

# **Selective exhaust gas recirculation in combined cycle gas turbine power plants with post-combustion CO<sub>2</sub> capture**

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## Appendix A. Design Basis and description of the CCGT power plant and the PCC system for the reference configuration.

### A.1.- Ambient conditions

Table A.1.- Ambient conditions

Pressure (ISO)	kPa	101.325
Temperature (ISO)	°C	15
Relative humidity, $\phi$	%	60
<i>Composition, dry molar fraction (%)</i>		
N <sub>2</sub>	%vol	78.08
CO <sub>2</sub>	%vol	0.03
H <sub>2</sub> O	%vol	0.00
Ar	%vol	0.94
O <sub>2</sub>	%vol	20.95
Molar Mass	g/mol	28.86

### A.2.- Natural gas

Table A.2.- Natural gas supply specifications

Fuel type	Natural Gas	
Supply temperature	°C	10
Supply pressure	MPa	7
<i>Composition, molar fraction (%)</i>		
CH <sub>4</sub>	%vol	89
C <sub>2</sub> H <sub>6</sub>	%vol	7
C <sub>3</sub> H <sub>8</sub>	%vol	1
C <sub>4</sub> H <sub>10</sub>	%vol	0.11
N <sub>2</sub>	%vol	0.89
CO <sub>2</sub>	%vol	2
S	ppm	<5
Molar mass	g/mol	17.84
CO <sub>2</sub> emissions	g/kWh LHV	208.00
<i>Heating values</i>		
Low heat value (LHV) @ 25 °C	MJ/kg	46.94
High heat value (HHV) @ 25 °C	MJ/kg	51.58

### A.3.- Design and operating parameters for the CCGT plant in the reference configuration.

The reference plant is an air-based combustion CCGT with a 2-on-1 configuration: two GE Class F (GE9371B) gas turbines with the flue gas exiting into two HRSGs, which jointly supply steam to a subcritical triple pressure steam cycle. The reference power plant is designed according to the configuration of a report commissioned by the International Energy Agency Greenhouse Gas R&D programme (IEA 2012). The net power output and the thermal efficiency of the CCGT plant equipped with PCC is presented in Table A.3. Design and operating parameters of the gas turbine engine are presented in Table A.4 and have been obtained from General Electric technical specifications (General Electric Power Generation 2016) and data from journal publications using the same class of gas turbine

engine (Sánchez et al. 2010, Jordal et al. 2012, Jonshagen et al. 2011, Chacartegui et al. 2012). Design and operating parameters of the HRSG are presented in Table A.5 and technical details of the steam turbines are presented in Table A.6.

Table A.3.- Power output and thermal efficiency for the air-based combustion CCGT in the reference configuration

<b>Combined Cycle Gas Turbine plant</b>			
Gas turbine net power per GT-HRSG train	MWe		285.8
Gas turbines net power	MWe		571.6
Open cycle thermal efficiency	% LHV		38.18
Steam turbine power	MWe		256.0
Fuel heat input per GT-HRSG train	MWth		748.0
CCGT gross power	MWe		827.6
CCGT gross thermal efficiency	% LHV		55.50
<b>CCGT net power output</b>	<b>MWe</b>		<b>777.44</b>
<b>CCGT net thermal efficiency</b>	<b>%</b>		<b>51.94</b>

Table A.4.- Gas turbine engine design parameters per GT-HRSG train in the reference configuration

<b>Gas turbine</b>			
<b>Model</b>	<b>GE 9371FB</b>		
Air mass flow rate	kg/s		641.81
Fuel mass flow rate	kg/s		16.10
<b>Compressor</b>			
No. stages	--		18
Inlet pressure drop	kPa		1
Pressure ratio (PR)	--		18.1
Compressor isentropic efficiency	%		79.5%
<b>Combustor</b>			
Pressure drop	%		5
Combustor efficiency	%		99.8%
P fuel in	bar		27.20
T fuel in	°C		117
<b>Turbine</b>			
No. stages	--		4
Turbine inlet temperature (TIT)	°C		1371
Turbine inlet temperature (TIT)	K		1644
Turbine back pressure	bar		1.039
Turbine isentropic efficiency	%		90.5%
<b>Exhaust flue gas (EFG)</b>			
Pressure	kPa		103.9
Temperature	°C		643.29
Exhaust flue gas flow rate	kg/s		657.92
Composition:			
	CO <sub>2</sub>	% vol	4.21%
	H <sub>2</sub> O	% vol	8.82%
	N <sub>2</sub>	% vol	74.21%
	O <sub>2</sub>	% vol	11.87%
	Ar	% vol	0.89%
Molar mass		g/mol	28.38
<b>Gas turbine performance</b>			
Mechanical efficiency compressor	%		99.6
Net Power	MW		285.76
Net Thermal efficiency	%		38.18
Net Heat Rate		kJ/kWh	9428.07

