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Reduction of Carbon Dioxide Emissions from a SCGT/CC by Ammonia Solution Absorption – Preliminary Results

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Abstract

The reduction of carbon dioxide from the flue gases of a semi-closed gas turbine combined cycle (SCGT/CC) by means of absorption in ammonia aqueous solutions has been studied. The absorption system has been simulated by means of Aspen Plus^{TM}. The main variables of the removal system have been varied in order to understand their influence on system performance. With reference to the SCGT/CC case study, the removal of CO_2, considering a removal efficiency of 89%, dramatically decreases the overall cycle efficiency from 53 to 41%, with the main contribution to this decrease being due to the power consumption for flue gas compression up to the absorption unit pressure. CO_2 specific emissions pass from 390 to 57 kg/MWh.

Keywords: Greenhouse effect, carbon dioxide, CO_2, absorption, ammonia, semi-closed gas turbine combined cycle.

1. Introduction

The contribution of CO_2 release with respect to greenhouse effects and long-term climate alteration is becoming more relevant every year. New concepts for energy conversion with very low or null release of CO_2 to the environment are at a developing stage and, in the transition phase techniques for CO_2 separation from the stack, can be applied to existing power plants with minor modifications.

The reduction of carbon dioxide from flue gases of a semi-closed gas turbine combined cycle (SCGT/CC), by means of absorption in ammonia aqueous solutions, has been studied by means of simulations using Aspen Plus^{TM} (Aspen Tech, Inc., 1996).

Information found in the literature indicated that NH_3 has the potential to be an interesting alternative to other more readily known processes such as amine absorption (Yeh and Bai, 1996, Yeh and Bai, 1998). Hence, general studies on the possibility of carbon dioxide absorption by ammonia solution have already been performed, while the contribution of the present work is the study of the application of this absorption system to a power cycle selected for the desirable feature of high carbon dioxide mass concentration in the flue gases, which facilitates the removal process.

The process scheme is based first on an absorption unit, where the flue gas and absorbing solution come in contact and CO_2 absorption takes place, and a second unit (desorption unit) where the CO_2 load solution is thermally regenerated. Furthermore, ammonia and carbon dioxide, which can form a salt under certain conditions at low temperatures, make sequestration of carbon dioxide in solid form of interest. The problem, however, is that the process cannot reach a high yield of salt formation. Hence, it was decided to investigate conditions that avoid salt formation, in order to set up a removal system where it is possible to regenerate the CO_2 load solution and recirculate the regenerated solution to the absorption column. In this way, the effective consumption of ammonia could be reduced with respect to producing a final solid compound containing the ammonia.

The sensitivity of the absorption and desorption processes with respect to some parameters considered to be relevant - pressure,

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*Author to whom correspondence should be addressed temperature, ammonia concentration - was carried out. Ammonia, near ambient conditions, tends to pass from the liquid to the gaseous phase very easily, causing some problems with the presence of ammonia in the gaseous stream (purified gas) leaving the absorber (ammonia slipping). This problem can be partly overcome by increasing the absorber pressure above the atmospheric value. For the same reason, in order to improve the separation of CO_2 and ammonia (the aim is to keep ammonia in an aqueous phase) in the desorption unit, pressurised conditions are required. With respect to the ammonia solution concentration, in order to reduce the ammonia slip and to increase the desorption process efficiency, low concentrations are investigated.

The basic process scheme was further improved with heat recovery devices, ammonia recovery from absorption unit stack gas, and flow rate and composition control.

2. SCGT/CC Basic Concepts

The first step in decreasing the overall emissions of CO_2 is to develop power plants with improved efficiencies in order to reduce the amount of CO_2 emitted per unit energy produced. As a matter of fact, this feature is retained by the SCGT/CC (*Figure 1*) that was previously studied and described (Facchini, et al., 1997; Fiaschi et al., 1998; Corti et al., 1999) and whose efficiency is comparable to modern open combined cycles. However, the SCGT/CC adds some very desirable features, mainly the possibility of applying CO_2 removal techniques, because of the relatively high concentration of CO_2 in the flue gases, which is a consequence of the extensive recycle of the cooled flue gases at the compressor inlet.

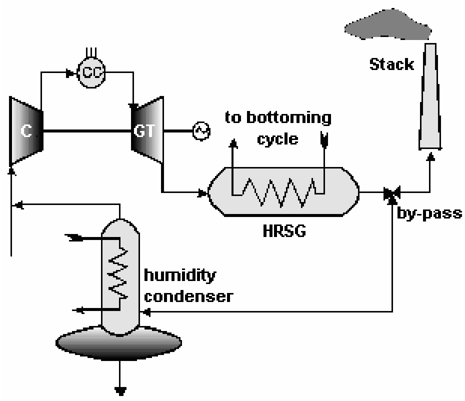

Figure 1. SCGT/CC scheme.

