performed to be able to rate the most potential process flow sheet modifications.

Especially for solvents with higher specific heat duties, the positive effect of the process modifications is expected to be higher.

From the economic point of view, the increase of the Cost of Electricity for a SCPC power plant with post combustion CO_2 capture amount to +61.7% for the Base Case process. Some of the process flow sheet modifications yield cost reduction potential. The OHC heat integration flow sheet modification shows the lowest CoE increase with a value of +60.2% in comparison to a plant without CO_2 capture.

In the case of the NGCC power plant, the costs increase due to a Base Case CO_2 capture plant amounts to +29.1%, being so less than the half than for the SCPC power plant. Flow sheet modifications of the Base Case process lead to similar costs variations as for the SCPC power plant. Also for the NGCC power plant the OHC heat integration represents the most advantageous flow sheet modification, leading to additional costs of +27.3% in comparison to a power plant without CO_2 capture.

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Appendix

Table 63: Further economic indicators for SCPC power plant flow sheet modifications

Modification	Net efficiency penalty in %-pts.	CAPEX in Mio € / yr	OPEX in Mio € / yr	ΔCoE_output in €/MWh	ΔCoE_CAPEX in €/MWh	ΔCoE_OPEX in €/MWh	ΔCoE_T&S in €/MWh
Base Case	6.11	40.50	27.26	6.59	7.57	5.10	6.80
Vapour recompression	6.09	41.25	27.45	6.57	7.71	5.13	6.80
Multi-pressure stripper	6.25	44.88	28.35	6.77	8.42	5.32	6.80
Heat-integrated stripping column	6.18	40.47	27.25	6.68	7.58	5.10	6.80
Improved split flow process	5.99	39.69	26.81	6.45	7.40	5.00	6.80
Matrix stripping	6.34	41.77	27.60	6.88	7.86	5.19	6.80
OHC integration	5.84	39.81	26.74	6.26	7.39	4.97	6.81
Vapour recompression + split flow	5.99	39.24	26.82	6.45	7.31	5.00	6.80
Heat-integrated stripper + OHC integration	5.88	39.81	26.75	6.31	7.40	4.97	6.81

Table 64: Further economic indicators for NGCC power plant flow sheet modifications

Modification	Net efficiency penalty in %-pts.	CAPEX in Mio € / yr	OPEX in Mio € / yr	ΔCoE_output in €/MWh	ΔCoE_CAPEX in €/MWh	ΔCoE_OPEX in €/MWh	ΔCoE_T&S in €/MWh
Base Case	5.93	26.46	16.21	6.85	4.53	2.78	3.16
Vapour recompression	5.86	27.70	16.51	6.76	4.74	2.82	3.16
Multi-pressure stripper	5.86	29.91	17.05	6.76	5.12	2.92	3.16
Heat-integrated stripping column	5.92	25.14	15.80	6.83	4.31	2.71	3.16
Improved split flow process	5.46	25.47	15.79	6.25	4.32	2.68	3.17
Matrix stripping	6.04	28.30	16.77	6.99	4.86	2.88	3.16
OHC integration	5.28	25.78	15.80	6.02	4.36	2.67	3.17
Reboiler condensate integration	5.83	26.68	16.29	6.72	4.56	2.79	3.16
Vapour recompression + split flow	5.46	25.60	15.83	6.25	4.35	2.69	3.17
Heat-integrated stripper + OHC integration	5.34	25.74	15.80	6.10	4.36	2.68	3.17

Table 65: List of Equipment & PEC – CO_2 capture plant Base Case in combination with SCPC power plant

List of Equipment	Ref. 2010						
Number of absorber trains	2						
Component	**Type**	**Material (S: steal; SS stainless steal; CI: cast iron)**	**Number per train**	**Total number**	**Reference value**	**PEC per train (€)**	**PEC (k€)**
Absorber shell	incl. collectors and distributors	S	1	2	547989 kg	1,272,409	2,545
Absorber packing	Mellapak Plus 252 Y	-	1	2	2433 m3	3,034,986	6,070
Absorber extras	Platforms and ladders	-	1	2	17.6 m	173,071	346
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	4	8	0.928 m3/s	520,707	1,041
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	4	8	606 kW	181,946	364
RL heat exchanger	Plates, sealed	SS	29	58	1075 m2	1,172,997	2,346
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	4	8	971 m2	149,310	299
Desorber overhead condenser	Pipe bundle	SS	1	2	648 m2	157,460	315
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.685 m	19,045	38
Desorber shell	incl. collectors and distributors	S	1	2	288387 kg	755,917	1,512
Desorber packing	Mellapak Plus 252 Y	-	1	2	650 m3	811,361	1,623
Desorber extras	Platforms and ladders	-	1	2	9.1 m	85,820	172
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	7	14	2859 m2	2,903,183	5,806
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	1001 m2	408,539	817
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.073 m3/s	23,671	47
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	70 kW	8,953	18
Activated-C filter	Inlet filter	SS	2	4	77 m2	237,009	474
Mechanical filter	Vertical plates	SS	4	8	77 m2	29,925	60
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	31 m3	65,833	132
Surge tank	Small field errected tank, incl. stairs etc.	S	5	10	1077 m3	796,907	1,594
ID fan	Axial fan with guide vane	S	1	2	328 m3/s	1,847,421	3,695
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	3627 kW	96,580	193
Heater to stack	Gasketed plate & frame	S	2	4	631 m2	26,577	53
DCC	incl. collectors and distributors	S	1	2	277411 kg	732,939	1,466
DCC surfaces	Plates, sealed	S	3	6	790 m2	47,571	95
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.558 m3/s	32,838	66
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	66 kW	17,124	34
Washing section (cooler)	Plates, sealed	S	1	2	93 m2	2,940	6
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.050 m3/s	15,026	30
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9 kW	3,389	7
Intercooler	Plates, sealed	S	3	6	752 m2	45,784	92
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	4	8	0.928 m3/s	77,475	155
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	4	8	60 kW	31,527	63
CO2 compressor	Integrally geared, 6 stages, intercooled incl. Driving engine	-	4	8	19.91 kg/s	15,429,312	30,859
Overall PEC (2010)							**62,431**
						Year of cost analysis	2010
						CEPCI (2012)	1.04
Overall PEC (2012)							**64,787**

Table 66: List of Equipment & PEC – CO_2 capture plant with Vapour Recompression in combination with SCPC power plant

List of Equipment	**Ref. 2010**								
Number of absorber trains	**2**								
Component	**Type**	**Material (S: steal; SS stainless steal; CI: cast iron)**	**Number per train**	**Total number**	**Reference value**	**PEC per train (€)**	**PEC (k€)**	**PEC (Base Case) (k€)**	**PEC difference to Base Case (k€)**
Absorber shell	incl. collectors and distributors	S	1	2	547989 kg	1,272,409	2,545	2,545	0
Absorber packing	Mellapak Plus 252 Y	-	1	2	2433 m3	3,034,986	6,070	6,070	0
Absorber extras	Platforms and ladders	-	1	2	17.6 m	173,071	346	346	0
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	4	8	0.928 m3/s	520,677	1,041	1,041	0
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	4	8	606 kW	181,933	364	364	0
RL heat exchanger	Plates, sealed	SS	27	54	1065 m2	1,084,240	2,168	2,346	-178
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	4	8	1027 m2	156,093	312	299	14
Flash tank	Horizontal storage vessel, D 2m, 2 bar	SS	1	2	3 m	35,736	71	0	71
Desorber overhead condenser	Pipe bundle	SS	1	2	558 m2	144,268	289	315	-26
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.564 m	16,431	33	38	-5
Desorber shell	incl. collectors and distributors	S	1	2	277708 kg	733,561	1,467	1,512	-45
Desorber packing	Mellapak Plus 252 Y	-	1	2	636 m3	793,627	1,587	1,623	-35
Desorber extras	Platforms and ladders	-	1	2	9 m	84,975	170	172	-2
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	7	14	2818 m2	2,875,578	5,751	5,806	-55
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	986 m2	404,655	809	817	-8
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.072 m3/s	23,563	47	47	0
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	69 kW	8,848	18	18	0
Activated-C filter	Inlet filter	SS	2	4	77 m2	236,979	474	474	0
Mechanical filter	Vertical plates	SS	4	8	77 m2	29,921	60	60	0
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	31 m3	65,836	132	132	0
Surge tank	Small field errected tank, incl. stairs etc.	S	5	10	1070 m3	794,488	1,589	1,594	-5
ID fan	Axial fan with guide vane	S	1	2	328 m3/s	1,847,313	3,695	3,695	0
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	3626 kW	96,577	193	193	0
Heater to stack	Gasketed plate & frame	S	2	4	634 m2	26,671	53	53	0
DCC	incl. collectors and distributors	S	1	2	277411 kg	732,939	1,466	1,466	0
DCC surfaces	Plates, sealed	S	3	6	790 m2	47,571	95	95	0
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.558 m3/s	32,838	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	66 kW	17,124	34	34	0
Washing section (cooler)	Plates, sealed	S	1	2	93 m2	2,937	6	6	0
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.050 m3/s	15,026	30	30	0
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9 kW	3,389	7	7	0
Intercooler	Plates, sealed	S	3	6	752 m2	45,764	92	92	0
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	4	8	0.928 m3/s	77,471	155	155	0
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	4	8	60 kW	31,522	63	63	0
CO2 compressor	Integrally geared, 6 stages, intercooled incl. Driving engine	-	4	8	19.91 kg/s	15,430,072	30,860	30,859	2
Lean vapour CO2-Compressor	Compressor, centrifugal, motor	SS	1	2	493.1 kW	725,803	1,452		1,452
Overall PEC							**63,610**	**62,431**	**1,179**
						Year of cost analysis	2010		
						CEPCI (2012)	1.04		
Overall PEC (reference capture plant)							**66,011**	**64,787**	**1,223**

Table 67: List of Equipment & PEC – CO_2 capture plant with Multi-pressure Stripper in combination with SCPC power plant

List of Equipment	Ref. 2010								
Number of absorber trains	2								
Component	**Type**	**Material (S: steal; SS stainless steal; CI: cast iron)**	**Number per train**	**Total number**	**Reference value**	**PEC per train (€)**	**PEC**	**PEC (Base Case)**	**PEC diff to Base Case**
Absorber shell	incl. collectors and distributors	S	1	2	533612 kg	1,244,794	2,490	2,545	-55
Absorber packing	Mellapak Plus 252 Y	-	1	2	2351 m3	2,932,402	5,865	6,070	-205
Absorber extras	Platforms and ladders	-	1	2	17.3 m	170,232	340	346	-6
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	4	8	0.930 m3/s	521,073	1,042	1,041	1
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	4	8	607 kW	182,112	364	364	0
RL heat exchanger	Plates, sealed	SS	27	54	1073 m2	1,090,024	2,180	2,346	-166
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	4	8	1030 m2	156,463	313	299	14
Desorber overhead condenser	Pipe bundle	SS	1	2	563 m2	145,059	290	315	-25
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.571 m	16,585	33	38	-5
Desorber shell	incl. collectors and distributors	S	1	2	183675 kg	530,229	1,060	1,512	-451
Desorber packing	Mellapak Plus 252 Y	-	1	2	255 m3	318,592	637	1,623	-986
Desorber extras	Platforms and ladders	-	1	2	9.4 m	68,298	137	172	-35
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	7	14	2801 m2	2,863,490	5,727	5,806	-79
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	980 m2	402,954	806	817	-11
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.071 m3/s	23,515	47	47	0
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	68 kW	8,802	18	18	0
Activated-C filter	Inlet filter	SS	2	4	77 m2	237,378	475	474	1
Mechanical filter	Vertical plates	SS	4	8	78 m2	29,970	60	60	0
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	31 m3	65,835	132	132	0
Surge tank	Small field errected tank, incl. stairs etc.	S	5	10	1073 m3	795,589	1,591	1,594	-3
ID fan	Axial fan with guide vane	S	1	2	326 m3/s	1,840,382	3,681	3,695	-14
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	4388 kW	104,645	209	193	16
Heater to stack	Gasketed plate & frame	S	1	2	246 m2	6,329	13	53	-40
DCC	incl. collectors and distributors	S	1	2	277411 kg	732,939	1,466	1,466	0
DCC surfaces	Plates, sealed	S	3	6	790 m2	47,571	95	95	0
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.558 m3/s	32,838	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	66 kW	17,124	34	34	0
Washing section (cooler)	Plates, sealed	S	1	2	93 m2	2,941	6	6	0
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.050 m3/s	15,026	30	30	0
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9 kW	3,389	7	7	0
Intercooler	Plates, sealed	S	3	6	761 m2	46,198	92	92	1
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	4	8	0.930 m3/s	77,529	155	155	0
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	4	8	445 kW	166,815	334	63	271
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	19.91 kg/s	15,429,727	30,859	30,859	1
Desorber1 shell	incl. collectors and distributors	S	1	2	209225 kg	586,826	1,174		1,174
Desorber1 packing	Mellapak Plus 252 Y	-	1	2	219 m3	273,574	547		547
Desorber1 extras	Platforms and ladders	-	1	2	9.400 m	66,543	133		133
Desorber2 shell	incl. collectors and distributors	S	1	2	179336 kg	520,493	1,041		1,041
Desorber2 packing	Mellapak Plus 252 Y	-	1	2	219 m3	273,574	547		547
Desorber2 extras	Platforms and ladders	-	1	2	9.400 m	66,543	133		133
Multi-stripper compressor1	Compressor, centrifugal, motor	Stainless Steel	1	2	872 kW	1,234,524	2,469		2,469
Multi-stripper compressor2	Compressor, centrifugal, motor	Stainless Steel	1	2	955 kW	1,343,286	2,687		2,687
Overall PEC							**69,384**	**62,431**	**6,953**
						Year of cost analysis	2010		
						CEPCI (2012)	1.04		
Overall PEC (reference capture plant)							**72,003**	**64,787**	**7,216**

Table 68: List of Equipment & PEC – CO_2 capture plant with Heat-integrated Stripper in combination with SCPC power plant

List of Equipment — Ref. 2010

Number of absorber trains: 2

Component	Type	Material (S: steal; SS stainless steal; CI: cast iron)	Number per train	Total number	Reference value	PEC per train (€)	PEC (k€)	PEC (Base Case) (k€)	PEC difference to Base Case (k€)
Absorber shell	incl. collectors and distributors	S	1	2	547989 kg	1,272,409	2,545	2,545	0
Absorber packing	Mellapak Plus 252 Y	-	1	2	2433 m3	3,034,986	6,070	6,070	0
Absorber extras	Platforms and ladders	-	1	2	17.6 m	173,071	346	346	0
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	4	8	0.928 m3/s	520,717	1,041	1,041	0
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	4	8	606 kW	181,951	364	364	0
RL heat exchanger	Plates, sealed	SS	28	56	1083 m2	1,139,168	2,278	2,346	-68
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	4	8	991 m2	151,662	303	299	5
Desorber overhead condenser	Pipe bundle	SS	1	2	614 m2	152,646	305	315	-10
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.639 m	18,080	36	38	-2
Desorber shell	incl. collectors and distributors	S	1	2	239455 kg	755,917	1,512	1,512	0
Desorber packing	Mellapak Plus 252 Y	-	1	2	342 m3	811,361	1,623	1,623	0
Desorber extras	Platforms and ladders	-	1	2	9.1 m	85,820	172	172	0
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	7	14	2842 m2	2,892,033	5,784	5,806	-22
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	995 m2	406,970	814	817	-3
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.072 m3/s	23,627	47	47	0
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	69 kW	8,910	18	18	0
Activated-C filter	Inlet filter	SS	2	4	77 m2	237,019	474	474	0
Mechanical filter	Vertical plates	SS	4	8	77 m2	29,926	60	60	0
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	31 m3	65,835	132	132	0
Surge tank	Small field errected tank, incl. stairs etc.	S	5	10	1077 m3	797,064	1,594	1,594	0
ID fan	Axial fan with guide vane	S	1	2	328 m3/s	1,847,386	3,695	3,695	0
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	3627 kW	96,579	193	193	0
Heater to stack	Gasketed plate & frame	S	2	4	632 m2	26,604	53	53	0
DCC	incl. collectors and distributors	S	1	2	277411 kg	732,939	1,466	1,466	0
DCC surfaces	Plates, sealed	S	3	6	789 m2	47,548	95	95	0
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.558 m3/s	32,832	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	66 kW	17,115	34	34	0
Washing section (cooler)	Plates, sealed	S	1	2	93 m2	2,937	6	6	0
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.050 m3/s	15,026	30	30	0
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9 kW	3,389	7	7	0
Intercooler	Plates, sealed	S	3	6	752 m2	45,777	92	92	0
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	4	8	0.928 m3/s	77,477	155	155	0
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	4	8	60 kW	31,528	63	63	0
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	19.91 kg/s	15,429,312	30,859	30,859	0
Stripper interheater		SS	1	2	800 m2	32,039	64		64
Overall PEC							**62,395**	**62,431**	**-36**
						Year of cost analysis	2010		
						CEPCI (2012)	1.04		
Overall PEC (reference capture plant)							**64,750**	**64,787**	**-37**

Table 69: List of Equipment & PEC – CO_2 capture plant with Improved split flow process in combination with SCPC power plant