Table A.5.- Heat recovery steam generator design parameters per GT-HRSG train for the reference configuration

<b>Heat recovery steam generator</b>		
<b>Flue gas</b>		
Flue gas mass flow rate	kg/s	657.92
Gas Inlet temperature	°C	643.29
Gas Outlet temperature	°C	117.87
Feed water temperature	°C	32.28
Gas inlet pressure	bar	1.039
Gas outlet pressure	bar	1.013
HRSG efficiency	%	99.7
<b>Pressure levels</b>		
HP drum pressure	bar	173
IP drum pressure	bar	43
LP drum pressure	bar	4
<b>Temperature differences</b>		
$\Delta T$ approach main steam	°C	42.5
$\Delta T$ approach hot reheated steam	°C	34
$\Delta T$ pinch gas boiling - liquid in evaporator	°C	10
$\Delta T$ subcooling economiser	°C	3
<b>Pressure losses</b>		
$\Delta P$ gas side	kPa	2.6
$\Delta P$ HP SH	%	3.5
$\Delta P$ HP ECO	%	2.6
$\Delta P$ IP RH	%	3
$\Delta P$ IP SH	%	2
$\Delta P$ IP ECO	%	3
$\Delta P$ LP SH	%	2
$\Delta P$ LP ECO	%	1.3
$\Delta P$ system - LIVE STEAM		
$\Delta P$ system - COLD RH	%	7
$\Delta P$ system- HOT RH	%	9
$\Delta P$ system - LP STEAM	%	9
<b>Temperature losses</b>		
From superheater / reheater to turbine (approximately 0.5 K)	kJ/kg	1
<b>Condenser</b>		
Pressure condenser	kPa	4.814
Saturation temperature	°C	32.20
Temperature pinch	°C	3.2
Feed water temperature (after feed water pump)	°C	32.2
Cooling water supply (CWS) temperature	°C	18
Cooling water return (CWR) temperature	°C	29
Condenser duty	MW	224.3
Power to thermal duty (Electric consumption for heat rejection of rejected thermal power)	%	0.8
<b>Fuel heater</b>		
Fuel Inlet Temperature	°C	9.00
Fuel Outlet Temperature	°C	117.00
IP water inlet temperature	°C	252.85
Temperature pinch	°C	26.98
<b>Efficiency calculations</b>		
Pump efficiency	%	70
Generator efficiency	%	98.5

Table A.6.- Steam turbines design and operating parameters for the two GT-HRSG trains for the reference configuration.

<b>Steam turbines</b>		
<b><i>HP STEAM TURBINE</i></b>		
Live steam molar flow	mol/s	9587
Live steam mass flow	tn/h	621
P in	bar	170
T in	°C	600
P out	bar	45.2
T out	°C	394
Isentropic efficiency	%	88.1%
<b><i>IP STEAM TURBINE</i></b>		
Hot reheated molar flow	mol/s	10445
Hot reheated mass flow	tn/h	677
P in	bar	40
T in	°C	600
P out	bar	3.75
T out	°C	266.8
Isentropic efficiency	%	92.4%
<b><i>LP STEAM TURBINE</i></b>		
Turbine exhaust molar flow	mol/s	5595
Turbine exhaust mass flow	tn/h	363
P in	bar	3.75
T in	°C	266.93
P out	bar	0.0481
Exhaust steam quality	%	91.9%
Isentropic efficiency	%	88.0%

#### ***A.4.- Design and operating parameters for the CO<sub>2</sub> capture plant in the reference configuration***

The carbon capture process is modelled in Aspen Plus V7.0 commercial software (Aspen Tech). The process flow diagram of the conventional MEA based chemical absorption process, as set up in Aspen Plus, is illustrated in Figure A.1. Design parameters are presented in Table A.7 and results from the model are presented in Table A.8. The electrolyte non-random two-liquid (E-RN2L) thermodynamic package is selected to describe the CO<sub>2</sub>-H<sub>2</sub>O-MEA chemistry in the liquid phase, and the Peng-Robinson equation of state for the gas phase.