The SCGT/CC scheme is based on a combined cycle configuration with topping gas turbine (GT) and bottoming steam turbine (ST) cycles. A bypass is added downstream of the heat recovery steam generator (HRSG) so that part of the flue gases is driven to the stack and the rest recirculated to the compressor inlet (C). Before entering the compressor, the recirculated gas stream needs cooling, achieving temperatures below the dew point - in the humidity condenser - so that a large part of the water vapour is condensed.

The study case for the SCGT/CC (Corti et al., 1999) with CO_2 removal done here is based on gas turbine operating data corresponding to those of the LM501F gas turbine jointly developed by Mitsubishi and Westinghouse, a heavy duty gas turbine well suited for operation with combined cycles. The SCGT/CC characteristics are reported in TABLE I. Results were obtained by means of an in-house developed FORTRAN code of its thermodynamic model (Facchini et al., 1997).

TABLE I. SCGT/CC WORKING CONDITIONS WITHOUT CO_2 REMOVAL

GT inlet temperature [°C]	1349
GT compression ratio	14
GT power [kW]	151590
Bottoming cycle high pres. steam [bar]	120
Bottoming cycle low pres. steam [bar]	12
ST power [kW]	92769
CC net power [kW]	244359
Fuel flow rate [kg/s]	9.65
Fuel LHV [kJ/kg]	47.766
Efficiency [%]	53
CO_2 specific emissions [kg/MWh]	390

TABLE II. SCGT/CC FLUE GAS FEATURES

Composition	Mass fraction
H_2O	0.0599
CO_2	0.147
N_2	0.773
NO_x	7.00E-05
O_2	0.02
Mass flow rate [kg/s]	180
Temperature [°C]	120

The flue gas mass flow rate, temperature and composition are reported in TABLE II. The flue gas results from the FORTRAN model in terms of mass flow rate, temperature, pressure and composition were used as input to the absorption model developed in Aspen Plus^{TM}. Once calculated in Aspen Plus^{TM}, the thermal request for regeneration in the absorption system, which is supposed to be supplied by steam bleed from power cycle, is calculated and the FORTRAN code is run again taking this consumption in account in order to determine the power and efficiency reduction of the cycle.

3. CO_2 Ammonia Absorption System

The CO_2 absorption system is based on two main units: the absorption column where the CO_2 absorption takes place and the desorption column where the CO_2 load solution is thermally regenerated. The absorption and desorption columns are simulated with the Aspen Plus^{TM} unit RADFRAC, in both cases using two equilibrium stages. The absorber unit has no condenser or reboiler. The desorber unit has one inlet stream and two exit streams (regenerated solution and carbon dioxide-rich gaseous stream) and is provided with external heat duty. The heat duty amount is calculated imposing the stream exiting temperature. The adopted property set in Aspen Plus^{TM} is ELECNRTL (Aspen Tech, Inc., 1996). Actually, the basic scheme for the absorption/ desorption process has been improved as shown in *Figure 2*.

The flue gas entering stream needs to be compressed to the absorber working pressure (flue gas compressor) and is then cooled down in a recovery heat exchanger (HE-1) where part of its enthalpy is used to heat up the CO_2 load solution before entering the regeneration column.

Flue gases are further cooled down in a second heat exchanger (HE-2), heating the purified gaseous stream and increasing the buoyancy at the stack; and finally they are cooled down to the defined absorber inlet temperature (HE-3) before entering the bottom stage of the absorption column.

The ammonia aqueous solution enters the top stage of the absorption column, after recirculation, fresh ammonia addition - in the required amount to keep a constant ammonia concentration in the absorber entering solution - and re-pressurisation. The CO_2 load solution leaves the absorption column from the bottom stage. It is pressurised up to the desorber working pressure and is heated by means of energy recovery from the lean solution (HE-4) and from the flue gases (HE-1)

The CO_2 load solution enters the desorption column, where it is thermally regenerated by an external heat duty. In the desorption unit, the gaseous CO_2 is released and separated from the solid phase: a stream of almost pure CO_2 is obtained. This lean solution is cooled in the HE-4 and further in an externally cooled heat exchanger (HE-5) in order to reach the defined absorber inlet temperature.