List of Equipment	**Ref. 2010**								
Number of absorber trains	**2**								
Component	**Type**	**Material (S: steal; SS stainless steal; CI: cast iron)**	**Number per train**	**Total number**	**Reference value**	**PEC per train (€)**	**PEC (k€)**	**PEC (Base Case) (k€)**	**PEC difference to Base Case (k€)**
Absorber shell	incl. collectors and distributors	S	1	2	547989 kg	1,272,409	2,545	2,545	0
Absorber packing	Mellapak Plus 252 Y	-	1	2	2433 m3	3,034,986	6,070	6,070	0
Absorber extras	Platforms and ladders	-	1	2	17.6 m	173,071	346	346	0
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	4	8	0.871 m3/s	510,101	1,020	1,041	-21
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	4	8	569 kW	177,167	354	364	-10
RL heat exchanger	Plates, sealed	SS	28	56	1070 m2	1,128,436	2,257	2,346	-89
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	4	8	984 m2	150,874	302	299	3
Desorber overhead condenser	Pipe bundle	SS	1	2	253 m2	90,928	182	315	-133
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.187 m	7,098	14	38	-24
Desorber shell	incl. collectors and distributors	S	1	2	288387 kg	755,917	1,512	1,512	0
Desorber packing	Mellapak Plus 252 Y	-	1	2	650 m3	811,361	1,623	1,623	0
Desorber extras	Platforms and ladders	-	1	2	9.1 m	85,820	172	172	0
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	6	12	2889 m2	2,506,381	5,013	5,806	-794
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	867 m2	370,844	742	817	-75
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.063 m3/s	22,592	45	47	-2
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	60 kW	7,947	16	18	-2
Activated-C filter	Inlet filter	SS	2	4	72 m2	226,440	453	474	-21
Mechanical filter	Vertical plates	SS	4	8	73 m2	28,630	57	60	-3
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	31 m3	65,861	132	132	0
Surge tank	Small field errected tank, incl. stairs etc.	S	5	10	1017 m3	773,422	1,547	1,594	-47
ID fan	Axial fan with guide vane	S	1	2	330 m3/s	1,851,930	3,704	3,695	9
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	3641 kW	96,741	193	193	0
Heater to stack	Gasketed plate & frame	S	2	4	495 m2	21,933		53	-53
DCC	incl. collectors and distributors	S	1	2	277411 kg	732,939	1,466	1,466	0
DCC surfaces	Plates, sealed	S	3	6	790 m2	47,571	95	95	0
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.558 m3/s	32,838	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	66 kW	17,124	34	34	0
Washing section (cooler)	Plates, sealed	S	1	2	98 m2	3,054	6	6	0
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.051 m3/s	15,031	30	30	0
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9 kW	3,392	7	7	0
Intercooler	Plates, sealed	S	3	6	762 m2	46,261	93	92	1
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	4	8	0.871 m3/s	75,911	152	155	-3
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	4	8	56 kW	29,928	60	63	-3
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	19.93 kg/s	15,437,879	30,876	30,859	17
Overall PEC							**61,182**	**62,431**	**-1,249**
						Year of cost analysis	2010		
						CEPCI (2012)	1.04		
Overall PEC (reference capture plant)							**63,491**	**64,787**	**-1,296**

Table 70: List of Equipment & PEC – CO_2 capture plant with Matrix stripping in combination with SCPC power plant

List of Equipment Ref. 2010

Number of absorber trains: 2

Component	Type	Material (S: steal; SS stainless steal; CI: cast iron)	Number per train	Total number	Reference value	PEC per train (€)	PEC (k€)	PEC (Base Case) (k€)	PEC difference to Base Case
Absorber shell	incl. collectors and distributors	S	1	2	547989 kg	1,272,409	2,545	2,545	0
Absorber packing	Mellapak Plus 252 Y	-	1	2	2433 m3	3,034,986	6,070	6,070	0
Absorber extras	Platforms and ladders	-	1	2	17.6 m	173,071	346	346	0
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	4	8	0.928 m3/s	520,714	1,041	1,041	0
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	4	8	606 kW	181,950	364	364	0
RL heat exchanger	Plates, sealed	SS	27	54	1071 m2	1,088,674	2,177	2,346	-169
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	4	8	1023 m2	155,622	311	299	13
Desorber overhead condenser	Pipe bundle	SS	1	2	245 m2	89,143	178	315	-137
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.248 m	8,804	18	38	-20
Desorber shell	incl. collectors and distributors	S	1	2	121936 kg	387,370	775	1,512	-737
Desorber packing	Mellapak Plus 252 Y	-	1	2	192 m3	240,048	480	1,623	-1,143
Desorber extras	Platforms and ladders	-	1	2	7 m	55,242	110	172	-61
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	1	2	2113 m2	338,276	677	5,806	-5,130
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	106 m2	89,665	179	817	-638
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.073 m3/s	23,749	47	47	0
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	70 kW	9,028	18	18	0
Activated-C filter	Inlet filter	SS	2	4	77 m2	237,017	474	474	0
Mechanical filter	Vertical plates	SS	4	8	77 m2	29,926	60	60	0
Solvent storage tank	Small field errected tank, incl. stairs etc	SS	1	2	31 m3	65,834	132	132	0
Surge tank	Small field errected tank, incl. stairs etc	S	5	10	1071 m3	794,520	1,589	1,594	-5
ID fan	Axial fan with guide vane	S	1	2	328 m3/s	1,847,402	3,695	3,695	0
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	3627 kW	96,580	193	193	0
Heater to stack	Gasketed plate & frame	S	2	4	631 m2	26,588	53	53	0
DCC	incl. collectors and distributors	S	1	2	277411 kg	732,939	1,466	1,466	0
DCC surfaces	Plates, sealed	S	3	6	790 m2	47,571	95	95	0
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.558 m3/s	32,838	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	66 kW	17,124	34	34	0
Washing section (cooler)	Plates, sealed	S	1	2	93 m2	2,939	6	6	0
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.050 m3/s	15,026	30	30	0
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9 kW	3,389	7	7	0
Intercooler	Plates, sealed	S	3	6	752 m2	45,787	92	92	0
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	4	8	0.928 m3/s	77,476	155	155	0
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	4	8	60 kW	31,528	63	63	0
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	19.91 kg/s	15,429,565	30,859	30,859	1
Desorber overhead condenser1	Pipe bundle	SS	1	2	87 m2	48,506	97		97
Condensate return tank1	Vertical tank, D 4 m, 5 bar	SS	1	2	0 m	4,024	8		8
Desorber1 shell	incl. collectors and distributors	S	1	2	78717 kg	279,370	559		559
Desorber1 packing	Mellapak Plus 252 Y	-	1	2	238 m3	296,385	593		593
Desorber1 extras	Platforms and ladders	-	1	2	6 m	55,426	111		111
Reboiler1	Pipe bundles, onesided fixed, 7 bar	S / SS	1	2	1526 m2	271,600	543		543
Reclaimer1	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	76 m2	71,992	144		144
Overhead condenser2	Pipe bundle	SS	1	2	291 m2	98,535	197		197
Reflux drum2	Vertical tank, D 4 m, 5 bar	SS	1	2	0 m	10,190	20		20
Desorber2 shell	incl. collectors and distributors	S	1	2	138586 kg	426,936	854		854
Desorber2 packing	Mellapak Plus 252 Y	-	1	2	442 m3	551,130	1,102		1,102
Desorber2 extras	Platforms and ladders	-	1	2	8 m	72,325	145		145
Reboiler2	Pipe bundles, onesided fixed, 7 bar	S / SS	6	12	16573 m2	2,431,480	4,863		4,863
Reclaimer2	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	4972 m2	359,761	720		720
Overall PEC							**64,361**	**62,431**	**1,930**
						Year of cost analysis	2010		
						CEPCI (2012)	1.04		
Overall PEC (reference capture plant)							**66,790**	**64,787**	**2,003**

Table 71: List of Equipment & PEC – CO_2 capture plant with OHC integration in combination with SCPC power plant

List of Equipment	Ref. 2010									
Number of absorber trains	2									
Component	Type	Material (S: steal; SS stainless steal; CI: cast iron)	Number per train	Total number	Reference value		PEC per train (€)	PEC (k€)	PEC (Base Case) (k€)	PEC difference to Base Case (k€)
Absorber shell	incl. collectors and distributors	S	1	2	543178	kg	1,263,177	2,526	2,545	-18
Absorber packing	Mellapak Plus 252 Y	-	1	2	2405	m3	3,000,596	6,001	6,070	-69
Absorber extras	Platforms and ladders	-	1	2	17.5	m	172,124	344	346	-2
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	4	8	0.871	m3/s	510,044	1,020	1,041	-21
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	4	8	568	kW	177,141	354	364	-10
RL heat exchanger	Plates, sealed	SS	27	54	1099	m2	1,111,014	2,222	2,346	-124
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	4	8	1010	m2	154,029	308	299	9
Desorber overhead condenser	Pipe bundle	SS	1	2	76	m2	44,823	90	315	-225
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.833	m	22,106	44	38	6
Desorber shell	incl. collectors and distributors	S	1	2	288387	kg	755,917	1,512	1,512	0
Desorber packing	Mellapak Plus 252 Y	-	1	2	650	m3	811,361	1,623	1,623	0
Desorber extras	Platforms and ladders	-	1	2	9.1	m	85,820	172	172	0
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	6	12	2726	m2	2,409,919	4,820	5,806	-987
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	818	m2	356,571	713	817	-104
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.059	m3/s	22,168	44	47	-3
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	57	kW	7,573	15	18	-3
Activated-C filter	Inlet filter	SS	2	4	72	m2	226,384	453	474	-21
Mechanical filter	Vertical plates	SS	4	8	73	m2	28,624	57	60	-3
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	31	m3	65,837	132	132	0
Surge tank	Small field errected tank, incl. stairs etc.	S	5	10	1011	m3	771,233	1,542	1,594	-51
ID fan	Axial fan with guide vane	S	1	2	330	m3/s	1,852,166	3,704	3,695	9
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	3642	kW	96,749	193	193	0
Heater to stack	Gasketed plate & frame	S	2	4	491	m2	21,811	44	53	-10
DCC	incl. collectors and distributors	S	1	2	277411	kg	732,939	1,466	1,466	0
DCC surfaces	Plates, sealed	S	3	6	790	m2	47,571	95	95	0
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.558	m3/s	32,838	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	66	kW	17,124	34	34	0
Washing section (cooler)	Plates, sealed	S	1	2	98	m2	3,057	6	6	0
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.051	m3/s	15,031	30	30	0
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9	kW	3,392	7	7	0
Intercooler	Plates, sealed	S	3	6	762	m2	46,243	92	92	1
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	4	8	0.871	m3/s	75,902	152	155	-3
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	4	8	56	kW	29,919	60	63	-3
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	19.91	kg/s	15,430,546	30,861	30,859	2
OH rich split heat exchanger	Gasketed plate & frame	SS	7	14	1086.903342	m2	285,571	571		571
Overall PEC								**61,375**	**62,431**	**-1,056**
							Year of cost analysis	2010		
							CEPCI (2012)	1.04		
Overall PEC (reference capture plant)								**63,691**	**64,787**	**-1,096**

Table 72: List of Equipment & PEC – CO_2 capture plant with Vapour recompression and Improved split flow process in combination with SCPC power plant

List of Equipment	Ref. 2010									
Number of absorber trains	2									
Component	**Type**	**Material (S: steal; SS stainless steal; CI: cast iron)**	**Number per train**	**Total number**	**Reference value**		**PEC per train (€)**	**PEC (k€)**	**PEC (Base Case) (k€)**	**PEC difference to Base Case (k€)**
Absorber shell	incl. collectors and distributors	S	1	2	524125	kg	1,226,522	2,453	2,545	-92
Absorber packing	Mellapak Plus 252 Y	-	1	2	2297	m3	2,864,993	5,730	6,070	-340
Absorber extras	Platforms and ladders	-	1	2	17.1	m	168,343	337	346	-9
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	4	8	0.739	m3/s	483,521	967	1,041	-74
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	4	8	482	kW	165,317	331	364	-33
RL heat exchanger	Plates, sealed	SS	25	50	1073	m2	1,009,852	2,020	2,346	-326
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	4	8	857	m2	135,351	271	299	-28
Flash tank	Horizontal storage vessel, D 2m, 2 bar	SS	1	2	3	m	35,736	71	0	71
Desorber overhead condenser	Pipe bundle	SS	1	2	233	m2	86,569	173	315	-142
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.165	m	6,452	13	38	-25
Desorber shell	incl. collectors and distributors	S	1	2	273327	kg	724,354	1,449	1,512	-63
Desorber packing	Mellapak Plus 252 Y	-	1	2	622	m3	776,089	1,552	1,623	-71
Desorber extras	Platforms and ladders	-	1	2	8.9	m	84,131	168	172	-3
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	6	12	2851	m2	2,484,197	4,968	5,806	-838
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	855	m2	367,561	735	817	-82
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.062	m3/s	22,495	45	47	-2
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	60	kW	7,861	16	18	-2
Activated-C filter	Inlet filter	SS	2	4	61	m2	201,109	402	474	-72
Mechanical filter	Vertical plates	SS	4	8	62	m2	25,520	51	60	-9
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	31	m3	65,838	132	132	0
Surge tank	Small field errected tank, incl. stairs etc.	S	4	8	1083	m3	639,527	1,279	1,594	-315
ID fan	Axial fan with guide vane	S	1	2	335	m3/s	1,869,413	3,739	3,695	44
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	3697	kW	97,362	195	193	2
Heater to stack	Gasketed plate & frame	S	1	2	0	m2	17	0	53	-53
DCC	incl. collectors and distributors	S	1	2	277411	kg	732,939	1,466	1,466	0
DCC surfaces	Plates, sealed	S	3	6	790	m2	47,571	95	95	0
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.558	m3/s	32,838	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	66	kW	17,124	34	34	0
Washing section (cooler)	Plates, sealed	S	1	2	113	m2	3,422	7	6	1
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.051	m3/s	15,051	30	30	0
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9	kW	3,403	7	7	0
Intercooler	Plates, sealed	S	3	6	765	m2	46,398	93	92	1
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	4	8	0.740	m3/s	71,986	144	155	-11
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	4	8	354	kW	138,047	276	63	213
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	19.91	kg/s	15,430,710	30,861	30,859	3
										0
Lean vapour CO2-Compressor	Compressor, centrifugal, motor	SS	1	2	59.1315439	kW	100,654	201		201
Overall PEC								**60,377**	**62,431**	**-2,055**
						Year of cost analysis		2010		
						CEPCI (2012)		1.04		
Overall PEC (reference capture plant)								**62,655**	**64,787**	**-2,132**

Table 73: List of Equipment & PEC – CO_2 capture plant with Heat-integrated stripper and OHC heat integration in combination with SCPC power plant

List of Equipment Ref. 2010

Number of absorber trains: 2

Component	Type	Material (S: steal; SS stainless steal; CI: cast iron)	Number per train	Total number	Reference value	PEC per train (€)	PEC (k€)	PEC (Base Case) (k€)	PEC difference to Base Case (k€)
Absorber shell	incl. collectors and distributors	S	1	2	519410 kg	1,217,427	2,435	2,545	-110
Absorber packing	Mellapak Plus 252 Y	-	1	2	2270 m3	2,831,582	5,663	6,070	-407
Absorber extras	Platforms and ladders	-	1	2	17 m	167,399	335	346	-11
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	4	8	0.703 m3/s	475,750	952	1,041	-90
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	4	8	459 kW	161,888	324	364	-40
RL heat exchanger	Plates, sealed	SS	24	48	1059 m2	959,433	1,919	2,346	-427
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	4	8	827 m2	131,523	263	299	-36
Desorber overhead condenser	Pipe bundle	SS	1	2	83 m2	47,228	94	315	-220
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	1.089 m	27,118	54	38	16
Desorber shell	incl. collectors and distributors	S	1	2	268975 kg	715,186	1,430	1,512	-81
Desorber packing	Mellapak Plus 252 Y	-	1	2	608 m3	758,747	1,517	1,623	-105
Desorber extras	Platforms and ladders	-	1	2	8.8 m	83,286	167	172	-5
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	6	12	2679 m2	2,381,866	4,764	5,806	-1,043
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	804 m2	352,421	705	817	-112
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.058 m3/s	22,044	44	47	-3
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	56 kW	7,464	15	18	-3
Activated-C filter	Inlet filter	SS	2	4	58 m2	194,014	388	474	-86
Mechanical filter	Vertical plates	SS	4	8	59 m2	24,647	49	60	-11
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	31 m3	65,806	132	132	0
Surge tank	Small field errected tank, incl. stairs etc.	S	4	8	1025 m3	621,475	1,243	1,594	-351
ID fan	Axial fan with guide vane	S	1	2	337 m3/s	1,876,503	3,753	3,695	58
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	3720 kW	97,614	195	193	2
DCC	incl. collectors and distributors	S	1	2	277411 kg	732,939	1,466	1,466	0
DCC surfaces	Plates, sealed	S	3	6	790 m2	47,571	95	95	0
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.558 m3/s	32,838	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	66 kW	17,124	34	34	0
Washing section (cooler)	Plates, sealed	S	1	2	118 m2	3,551	7	6	1
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.051 m3/s	15,059	30	30	0
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9 kW	3,408	7	7	0
Intercooler	Plates, sealed	S	3	6	758 m2	46,055	92	92	1
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	4	8	0.704 m3/s	70,839	142	155	-13
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	4	8	337 kW	132,500	265	63	202
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	19.89 kg/s	15,420,733	30,841	30,859	-17
Stripper interheater	Gasketed plate & frame	SS	1	2	800 m2	32,039	64		64
OH rich split heat exchanger	Gasketed plate & frame	SS	9	18	1092 m2	368,570	737		737
Overall PEC							**60,287**	**62,431**	**-2,144**
						Year of cost analysis	2010		
						CEPCI (2012)	1.04		
Overall PEC (reference capture plant)							**62,563**	**64,787**	**-2,225**

Table 74: List of Equipment & PEC – CO_2 capture plant Base Case in combination with NGCC power plant