The absorption process consists of three unit operations: direct contact cooler, absorber and water wash section. The absorber is simulated with a rate-based model which includes reaction kinetics, thermodynamic equilibrium, mass transfer and heat transfer phenomena. The reactions between CO<sub>2</sub> and MEA are very fast and take place in the liquid phase with no reaction in the gas phase (Razi et al. 2013). Mass and heat transfer rates between the contacting phase are therefore based on the combined diffusion-reaction process in the liquid phase, while phase equilibrium exists at the interface. For the gas phase only diffusional resistance is taken into account. Effective interfacial area and liquid side mass transfer coefficients are determined using Bravo-Rocha-Fair correlations (Bravo et al 1982, Bravo et al 1985) which are valid for structured packing Sulzer Mellapak 250Y, and heat transfer coefficients are predicted by the Chilton and Colburn analogy. The bulk properties are evaluated at the outlet conditions for the liquid phase (well mixed flow model) and at average conditions for the vapor phase (plug flow model). The kinetic model of the CO<sub>2</sub>-MEA-H<sub>2</sub>O system considered in Razi's work, where results from the absorber model were in good agreement with experiment data sets from the Esbjerg CESAR (CO<sub>2</sub> Enhanced Separation and Recovery) pilot plant, for CO<sub>2</sub> capture with aqueous MEA solution at 30 wt% from a flue gas slip of a 400 MWe pulverized coal-fired power plant (Razi et al. 2013; Sanchez Fernandez et al. 2014). The modelling results for the reference case, a CCGT plant with PCC, are compared to data from a test campaign with 30 wt% MEA at the CO<sub>2</sub> Technology Centre, in Mongstad, Norway (Hamborg et al. 2014).

The lean-rich solvent heat exchanger is designed on the basis of a fixed overall heat transfer coefficient and a temperature approach of 8 °C (cold out – hot in approach).

The stripper is simulated with an equilibrium model. Due to the high operating temperature, the kinetic of the reactions is fast and the process is thermodynamically controlled. Moreover, a high temperature enhances the mass transfer and mass transfer resistance is neither the limiting step. The stripper is designed to achieve a fixed molar CO<sub>2</sub> recovery ratio and the number of equilibrium stages is consequently defined. The CO<sub>2</sub> recovery ratio is set to control the lean solvent CO<sub>2</sub> loading at which the stripper column operates and this value is selected to bring the specific heat consumption to a minimum.

Table A.7.- Design parameters of the PCC system for the reference configuration

<b><i>Flue gas conditioning system</i></b>		
Booster fan pressure ratio	bar	0.1
Booster fan isentropic efficiency	%	85
Booster fan mechanical efficiency	%	95
Direct contact cooler flue gas outlet temperature	°C	45
Direct contact cooler + rotary heat exchanger pressure drop	bar	0.05
<b><i>Absorber</i></b>		
Solvent		MEA
Concentration	% wt.	30
Lean solvent temperature	°C	40
Flue gas inlet temperature	°C	45
Flue gas temperature water wash outlet	°C	45
Absorber column pressure	bar	1.1
Flooding factor	%	80
Packing		Mellapak 250 Y
Absorber column pressure drop	mbar	50
<b><i>Lean/rich heat exchanger</i></b>		
Lean/rich heat exchanger DT min	°C	8
Lean/rich stream heat exchanger pressure drop	kPa	20
Rich pump outlet pressure	bar	
<b><i>Stripper</i></b>		
Stripper column pressure drop	mbar	300
Stripper column number of stages	--	8
Flooding factor	%	80
<b><i>Reboiler</i></b>		
Reboiler pressure / saturated	bar	3
Saturated steam inlet temperature	°C	133
Lean solvent outlet temperature	°C	120
Pinch temperature	°C	13
<b><i>Lean/ Rich solvent pumps</i></b>		
Pumps hydraulic efficiency	%	75
Pumps driver efficiency	%	95

Table A.8.- Operating parameters of the PCC plant for the reference configuration per GT-HRSG train



<b>CASE</b>	<b>Air-based combustion (reference case)</b>	
Overall CO <sub>2</sub> capture level	%	90
<b>Absorber</b>		
<i>Flue gas BOTTOM</i>		
Mass flow rate	kg/s	658
CO <sub>2</sub> conc.	% vol	4.21
<i>Flue gas TOP</i>		
CO <sub>2</sub> conc.	% vol	0.43
Absorber efficiency	%	90
Rich solvent loading	mol <sub>CO<sub>2</sub></sub> /mol <sub>MEA</sub>	0.26
Lean solvent loading	mol <sub>CO<sub>2</sub></sub> /mol <sub>MEA</sub>	0.458
Lean solvent molar flow rate	mol/s	886
<i>Absorber dimensions</i>		
No. Absorbers	--	2
Diameter	m	12
Packing height	m	19
Total packing volume per GT+HRSG train	m <sup>3</sup>	4190
<b>Stripper</b>		
Stripper pressure	bar	1.84
Steam specific consumption	kg/kg CO <sub>2</sub>	1.71
Specific reboiler duty	MJ/kg CO <sub>2</sub>	3.746
CO <sub>2</sub> to pipeline	kg/s	76.5
<b>CO<sub>2</sub> compression train</b>		
CO <sub>2</sub> final pressure	bar	110
Specific compression work	kJ/kg CO <sub>2</sub>	334