After a mass flow rate control system (adding fresh water or blowing down part of the solution) and a fresh ammonia make-up, the regenerated solution is recirculated back to the absorber. The gaseous stream exiting from the absorber - the purified flue gases -may contain a high amount of ammonia. Hence water scrubbing may be required.

4. Chemical Reactions in the H_2O-NH_3-CO_2 System

The absorption-desorption system, realised by means of Aspen Plus^{TM} (Aspen Tech, Inc., 1996), is based on chemical equilibrium units in which the following reactions are taken into account for the carbon dioxide-ammonia-water system:

$$2H_2O \leftrightarrow H_3O^+ + OH^- \qquad (1)$$

$$CO_2 + 2\,H_2O \leftrightarrow H_3O^+ + HCO_3^- \qquad (2)$$

$$HCO_3^- + H_2O \leftrightarrow H_3O^+ + CO_3^{2-} \qquad (3)$$

$$NH_3 + H_2O \leftrightarrow NH_4^+ + OH^- \qquad (4)$$

$$NH_3 + HCO_3^- \leftrightarrow NH_2COO^- + H_2O \qquad (5)$$

$$NH_4^+ + HCO_3^- \leftrightarrow NH_4HCO_3\ (s) \qquad (6)$$

The set of reactions for the CO_2-NH_3-H_2O system and the reaction constants provided by Aspen Plus^{TM} (Aspen Tech Inc, 1996) were checked and used. Actually, reaction (6) could be an interesting solution for solid capture of carbon dioxide, forming a solid product, but the low yield of formation of the salt at the considered working conditions (pressure: 1-3 bar; temperature < 50°C) (Aspen Tech, Inc., 1996) does not allow this feature to be exploited.

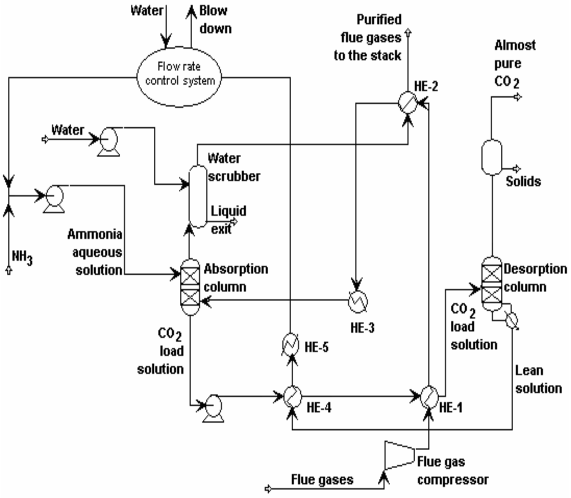

Figure 2. CO_2 removal system.

Reaction (5) is the main one on which the chemical absorption process is based. Ammonia reacts with the HCO_3^- ions in aqueous solutions, increasing the CO_2 mass transfer from the gaseous to the liquid phase (chemical absorption). Since this reaction is reversibly supplying heat, the absorbing solution can be thermally regenerated, at least in part, obtaining ammonia and HCO_3^- ions again. These ions, at the regeneration temperatures, again generate gaseous carbon dioxide.

5. Sensitivity of the Absorption System to the Main Working Parameters

In the system described above, the working parameters of primary importance are the ammonia solution concentration, the absorber temperature and pressure, the desorber temperature and pressure, and the flue gas temperature. From a preliminary evaluation of wider ranges for each variable, the reference values for the base case reported in TABLE III have been assumed for the primary parameters. The analysis is now focused on changing these values within relatively small ranges.

Generally, the regeneration of a solvent can be achieved by means of both temperature increase and pressure decrease with respect to the absorber conditions. In this case, the regeneration is operated by exploiting only the thermal mechanism of breaking the chemical bonds formed by CO_2 in an aqueous solution. In fact, contrary to the common use, the pressure in the desorber is increased with respect to the absorber level. This is necessary in order to improve the separation between ammonia and carbon dioxide. As a matter of fact, both compounds tend to be released in the gaseous phase during the regeneration process: both pure compounds are above the saturation temperature for the pressure range of interest (1 to 30 bar) and the temperature levels (higher than 100°C) (*Figure 3*) (Klein and Alvarado, 2002), while the aim of regeneration is to obtain gaseous CO_2 and an NH_3-rich aqueous solution to be recirculated.