List of Equipment	Ref. 2010						
Number of absorber trains	2						
Component	**Type**	**Material (S: steal; SS stainless steal; CI: cast iron)**	**Number per train**	**Total number**	**Reference value**	**PEC per train (€)**	**PEC (k€)**
Absorber shell	incl. collectors and distributors	S	1	2	407850 kg	998,975	1,998
Absorber packing	Mellapak Plus 252 Y	-	1	2	1651 m3	2,060,001	4,120
Absorber extras	Platforms and ladders	-	1	2	14.5 m	144,006	288
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	3	6	0.768 m3/s	367,210	734
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	3	6	501 kW	126,012	252
RL heat exchanger	Plates, sealed	SS	17	34	1073 m2	686,515	1,373
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	2	4	1028 m2	78,059	156
Desorber overhead condenser	Pipe bundle	SS	1	2	388 m2	116,639	233
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.432 m	13,422	27
Desorber shell	incl. collectors and distributors	S	1	2	142081 kg	435,130	870
Desorber packing	Mellapak Plus 252 Y	-	1	2	363 m3	453,053	906
Desorber extras	Platforms and ladders	-	1	2	6.8 m	66,424	133
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	4	8	2720 m2	1,604,263	3,209
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	544 m2	270,844	542
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.039 m3/s	19,416	39
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	38 kW	5,399	11
Activated-C filter	Inlet filter	SS	1	2	85 m2	127,216	254
Mechanical filter	Vertical plates	SS	2	4	85 m2	16,028	32
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	15 m3	45,027	90
Surge tank	Small field errected tank, incl. stairs etc.	S	3	6	995 m3	458,903	918
ID fan	Axial fan with guide vane	S	1	2	258 m3/s	1,590,662	3,181
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	2847 kW	87,225	174
Heater to stack	Gasketed plate & frame	S	26	52	688 m2	369,758	740
DCC	incl. collectors and distributors	S	1	2	240098 kg	653,767	1,308
DCC surfaces	Plates, sealed	S	3	6	807 m2	48,404	97
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.563 m3/s	32,926	66
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	67 kW	17,269	35
Washing section (cooler)	Plates, sealed	S	1	2	96 m2	3,021	6
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.051 m3/s	15,041	30
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9 kW	3,398	7
Intercooler	Plates, sealed	S	2	4	582 m2	24,934	50
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	3	6	0.768 m3/s	54,643	109
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	3	6	49 kW	20,213	40
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	9.75 kg/s	9,363,665	18,727
Overall PEC							**40,755**
						Year of cost analysis	2010
						CEPCI (2012)	1.04
Overall PEC (reference capture plant)							**42,293**

Table 75: List of Equipment & PEC – CO_2 capture plant with Vapour recompression in combination with NGCC power plant

List of Equipment	Ref. 2010								
Number of absorber trains	2								
Component	**Type**	**Material (S: steal; SS stainless steal; CI: cast iron)**	**Number per train**	**Total number**	**Reference value**	**PEC per train (€)**	**PEC (k€)**	**PEC (Base Case) (k€)**	**PEC difference to Base Case (k€)**
Absorber shell	incl. collectors and distributors	S	1	2	407850 kg	998,975	1,998	1,998	0
Absorber packing	Mellapak Plus 252 Y	-	1	2	1651 m3	2,060,001	4,120	4,120	0
Absorber extras	Platforms and ladders	-	1	2	14.5 m	144,006	288	288	0
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	3	6	0.768 m3/s	367,200	734	734	0
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	3	6	501 kW	126,008	252	252	0
Solvent pump (lean)	Radial pump w/o motor, 10 bar	SS	3	6	0.82170241 m3/s	375,358	751	0	751
Solvent pump motor (lean)	E-motor, capsulated, air-cooled	-	3	6	48.57832735 kW	19,915	40	0	40
RL heat exchanger	Plates, sealed	SS	15	30	1069 m2	603,997	1,208	1,373	-165
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	3	6	765 m2	92,786	186	156	29
Flash tank	Horizontal storage vessel, D 2m, 2 bar	SS	1	2	3 m	35,736	71	0	71
Desorber overhead condenser	Pipe bundle	SS	1	2	300 m2	100,302	201	233	-33
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.311 m	10,461	21	27	-6
Desorber shell	incl. collectors and distributors	S	1	2	132437 kg	412,429	825	870	-45
Desorber packing	Mellapak Plus 252 Y	-	1	2	342 m3	426,795	854	906	-53
Desorber extras	Platforms and ladders	-	1	2	6.6 m	64,737	129	133	-3
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	4	8	2650 m2	1,576,441	3,153	3,209	-56
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	530 m2	266,147	532	542	-9
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.038 m3/s	19,253	39	39	0
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	37 kW	5,284	11	11	0
Activated-C filter	Inlet filter	SS	1	2	85 m2	127,209	254	254	0
Mechanical filter	Vertical plates	SS	2	4	85 m2	16,027	32	32	0
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	15 m3	45,026	90	90	0
Surge tank	Small field errected tank, incl. stairs etc.	S	3	6	986 m3	456,732	913	918	-4
ID fan	Axial fan with guide vane	S	1	2	258 m3/s	1,590,680	3,181	3,181	0
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	2847 kW	87,226	174	174	0
Heater to stack	Gasketed plate & frame	S	2	4	450 m2	20,357	41	740	-699
DCC	incl. collectors and distributors	S	1	2	240224 kg	654,037	1,308	1,308	1
DCC surfaces	Plates, sealed	S	3	6	808 m2	48,435	97	97	0
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.563 m3/s	32,935	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	67 kW	17,281	35	35	0
Washing section (cooler)	Plates, sealed	S	1	2	96 m2	3,021	6	6	0
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.051 m3/s	15,041	30	30	0
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9 kW	3,398	7	7	0
Intercooler	Plates, sealed	S	2	4	582 m2	24,930	50	50	0
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	3	6	0.768 m3/s	54,641	109	109	0
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	3	6	49 kW	20,211	40	40	0
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	9.75 kg/s	9,363,239	18,726	18,727	-1
Lean vapour CO2-Compressor	Compressor, centrifugal, motor	SS	1	2	737 kW	1,055,225	2,110		2,110
Overall PEC							**42,683**	**40,755**	**1,928**
						Year of cost analysis	2010		
						CEPCI (2012)	1.04		
Overall PEC (reference capture plant)							**44,294**	**42,293**	**2,001**

Table 76: List of Equipment & PEC – CO_2 capture plant with Multi-pressure stripper in combination with NGCC power plant

List of Equipment	Ref. 2010								
Number of absorber trains	2								
Component	**Type**	**Material (S: steal; SS stainless steal; CI: cast iron)**	**Number per train**	**Total number**	**Reference value**	**PEC per train (€)**	**PEC (k€)**	**PEC (Base Case) (k€)**	**PEC difference to Base Case (k€)**
Absorber shell	incl. collectors and distributors	S	1	2	407850 kg	998,975	1,998	1,998	0
Absorber packing	Mellapak Plus 252 Y	-	1	2	1651 m3	2,060,001	4,120	4,120	0
Absorber extras	Platforms and ladders	-	1	2	14.5 m	144,006	288	288	0
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	3	6	0.768 m3/s	367,146	734	734	0
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	3	6	501 kW	125,984	252	252	0
Solvent pump (lean)	Radial pump w/o motor, 10 bar	SS	3	6	0.821703935 m3/s	375,358	751	0	751
Solvent pump motor (lean)	E-motor, capsulated, air-cooled	-	3	6	42.30608715 kW	44,529	89	0	89
RL heat exchanger	Plates, sealed	SS	15	30	1076 m2	607,069	1,214	1,373	-159
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	3	6	760 m2	92,356	185	156	29
Desorber overhead condenser	Pipe bundle	SS	1	2	304 m2	101,056	202	233	-31
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.317 m	10,593	21	27	-6
Desorber shell	incl. collectors and distributors	S	1	2	65660 kg	244,611	489	870	-381
Desorber packing	Mellapak Plus 252 Y	-	1	2	126 m3	157,061	314	906	-592
Desorber extras	Platforms and ladders	-	1	2	6.6 m	48,908	98	133	-35
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	4	8	2628 m2	1,567,305	3,135	3,209	-74
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	526 m2	264,604	529	542	-12
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.038 m3/s	19,199	38	39	0
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	37 kW	5,246	10	11	0
Activated-C filter	Inlet filter	SS	1	2	85 m2	127,167	254	254	0
Mechanical filter	Vertical plates	SS	2	4	85 m2	16,022	32	32	0
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	15 m3	45,032	90	90	0
Surge tank	Small field errected tank, incl. stairs etc.	S	3	6	986 m3	456,733	913	918	-4
ID fan	Axial fan with guide vane	S	1	2	258 m3/s	1,590,768	3,182	3,181	0
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	2847 kW	87,229	174	174	0
Heater to stack	Gasketed plate & frame	S	2	4	448 m2	20,273	41	740	-699
DCC	incl. collectors and distributors	S	1	2	239027 kg	651,468	1,303	1,308	-5
DCC surfaces	Plates, sealed	S	3	6	794 m2	47,779	96	97	-1
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.554 m3/s	32,752	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	66 kW	17,037	34	35	0
Washing section (cooler)	Plates, sealed	S	1	2	97 m2	3,024	6	6	0
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.051 m3/s	15,041	30	30	0
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9 kW	3,398	7	7	0
Intercooler	Plates, sealed	S	2	4	583 m2	24,966	50	50	0
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	3	6	0.768 m3/s	54,633	109	109	0
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	3	6	49 kW	20,204	40	40	0
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	9.76 kg/s	9,364,929	18,730	18,727	3
Desorber1 shell	incl. collectors and distributors	S	1	2	72498 kg	262,975	526		526
Desorber1 packing	Mellapak Plus 252 Y	-	1	2	108 m3	134,867	270		270
Desorber1 extras	Platforms and ladders	-	1	2	7 m	47,514	95		95
Desorber2 shell	incl. collectors and distributors	S	1	2	95698 kg	322,897	646		646
Desorber2 packing	Mellapak Plus 252 Y	-	1	2	108 m3	134,867	270		270
Desorber2 extras	Platforms and ladders	-	1	2	7 m	47,514	95		95
Multi-stripper compressor1	Compressor, centrifugal, motor	Stainless Steel	1	2	557 kW	813,430	1,627		1,627
Multi-stripper compressor2	Compressor, centrifugal, motor	Stainless Steel	1	2	1036 kW	1,450,041	2,900		2,900
Overall PEC							**46,053**	**40,755**	**5,298**
						Year of cost analysis	2010		
						CEPCI (2012)	1.04		
Overall PEC (reference capture plant)							**47,791**	**42,293**	**5,498**

Table 77: List of Equipment & PEC – CO_2 capture plant with Heat-integrated stripper in combination with NGCC power plant

List of Equipment	Ref. 2010								
Number of absorber trains	2								
Component	**Type**	**Material (S: steal; SS stainless steal; CI: cast iron)**	**Number per train**	**Total number**	**Reference value**	**PEC per train (€)**	**PEC (k€)**	**PEC (Base Case) (k€)**	**PEC difference to Base Case (k€)**
Absorber shell	incl. collectors and distributors	S	1	2	391126 kg	965,607	1,931	1,998	-67
Absorber packing	Mellapak Plus 252 Y	-	1	2	1561 m3	1,947,913	3,896	4,120	-224
Absorber extras	Platforms and ladders	-	1	2	14.1 m	140,296	281	288	-7
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	3	6	0.586 m3/s	336,314	673	734	-62
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	3	6	383 kW	112,470	225	252	-27
RL heat exchanger	Plates, sealed	SS	12	24	1098 m2	493,546	987	1,373	-386
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	2	4	807 m2	64,514	129	156	-27
Desorber overhead condenser	Pipe bundle	SS	1	2	346 m2	109,140	218	233	-15
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.375 m	12,039	24	27	-3
Desorber shell	incl. collectors and distributors	S	1	2	118074 kg	378,055	756	870	-114
Desorber packing	Mellapak Plus 252 Y	-	1	2	302 m3	376,630	753	906	-153
Desorber extras	Platforms and ladders	-	1	2	6.2 m	61,358	123	133	-10
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	4	8	2634 m2	1,569,811	3,140	3,209	-69
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	527 m2	265,028	530	542	-12
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.038 m3/s	19,213	38	39	0
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	37 kW	5,257	11	11	0
Activated-C filter	Inlet filter	SS	1	2	65 m2	104,695	209	254	-45
Mechanical filter	Vertical plates	SS	2	4	65 m2	13,269	27	32	-6
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	15 m3	45,040	90	90	0
Surge tank	Small field errected tank, incl. stairs etc.	S	2	4	1149 m3	329,649	659	918	-259
ID fan	Axial fan with guide vane	S	1	2	262 m3/s	1,607,732	3,215	3,181	34
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	2896 kW	87,862	176	174	1
Heater to stack	Gasketed plate & frame	S	1	2	7 m2	377	1	740	-739
DCC	incl. collectors and distributors	S	1	2	238966 kg	651,338	1,303	1,308	-5
DCC surfaces	Plates, sealed	S	3	6	794 m2	47,779	96	97	-1
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.554 m3/s	32,752	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	66 kW	17,037	34	35	0
Washing section (cooler)	Plates, sealed	S	1	2	243 m2	6,271	13	6	7
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.102 m3/s	18,911	38	30	8
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	19 kW	6,094	12	7	5
Intercooler	Plates, sealed	S	2	4	552 m2	23,905	48	50	-2
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	3	6	0.587 m3/s	50,060	100	109	-9
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	3	6	38 kW	16,164	32	40	-8
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	9.76 kg/s	9,367,116	18,734	18,727	7
Stripper interheater		SS	1	2	1000 m2	76,404	153		153
Overall PEC							**38,719**	**40,755**	**-2,036**
						Year of cost analysis	2010		
						CEPCI (2012)	1.04		
Overall PEC (reference capture plant)							**40,181**	**42,293**	**-2,112**

Table 78: List of Equipment & PEC – CO_2 capture plant with Improved split flow process in combination with NGCC power plant

List of Equipment	Ref. 2010									
Number of absorber trains	2									
Component	Type	Material (S: steal; SS stainless steal; CI: cast iron)	Number per train	Total number	Reference value		PEC per train (€)	PEC (k€)	PEC (Base Case) (k€)	PEC difference to Base Case (k€)
Absorber shell	incl. collectors and distributors	S	1	2	403640	kg	990,592	1,981	1,998	-17
Absorber packing	Mellapak Plus 252 Y	-	1	2	1629	m3	2,031,685	4,063	4,120	-57
Absorber extras	Platforms and ladders	-	1	2	14.4	m	143,078	286	288	-2
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	3	6	0.713	m3/s	358,406	717	734	-18
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	3	6	465	kW	122,118	244	252	-8
RL heat exchanger	Plates, sealed	SS	16	32	1077	m2	648,126	1,296	1,373	-77
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	2	4	1070	m2	80,561	161	156	5
Desorber overhead condenser	Pipe bundle	SS	1	2	126	m2	60,403	121	233	-112
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.094	m	4,192	8	27	-18
Desorber shell	incl. collectors and distributors	S	1	2	139255	kg	428,506	857	870	-13
Desorber packing	Mellapak Plus 252 Y	-	1	2	353	m3	439,826	880	906	-26
Desorber extras	Platforms and ladders	-	1	2	6.7	m	65,581	131	133	-2
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	4	8	2295	m2	1,430,397	2,861	3,209	-348
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	459	m2	241,491	483	542	-59
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.033	m3/s	18,371	37	39	-2
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	32	kW	4,688	9	11	-1
Activated-C filter	Inlet filter	SS	1	2	79	m2	120,553	241	254	-13
Mechanical filter	Vertical plates	SS	2	4	79	m2	15,213	30	32	-2
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	15	m3	45,031	90	90	0
Surge tank	Small field errected tank, incl. stairs etc.	S	3	6	930	m3	442,949	886	918	-32
ID fan	Axial fan with guide vane	S	1	2	260	m3/s	1,597,339	3,195	3,181	13
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	2866	kW	87,474	175	174	0
Heater to stack	Gasketed plate & frame	S	1	2	547	m2	11,879		740	-740
DCC	incl. collectors and distributors	S	1	2	239595	kg	652,688	1,305	1,308	-2
DCC surfaces	Plates, sealed	S	3	6	801	m2	48,100	96	97	-1
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.558	m3/s	32,841	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	66	kW	17,156	34	35	0
Washing section (cooler)	Plates, sealed	S	1	2	82	m2	2,657	5	6	-1
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.040	m3/s	13,986	28	30	-2
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	8	kW	2,822	6	7	-1
Intercooler	Plates, sealed	S	2	4	584	m2	25,003	50	50	0
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	3	6	0.713	m3/s	53,343	107	109	-3
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	3	6	46	kW	19,009	38	40	-2
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	9.76	kg/s	9,364,610	18,729	18,727	2
Overall PEC								**39,218**	**40,755**	**-1,537**
							Year of cost analysis	2010		
							CEPCI (2012)	1.04		
Overall PEC (reference capture plant)								**40,698**	**42,293**	**-1,595**

Table 79: List of Equipment & PEC – CO_2 capture plant with Matrix stripping in combination with NGCC power plant