#### ***A.5.- Design and operating parameters for the CO<sub>2</sub> compression train in the reference configuration***

The CO<sub>2</sub>-rich gas stream leaves the condenser of the stripper column at 40 °C and 2 bar, with a CO<sub>2</sub> purity of 95 vol% and needs to be conditioned prior to transport and storage/utilization. The CO<sub>2</sub>-rich stream is compressed up to around the critical pressure (73.8 bar) in the compression train, which consists of three stages with intercooling and water separation between stages. Liquid phase CO<sub>2</sub> at 73 bar and 28 °C is pumped up to 110 bar for transfer and storage in supercritical/dense phase. The operating parameters are based on the Common Framework Definition Document published by the European Benchmarking Task Force (EBTF 2011). The process flow diagram of the compressor train is illustrated in Figure A.2. Design and operating parameters are presented in Table A.9.

Table A.9.- Design parameters and operating variables of the CO<sub>2</sub> compression train for the reference configuration.

#### ***CO<sub>2</sub> compression plant***

Final delivery pressure	bar	110
CO <sub>2</sub> purity	%	99.8
Compression stages pressure ratios	--	4.3 / 4.3 / 3.7
Compressor isentropic efficiency	%	85
Compressor mechanical efficiency	%	95
Intercooler stages	--	3
Intercoolers outlet temperature	°C	28
Pump efficiency	%	75
Pump delivery pressure	bar	110

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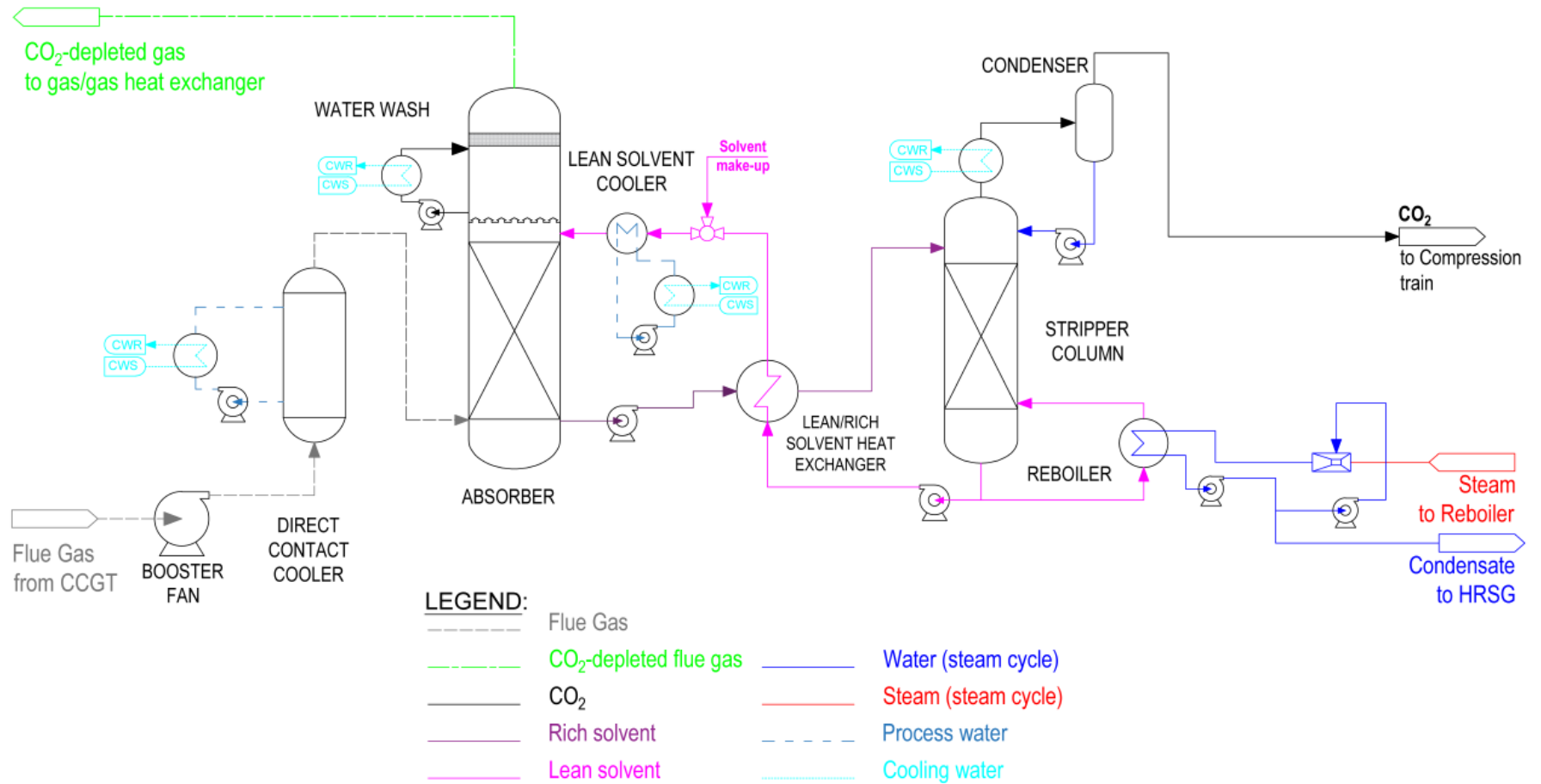