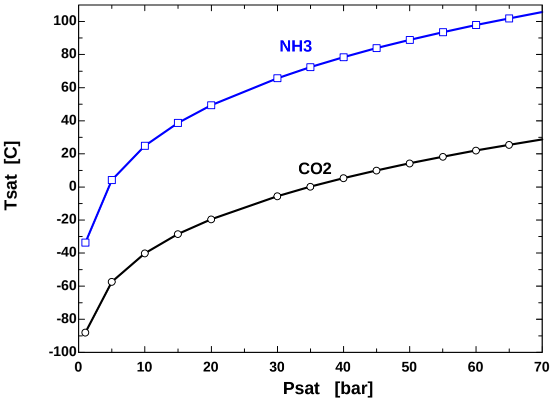

Figure 3. NH_3 and CO_2 saturation temperature and pressure trends.

If the desorber pressure is increased – as will be shown in the sensitivity analysis – the ratio between CO_2 and NH_3 transferred in the gaseous phase increases. Referring to pure compounds (*Figure 3*), it is possible to highlight the higher carbon dioxide amount in the gaseous phase with respect to the ammonia amount and the increasing value of this difference (CO_2 amount - NH_3 amount) with increasing pressure.

In the sensitivity analysis, each of the parameters have been varied within a defined range, keeping constant the others and keeping a constant carbon dioxide removal efficiency of 80% (considering for the moment only the absorber and not the water scrubber). The latter is accomplished by means of adjusting the aqueous ammonia solution mass flow rate and evaluating the response of the system in terms of the variables considered to be of primary importance as reported in TABLE IV.

A constant temperature of 150°C is kept for the desorber entering solution supplied by energy recovery. Also, the desorber exiting solution temperature is kept constant at 165°C, representing the primary parameter that influences CO_2 desorption and directly affects the external heat duty requirement.

TABLE III. BASE CASE REFERENCE VALUES FOR THE PRIMARY PARAMETERS

Absorber pressure [bar]	3
Desorber pressure [bar]	16
Absorber inlet solution temperature [°C]	40
Flue gas temperature [°C]	25
Ammonia mass fraction in the absorber inlet solution [%]	3.4

TABLE IV. VARIABLES OF PRIMARY IMPORTANCE IN THE SENSITIVITY ANALYSIS

Description of the variable	Name
Solution mass flow rate entering the absorber [kg/s]	Flow
NH_3 mass fraction in the gas leaving the absorber	$NH_{3\ abs}$
NH_3 mass fraction in the CO_2-rich gas leaving the desorber	$NH_{3\ des}$
CO_2 mass fraction in the CO_2-rich gas leaving the desorber	CO_2-$_{purity}$
The mass ratio between the amount of NH_3 regenerated and the total amount entering the absorber	$NH_{3\ reg}$
The mass ratio between the amounts of CO_2 and NH_3 in the CO_2-rich gas leaving the desorber	CO_2/ $NH_{3\ sep}$
The mass ratio between the amounts of CO_2 leaving the desorber and the CO_2 entering the desorber	CO_{2out}/ $CO_{2in\ des}$
External heat duty [kW]	EHD

5.1 Ammonia solution concentration

The system sensitivity to the ammonia concentration in the absorbing solution was tested over a wide range of values varying from 2 to 35% of NH_3 by weight. Actually, higher ammonia concentrations allow one to work with a lower solution mass flow rate, with a minimum corresponding to about 15% of ammonia. In contrast, NH_3 regeneration - and NH_3/CO_2 separation in the desorber - increases with lower concentration and stack loss of ammonia decreases with lower concentration.

In order to minimise ammonia losses and improve regeneration and CO_2/NH_3 separation, the range for a more detailed investigation was restricted to 2-4% of ammonia mass fraction in the solution. TABLE V shows the results of the primary variables with respect to varying ammonia solution concentrations.

The change in NH_3 leaving the absorber with the gas (NH_{3abs}) is lower at lower concentrations. Also the amount of NH_3 leaving in the gas phase from the desorber (NH_{3des}) is almost linear at these concentrations, decreasing as the concentration decreases. Hence, lowering the concentration as much as possible should be beneficial.

In order to decrease the greatest possible the ammonia losses in the gas leaving the absorber and to increase the separated CO_2 purity, it is suggested to work with a low concentration, even if this increases the solution mass flow rate. Changing the concentration from 2 to 3% reduces the flow rate by about 26%, while the step from 3 to 4% in concentration decreases the flow rate by about 16%. With the solution mass flow rate reduction being smaller for higher concentrations, the concentration value can be set a bit above 3% (3.4%).