List of Equipment	Ref. 2010								
Number of absorber trains	2								
Component	**Type**	**Material (S: steal; SS stainless steal; CI: cast iron)**	**Number per train**	**Total number**	**Reference value**	**PEC per train (€)**	**PEC (k€)**	**PEC (Base Case) (k€)**	**PEC difference to Base Case (k€)**
Absorber shell	incl. collectors and distributors	S	1	2	416328 kg	1,015,822	2,032	1,998	34
Absorber packing	Mellapak Plus 252 Y	-	1	2	1697 m3	2,117,220	4,234	4,120	114
Absorber extras	Platforms and ladders	-	1	2	14.7 m	145,864	292	288	4
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	3	6	0.838 m3/s	377,694	755	734	21
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	3	6	547 kW	130,686	261	252	9
RL heat exchanger	Plates, sealed	SS	15	30	1091 m2	613,732	1,227	1,373	-146
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	3	6	888 m2	104,349	209	156	53
Desorber overhead condenser	Pipe bundle	SS	1	2	66 m2	41,383	83	233	-151
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.066 m	3,221	6	27	-20
Desorber shell	incl. collectors and distributors	S	1	2	49137 kg	198,430	397	870	-473
Desorber packing	Mellapak Plus 252 Y	-	1	2	90 m3	112,871	226	906	-680
Desorber extras	Platforms and ladders	-	1	2	4.8 m	39,524	79	133	-54
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	1	2	587 m2	142,557	285	3,209	-2,923
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	29 m2	37,787	76	542	-466
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.043 m3/s	19,904	40	39	1
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	41 kW	5,753	12	11	1
Activated-C filter	Inlet filter	SS	1	2	92 m2	135,408	271	254	16
Mechanical filter	Vertical plates	SS	2	4	93 m2	17,027	34	32	2
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	15 m3	45,043	90	90	0
Surge tank	Small field errected tank, incl. stairs etc.	S	3	6	1070 m3	476,501	953	918	35
ID fan	Axial fan with guide vane	S	1	2	257 m3/s	1,585,916	3,172	3,181	-9
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	2833 kW	87,048	174	174	0
Heater to stack	Gasketed plate & frame	S	26	52	688 m2	369,758	740	740	0
DCC	incl. collectors and distributors	S	1	2	240098 kg	653,767	1,308	1,308	0
DCC surfaces	Plates, sealed	S	3	6	807 m2	48,404	97	97	0
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.563 m3/s	32,926	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	67 kW	17,269	35	35	0
Washing section (cooler)	Plates, sealed	S	1	2	91 m2	2,879	6	6	0
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.051 m3/s	15,035	30	30	0
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9 kW	3,394	7	7	0
Intercooler	Plates, sealed	S	2	4	567 m2	24,419	49	50	-1
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	3	6	0.837 m3/s	56,189	112	109	3
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	3	6	54 kW	21,705	43	40	3
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	9.76 kg/s	9,367,983	18,736	18,727	9
Desorber overhead condenser1	Pipe bundle	SS	1	2	60 m2	39,341	79		79
Condensate return tank1	Vertical tank, D 4 m, 5 bar	SS	1	2	0 m	3,094	6		6
Desorber1 shell	incl. collectors and distributors	S	1	2	46243 kg	190,017	380		380
Desorber1 packing	Mellapak Plus 252 Y	-	1	2	173 m3	216,435	433		433
Desorber1 extras	Platforms and ladders	-	1	2	5 m	48,597	97		97
Reboiler1	Pipe bundles, onesided fixed, 7 bar	S / SS	1	2	587 m2	142,557	285		285
Reclaimer1	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	29 m2	37,787	76		76
Overhead condenser2	Pipe bundle	SS	1	2	256 m2	91,498	183		183
Reflux drum2	Vertical tank, D 4 m, 5 bar	SS	1	2	0 m	9,443	19		19
Desorber2 shell	incl. collectors and distributors	S	1	2	66416 kg	246,660	493		493
Desorber2 packing	Mellapak Plus 252 Y	-	1	2	302 m3	376,630	753		753
Desorber2 extras	Platforms and ladders	-	1	2	6 m	61,358	123		123
Reboiler2	Pipe bundles, onesided fixed, 7 bar	S / SS	4	8	2643 m2	1,573,301	3,147		3,147
Reclaimer2	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	529 m2	265,617	531		531
Overall PEC							**43,587**	**40,755**	**2,832**
						Year of cost analysis	2010		
						CEPCI (2012)	1.04		
Overall PEC (reference capture plant)							**45,233**	**42,293**	**2,939**

Table 80: List of Equipment & PEC – CO_2 capture plant with OHC heat integration in combination with NGCC power plant

List of Equipment	Ref. 2010									
Number of absorber trains	2									
Component	**Type**	**Material (S: steal; SS stainless steal; CI: cast iron)**	**Number per train**	**Total number**	**Reference value**		**PEC per train (€)**	**PEC (k€)**	**PEC (Base Case) (k€)**	**PEC difference to Base Case (k€)**
Absorber shell	incl. collectors and distributors	S	1	2	399449	kg	982,237	1,964	1,998	-33
Absorber packing	Mellapak Plus 252 Y	-	1	2	1606	m3	2,003,565	4,007	4,120	-113
Absorber extras	Platforms and ladders	-	1	2	14.3	m	142,150	284	288	-4
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	3	6	0.666	m3/s	350,552	701	734	-33
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	3	6	435	kW	118,667	237	252	-15
RL heat exchanger	Plates, sealed	SS	15	30	1088	m2	612,358	1,225	1,373	-148
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	2	4	1009	m2	76,964	154	156	-2
Desorber overhead condenser	Pipe bundle	SS	1	2	42	m2	31,787	64	233	-170
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.587	m	16,947	34	27	7
Desorber shell	incl. collectors and distributors	S	1	2	139255	kg	428,506	857	870	-13
Desorber packing	Mellapak Plus 252 Y	-	1	2	353	m3	439,826	880	906	-26
Desorber extras	Platforms and ladders	-	1	2	6.7	m	65,581	131	133	-2
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	3	6	2850	m2	1,241,773	2,484	3,209	-725
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	428	m2	230,219	460	542	-81
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.031	m3/s	17,952	36	39	-3
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	30	kW	4,420	9	11	-2
Activated-C filter	Inlet filter	SS	1	2	74	m2	114,775	230	254	-25
Mechanical filter	Vertical plates	SS	2	4	74	m2	14,506	29	32	-3
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	15	m3	45,029	90	90	0
Surge tank	Small field errected tank, incl. stairs etc.	S	3	6	864	m3	426,312	853	918	-65
ID fan	Axial fan with guide vane	S	1	2	261	m3/s	1,601,753	3,204	3,181	22
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	2879	kW	87,639	175	174	1
Heater to stack	Gasketed plate & frame	S	26	52	688	m2	369,758	740	740	0
DCC	incl. collectors and distributors	S	1	2	240098	kg	653,767	1,308	1,308	0
DCC surfaces	Plates, sealed	S	3	6	806	m2	48,366	97	97	0
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.562	m3/s	32,915	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	67	kW	17,255	35	35	0
Washing section (cooler)	Plates, sealed	S	1	2	108	m2	3,313	7	6	1
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.051	m3/s	15,056	30	30	0
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9	kW	3,407	7	7	0
Intercooler	Plates, sealed	S	2	4	578	m2	24,800	50	50	0
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	3	6	0.667	m3/s	52,180	104	109	-5
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	3	6	43	kW	17,969	36	40	-4
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	9.76	kg/s	9,364,205	18,728	18,727	1
OH rich split heat exchanger	Gasketed plate & frame	SS	5	10	1055.2916	m2	199,287	399		399
Overall PEC								**39,712**	**40,755**	**-1,043**
							Year of cost analysis	2010		
							CEPCI (2012)	1.04		
Overall PEC (reference capture plant)								**41,210**	**42,293**	**-1,083**

Table 81: List of Equipment & PEC – CO_2 capture plant with Reboiler condensate integration in combination with NGCC power plant

List of Equipment	Ref. 2010								
Number of absorber trains	2								
Component	**Type**	**Material (S: steal; SS stainless steal; CI: cast iron)**	**Number per train**	**Total number**	**Reference value**	**PEC per train (€)**	**PEC (k€)**	**PEC (Base Case) (k€)**	**PEC difference to Base Case (k€)**
Absorber shell	incl. collectors and distributors	S	1	2	416328 kg	1,015,822	2,032	1,998	34
Absorber packing	Mellapak Plus 252 Y	-	1	2	1697 m3	2,117,220	4,234	4,120	114
Absorber extras	Platforms and ladders	-	1	2	14.7 m	145,864	292	288	4
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	3	6	0.837 m3/s	377,589	755	734	21
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	3	6	546 kW	130,639	261	252	9
RL heat exchanger	Plates, sealed	SS	18	36	1072 m2	726,480	1,453	1,373	80
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	3	6	776 m2	93,872	188	156	32
Desorber overhead condenser	Pipe bundle	SS	1	2	366 m2	112,660	225	233	-8
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.401 m	12,684	25	27	-1
Desorber shell	incl. collectors and distributors	S	1	2	149192 kg	451,694	903	870	33
Desorber packing	Mellapak Plus 252 Y	-	1	2	374 m3	466,476	933	906	27
Desorber extras	Platforms and ladders	-	1	2	6.9 m	67,268	135	133	2
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	4	8	2637 m2	1,571,078	3,142	3,209	-66
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	527 m2	265,241	530	542	-11
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.038 m3/s	19,221	38	39	0
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	37 kW	5,262	11	11	0
Activated-C filter	Inlet filter	SS	1	2	92 m2	135,325	271	254	16
Mechanical filter	Vertical plates	SS	2	4	93 m2	17,017	34	32	2
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	15 m3	45,024	90	90	0
Surge tank	Small field errected tank, incl. stairs etc.	S	3	6	1083 m3	479,469	959	918	41
ID fan	Axial fan with guide vane	S	1	2	257 m3/s	1,585,987	3,172	3,181	-9
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	2833 kW	87,050	174	174	0
Heater to stack	Gasketed plate & frame	S	26	52	688 m2	369,758	740	740	0
DCC	incl. collectors and distributors	S	1	2	240098 kg	653,767	1,308	1,308	0
DCC surfaces	Plates, sealed	S	3	6	807 m2	48,404	97	97	0
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.563 m3/s	32,926	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	67 kW	17,269	35	35	0
Washing section (cooler)	Plates, sealed	S	1	2	91 m2	2,880	6	6	0
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.051 m3/s	15,035	30	30	0
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9 kW	3,394	7	7	0
Intercooler	Plates, sealed	S	2	4	567 m2	24,410	49	50	-1
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	3	6	0.836 m3/s	56,174	112	109	3
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	3	6	54 kW	21,690	43	40	3
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	9.75 kg/s	9,362,689	18,725	18,727	-2
Stripper interheater (condens. HX)	Gasketed plate & frame	SS	1	2	184 m2	10,071	20		20
Overall PEC							**41,095**	**40,755**	**340**
						Year of cost analysis	2010		
						CEPCI (2012)	1.04		
Overall PEC (reference capture plant)							**42,646**	**42,293**	**353**

Table 82: List of Equipment & PEC – CO_2 capture plant with Vapour recompression and Improved split flow process in combination with NGCC power plant

List of Equipment	**Ref. 2010**									
Number of absorber trains	**2**									
Component	**Type**	**Material (S: steal; SS stainless steal; CI: cast iron)**	**Number per train**	**Total number**	**Reference value**		**PEC per train (€)**	**PEC (k€)**	**PEC (Base Case) (k€)**	**PEC difference to Base Case (k€)**
Absorber shell	incl. collectors and distributors	S	1	2	403640	kg	990,592	1,981	1,998	-17
Absorber packing	Mellapak Plus 252 Y	-	1	2	1629	m3	2,031,685	4,063	4,120	-57
Absorber extras	Platforms and ladders	-	1	2	14.4	m	143,078	286	288	-2
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	3	6	0.713	m3/s	358,418	717	734	-18
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	3	6	465	kW	122,123	244	252	-8
RL heat exchanger	Plates, sealed	SS	16	32	1057	m2	638,495	1,277	1,373	-96
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	2	4	1089	m2	81,722	163	156	7
Flash tank	Horizontal storage vessel, D 2m, 2 bar	SS	1	2	3	m	35,736	71	0	71
Desorber overhead condenser	Pipe bundle	SS	1	2	119	m2	58,467	117	233	-116
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.086	m	3,936	8	27	-19
Desorber shell	incl. collectors and distributors	S	1	2	139255	kg	428,506	857	870	-13
Desorber packing	Mellapak Plus 252 Y	-	1	2	353	m3	439,826	880	906	-26
Desorber extras	Platforms and ladders	-	1	2	6.7	m	65,581	131	133	-2
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	4	8	2306	m2	1,434,967	2,870	3,209	-339
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	461	m2	242,262	485	542	-57
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.033	m3/s	18,399	37	39	-2
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	32	kW	4,707	9	11	-1
Activated-C filter	Inlet filter	SS	1	2	79	m2	120,562	241	254	-13
Mechanical filter	Vertical plates	SS	2	4	79	m2	15,214	30	32	-2
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	15	m3	45,027	90	90	0
Surge tank	Small field errected tank, incl. stairs etc.	S	3	6	928	m3	442,611	885	918	-33
ID fan	Axial fan with guide vane	S	1	2	259	m3/s	1,595,914	3,192	3,181	11
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	2862	kW	87,421	175	174	0
Heater to stack	Gasketed plate & frame	S	1	2	622	m2	13,139	26	740	-713
DCC	incl. collectors and distributors	S	1	2	240224	kg	654,037	1,308	1,308	1
DCC surfaces	Plates, sealed	S	3	6	808	m2	48,435	97	97	0
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.563	m3/s	32,935	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	67	kW	17,281	35	35	0
Washing section (cooler)	Plates, sealed	S	1	2	102	m2	3,166	6	6	0
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.051	m3/s	15,048	30	30	0
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9	kW	3,402	7	7	0
Intercooler	Plates, sealed	S	2	4	583	m2	24,983	50	50	0
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	3	6	0.713	m3/s	53,343	107	109	-3
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	3	6	46	kW	19,009	38	40	-2
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	9.75	kg/s	9,363,653	18,727	18,727	0
				0						
Lean vapour CO2-Compressor	Compressor, centrifugal, motor	SS	1	2	35.0	kW	61,713	123		123
Overall PEC								**39,431**	**40,755**	**-1,324**
							Year of cost analysis	2010		
							CEPCI (2012)	1.04		
Overall PEC (reference capture plant)								**40,919**	**42,293**	**-1,374**

Table 83: List of Equipment & PEC – CO_2 capture plant with Heat-integrated stripping column and OHC heat integration in combination with NGCC power plant

List of Equipment	**Ref. 2010**								
Number of absorber trains	**2**								
Component	**Type**	**Material (S: steal; SS stainless steal; CI: cast iron)**	**Number per train**	**Total number**	**Reference value**	**PEC per train (€)**	**PEC (k€)**	**PEC (Base Case) (k€)**	**PEC difference to Base Case (k€)**
Absorber shell	incl. collectors and distributors	S	1	2	424884 kg	1,032,779	2,066	1,998	68
Absorber packing	Mellapak Plus 252 Y	-	1	2	1744 m3	2,175,224	4,350	4,120	230
Absorber extras	Platforms and ladders	-	1	2	14.9 m	147,724	295	288	7
Solvent pump (rich)	Radial pump w/o motor, 10 bar	SS	3	6	0.666 m3/s	350,494	701	734	-33
Solvent pump motor (rich)	E-motor, capsulated, air-cooled	-	3	6	434 kW	118,642	237	252	-15
RL heat exchanger	Plates, sealed	SS	15	30	1031 m2	587,156	1,174	1,373	-199
Solvent cooler (lean)	U-pipe bundles, 1 bar	SS	2	4	1044 m2	79,055	158	156	2
Desorber overhead condenser	Pipe bundle	SS	1	2	39 m2	30,620	61	233	-172
Condensate return tank	Vertical tank, D 4 m, 5 bar	SS	1	2	0.493 m	14,831	30	27	3
Desorber shell	incl. collectors and distributors	S	1	2	159556 kg	475,594	951	870	81
Desorber packing	Mellapak Plus 252 Y	-	1	2	396 m3	493,910	988	906	82
Desorber extras	Platforms and ladders	-	1	2	7.1 m	68,954	138	133	5
Reboiler	Pipe bundles, onesided fixed, 7 bar	S / SS	3	6	2911 m2	1,259,420	2,519	3,209	-690
Reclaimer	Pipe bundles, onesided fixed, 7 bar	S / SS	2	4	437 m2	233,490	467	542	-75
Condensate pump	Radial pump w/o motor, 10 bar	CI	1	2	0.032 m3/s	18,075	36	39	-3
Condensate pump motor	E-motor, capsulated, air-cooled	-	1	2	30 kW	4,498	9	11	-2
Activated-C filter	Inlet filter	SS	1	2	74 m2	114,732	229	254	-25
Mechanical filter	Vertical plates	SS	2	4	74 m2	14,501	29	32	-3
Solvent storage tank	Small field errected tank, incl. stairs etc.	SS	1	2	15 m3	45,043	90	90	0
Surge tank	Small field errected tank, incl. stairs etc.	S	3	6	865 m3	426,516	853	918	-65
ID fan	Axial fan with guide vane	S	1	2	261 m3/s	1,601,960	3,204	3,181	23
ID fan motor	E-motor, capsulated, air-cooled	S	1	2	2880 kW	87,647	175	174	1
DCC	incl. collectors and distributors	S	1	2	239027 kg	651,468	1,303	1,308	-5
DCC surfaces	Plates, sealed	S	3	6	794 m2	47,779	96	97	-1
DCC pump	Radial pump w/o motor, 1 bar	CI	2	4	0.554 m3/s	32,752	66	66	0
DCC pump motor	E-motor, capsulated, air-cooled	-	2	4	66 kW	17,037	34	35	0
Washing section (cooler)	Plates, sealed	S	1	2	109 m2	3,318	7	6	1
Washing section pump	Radial pump w/o motor, 1 bar	CI	2	4	0.051 m3/s	15,056	30	30	0
Washing section pump motor	E-motor, capsulated, air-cooled	-	2	4	9 kW	3,407	7	7	0
Intercooler	Plates, sealed	S	2	4	579 m2	24,822	50	50	0
Intercooler Pump	Radial pump w/o motor, 1 bar	CI	3	6	0.666 m3/s	52,171	104	109	-5
Intercooler Pump Motor	E-motor, capsulated, air-cooled	-	3	6	43 kW	17,961	36	40	-5
CO2 compressor	Integrally geared, intercooled incl. Driving engine	-	4	8	9.76 kg/s	9,367,987	18,736	18,727	9
Stripper interheater	Gasketed plate & frame	SS	1	2	800 m2	32,039	64		64
OH rich split heat exchanger	Gasketed plate & frame	SS	5	10	892 m2	174,616	349		349
Overall PEC							**39,643**	**40,755**	**-1,112**
						Year of cost analysis	2010		
						CEPCI (2012)	1.04		
Overall PEC (reference capture plant)							**41,139**	**42,293**	**-1,154**