Figure A.1.- Process flow diagram of the post-combustion CO<sub>2</sub> capture system with amine-based chemical absorption technology.

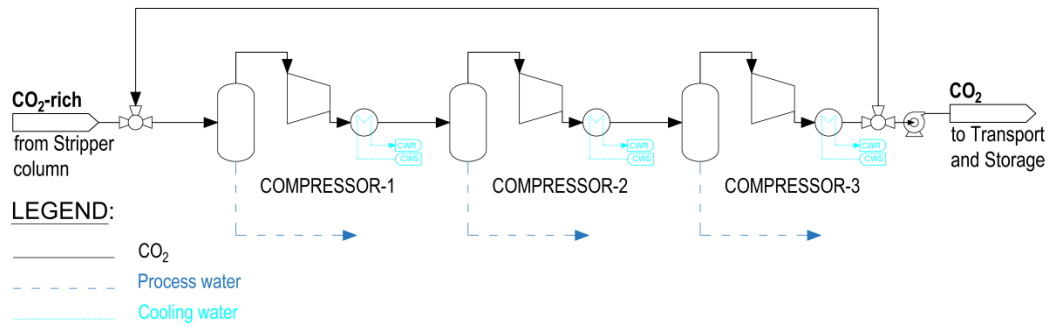


Figure A.2.- Process flow diagram of the CO<sub>2</sub> compression train.

## **Appendix B. Sizing and optimisation of the PCC system with SEGR**

The procedure followed to size and optimise the PCC system is described here and illustrated in Figures B.1, for SEGR in parallel operating at 97% PCC efficiency, and in Figure B.2, for SEGR in series operating at 31% PCC efficiency.

### Absorber sizing

Operating and design parameters in the CO<sub>2</sub> capture system are evaluated following the procedure described by Freguia and Rochelle (Freguia and Rochelle 2003) to achieve the CO<sub>2</sub> capture efficiency in the PCC process that is required in each configuration for a 90% overall CO<sub>2</sub> capture level. The absorber diameter is determined for a flue gas velocity that corresponds to 80% of the velocity at the flood-point (Oexmann, Hensel and Kather, 2008). The absorber packing height is then increased for a constant diameter. A larger packing volume results in a larger contact surface area and a longer residence time, which enhances the CO<sub>2</sub> absorption rate and, thus, the CO<sub>2</sub> loading of the solvent at the bottom of the absorber (rich solvent). For a constant lean solvent CO<sub>2</sub> loading, it results in an increased solvent capacity and, thus, less amount of solvent is required to achieve certain CO<sub>2</sub> absorption efficiency. The absorber packing height is augmented up to a value at which a further increase results in a marginal gain in the rich solvent CO<sub>2</sub> loading (< 0.2% of the previous value) and in a marginal reduction of the reboiler duty. The absorber packing height is optimised for each one of the considered configurations.