TABLE V. SENSTIVITY OF THE PRIMARY VARIABLES TO AMMONIA SOLUTION CONCENTRATION CHANGES

NH_3 solution mass fraction	Flow	$NH_{3\ abs}$	$NH_{3\ des}$	CO_2-$_{purity}$
0.020	585	0.004	0.029	0.810
0.030	433	0.007	0.041	0.800
0.034	403	0.009	0.045	0.800
0.040	365	0.012	0.052	0.790
NH_3 solution mass fraction	$NH_{3\ reg}$	$CO_2/NH_{3\ sep}$	EHD	CO_{2out}/CO_{2indes}
0.020	0.88	28.46	54	0.997
0.030	0.83	19.68	44	0.997
0.034	0.81	17.60	43	0.998
0.040	0.78	1.22	41	0.998

5.2 Absorption column pressure

The absorption column pressure directly affects the NH_3 amount leaving the absorber with the gases, the solution mass flow rate, and consequently, the external heat duty. *Figure 4* shows the decrease in NH_3 mass fraction of the gas leaving the absorber and the percentage decerease in solution mass flow rate - the system is improved - as the pressure is increased. However, at higher pressures, the improvement is less; from 2 to 3 bar a percentage reduction of 36% is obtained for the NH_3 mass fraction in the gas and a percentage reduction of 11% for the solution mass flow rate. When changing the pressure from 3 to 4 bar, the NH_3 mass fraction in the gas is further reduced by 6 percentage points and the solution mass flow rate by a further 14 percentage points. Hence, a pressure of 3 bar could be a good compromise, also considering that higher pressures imply higher flue gas compression power consumption.

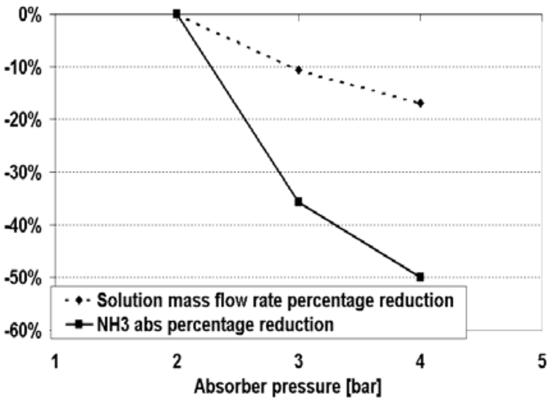

Figure 4. Percentage reductions of solution mass flow rate and NH_3 mass fraction in the gas leaving the absorber versus absorber pressure.

TABLE VI summarises the various results for the different values of the absorber pressure.

TABLE VI. SENSITIVITY OF THE PRIMARY VARIABLES TO ABSORBER PRESSURE CHANGES

Absorber pressure [bar]	Flow	$NH_{3\ abs}$	$NH_{3\ des}$	CO_2-$_{purity}$
2	451	0.014	0.040	0.790
3	403	0.009	0.050	0.800
4	375	0.007	0.050	0.770
Absorber pressure [bar]	$NH_{3\ reg}$	$CO_2/NH_{3\ sep}$	EHD	$CO_{2out}/CO_{2in\ des}$
2	0.78	18.00	46	0.999
3	0.81	17.60	43	0.998
4	0.80	15.18	31	0.997

5.3 Desorption column pressure

The desorber pressure has been varied between 8 and 32 bar. TABLE VII shows the main variable results, indicating that increasing the desorber pressure reduces the amount of ammonia in the gaseous stream exiting from the desorber. Consequently, the ammonia regeneration and the CO_2/NH_3 separation increase, as well as the CO_2-purity. The percentage variation of the primary variables is shown in *Figure 5* with respect to the base case of 8 bar at the desorber and indicates that a pressure level of about 20 bar would be a reasonable value, since the percent changes in the parameters $NH_{3\ des}$, $NH_{3\ reg}$, $CO_2/\ NH_{3\ sep}$, and $CO_{2\ purity}$, are smaller as the pressure is increased above 20 bar. The negative effect of increasing the pressure during the regeneration process is visible, in TABLE VII as the ratio between the amounts of CO_2 leaving the desorber and the CO_2 entering the desorber ($CO_{2out}/\ CO_{2in\ des}$), indicating a reduction in CO_2 regeneration efficiency that, however, is not sensitive.