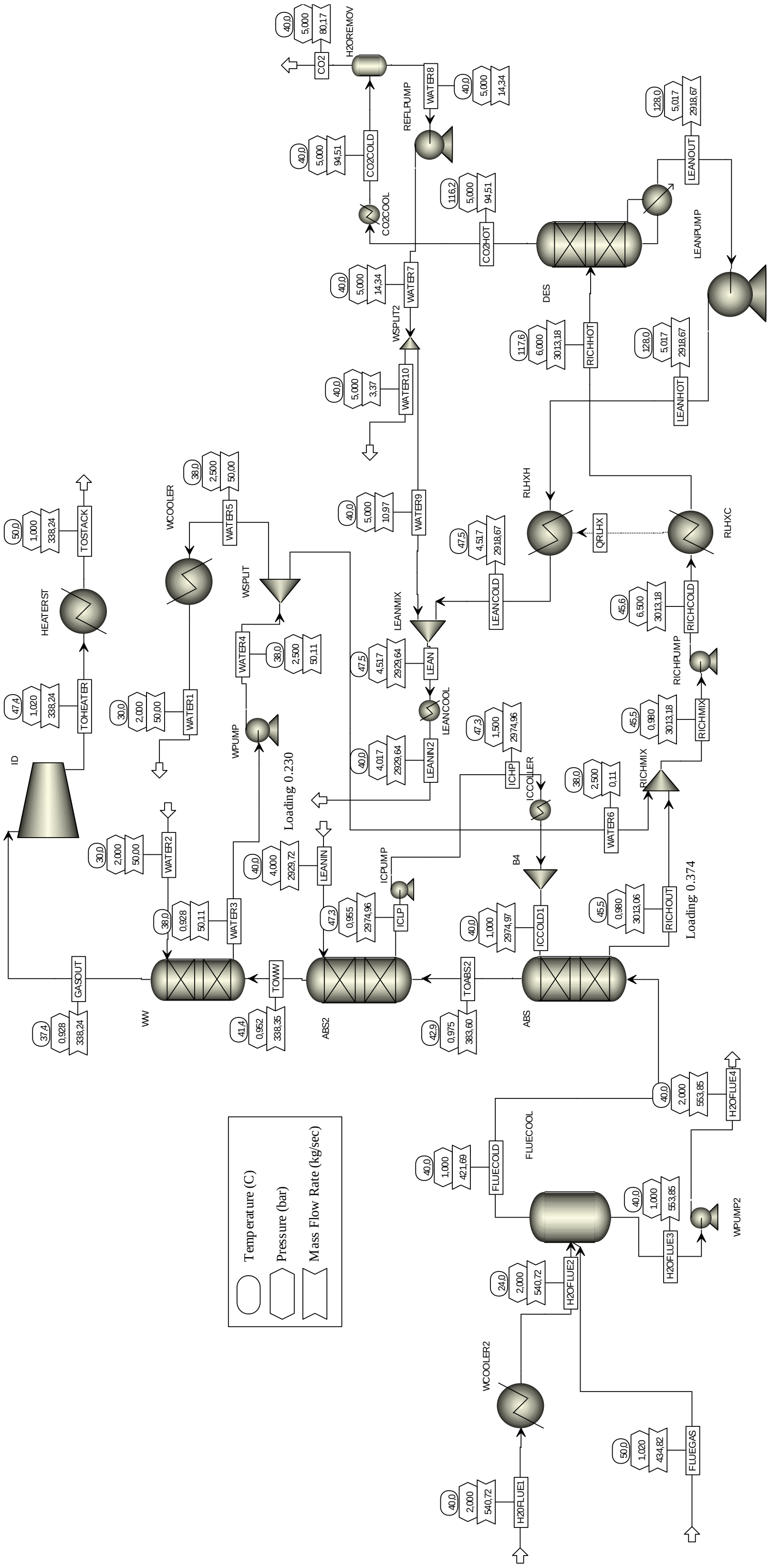

Figure 87: Flow sheet of capture plant in combination with SCPC for base case

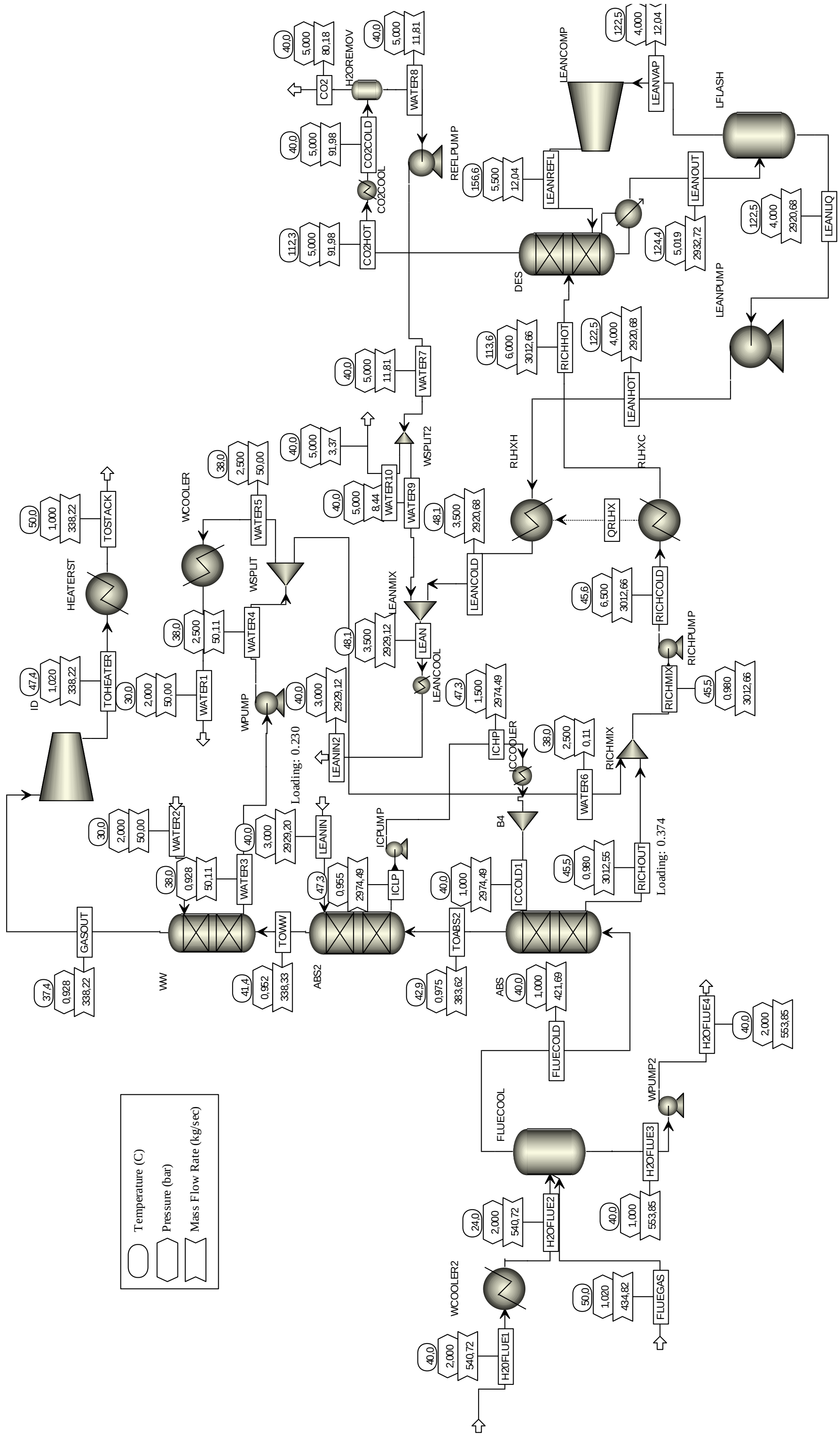

Figure 88: Flow sheet of capture plant with vapour recompression in combination with SCPC plant

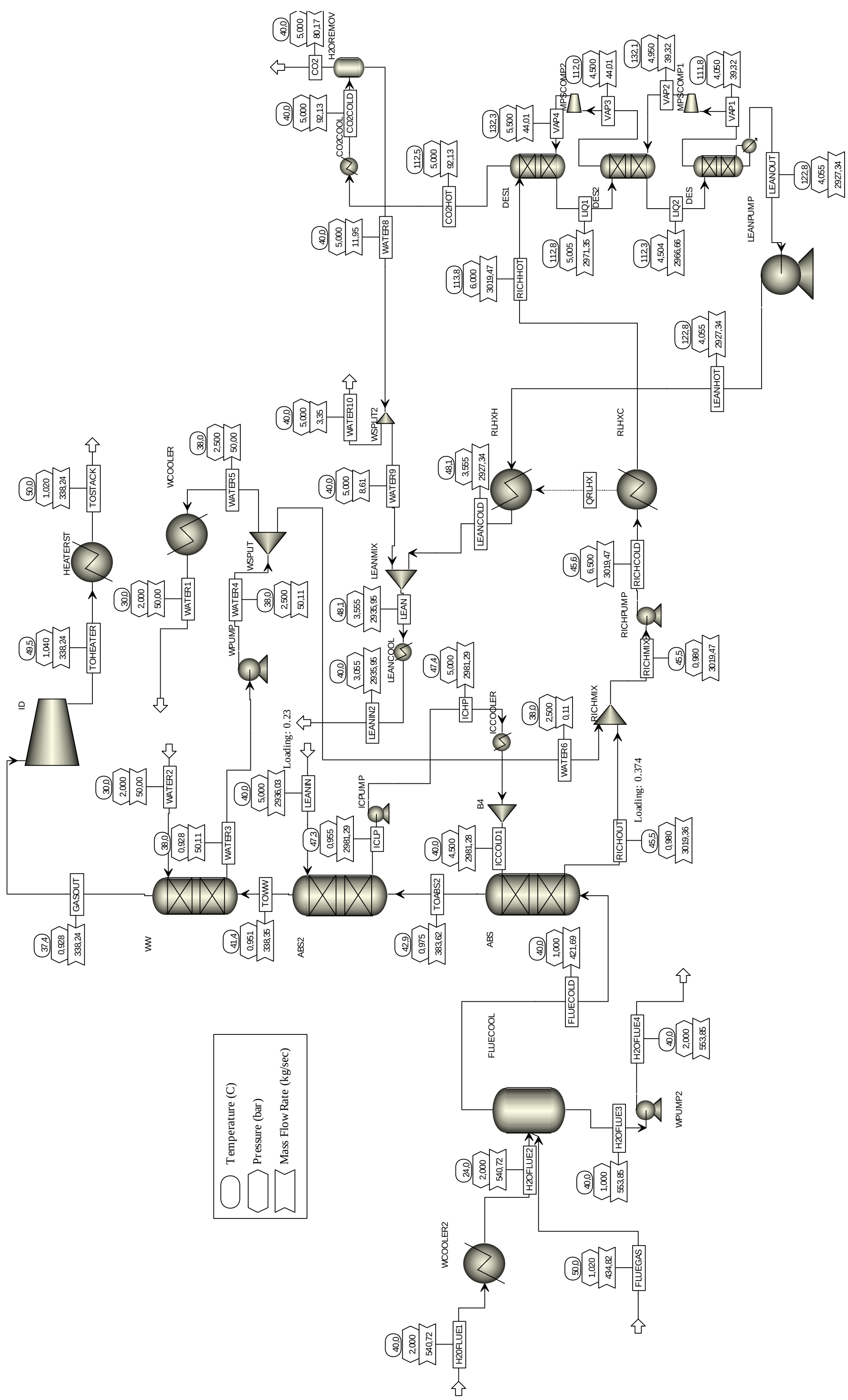

Figure 89: Flow sheet of capture plant with multi pressure stripper in combination with SCPC plant

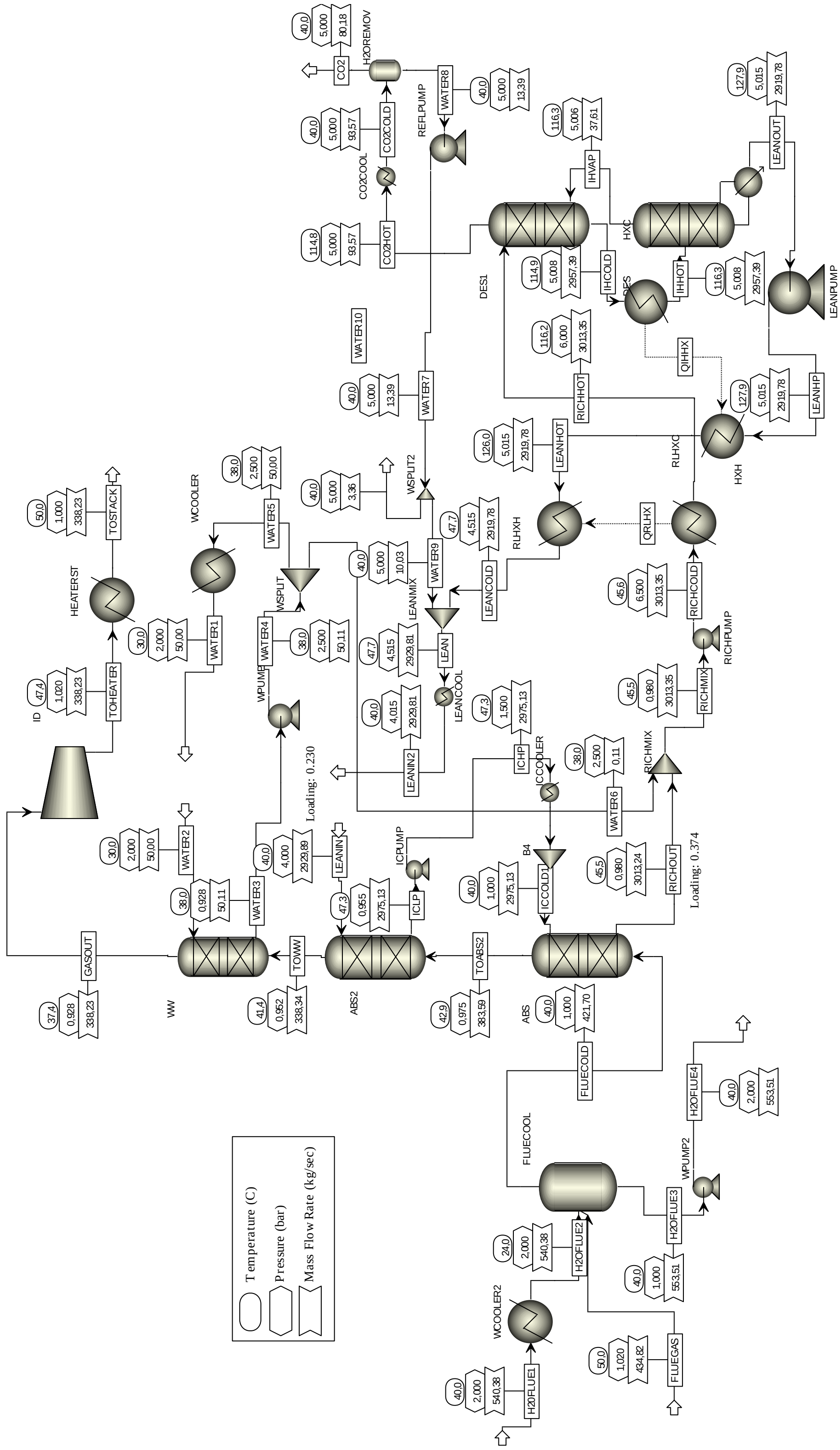

Figure 90: Flow sheet of capture plant with heat-integrated stripping column in combination with SCPC plant

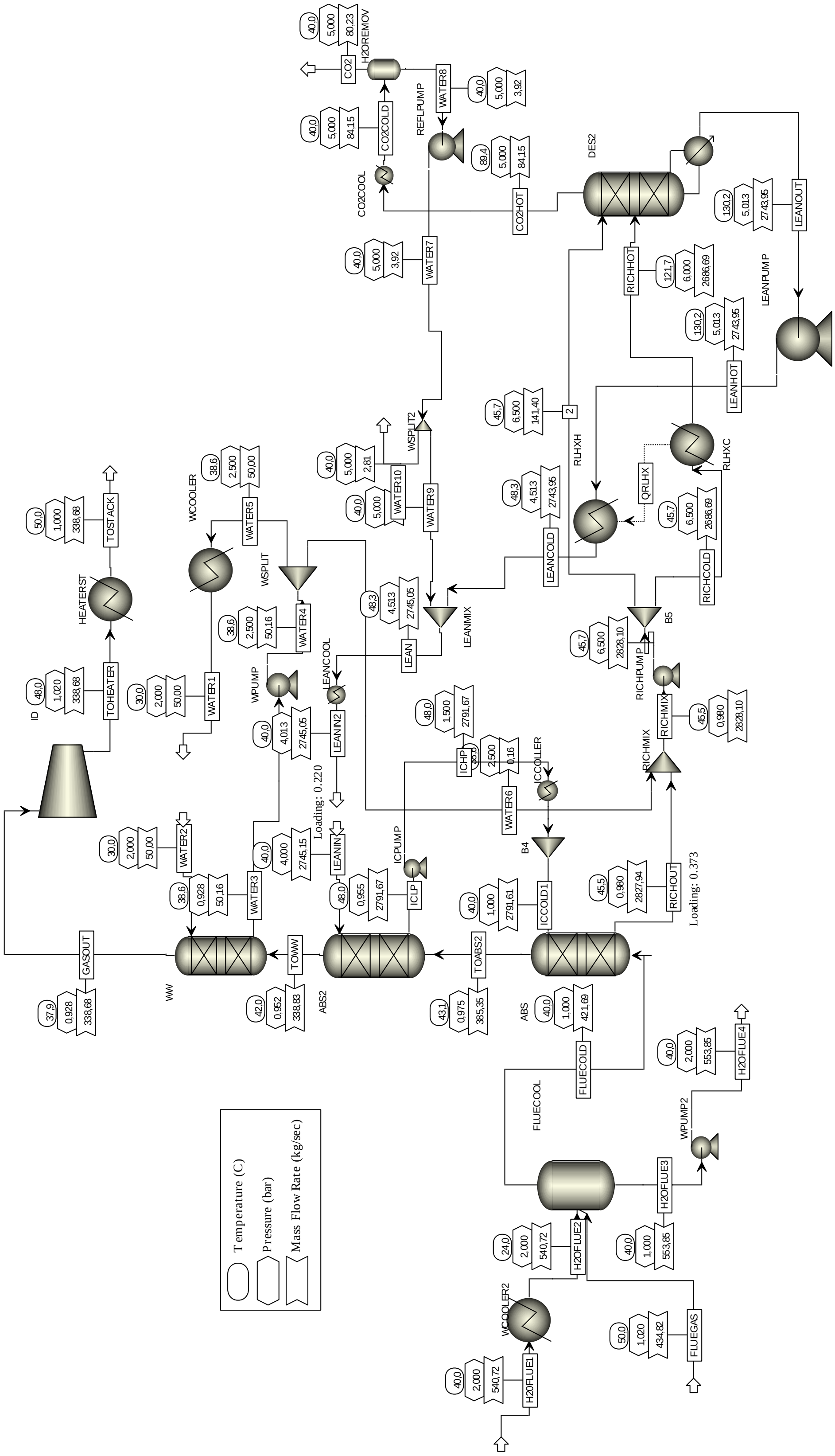

Figure 91: Flow sheet of capture plant with split flow in combination with SCPC plant