### Reboiler duty optimisation

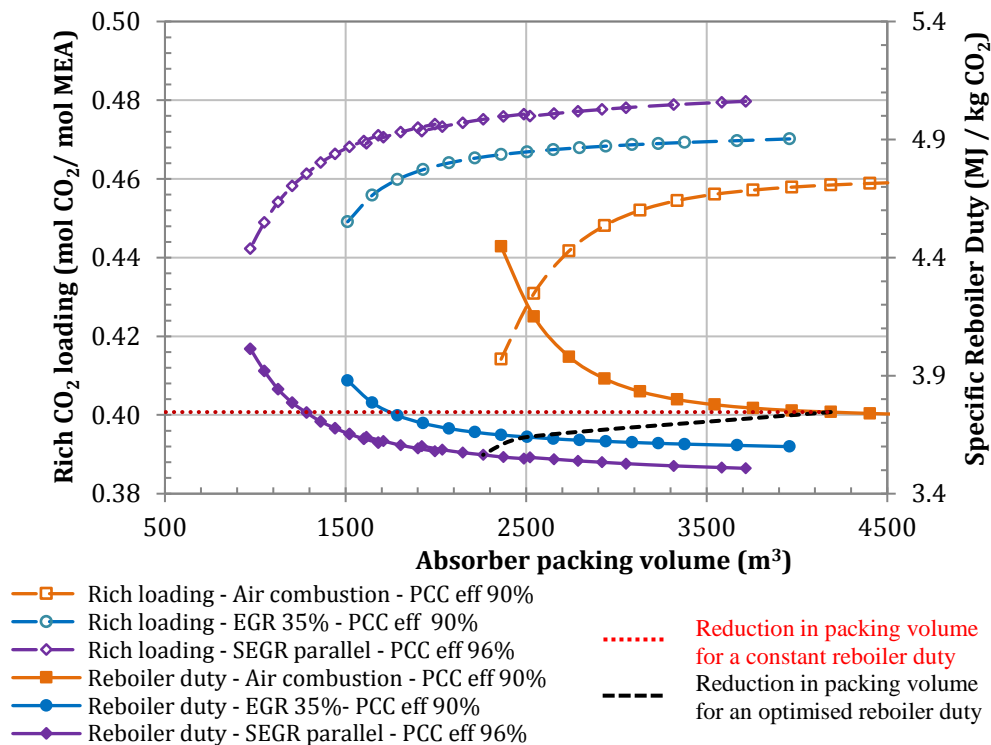
The thermal energy supplied in the reboiler by condensing steam is used in the stripper column to provide: the sensible heat required to heat up the solvent to the reboiler temperature, the heat of desorption of CO<sub>2</sub> required to reverse the chemical reactions between the amine and the CO<sub>2</sub> and release CO<sub>2</sub> into the vapor phase, and the latent heat for vaporisation of water to produce the stripping steam. The steam is condensed in the overhead condenser at 40 °C. The condensed water returns to the stripper as reflux and the CO<sub>2</sub> rich gas is sent to the compression train.

For conventional MEA solvent, the maximum allowable regeneration temperature is around 120 °C to prevent from thermal degradation, i.e. polymerisation (Rochelle, 2012). The temperature of the condensing steam in the reboiler exceeds the solvent temperature by the pinch temperature assumed in the reboiler design, i.e. 13 °C (Sanchez Fernandez et al. 2014). Saturated steam at 133 °C and 3 bar is supplied from the power plant steam cycle.