TABLE VII. SENSITIVITY OF THE PRIMARY VARIABLES TO THE DESORBER PRESSURE CHANGES

Desorber pressure [bar]	Flow	$NH_{3\ abs}$	$NH_{3\ des}$	CO_2-purity
8	400	0.008	0.070	0.700
12	400	0.008	0.050	0.750
16	403	0.009	0.050	0.790
20	405	0.009	0.040	0.820
32	416	0.010	0.040	0.850
Desorber pressure [bar]	$NH_{3\ reg}$	$CO_2/\ NH_{3\ sep}$	EHD	$CO_{2out}/\ CO_{2in\ des}$
8	0.76	10.37	49	1.000
12	0.79	14.14	45	0.999
16	0.81	17.60	43	0.998
20	0.81	19.94	42	0.996
32	0.82	24.26	36	0.997

5.4 Absorber inlet temperature

Lower temperatures in the absorber enhance the absorption process, obviously reducing the required solution flow rate. The temperature within the absorber is determined mainly by the inlet solution temperature and in part, by the flue gas temperature. In addition, a lower absorbing temperature favours the reduction of ammonia slip in the gas stream.

The values of the solution mass flow rate, external heat duty and ammonia mass fraction in the gas leaving the absorber, reported in TABLE VIII, obviously confirm that lowering the temperature improves the system. The other parameters remain essentially unchanged. Hence, the temperature in the absorber should be as low as possible.

An analogous trend is found for the solution mass flow rate, external heat duty, and ammonia mass fraction in the gas leaving the absorber when the flue gas temperature is changed, as shown in *Figure 6*. Theoretically, a temperature of 25°C for both flue gases and solution should be assumed; but for technical limits of heat exchange, a value of 40°C is suggested.

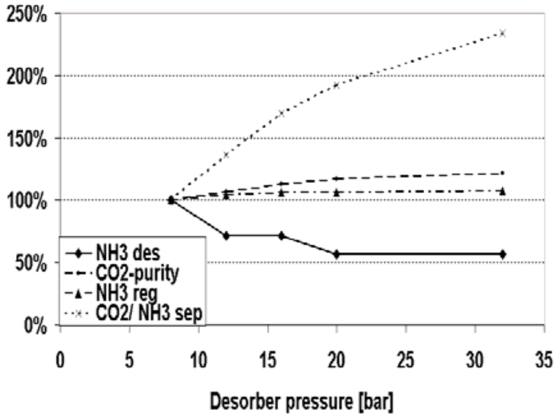

Figure 5. Percentage variation - with respect to an 8 bar desorber pressure - of the variables $NH_{3\ des}$, $NH_{3\ reg}$, $CO_2/\ NH_{3\ sep}$, and $CO_{2\ purity}$ versus desorber pressure.

TABLE VIII. SENSITIVITY OF THE PRIMARY VARIABLES TO THE ABSORBER INLET SOLUTION TEMPERATURE

Solution temp.[°C]	Flow	$NH_{3\ abs}$	$NH_{3\ des}$	CO_2-purity
25	335	0.004	0.050	0.800
30	355	0.006	0.050	0.820
40	403	0.009	0.050	0.790
60	644	0.025	0.050	0.760
Solution temp.[°C]	$NH_{3\ reg}$	$CO_2/\ NH_{3\ sep}$	EHD	$CO_{2out}/\ CO_{2in\ des}$
25	0.82	17.36	38	0.996
30	0.82	17.41	39	0.996
40	0.81	17.60	43	0.998
60	0.75	16.32	63	1.000

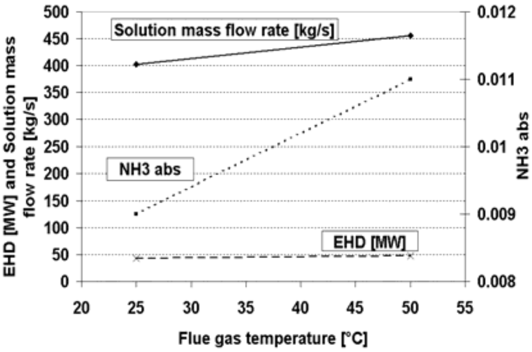

Figure 6. Trends of the solution mass flow rate, external heat duty, and ammonia mass fraction in the gas leaving the absorber versus flue gas temperature.

6. System Definition and Performance Calculation

From the sensitivity analysis, it has been possible to highlight the indicative values for the primary parameters that allow the reduction of some undesirable effects - mainly ammonia slip in the absorber and the desorber and the external heat duty requirement. On the basis of the previous analysis results, the set values for the studied parameters are reported in TABLE IX. Using the values in this table for the working parameters and keeping a CO_2 removal efficiency of 80%, a temperature of the desorber entering solution of 150°C, and a temperature of the desorber exiting solution of 165°C, the system simulation was run supplying the results of TABLE X. A fresh flow rate of ammonia of 2.9 kg/s is required to keep the defined constant concentration in the entering absorber solution.