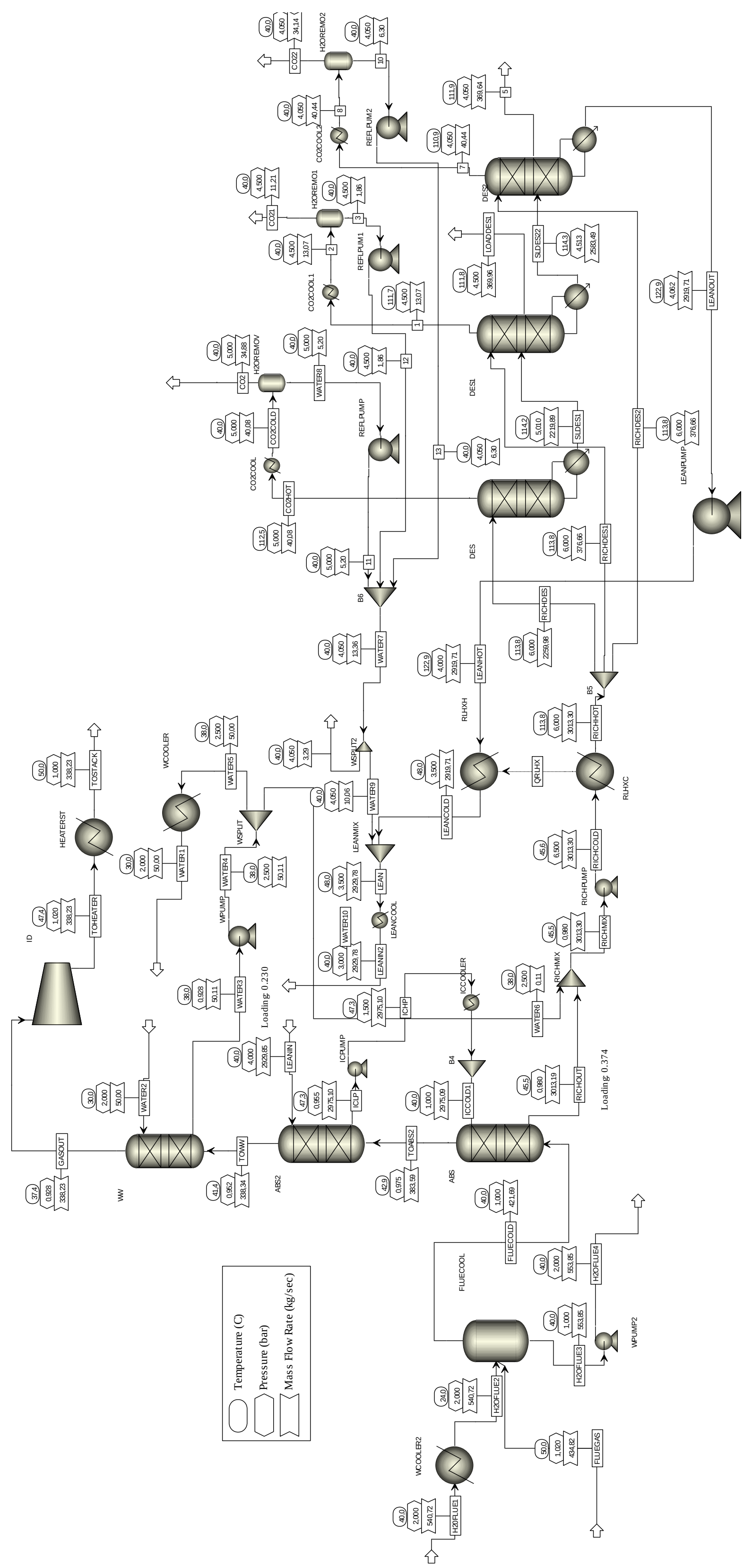

Figure 92: Flow sheet of capture plant with matrix stripping in combination with SCPC plant

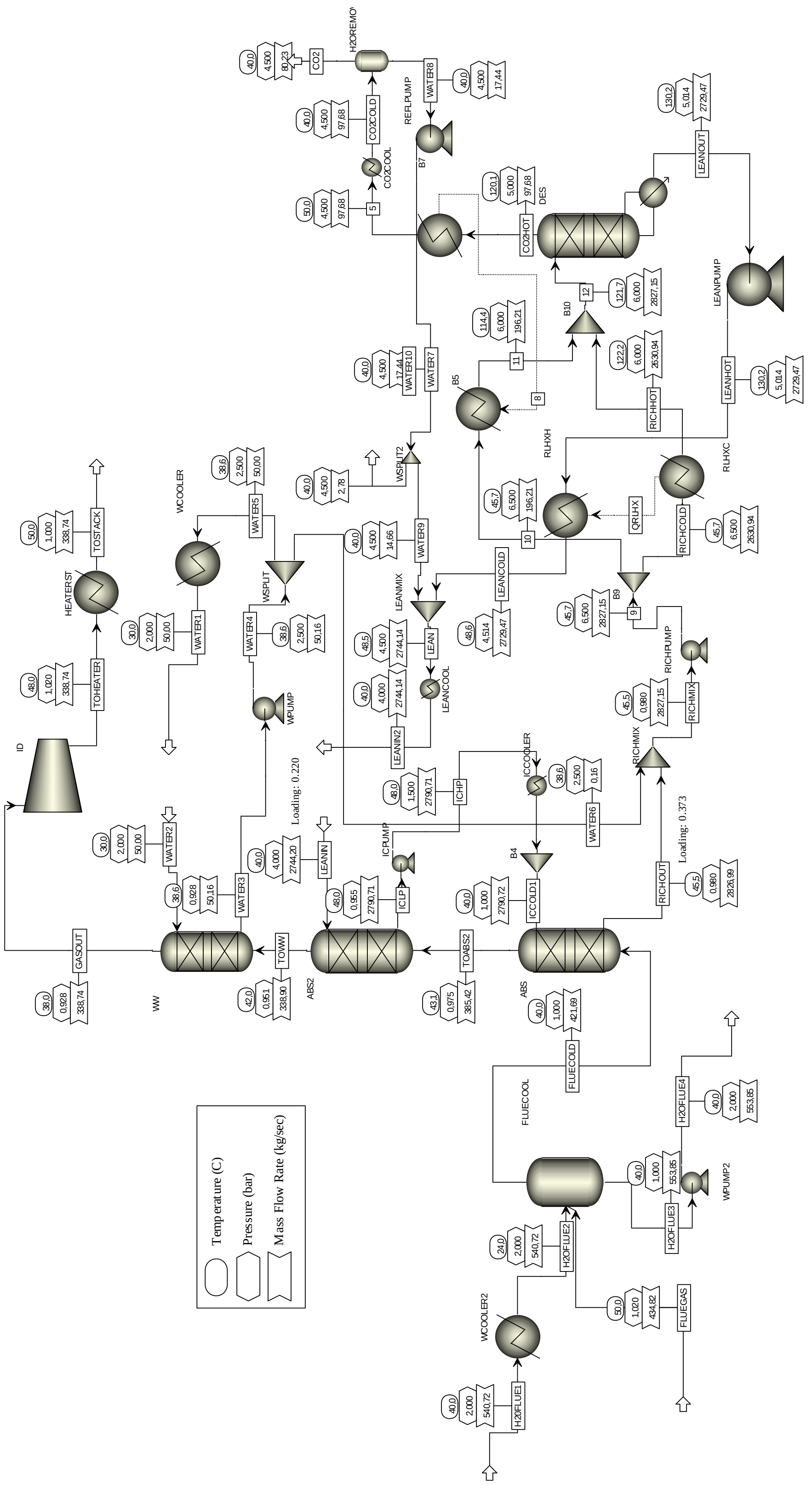

Figure 93: Flow sheet of capture plant with overhead condenser heat integration in combination with SCPC plant

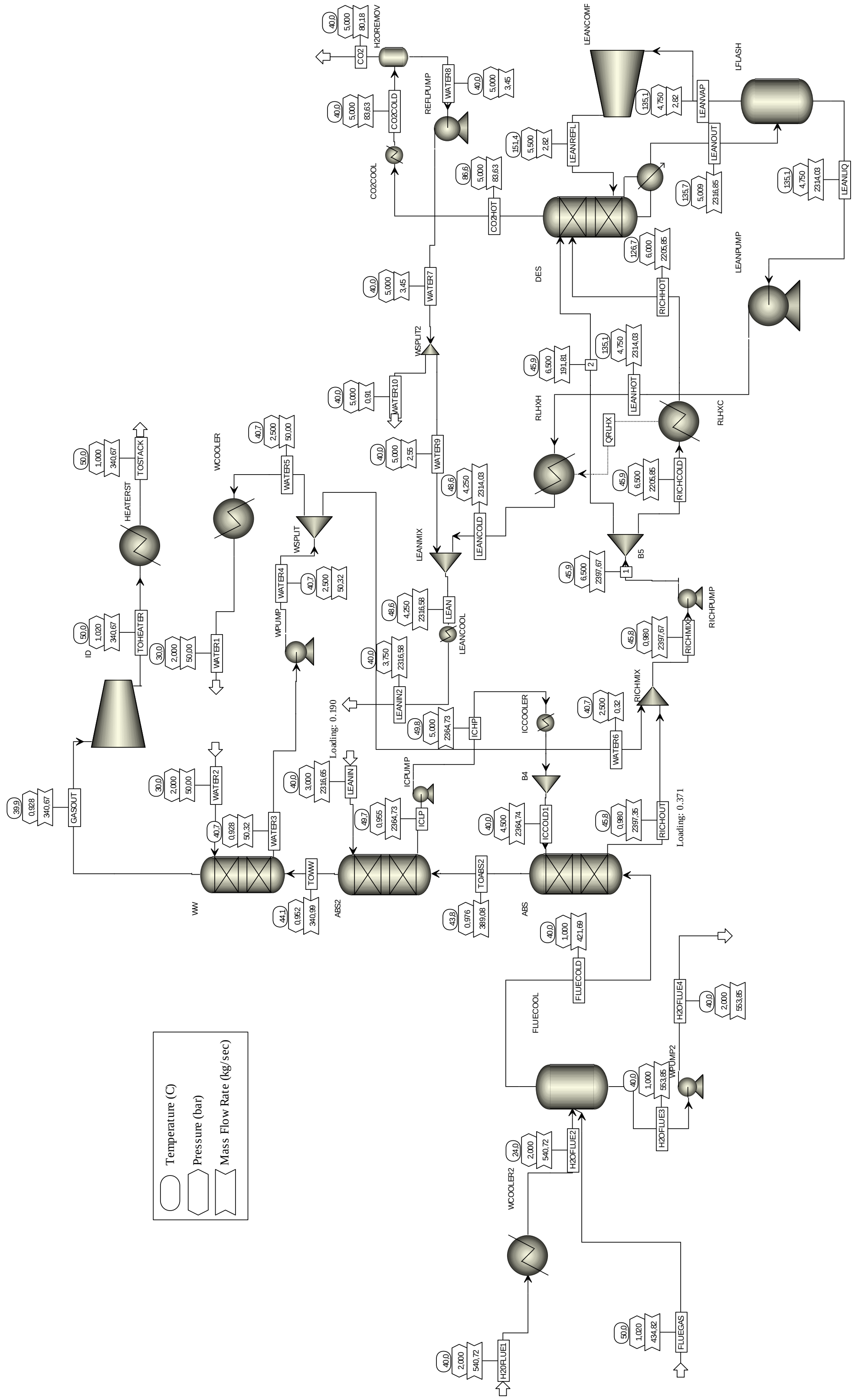

Figure 94: Flow sheet of capture plant with vapour recompression and split flow in combination with SCPC plant

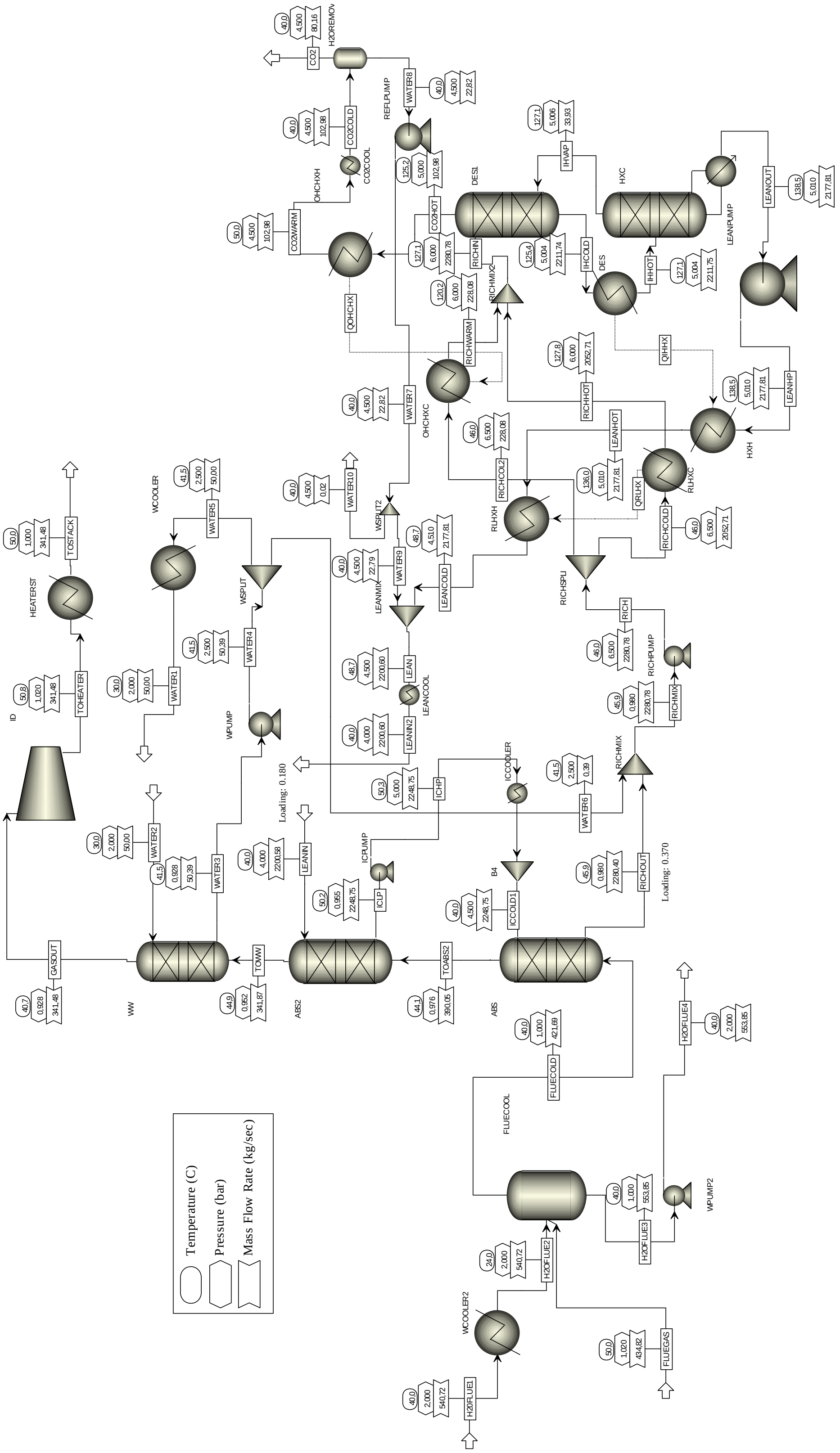

Figure 95: Flow sheet of capture plant with heat-integrated stripper and overhead condenser heat integration in combination with SCPC plant

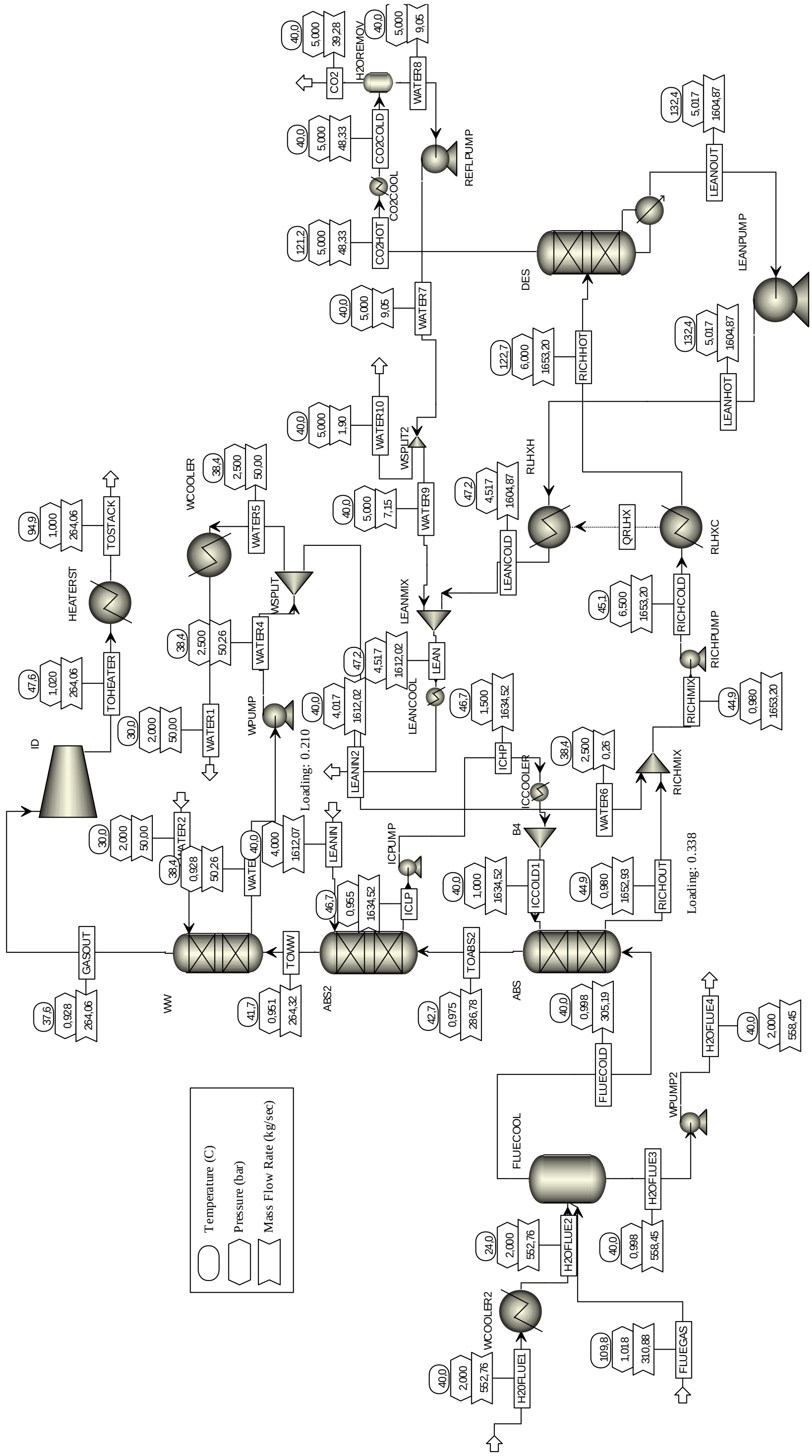

Figure 96: Flow sheet of capture plant in combination with NGCC plant for base case