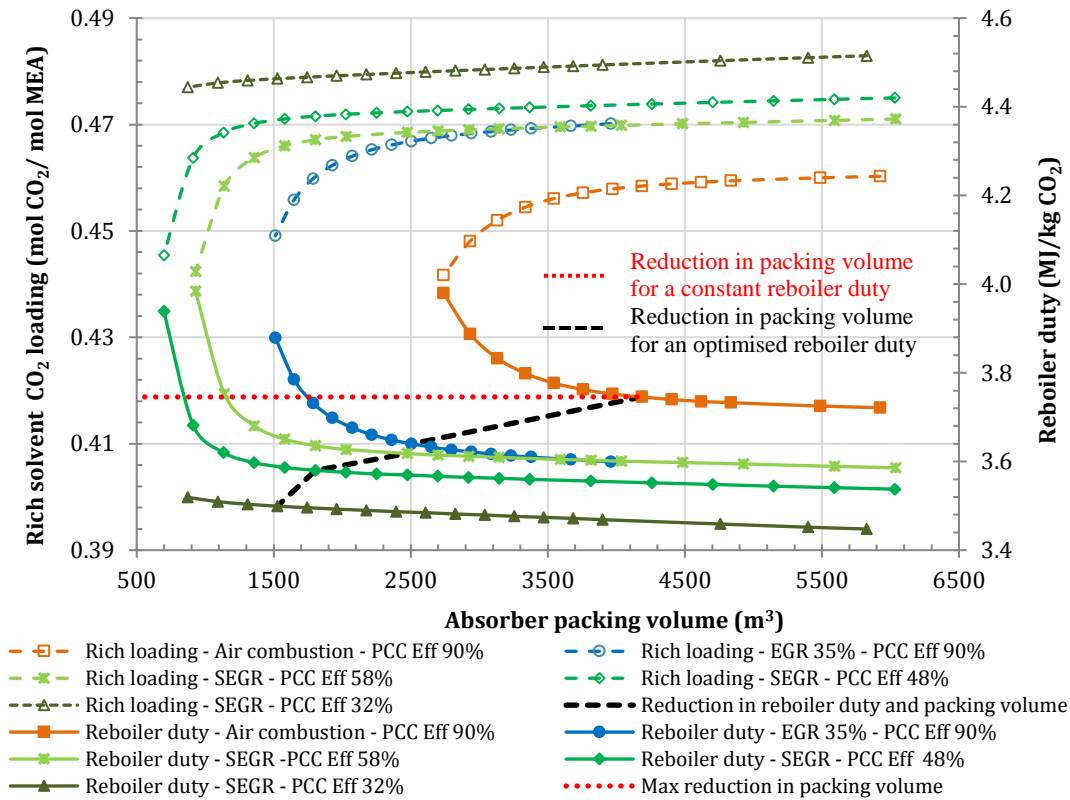
The optimum lean solvent flow rate entering the absorber which minimises the specific reboiler duty, in MJ/kg CO<sub>2</sub>, is evaluated for each configuration. The conditions in the stripper are set to achieve the lean solvent CO<sub>2</sub> loading resulting in the required CO<sub>2</sub> capture efficiency in the absorber to achieve an overall CO<sub>2</sub> capture level of 90%. The overall CO<sub>2</sub> capture level takes into account the amount of CO<sub>2</sub> exiting the boundaries of the plant.

The control strategy to vary the lean solvent CO<sub>2</sub> loading can be designed by varying either the temperature or the pressure in the stripper column. A constant temperature of the saturated stream supplied to the reboiler of 120 °C is assumed here and the total pressure in the stripper column is modified, i.e. with a back pressure valve. The saturated steam flow is evaluated to supply the reboiler duty necessary to achieve the lean solvent CO<sub>2</sub> loading. A smaller total pressure results in a smaller CO<sub>2</sub> partial pressure, as the water vapour partial pressure is almost constant at a given temperature. This fact displaces the equilibrium towards a lower CO<sub>2</sub> loading of the lean solvent, according to the vapour-liquid equilibrium curve. The larger solvent capacity results in a smaller solvent flow rate to capture 90% of the CO<sub>2</sub> generated in the combustion. An increase in the total pressure results in a higher CO<sub>2</sub> loading of the lean solvent and, thus, a larger solvent flow rate is necessary.

The effect on the reboiler duty is that the contribution of the sensible heat increases directly proportional to the solvent flow rate and, thus, becomes more significant at a higher lean solvent CO<sub>2</sub> loading. Meanwhile, the contribution of the latent heat for steam generation is more relevant at a lower solvent flow rate as more steam needs to be generated to strip off the CO<sub>2</sub> and achieve low lean solvent CO<sub>2</sub> loadings. There is therefore an optimal value of the lean solvent CO<sub>2</sub> loading that results in a minimum specific reboiler duty.