However, the ammonia mass fraction in the absorber gaseous stream is still too high (13800 mg/Nm^3), even if a great effort was made to reduce its amount, in comparison with the general emission limits for industrial sites stated for air quality protection (250 mg/Nm^3) according to Italian law (D.M. 12/07/1990). This implies adding a secondary scrubber for the gaseous stream leaving the absorber with the aim of reducing the very high ammonia concentration by means of water scrubbing.

Using a water stream of about 165 kg/s, it is possible to reduce the amount of ammonia below the legal limit (a value of 20% with respect to the general limit was assumed in order to consider an intermediate value among the different figures for very different typologies of plants - hence about 50 mg/Nm^3). Consequently, also the CO_2 concentration in the gaseous stream is reduced further for an overall removal efficiency of 89%.

The water stream loaded with CO_2 and ammonia needs further treatment for separation of carbon dioxide and ammonia recovery. As a preliminary solution, regeneration in an additional desorber reactor, operating at the same conditions of the main one (20 bar, inlet load solution temperature of 150 °C, outlet lean solution temperature of 165 °C), is suggested. In this way, an additional stream rich in CO_2 and a liquid stream rich in NH_3 are produced. The first follows the same pattern of the CO_2 rich stream from the main desorber (compression and final storage), while the second one can be used for the make up of ammonia. Additional energy consumption for the external heat duty in this additional desorber is needed.

For the overall efficiency calculation (SCGT/CC with the CO_2-Ammonia Absorption System), several consumption contributions have to be considered: flue gas compression power consumption - and to a lesser extent also pump power consumptions - that is directly subtracted from the SCGT/CC power output; external heat duty for load solutions regeneration (both the main and additional desorber) that is supplied by means of steam extraction from the low pressure steam turbine of the power cycle (11.2 bar pressure level, about 29 kg/s of steam extracted with respect to an overall bottoming steam mass flow of 78 kg/s; the steam bleed reduces the SCGT/CC efficiency from 53% to 50..6%); CO_2 stream compression up to 80 bar for transportation to disposal sites. Considering the values reported in TABLE IX for the external heat duty requirement and the flue gas compressor power (adding 1.1 MW for pump consumption) and a specific compression work of about 355 kJ per kg of CO_2 (Lombardi, 2003), an overall efficiency of 41% is calculated. TABLE XI summarises the performance of the SCGT/CC with the CO_2-Ammonia Absorption System.

TABLE IX. SET VALUES FOR THE STUDIED PARAMETERS

Absorber pressure [bar]	3
Desorber pressure [bar]	20
Absorber inlet solution temperature [°C]	40
Flue gas temperature [°C]	40
Ammonia mass fraction in the absorber inlet solution [%]	3.4

TABLE X. MAIN SIMULATION RESULTS

Solution mass flow rate [kg/s]	444
External heat duty [MW]	42.8
Flue gas compression power [MW]	36
NH_3 mass fraction in the absorber stack gas	0.011
NH_3 mass fraction in the desorber exit gaseous stream	0.04
CO_2 mass fraction in the desorber exit gaseous stream	0.83
NH_3 regeneration efficiency [%]	81
CO_2/NH_3 separation in the desorber [rate]	21.2

TABLE XI. PERFORMANCE OF THE SCGT/ CC WITH CO_2-AMMONIA ABSORPTION SYSTEM

GT power [kW]	151590
ST power [kW]	81756
CC power [kW]	233346
Flue gas compressor + pump power [kW]	37500
CO_2 compression power [kW]	6686
CC net power [kW]	189161
Efficiency [%]	41
CO_2 specific emission [kg/MWh]	57

For comparison, results previously obtained for the SCGT/CC with an Amine Absorption System (Lombardi, 2001; Lombardi, 2003) are reported. In that case, an 85% CO_2 removal efficiency was reached and specific emissions of CO_2 of 65 kg/MWh. Considering again steam extraction for the amine regeneration external heat duty and CO_2 compression up to 80 bar, an efficiency of 46% was calculated.

When the carbon dioxide reduction from SCGT/CC flue gases was studied applying semi-permeable membranes (Corti et al., 2002) with a removal efficiency of 85%, the calculated overall cycle efficiency was 42.7%. In this case, the carbon dioxide specific emissions were 70 kg/MWh.