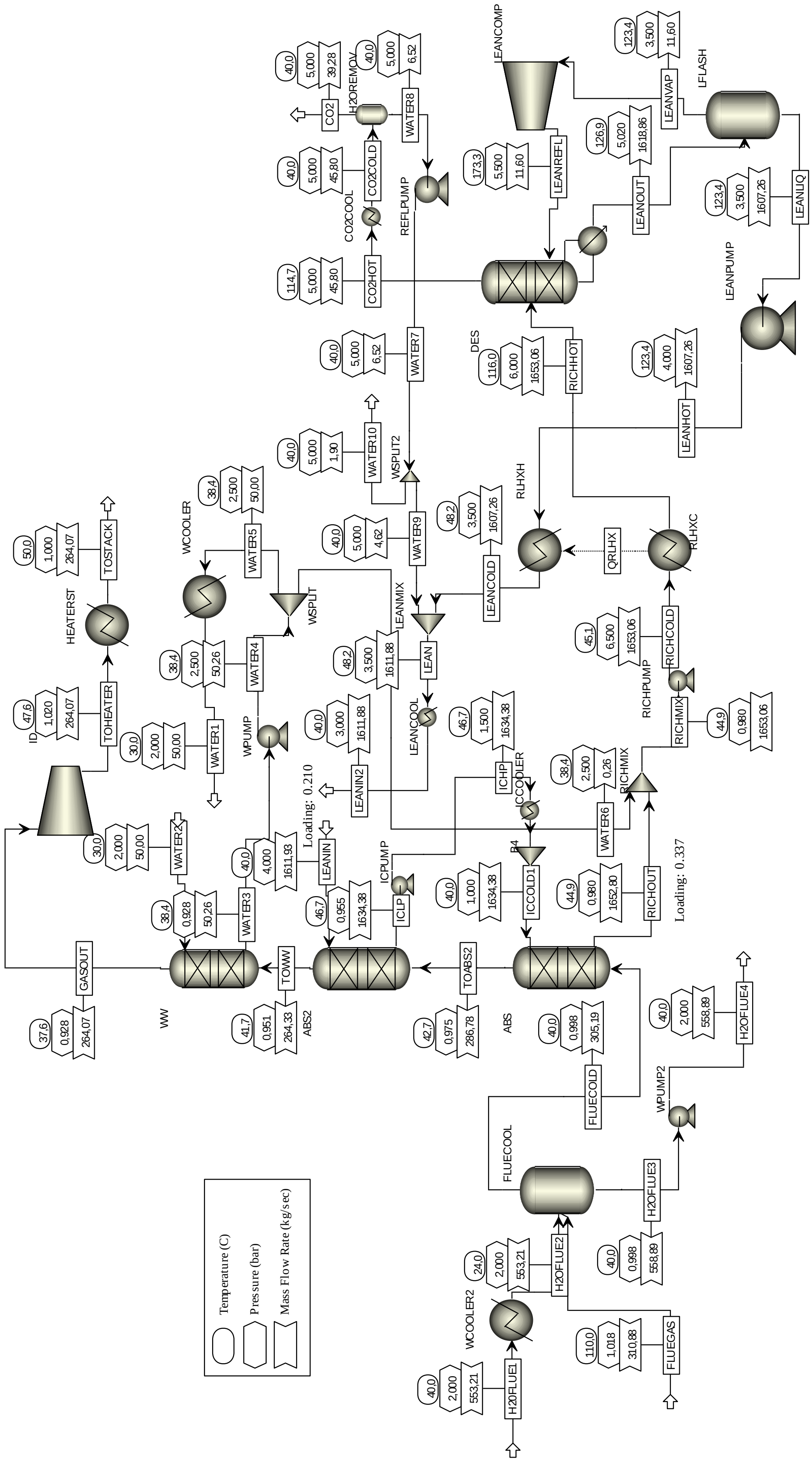

Figure 97: Flow sheet of capture plant with vapour recompression in combination with NGCC plant

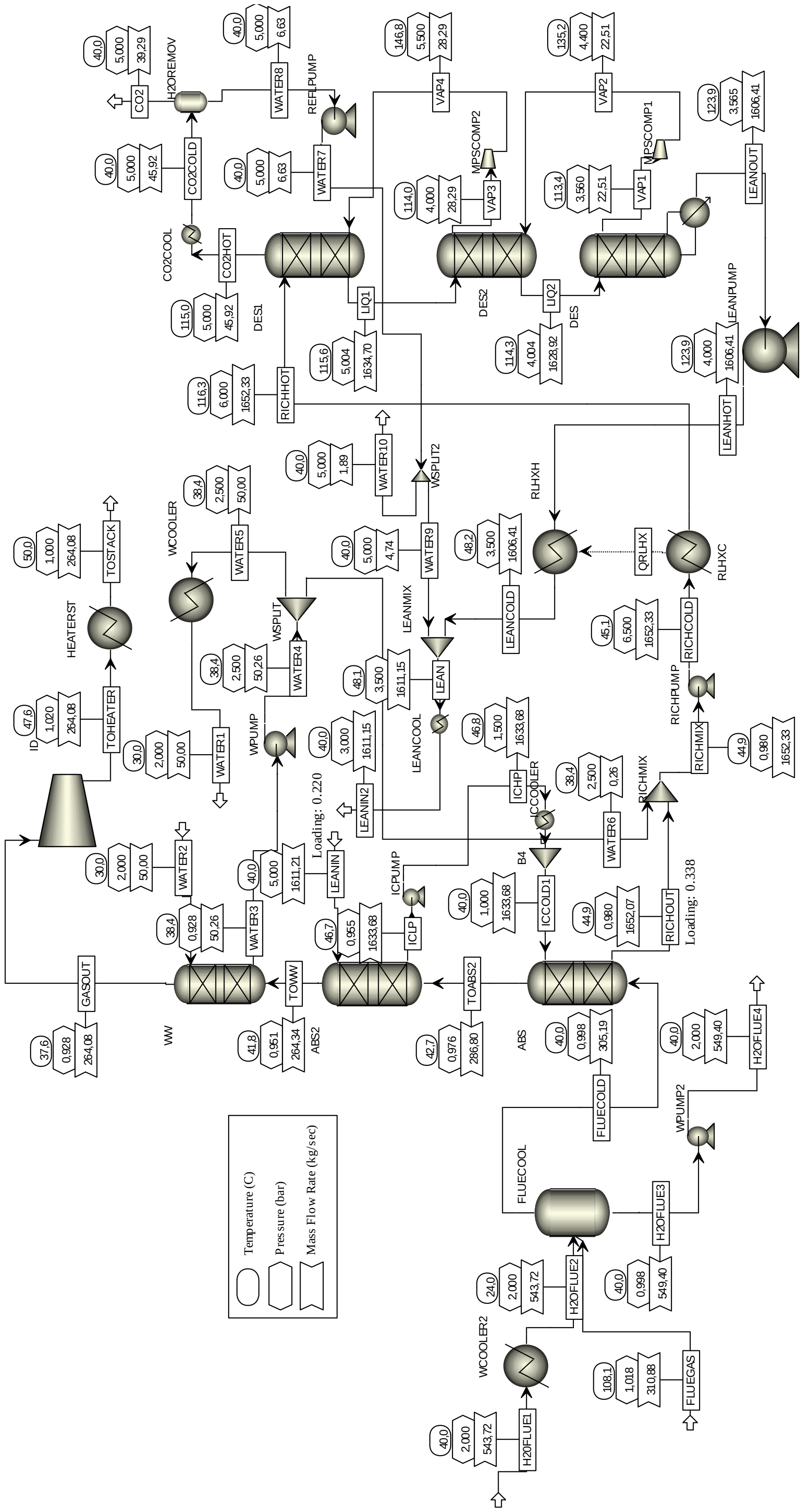

Figure 98: Flow sheet of capture plant with multi-pressure stripper in combination with NGCC plant

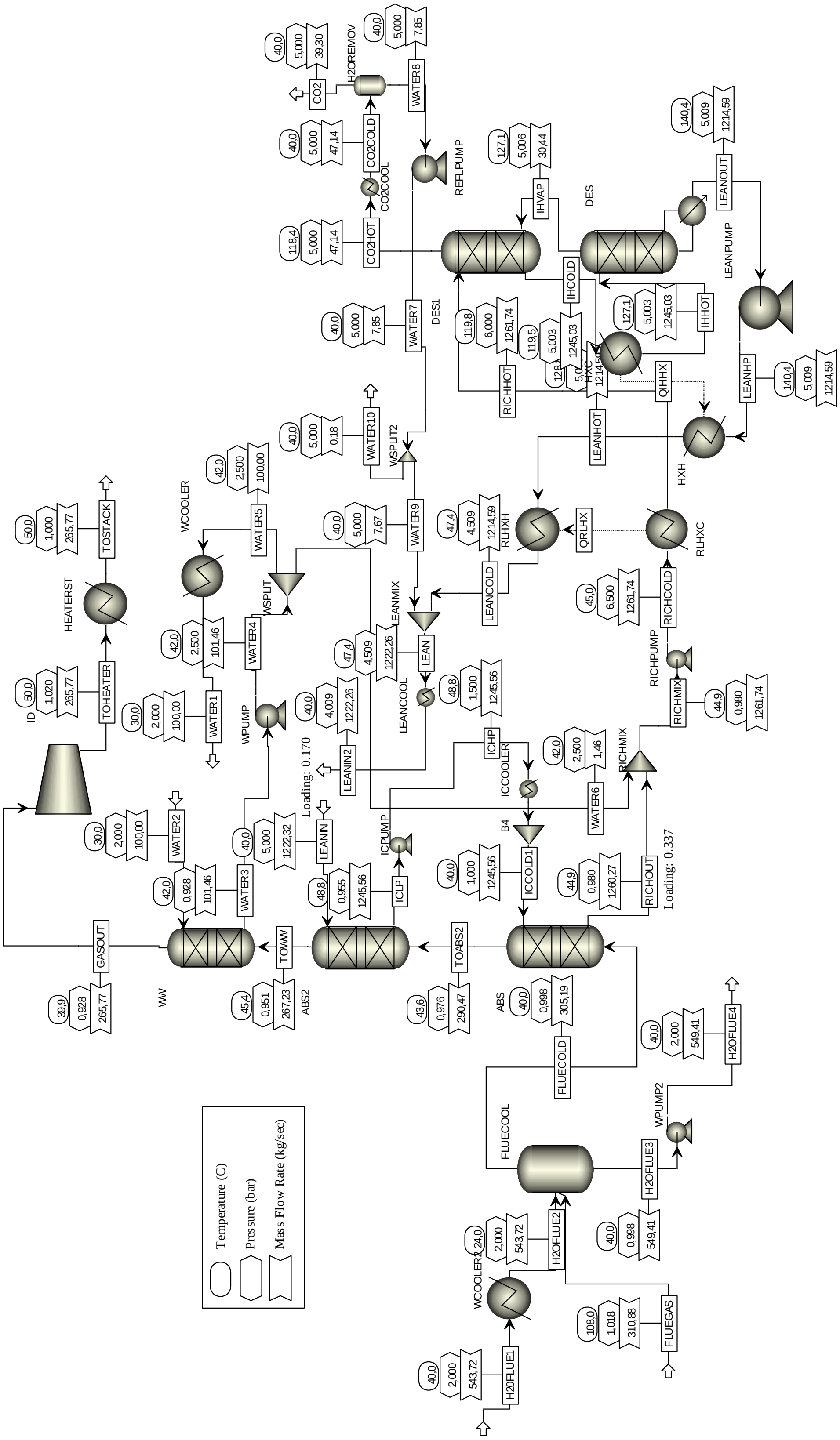

Figure 99: Flow sheet of capture plant with heat-integrated stripper in combination with NGCC plant

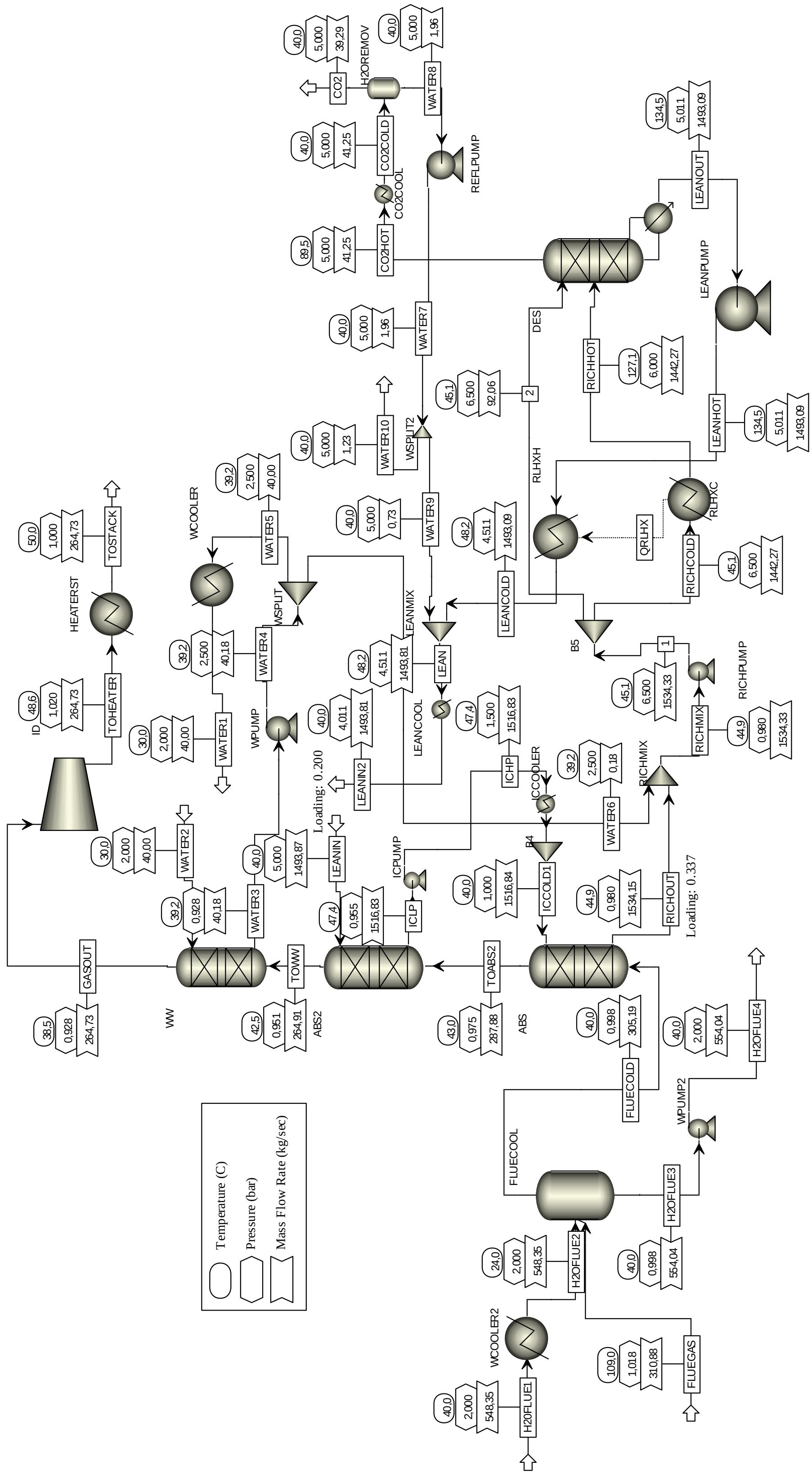

Figure 100: Flow sheet of capture plant with split flow in combination with NGCC plant