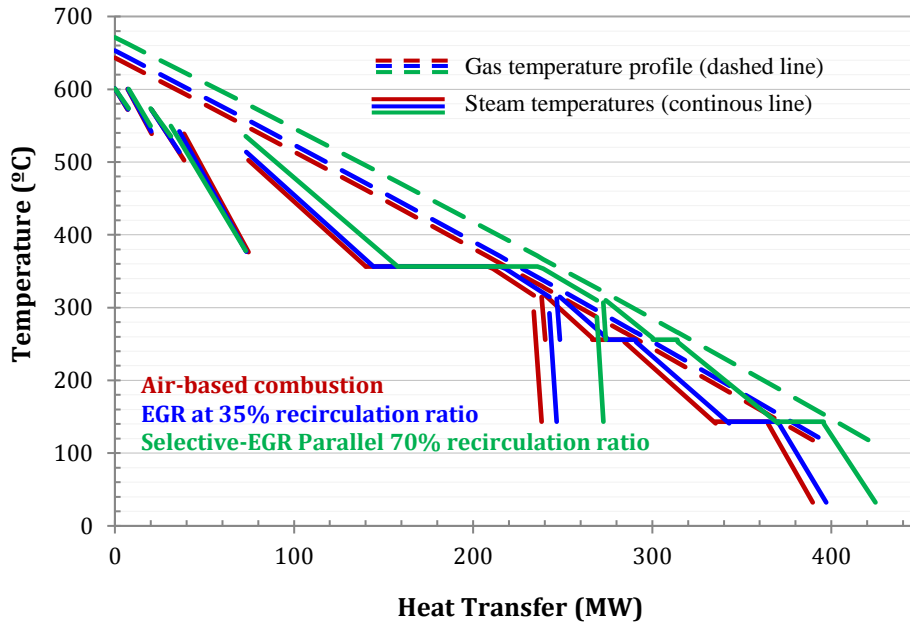
**Figure B.1.-** Sensitivity of the rich solvent CO<sub>2</sub> loading and the specific reboiler duty to the packing volume for S-EGR in parallel at 70% recirculation ratio and 96% post-combustion CO<sub>2</sub> capture efficiency. Compared to air-based combustion configuration and EGR at 35% recirculation ratio and 90% post-combustion CO<sub>2</sub> capture efficiency.



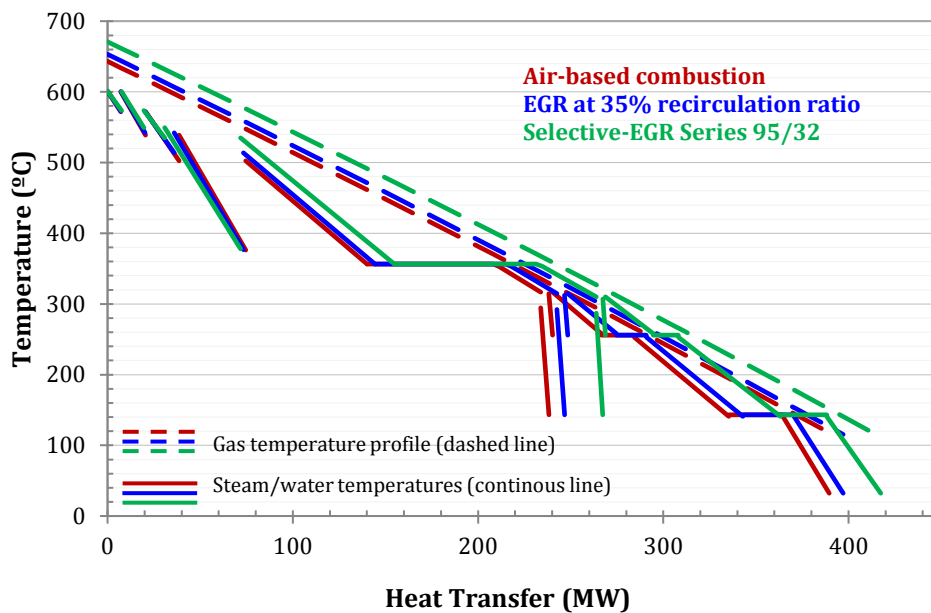
**Figure B.2.-** Sensitivity of the rich solvent CO<sub>2</sub> loading and the specific reboiler duty to the packing volume for S-EGR in series for the configurations operating at 32%, 48% and 58% CO<sub>2</sub> capture efficiency. Compared to air-based combustion configuration and EGR at 35% recirculation ratio for 90% CO<sub>2</sub> capture efficiency.

**Appendix C. Heat transfer and temperature diagram in the HRSGs for a CCGT plant with SEGR and PCC.**

The gas and water/steam temperature profile as a function of the heat transfer flow rate is shown in Figure C.1, for SEGR in parallel 97/96, and Figure C.2, for SEGR in series 95/31.



**Figure C.1.-** Heat transfer – temperature diagram for S-EGR configuration in parallel at 70% recirculation ratio, 97% selective CO<sub>2</sub> transfer efficiency and 96% post-combustion CO<sub>2</sub> capture efficiency, compared to air-based combustion and EGR at 35% recirculation ratio configurations.



**Figure C.2.-** Heat transfer – temperature diagram for S-EGR in series configuration, with 95% selective CO<sub>2</sub> transfer efficiency and 32% post-combustion CO<sub>2</sub> capture efficiency, compared to air-based combustion and EGR at 35% recirculation ratio diagrams.