7. Conclusions

A CO_2-Ammonia Absorption System, based on solvent thermal regeneration, was simulated by means of Aspen Plus^{TM}. A sensitivity analysis with respect to the primary parameters of the system - absorber and desorber pressure and temperature and ammonia concentration in the absorbing solution - was carried out, mainly in order to minimise the entrainment of ammonia in the gaseous streams leaving, respectively, the absorber and the desorber. Nevertheless, an additional water scrubbing is required to comply with legal limits for ammonia emissions at industrial sites, concerning the gaseous stream leaving the absorber.

For the primary parameters for which the values found for best system performance were assumed, it was possible to calculate the performance of the overall SCGT/CC with the CO_2-Ammonia Absorption System. The overall efficiency was dramatically reduced from 53 to 41%. The reduction can be ascribed to steam extraction for the regeneration external heat duty with a contribution of about 2.4 percentage points; to CO_2 compression with a reduction of about 1.5 .percentage points; and primarily to the power consumption for flue gas compression, contributing to a reduction of about 8.1 percentage points.

Considering an overall CO_2 removal of 89%, the CO_2 specific emissions pass from 390 kg/MWh for the simple SCGT/CC to 57 kg/MWh, for the SCGT/CC with the CO_2-Ammonia Absorption System. In comparison with previous results obtained for the SCGT/CC with an Amine Absorption System and one with membrane separation, the efficiency reduction that must be paid in order to decrease carbon dioxide emissions is greater for the system that uses ammonia, due primarily to the power required for flue gas compression, since the ammonia system must work at a higher pressure level in the absorber. This consideration leads to the conclusion that the ammonia system is probably more suitable for carbon dioxide removal applications where the gas to be treated is already in a pressurised condition as, for example, in syngas purification applications (CO_2 pre-combustion removal for Integrated Gasification Combined Cycles or methane reforming cycles).

Finally, further investigations are required about the treatment of the water used for scrubbing the ammonia from the gas leaving the absorber, and also about different possibilities of reducing the ammonia entrainment in the gas leaving the absorber.

References

Aspen Tech, Inc., 1996, “*Aspen Plus User Guide Volume 1-2*”.

Corti A., Fiaschi D. and Manfrida G., 1999, “Thermoeconomic evaluation of the SCGT cycle”,
Energy Conversion and Management, Volume 40, Issue 18, December 1999, Pages 1917-1929

Corti A., Fiaschi D., Lombardi L., 2002, “Carbon dioxide removal in power generation using membrane technology” *Energy*, Volume 29, Issues 12-15, October-December 2002, Pages 2025-2043, Elsevier Science Limited, Oxford (UK)

D.M. 12/07/1990, “Linee guida per il contenimento delle emissioni inquinanti degli impianti industriali e” la fissazione dei valori minimi di emissione”, *G.U. Supplemento Ordinario n°* 176 del 30/07/90

Facchini, B., Fiaschi D. and Manfrida G., 1997, “SCGT/CC: An innovative cycle with advanced environmental and peakload shaving features”, *Energy Conversion and Management*, Volume 38, Issues 15-17, October-November 1997, Pages 1647-1653

Fiaschi D. and Manfrida G., 1998, “Exergy analysis of the semi-closed gas turbine combined cycle (SCGT/CC)”, *Energy Conversion and Managemen*t, Volume 39, Issues 16-18, November-December 1998, Pages 1643-1652

Klein S. A., Alvarado F. L., 2002, “*Engineering Equation Solver”* – Academic Version 6.548N – F-Chart Software, Middleton (USA).

Lombardi L., 2001, “Life cycle assessment (LCA) and exergetic life cycle assessment (ELCA) of a semi-closed gas turbine cycle with CO_2chemical absorption”, *Energy Conversion and Management*, Volume 42, Issue 1, 1 January 2001, Pages 101-114

L. Lombardi, 2003, “Life cycle assessment comparison of technical solutions for CO_2emissions reduction in power generation”, *Energy Conversion and Management*, Volume 44, Issue 1, January 2003, Pages 93-108

Yeh, A C., Bai H., 1996, “A novel study on the carbon dioxide removal by ammonia scrubbing”, presented at the 89th annual meeting & exhibition of the Air & Waste Management Association, Nashville, Tennessee, June 23-28, 1996

Yeh, A C., Bai H., 1998, “Comparison of ammonia and monoethanolamine solvents to reduce CO_2 greenhouse gas emissions”, *The Science of the total Environment* Vol. 228, issues 2-3. 5th April 1999