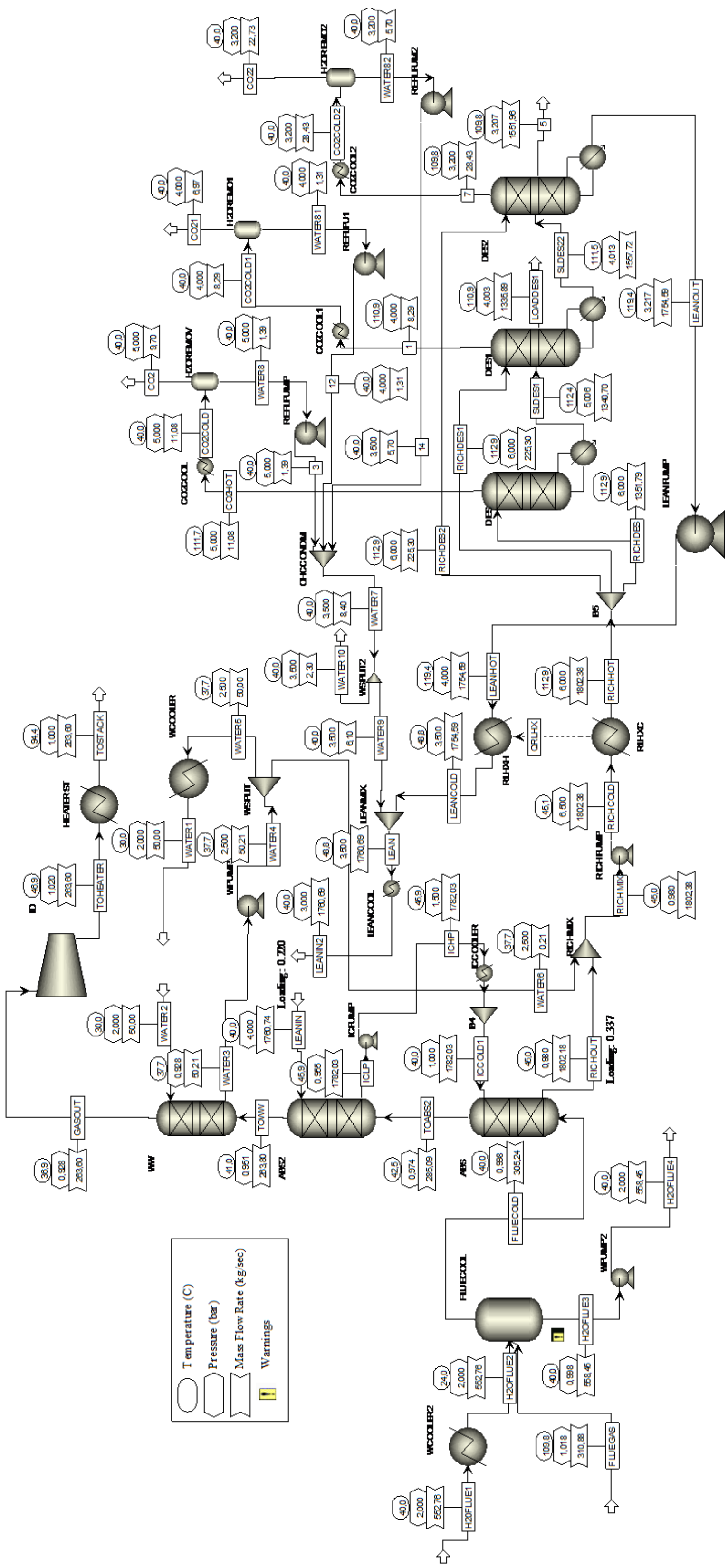

Figure 101: Flow sheet of capture plant with matrix stripping in combination with NGCC plant

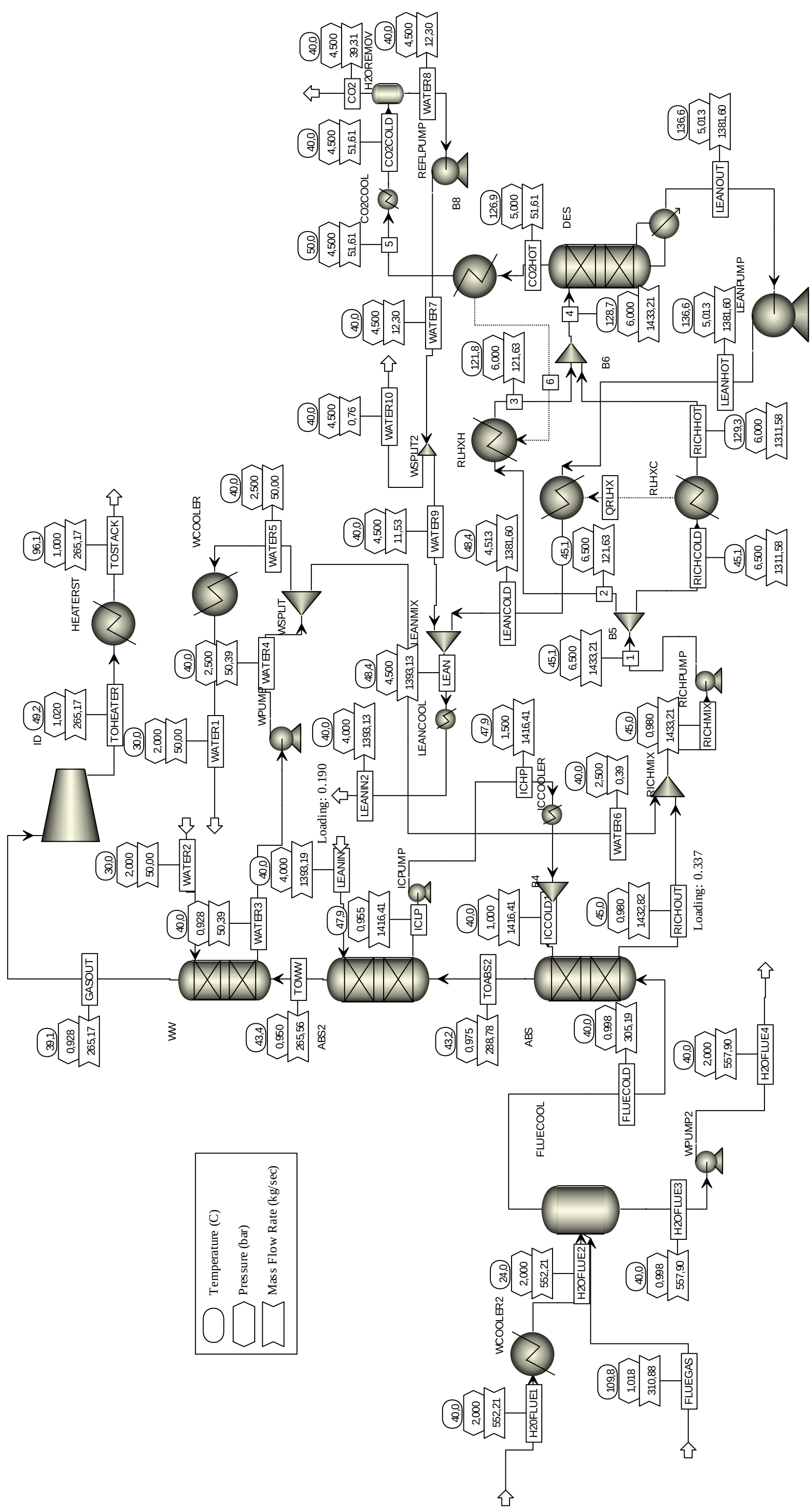

Figure 102: Flow sheet of capture plant with overhead condenser heat integration in combination with NGCC plant

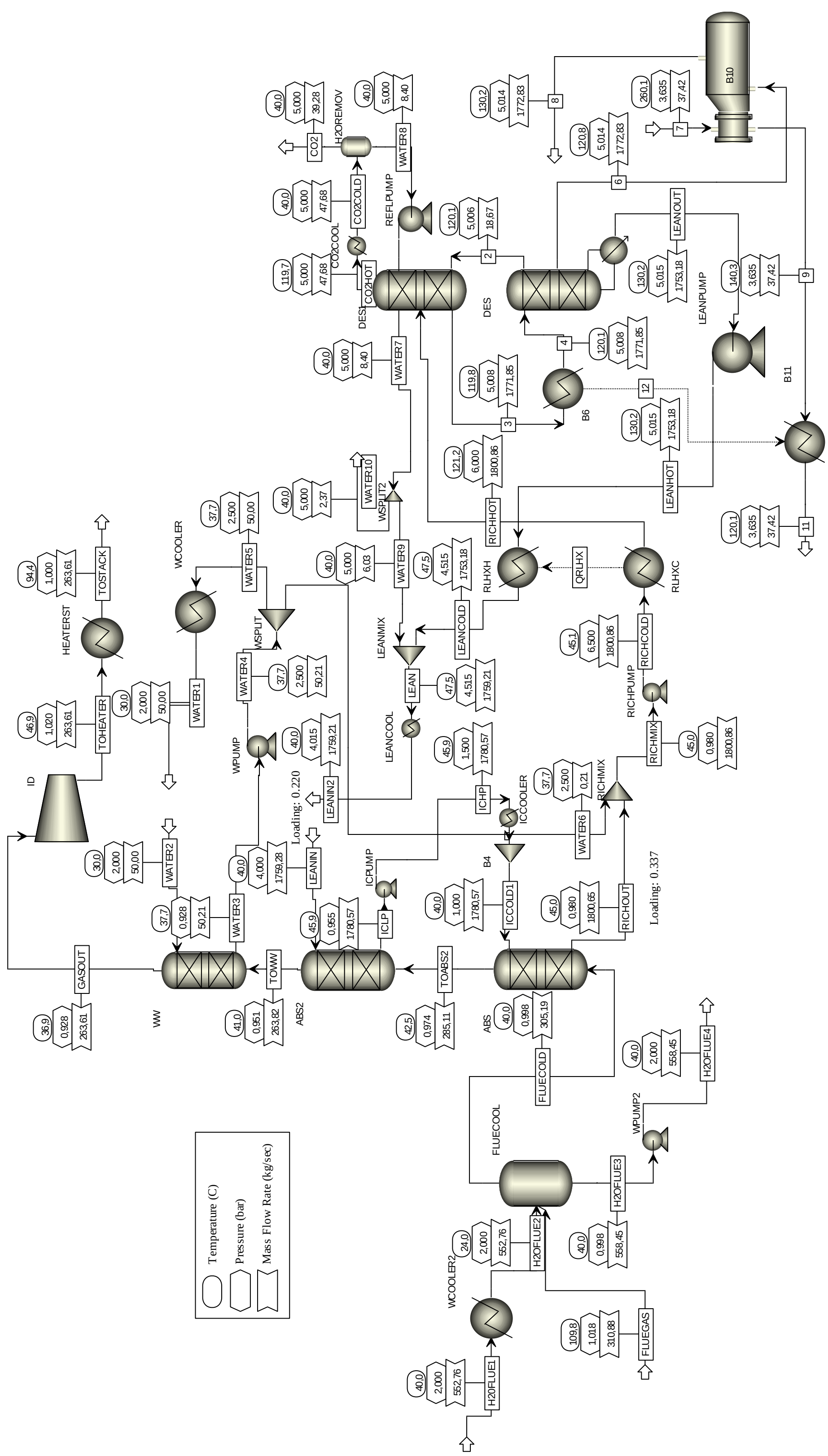

Figure 103: Flow sheet of capture plant with reboiler condensate heat integration in combination with NGCC plant

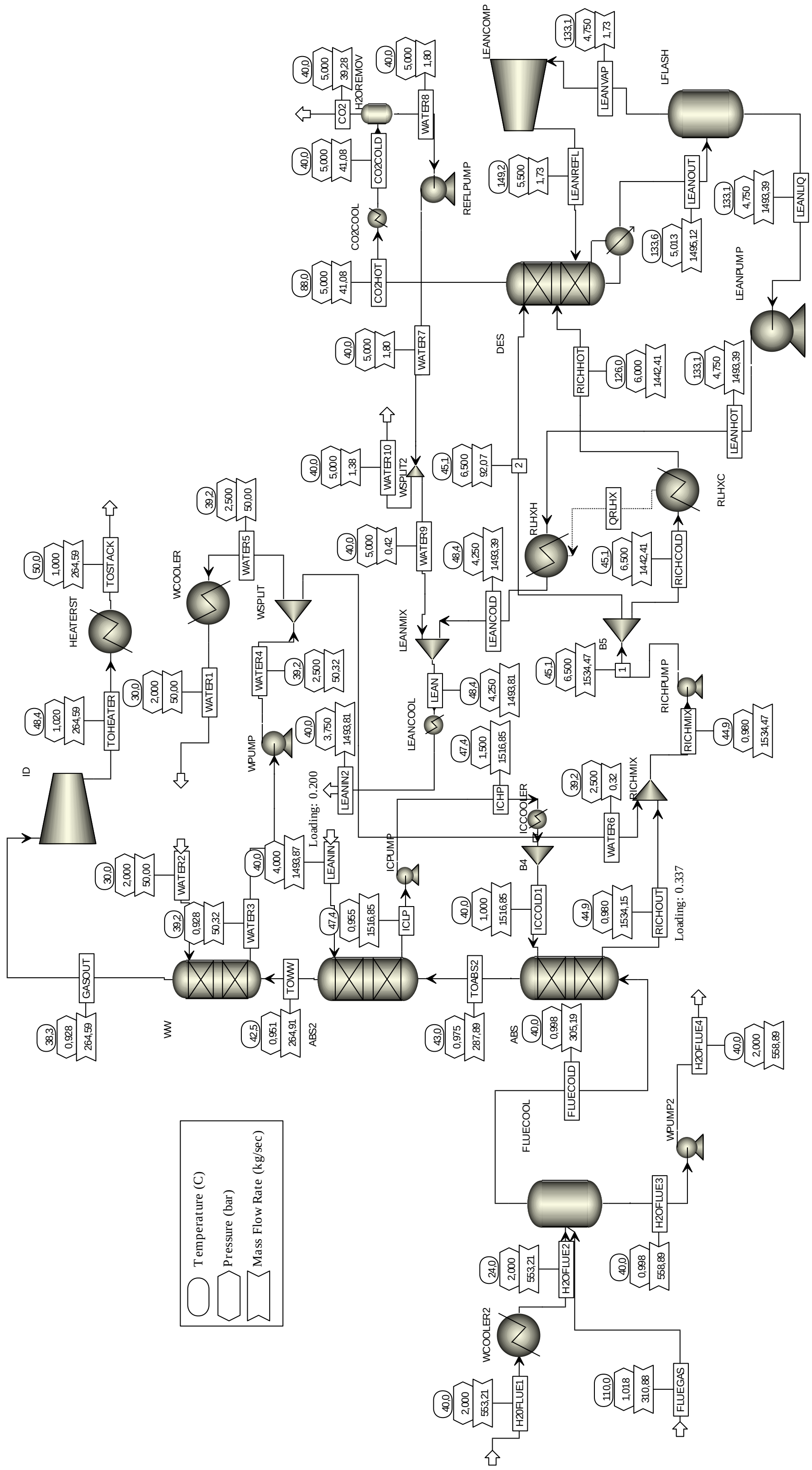

Figure 104: Flow sheet of capture plant with vapour recompression and split flow in combination with NGCC plant

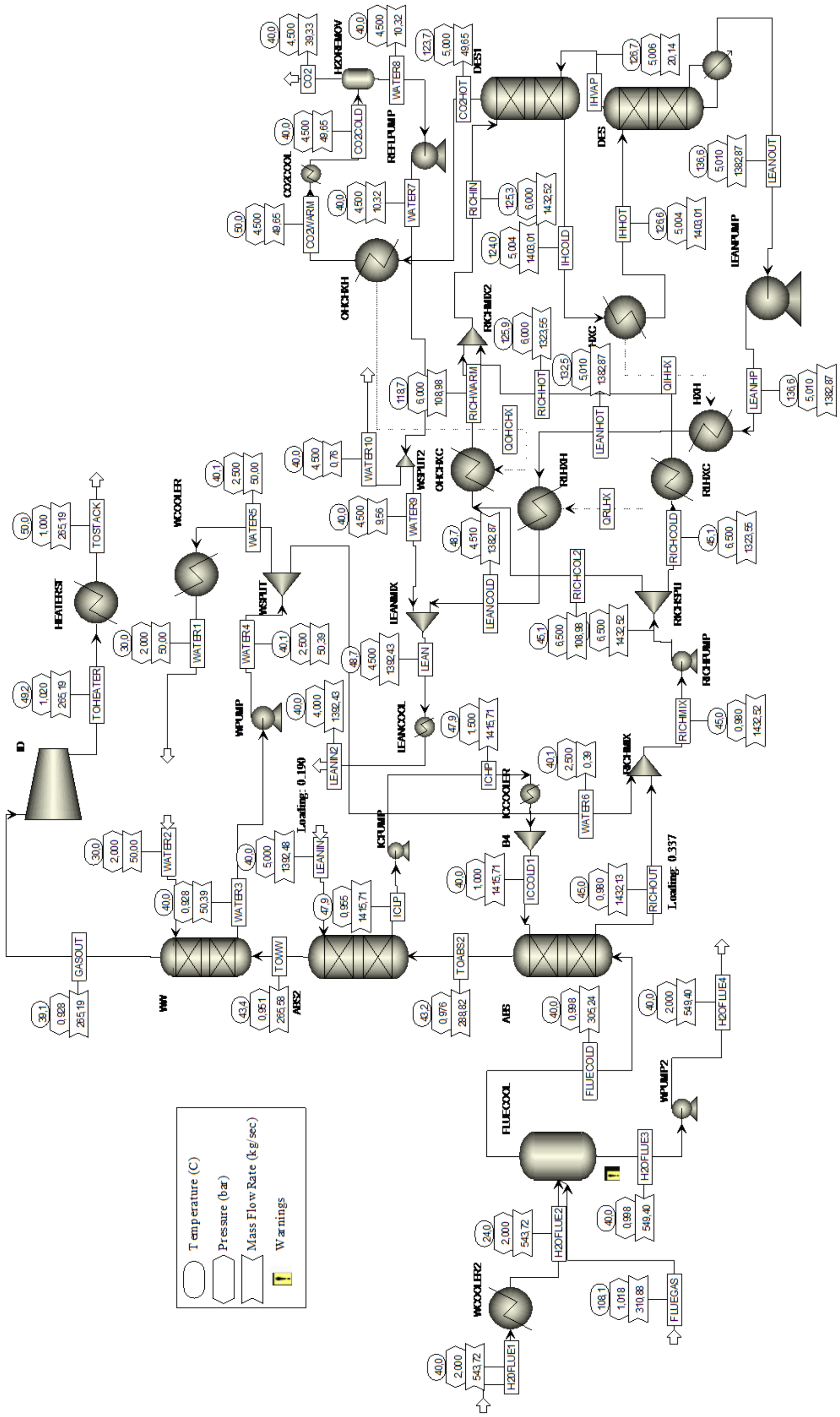

Figure 105: Flow sheet of capture plant with heat-integrated stripper and overhead condenser heat integration in combination with NGCC plant